INL/RPT-24-76435
Revision 1

Thermal Integration of Advanced Nuclear Reactors with a Reference Refinery, Methanol Synthesis, and a Wood Pulp Plant

September 2024

Vaclav Novotny, Kathleen P. Sweeney, Elizabeth K. Worsham,
Eliezer A. Reyes Molina, Sam J. Root, Frederick Joseck,
Byung-Hee Choi, Nipun Popli, Junyung Kim, Todd Knighton,
and Rami M. Saeed

Idaho National Laboratory

IES
Integrated Energy Systems

INL/RPT-24-76435
Revision 1

Thermal Integration of Advanced Nuclear Reactors with a Reference Refinery, Methanol Synthesis, and a Wood Pulp Plant

Vaclav Novotny, Kathleen P. Sweeney, Elizabeth K. Worsham,
Eliezer A. Reyes Molina, Sam J. Root, Frederick Joseck,
Byung-Hee Choi, Nipun Popli, Junyung Kim, Todd Knighton,
and Rami M. Saeed

Idaho National Laboratory

September 2024

Idaho National Laboratory
Integrated Energy Systems
Idaho Falls, Idaho 83415

http://www.ies.inl.gov

Prepared for the
U.S. Department of Energy
Office of Nuclear Energy
Under DOE Idaho Operations Office
Contract DE-AC07-05ID14517

Page intentionally left blank

EXECUTIVE SUMMARY

The present report is intended to provide process flow diagrams (PFDs) and energy and mass balance data sheets for a U.S. industrial sector subset with which nuclear heat and power could be integrated—a subset that includes the oil refining, methanol and pulp and paper industries. Coupling options for integrating nuclear energy into these industries are quantitatively outlined for reference systems, and future work will extend this analysis in greater detail.

Opportunities for integrating small modular nuclear reactors (SMNRs) were investigated for each of the industrial process configurations. Aspen HYSYS and Cycle-Tempo models for a high-temperature gas-cooled reactor were developed to evaluate the proposed integration. This introductory evaluation provides a general description and assessment of the operating principles, reactor coolant core outlet temperature, and reactor size to be integrated with industry.

The industrial processes of oil refining and the production of methanol, pulp and paper were simulated by using Aspen HYSYS, Aspen Plus, and the PRELIM (Petroleum Refinery Life Cycle Inventory Model) tool to develop process details. Cycle-Tempo models then extend the process modeling results to obtain net energy demands (e.g., heat, steam, and electricity) when accounting for process steam and waste heat recovery. This information is intended to foster the analysis of integrating an SMNR to decarbonize industrial facilities. The SMNR would provide reliable, competitive, and sustainable clean energy while reducing carbon emissions and other environmental impacts, such as water withdrawals, consumption, and contamination.

The refining industry, exhibited in Figure ES1, is a leading consumer of fossil -fuel-based heat, power, and hydrogen in the U.S. industrial sector, generating over 164 million metric tons (MMT) of CO_2 emissions in 2023 [1]. The overall mass and energy pertaining to a generalized complex refinery in the United States is reflected in Figure ES1, along with energy metrics regarding integration with a nuclear power plant (NPP). Data sheets were developed to indicate the energy requirements for the overall refinery and each refinery process. The data sheet for the overall refinery is shown in Table ES2.

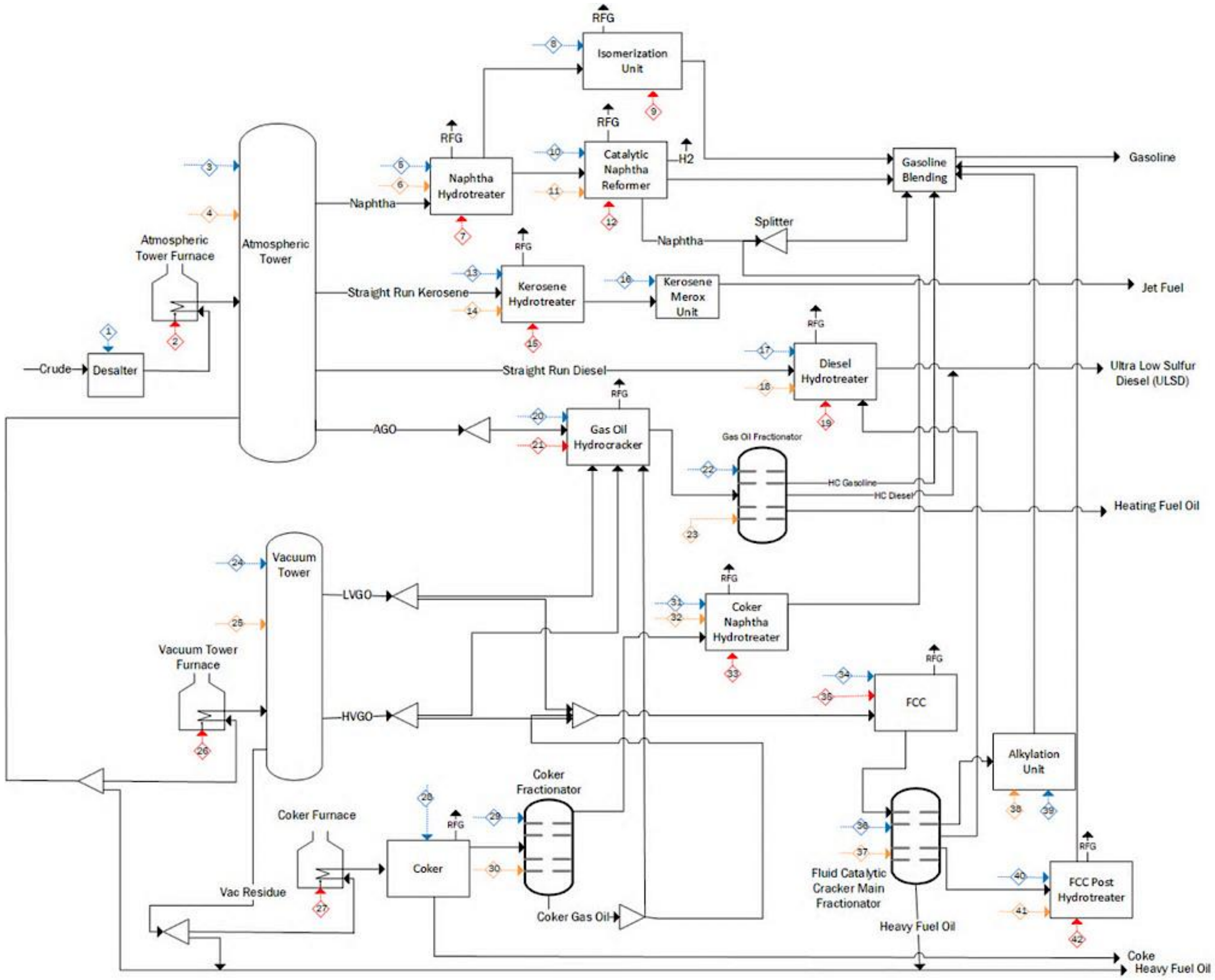
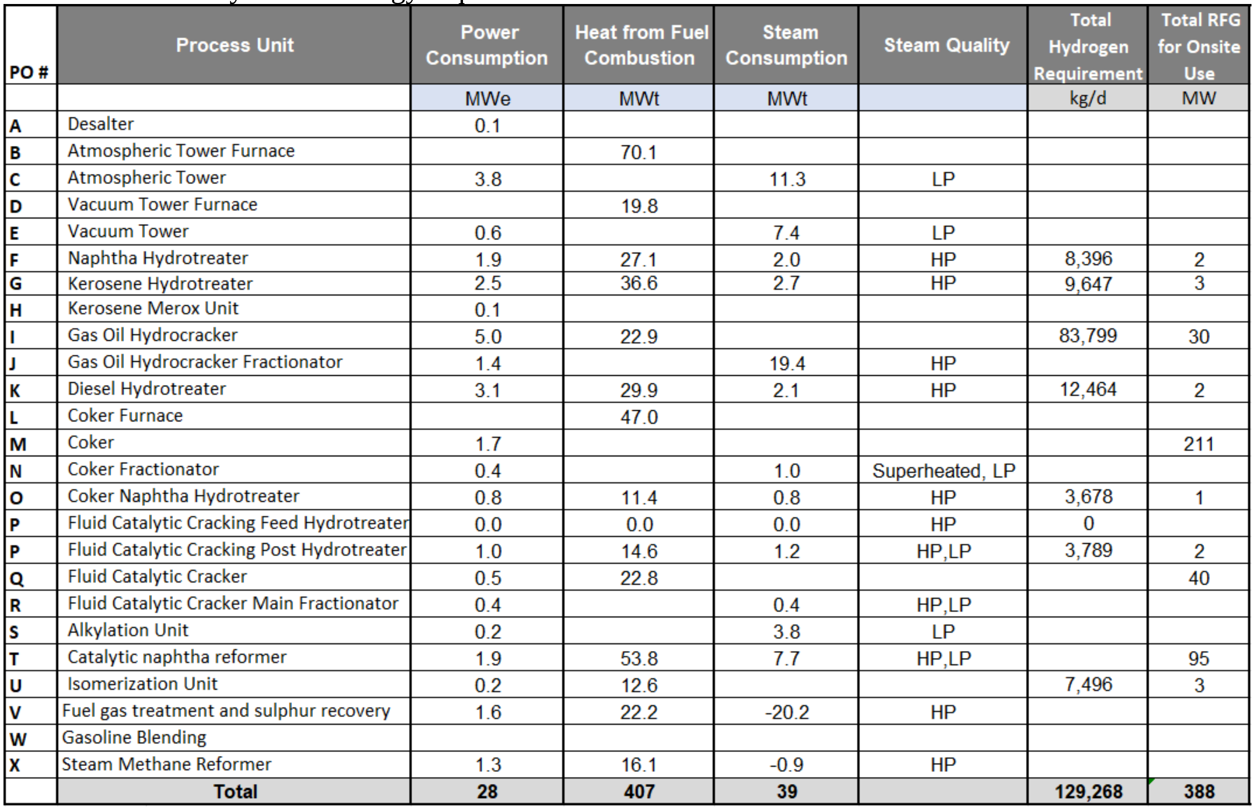


Figure ES1. Refinery overall energy PFD, where the numbers correspond to the energy input from electricity, heat in fuel, and steam, as listed in Appendix D.

Table ES1. Refinery overall energy requirements.



PO #	Process Unit	Power Consumption	Heat from Fuel Combustion	Steam Consumption	Steam Quality	Total Hydrogen Requirement	Total RFG for Onsite Use
		MWe	MWt	MWt		kg/d	MW
A	Desalter	0.1					
B	Atmospheric Tower Furnace		70.1				
C	Atmospheric Tower	3.8		11.3	LP		
D	Vacuum Tower Furnace		19.8				
E	Vacuum Tower	0.6		7.4	LP		
F	Naphtha Hydrotreater	1.9	27.1	2.0	HP	8,396	2
G	Kerosene Hydrotreater	2.5	36.6	2.7	HP	9,647	3
H	Kerosene Merox Unit	0.1					
I	Gas Oil Hydrocracker	5.0	22.9			83,799	30
J	Gas Oil Hydrocracker Fractionator	1.4		19.4	HP		
K	Diesel Hydrotreater	3.1	29.9	2.1	HP	12,464	2
L	Coker Furnace		47.0				
M	Coker	1.7					211
N	Coker Fractionator	0.4		1.0	Superheated, LP		
O	Coker Naphtha Hydrotreater	0.8	11.4	0.8	HP	3,678	1
P	Fluid Catalytic Cracking Feed Hydrotreater	0.0	0.0	0.0	HP	0	
P	Fluid Catalytic Cracking Post Hydrotreater	1.0	14.6	1.2	HP,LP	3,789	2
Q	Fluid Catalytic Cracker	0.5	22.8				40
R	Fluid Catalytic Cracker Main Fractionator	0.4		0.4	HP,LP		
S	Alkylation Unit	0.2		3.8	LP		
T	Catalytic naphtha reformer	1.9	53.8	7.7	HP,LP		95
U	Isomerization Unit	0.2	12.6			7,496	3
V	Fuel gas treatment and sulphur recovery	1.6	22.2	-20.2	HP		
W	Gasoline Blending						
X	Steam Methane Reformer	1.3	16.1	-0.9	HP		
	Total	**28**	**407**	**39**		**129,268**	**388**

The direct CO_2 emissions from this reference refinery total approximately 0.8 MMT/yr. Fractionation, alkylation, reforming, hydrotreating, hydrocracking, coking, isomerization, and fluid catalytic cracking operate at the temperatures listed in Table ES2.

Table ES2. Operating conditions of the refinery process units.

Process	Temperature	Pressure	Reaction Type
Crude Oil Atmospheric Fractionation	340–355°C	1.5–1.9 bar	NA
Crude Oil Vacuum Fractionation	340–355°C	28–30 mmHg absolute	NA
Alkylation	10–15°C	—	Exothermic
Reforming	510–526°C	9–10 bar	Endothermic
Hydrotreating	300–350°C	40–100 bar	Exothermic
Hydrocracking	300–500°C	80–204 bar	Exothermic
Fluid Catalytic Cracking	527–593°C	1.7–2.0 bar	Endothermic
Coking	485–505°C	0.7–2.0 bar	Endothermic
Isomerization	250–300°C	35–41 bar	Endothermic

The majority of the high-temperature heat is sourced from the combustion of refinery fuel gas, which is a byproduct off-gas that would otherwise be of very limited practical use. As a result, the emphasis is on supplying power, heat, and hydrogen via high-temperature steam electrolysis. When providing all three commodities, the reference plant would require three or four high-temperature gas-cooled reactor Xe-100 units, depending on the required net electricity for import and export.

Methanol is a chemical with versatile applications across a variety of industries, both as a feedstock and as a standalone product. The major downstream products derived from methanol are acetic acid, formaldehyde, and dimethyl ether. The derivative chemicals can be used to produce plastics, adhesives, and olefins, such as ethylene and propylene, and for the production of synthetic fuels, such as gasoline, diesel, and jet fuel. The annual methanol demand in the United States is forecasted to reach 8.4 MMT by 2027. The annual direct CO_2 emissions from the methanol sector total 2.85 MMT [2]. The methanol plants in the United States are energy intensive due to the high temperatures required for the endothermic reaction that produces synthesis gas (syngas) feedstock. These plants are heat integrated to recover the heat from the reforming process furnace and exothermic methanol reactor and to maximize plant energy efficiency. The nuclear integration potential may be limited for these conventional methanol plants. However, future plants (already commercialized at a smaller scale) capable of hydrogen production utilizing nuclear energy or other low-carbon power for high-temperature steam electrolysis can foster the decarbonization of the methanol production process.

Figure ES2 gives the overall mass and energy flow diagram for a modeled steam methane reforming methanol plant. The energy and mass metrics for each stream are provided in Appendix D. The energy requirements are provided in data sheets developed for the methanol plant. The data sheet for the methanol plant is shown in Table ES3. The CO_2 emissions from the reference methanol plant total ~464,072 MT/yr.

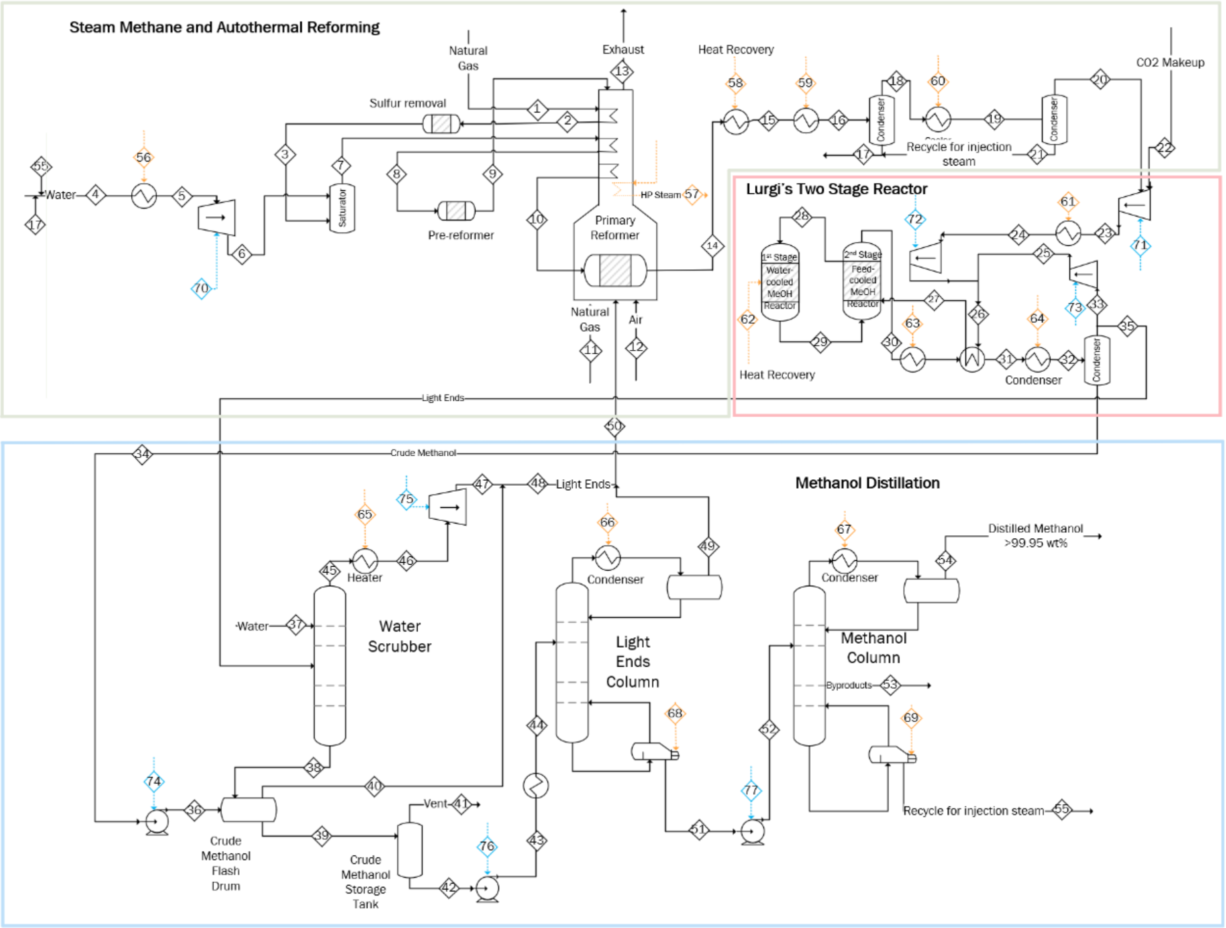


Figure ES2. Steam methane reforming methanol plant overall PFD.

Table ES3. Methanol plant overall energy requirements.

PO #	Process Unit	Power Consumption	Fuel heat content (LHV)	Steam Consumption	Steam Generation	Steam Quality (production)	Steam Quality (consumption)	Heat Released to Ambient
		MWe	MWt	MWt	MWt			MWt
A	**Steam Methane Reforming**	0	327	121 (48 kg/s)	151	LP, MP, HP	MP	64
B	**Lurgi's Two Stage Reactor & methanol cooling**	32			78	MP		80
C	**Methanol Distillation**	-3	-152	102			LP	94
E	**Process steam and power system**	-32		34	28		MP, LP	0
	Total	-2	175	257	257			238

Manufacturing of pulp and paper requires large amounts of low-pressure steam to digest and wash wood fibers and to press dry pulp into paper. Most of the low-pressure steam is taken from extraction or backpressure turbines that produce power from the high-pressure steam. High-pressure steam is generated from burning wood waste material; bark is burned in hog boilers and lignin is boiled in the black liquor recovery boiler. In a typical integrated pulp and paper mill, 50%–100% of the steam is produced from these sources, and additional steam is produced in a natural gas, fuel oil, or coal boiler. The other energy-intensive process in the plant is the lime kilns, which require high temperatures from natural gas combustion to convert the compounds needed for the chemical recovery process. An NPP could replace the steam and electricity generated from wood waste and additional fuel enabling lignin and bark to be processed into biobased chemicals or fuels. The annual CO_2 emissions from the pulp and paper sector total 96.6 MMT/yr [2].

This study evaluated the energy demand and heat material balances for the oil refining, methanol production and pulp and paper industries to assess integration with the energy production of SMNRs. Since SMNR integration with each of these industries is being considered, the inherent hazards associated with them and their operations—hazards that could impact the siting of an SMNR near an industrial manufacturing facility—were also considered.

Table ES4. Pulp and paper overall energy requirements.

PO #	Process Unit	Power Consumption	Heat from Fuel Combustion	Steam Consumption	Steam Quality	Byproduct Fuel Source	Heat of Combustion of Byproducts (LHV)
		MWe	MWt	MWt			MWt
A	Wood Processing	1.4					
B	Hog Fuel Boiler	2.0				bark & fines	23.8
C	Pulp Plant	8.6		29.8	LP, MP		
D	Bleaching	0.7		22.6	LP, MP		
E	Pulp Drying	5.2		47.4	LP		
F	Evaporation Plant	1.4		32.5	LP, MP		
G	Lime Kiln + Causticizing	0.9	28.1				
H	Recovery Boiler	3.1		24.1	LP, MP, HP	black liquor solids	161.0
I	Wastewater Plant	2.0					
J	CHP system	-26.6	45.0	-156.0	LP, MP, HP		-185.0
	Total	**-1**	**73**	**0**			**185**

Along with establishing a reference plant and identifying the utility requirements for nuclear integration for the three industries, this report also investigates strategies for nuclear integration. Table ES5 highlights possible options for each industry to integrate with advanced nuclear reactors, nuclear capacity needed and decarbonization potential. Nuclear integrated decarbonization is carried out in a step-wise manner, with the simplest cases requiring "drop-in" replacement of fossil fuel produced utilities (steam, heat, and electricity). More advanced tiers involve replacement of fossil fuel derived hydrogen to nuclear-electrolytic hydrogen, and, further, changes to the process to further support nuclear decarbonization. Brief discussion of the options presented in the table are included below, with details provided in Section 5.

Table ES5. Summary of step-wise decarbonization via nuclear integration by industry.

Industry		Drop-In	Process Change
Oil & Gas Refining	Case Study	Case 2/2a - Figure 33	Case 3/3a - Figure 34
	Nuclear Integrations	-NG CHP replaced with nuclear steam and electricity -SMR replaced with nuclear electrolytic hydrogen -Optional electrification of compressor drives to eliminate external HP steam demand (Case 2a)	-All previous integrations -Extract C3-C4 paraffin cut of fuel gas and sell as liquified petroleum gas -Make-up lost combustion heating by blending nuclear-electrolytic hydrogen into the fuel gas (Case 3a) or electrifying some of the fired heaters (Case 3)
	National Decarbonization Potential	52% reduction. Using nuclear steam and electricity has potential to remove 95 MMT/yr of CO_2e, while nuclear-electrolytic hydrogen can eliminate 68 MMT/yr of CO_2e when compared to SMR without CCS	Further 17% reduction. Diverting the LPG cut and replacing that combustion heating duty with clean hydrogen or electricity reduces national potential by 53 MMT/yr of CO_2e
	Reactor Thermal Power (scaled to reference plant)	605-631 MWth per 100 kbbl/day of crude oil processed	1102-1203 MWth per 100 kbbl/day of crude oil processed
	Further decarbonization opportunities	Fuel gas conversion to other products (market dependent)	CO_2 capture from refinery fuel gas combustion: pre-combustion, oxyfiring, post-combustion
Methanol	Case Study	Case 1/1a - Figure 40	Case 2 - Figure 42
	Nuclear Integrations	-Fuel switching: the SMR furnace has nuclear-electrolytic hydrogen substituted to back out natural gas usage (Case 1), or is electrified (Case 1a)	-Syngas is produced using the reverse water gas shift reaction. -Nuclear electrolytic hydrogen is used in a two-step process to convert an industrial CO_2 source to methanol
	National Decarbonization Potential	93% reduction. Substituting SMR furnace heating with a clean option can eliminate 2.6 MMT/yr of CO_2e emissions	557% reduction (CO_2 negative). 12.8 MMT/yr of CO_2 that would otherwise be emitted from other sources could be converted to methanol using nuclear-electrolytic hydrogen
	Reactor Thermal Power (scaled to reference plant)	367-488 MWth per 1 MMT/yr methanol product	2713 MWth per 1 MMT/yr methanol product
	Further decarbonization opportunities	CO_2 capture: pre-combustion, oxyfiring, post-combustion	Capture CO_2 from furnace and recycle to carbon feedstock
Pulp & Paper	Case Study	Case 1 - Figure 46	Case 2 - Figure 47
	Nuclear Integrations	-Electrify lime kiln	-Decommission NG auxiliary boiler and hog boiler

Industry		Drop-In	Process Change
		-Decommission NG auxiliary boiler -HTGR provides electricity for plant and electric lime kiln, and MP steam lost from existing CHP by decommissioning natural gas auxiliary boiler	-Convert recovery boiler and lime kiln to oxy-fuel combustion -Nuclear-HTSE with CO_2 sweep is used to produce oxygen for oxy-fuel combustion -Byproduct hydrogen sold at market
	National Decarbonization Potential	100% reduction of non-biogenic emissions. 73.2 MMT/yr of CO_2e emissions may be avoided nationally by replacing NG consumption with nuclear cogeneration	100% reduction of biogenic and non-biogenic emissions. With CCS, capturing particularly biogenic emissions, pulp mills can act as a delayed form of direct-air capture.
	Reactor Thermal Power (scaled to reference plant)	104.5 MWth per 400 kMT/yr pulp product	925.5 MWth per 400 kMT/yr pulp product
	Further decarbonization opportunities	CO_2 capture on lime kiln and recovery boiler: pre-combustion, oxyfiring, post-combustion	Lignin precipitation or other advanced processing could sequester or divert carbon in a non-volatile form

CO_2 emissions from the reference refinery can readily abated via simple utility substitution, where nuclear power presents a lower carbon-intensity alternative. This is notably achieved by using nuclear steam and electricity rather than importing electricity and/or operating natural gas fired CHP units. Further overall reductions are made by cleaning up the hydrogen supply chain. Much of the hydrogen consumed at the reference refinery to crack and upgrade certain fractionation streams is derived from natural gas through SMR. In practical refineries, this hydrogen is typically provided by a third-party provider over-the-fence, but a switch to nuclear-electrolytic hydrogen (whether owned and operated by the refinery or the third party) will have a great impact on the overall carbon footprint of the refinery. With utility switching completed, the remaining emissions primarily come from burning refinery fuel gas (RFG – a collection of light gases that have little market value) and regenerating the catalyst in the fluidized catalytic cracking unit. In markets where liquified petroleum gas (LPG) is a viable product, propane, butane, and isobutane can be extracted from these light gases after desulfurization. This reduces the amount of hydrocarbons that must be combusted on-site, and makes room for further nuclear decarbonization, as electrified furnaces or electrolytic hydrogen can be brought in to displace the caloric value lost to the new product. The remaining RFG is burned all over the refinery in a variety of process heaters, boilers, and furnaces. This would pose a logistical challenge to the operator if post-combustion CO_2 capture were proposed, as this would require a new system to combine all flue gases or a distributed network of scrubbers with a new system to bring the loaded sorbent to the regenerator. Because of this, pre-combustion CO_2 capture would be attractive, where the RFG is essentially reformed into hydrogen, isolating the GHGs at a single point. Some of the caloric value of the RFG would be consumed at the reformer rather than in the furnaces, presenting additional nuclear opportunity. The reference refinery uses less energy per barrel than the “average” refinery, so emission reductions at the reference refinery do not necessarily scale directly to the national potential. The reference refinery is still a useful tool, as energy efficiency upgrades would likely be made to ensure optimal operations before making the large investment of nuclear integrated decarbonization.

Thes integration cases with the refinery show that an NPP can provide up to 100% of considered external energy inputs. This excludes eventual auxiliary systems that may be added or are specific to each refinery configuration, such as an oxygen or nitrogen supply. Additionally, HP steam demand can be minimized, such that LP steam delivery may be sufficient in specific cases. In this case, a rather small amount of external demand for MP and HP steam, eventually for short-term situations such as startup operations, etc., can be supplied by steam compression or backup boilers. The analysis of reactor thermal power required per unit of product also illustrates the energy conversion efficiency between nuclear heat and all refinery demands supplied in each scenario. The best efficiency was found for Cases 1 and 1a, where the nuclear power plant delivered steam and power through the refinery's CHP systems. Efficiency decreased when supplying hydrogen (e.g., Cases 2 and 3a)—the majority of losses comes from the electricity needed for electrolyzers. Electrification (e.g., Case 3) has similar efficiency values to the hydrogen cases. The least efficient case was Case 3a, where hydrogen was supplied to the refinery as a fuel.

The reference methanol plant requires a different strategy for minimally invasive nuclear decarbonization. All the electricity and steam required at the plant are satisfied by advanced heat integration of intermediate and product cooling, so the only energy input to the process is fuel burned in the furnace. Fuel switching of this furnace is the main approach to lower the carbon intensity of the utilities, either by partial electrification of the furnace or by blending nuclear-electrolytic hydrogen into the fuel gas. As with the refinery, however, imperfect separations, suboptimal selectivity, and conversion limitations necessitate the combustion of some light gases, all of which have fossil origin. The advanced nuclear integration opportunity involves the conversion of an industrial CO_2 source to methanol, driven by nuclear-electrolytic hydrogen. This not only eliminates the carbon footprint of the methanol plant by making its entire energy profile nuclear-derived, but also makes the plant a negative emission facility.

These integration cases for the methanol plant show that an NPP can provide up to 100% of considered external energy inputs and can substitute feedstock NG with nuclear-generated hydrogen. The energy conversion efficiency was highest for Case 1a where the NG fuel demands from the SMR process are met by electrification. Case 1, where the demands are met by hydrogen, was less efficient. Case 2 using RWGS was the least efficient case, because of the losses associated with making and using hydrogen as both a feedstock and a fuel.

The reference kraft pulp mill contains a robust CHP system that recovers heat from byproducts (black liquor and hog fuel) and is augmented by a natural gas boiler. This offers a simple low-tier stepwise decarbonization plan, where the SMNR steps in to provide the auxiliary steam while also producing excess electricity to power the newly electrified lime kiln. In this way, all of the non-biogenic emissions can be abated, except for those that may come from flame stabilization in the recovery boiler. The remaining biogenic emissions are attractive for nuclear integrated CO_2 capture, as this is essentially an indirect, yet highly concentrated form of direct air capture. This is particularly true where oxy-fuel combustion based carbon capture is employed. The nuclear plant produces oxygen by electrolysis, resulting in a byproduct hydrogen stream – a valuable commodity to be produced alongside captured biogenic CO_2. This extensive nuclear integration is an attractive option for energy park applications, where a synfuels plant would be built next door, converting biogenic emissions and nuclear-electrolytic hydrogen into carbon neutral liquid fuels, crosslinking the efforts for stepwise decarbonization in the industrial and transportation sectors.

In the first scenario (Case 1) studied for a kraft pulp mill, a simple nuclear integration replaces all externally needed energy, displacing fossil fuels. Even though the mill is rather large, the demand of nuclear-sourced energy is less than a single nuclear reactor module. Case 1 is the most thermally efficient integration between the two cases. On the other hand, the complex integration approach (Case 2) using oxy-fuel combustion for carbon capture doesn't replace the fuel for the lime kiln but does capture all CO_2 emissions. This case is much less efficient, and the overall energy balance results in a large export in Case 2 due to the energy content of the hydrogen byproduct. As the system has carbon capture integration and most of the biogenic emissions are captured along with the lime kiln emissions, the system becomes CO_2 negative.

As the United States pursues its decarbonization goals, innovative integration methods will be needed to reduce the environmental impact of these significant industries. Replacing industrial carbon-emitting energy sources with integrated SMNRs will enable direct thermal substitution within processes, “green” electrification, and the production of green chemical processes to produce an identical product. This report will aid researchers and industry in determining how necessary energy streams may be replaced with clean energy sources from SMNRs, giving options for several levels of decarbonization.

CONTENTS

FIGURES

TABLES

ACRONYMS

ASU	air separation unit
BLRB	black liquor recovery boiler
CHP	combined heat and power
EIA	Energy Information Administration
EPA	Environmental Protection Agency
FCC	fluid catalytic cracking
FLIGHT	Facility Level Information on Greenhouse Gases Tool
GHG	greenhouse gas
HP	high pressure
HTGR	high-temperature gas-cooled reactor
HTSE	high-temperature steam electrolysis
LP	low pressure
LPG	liquefied petroleum gas
MP	medium pressure
NG	natural gas
NPP	nuclear power plant
PADD	Petroleum Administration for Defense District
PFD	process flow diagram
RFG	refinery fuel gas
SMNR	small modular nuclear reactor

Page intentionally left blank

Thermal Integration of Advanced Nuclear Reactors with a Reference Refinery, Methanol Synthesis, and a Wood Pulp Plant

1. INTRODUCTION

Though most decarbonization efforts have focused solely on electricity generation, electric power only accounts for 25% of the total U.S. greenhouse gas (GHG) emissions. In 2021, the industrial sector was the third largest source of direct U.S. GHG emissions (at 23%), just behind the transportation sector (at 28%). If electric power emissions are categorized by end use, industry accounts for 30% of emissions [3]. In 2022, industrial sector GHG emissions totaled 1,393 million metric tons (MMT). By 2050, this is expected to decrease by 7% to 1,282 MMT.

The industry sector consumes a large portion of energy in the form of electricity, heat, and steam. In 2022, the total U.S. energy consumption was ~106 EJ (100 quads), of which industry accounted for ~35 EJ (33 quads) [4]. The U.S. Energy Information Administration (EIA) projects that U.S. energy consumption will total about 104 EJ (98 quads) in 2027, with the industrial sector accounting for about 36 EJ (~34 quads) (i.e., 35%) of that total. Annual energy consumption in the industrial sector is expected to increase by 17% to 40.8 EJ (38.5 quads) by 2050, bringing energy use in the industrial sector to over 7% of the total U.S. energy consumption. Many low-carbon technologies for producing electricity are now available, but new technologies for clean fuel, heat, and steam generation must be developed to decarbonize the other sectors. Integrating nuclear power with industry represents a special opportunity, as unlike some clean energy technologies, nuclear power is a low-carbon source of both heat and electricity.

Achieving net-zero status in the United States by 2050 will require an additional ~550–770 GW of additional clean, reliable power. According to the report “Pathways to Commercial Liftoff: Advanced Nuclear” [5], this will necessitate over 200 GW of new nuclear capacity. Despite the deployment of renewable energy sources, system-level decarbonization modeling suggests that nuclear power remains one of the most feasible clean energy options, having proven itself at a large scale (i.e., around 100 GW worth of nuclear reactors currently operate in the United States, most constructed between 1970 and 1990). However, to meet national energy demands and foster the deep decarbonization of heavy industries, nuclear must also support applications beyond electricity production.

Figure 1 shows how various nuclear energy pathways interact within multiple industries to expand the role of nuclear energy to the generation of commodities. While many of these use cases stem from nuclear power’s firm, reliable thermal generation capabilities, the integration of nuclear power requires a deep evaluation of current industrial energy requirements and pathways across all industries.

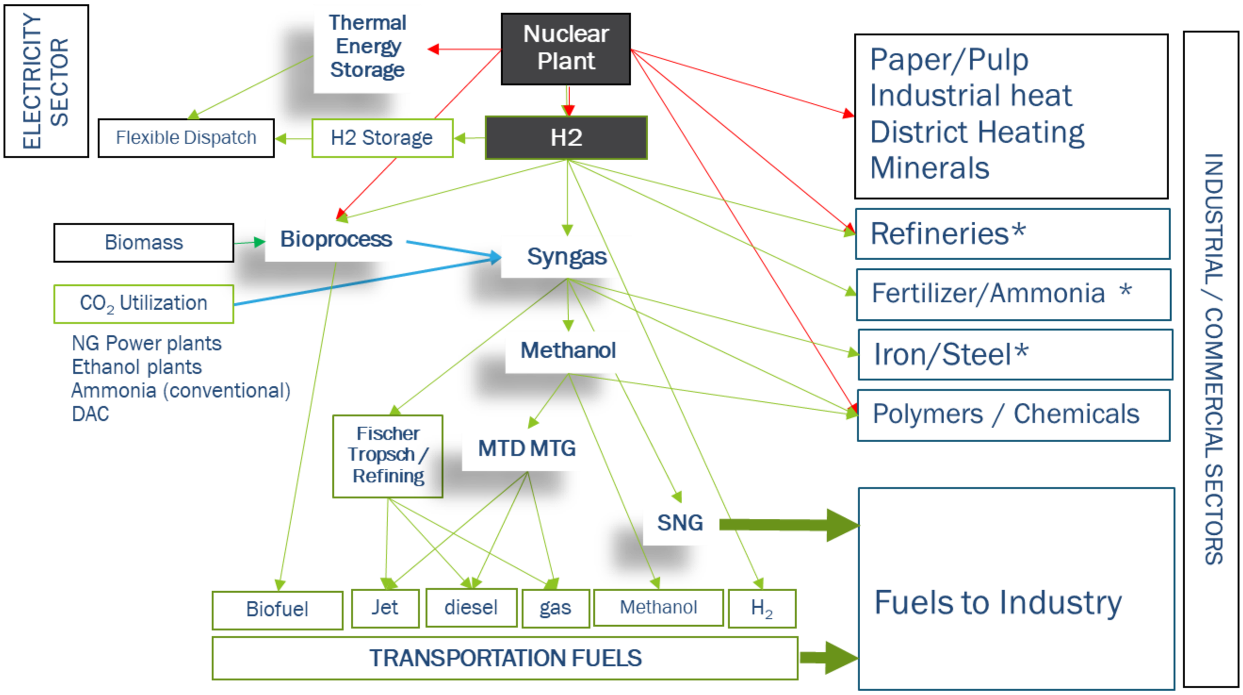


Figure 1. Diagram demonstrating additional nuclear applications and pathways by sector.

This work aims to support the Department of Energy Office of Nuclear Energy's vision for its Integrated Energy Systems Program by assessing the potential of nuclear-generated energy to displace petroleum, coal, and natural gas (NG) across various sectors of the economy. Building on prior efforts that led to the identification of five candidate industries [6]—ammonia production, refining, chlor-alkali production, methanol synthesis, and pulp and paper mills—the focus of this work is to provide, at the individual process level, an in-depth analysis that prioritizes oil refineries, methanol production, and pulp and paper plants.

To better understand the decarbonization and nuclear integration potential, this work compiles, for each industry studied, comprehensive information and original assessments pertaining to:

1. **Nationwide plant-level production capacities**, distributions, locations, and emissions.
2. **Process flow diagrams** (PFDs) of reference plants, detailing the main process unit operations and the energy and material flows.
3. Overall **balance data sheets** detailing each of the **main process unit operations**, along with the corresponding electric power consumption, heat demand from fuel combustion, steam consumption, steam generation, steam quality, heat loss, hydrogen demand, heating value of byproducts, and CO_2 emissions.
4. Identification of **multiple levels of nuclear integration opportunities**, which includes providing process steam, power, high temperature heat and hydrogen demands, eventually switching feedstock or including carbon capture. Specific scenarios are developed for a reference plant, going from drop-in energy source replacement to complex integrations with modified internal industrial processes.
5. Evaluation and assessment of the results of each **nuclear integration scenarios**. Each integration scenario is demonstrated with quantified performance parameters for an HTGR type SMNR plant.
6. Estimation of **national potential for nuclear integration** into the industries based on selected technically feasible integration scenario.

2. PETROLEUM REFINERIES

Figure 2 shows a PFD of a typical upgrading refinery, along with the refining unit operating conditions. The associated PFDs can be found in Appendix D. The overall energy, steam, electricity, and heat demands for each process unit are summarized in Table D. The unit numbers in the PFD are referenced in the data sheets provided in the appendices. The mass balances and energy inputs are also provided in Appendix D.

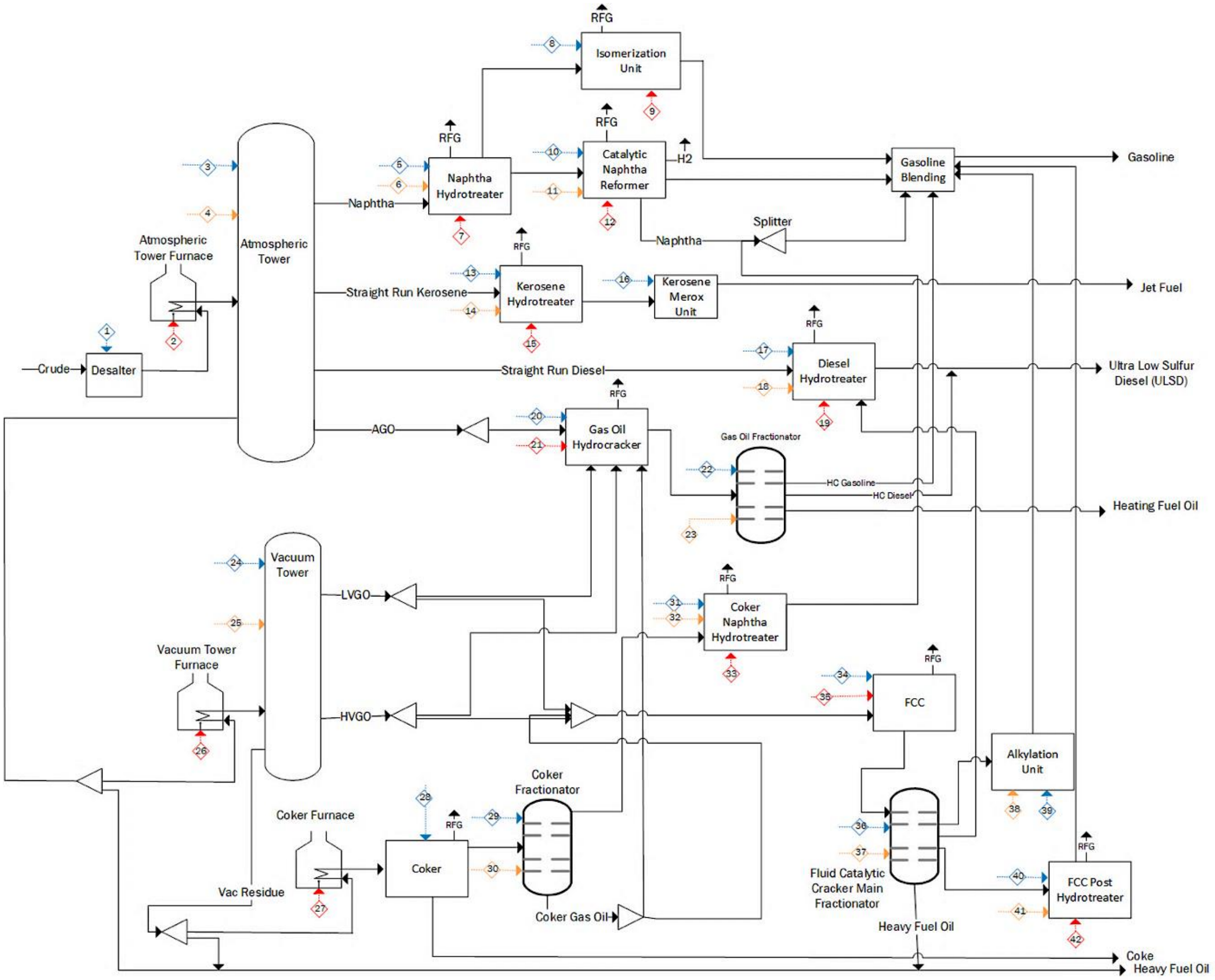


Figure 2. Refinery overall energy PFD.

A reference refinery is established in Section 2.4, which has emissions and utility requirements detailed in a block flow diagram (BFD) in Figure 3. Combustible gases are processed and mixed with NG to provide combustion heating for the plant and steam methane reforming (SMR), which itself provides the hydrogen required for crude oil upgrading. Some unit operations feedback process steam that can be used to heat other operations, as detailed in Section 2.2.6. The remaining process steam and electrical duties are satisfied with an NG-fired combined heat and power (CHP) system.

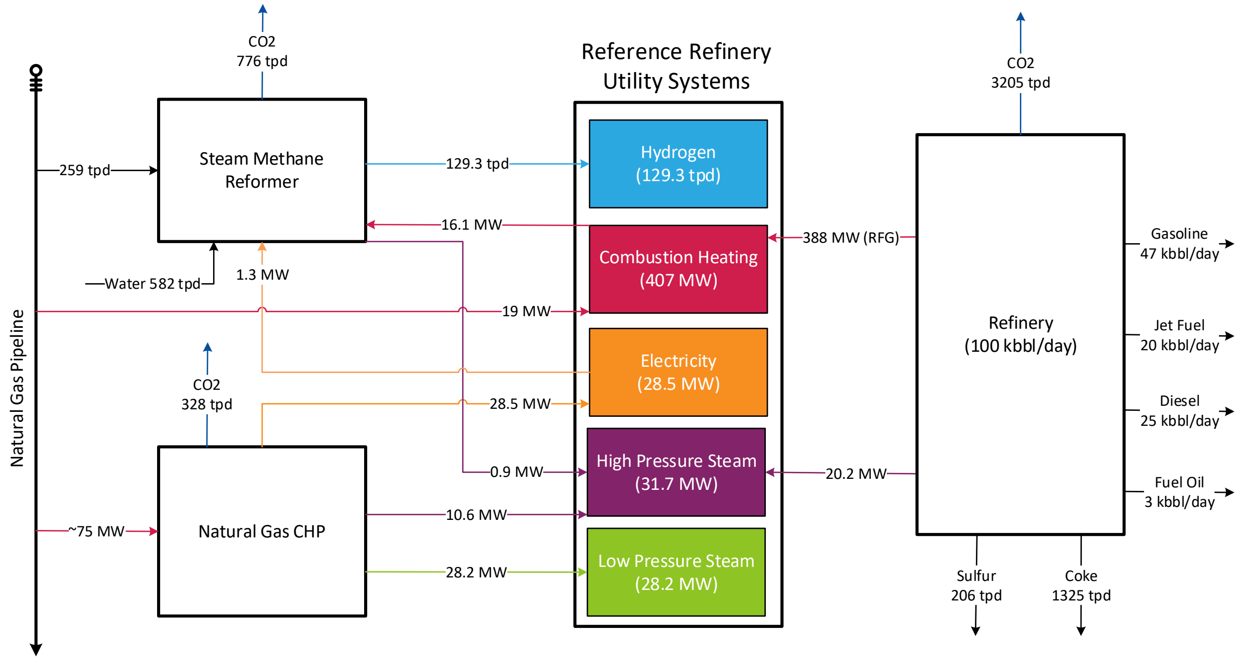


Figure 3. Utility system BFD for reference refinery.

2.1 Refinery Energy System

The refining systems require energy input, as shown in Figure 4. Heat is supplied from steam and the combustion of NG and refinery gas. Refinery gas is a byproduct of the refinery cracking operations consisting of light hydrocarbons, such as methane, ethane, ethylene, propane, and propylene. The energy from combustion is used for power and steam production and to separate hydrocarbon streams via fractionation.

The steam supplied to the refinery is produced through auxiliary boilers, heat exchange with refinery streams, and combined heat-power units. A typical refinery steam system encompasses three steam grades: high pressure (HP; 2.7–4 MPa, with 100–200°C of superheat), medium pressure (MP; 0.96–1.2 MPa), and low pressure (LP; 0.1–0.27 MPa). These various steam levels are employed for the following services:

- HP steam: To power turbines for compressors and blowers, provide reboiler heat for distillation columns, and high-temperature applications
- MP steam: To power turbines, provide reboiler heat for distillation columns, and enable other heat exchange applications
- LP steam: To provide reboiler heat for distillation columns, boiler water conditioning, and low-level heat for auxiliary applications

A typical refinery steam system is given in Figure 4. The refining and upgrading of crude oil into finished gasoline, diesel, and jet fuel products are discussed below.

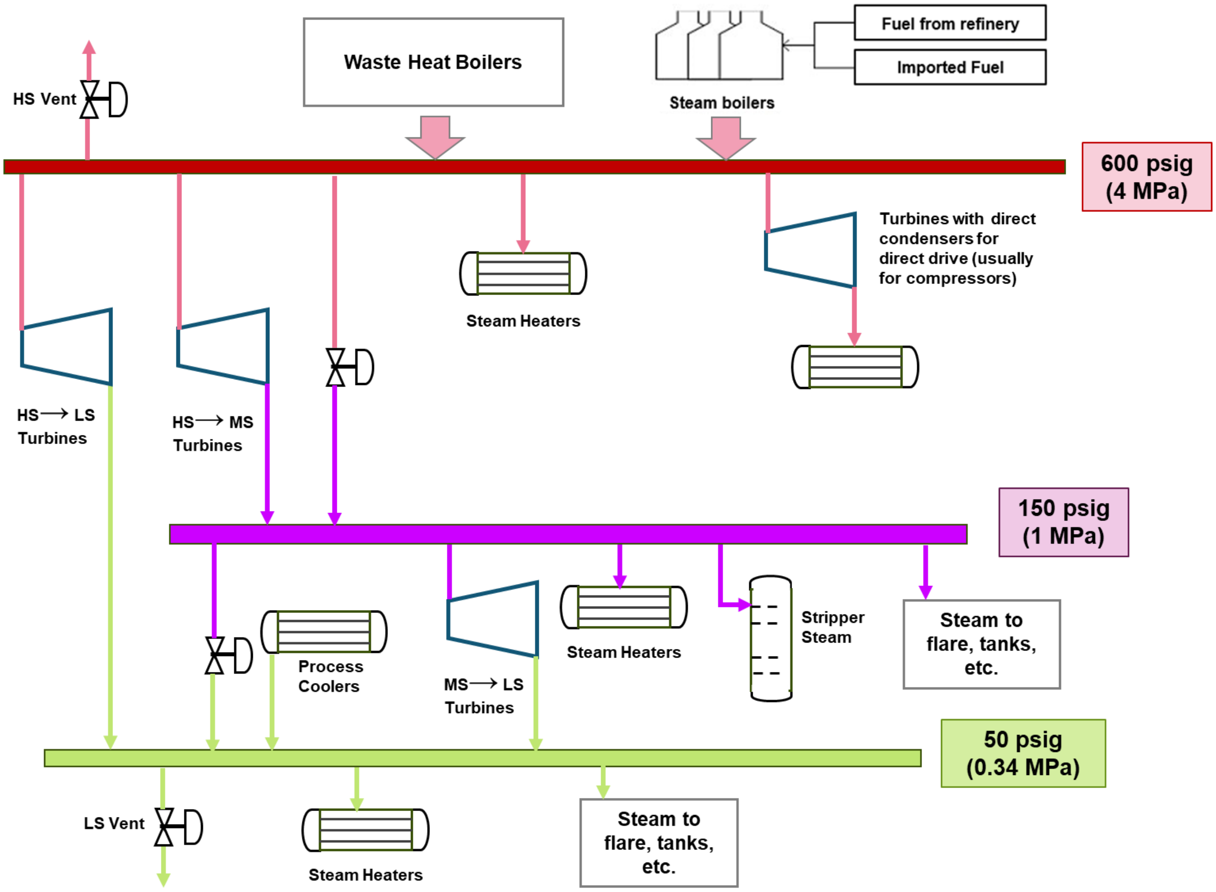


Figure 4. Generic refinery steam system.

2.2 Refining Foundational Processes

2.2.1 Desalting

Crude oil desalting is the first step of the refining process. Crude oil contains salt water, drilling chemicals, sediment, and other chemicals that must be removed before further processing the crude oil. The chemical components result from oil drilling and production. The water content stems from crude oil recovery and the salt water contained in oil formations. The sediment, which is mostly sand, comes from the oil formation and the transport tanker vessels. The crude oil refining process utilizes a series of water contact and electrical separation systems to remove residual water.

2.2.2 Fractionation

Fractionation separates components based on their boiling temperatures (i.e., crude oil and other intermediate streams) in fractionation and distillation columns. Heat is supplied to the columns by fired furnaces to enable the separation of lighter components from the heavier boiling fractions. The separated streams are sent for product blending or further conversion and upgrading in other refining units. The first step of fractionation in all refineries is an atmospheric distillation unit for the initial stream separation. In more complex refineries, a vacuum distillation column may be included to remove and upgrade heavier petroleum fractions in the refinery to higher-value products.

2.2.3 Upgrading and Conversion

Following the fractionation step, heavy, lower-value distillation fractions can be further processed into lighter, higher-value products, such as gasoline, diesel fuel, and other finished products. The most widely used conversion method is called cracking because it uses heat, pressure, catalysts, and sometimes hydrogen to “crack” (or reduce) heavy hydrocarbon molecules into lighter ones. Cracking processes upgrade these lower-value streams into lighter components that can be blended to meet the product specifications. Some cracking processes use hydrogen to catalytically reduce the heavy hydrocarbon streams. The heat energy generated in these cracking units is recovered using preheat heat exchange systems and steam generators. A cracking unit consists of one or more reactors and a network of furnaces, heat exchangers, and other vessels.

Petroleum refineries vary in complexity but include these basic steps to produce these products. In general, refining is an energy-intensive process. Intermediate oil streams require conversion and upgrading steps consisting of both endothermic and exothermic reactions. Table 1 lists different refining operations and their respective operating conditions.

Table 1. Operating conditions for select refinery unit operations.

Process	Temperature	Pressure	Reaction Type
Atmospheric Fractionation	340–355°C	1.5–1.9 bar	NA
Vacuum Fractionation	340–355°C	0.037-0.04 bar (28–30 mmHg)	NA
Alkylation	10–15°C	—	Exothermic
Reforming	510–526°C	9–10 bar	Endothermic
Hydrotreating	300–350°C	40–100 bar	Exothermic
Hydrocracking	300–350°C, 500°C	80–204 bar	Exothermic
Fluid Catalytic Cracking	527–593°C	1.7–2.0 bar	Endothermic
Coking	485–505°C	0.7–2.0 bar	Endothermic
Isomerization	250–300°C	35–41 bar	Endothermic

Note: Hydrocracking was not included in this refinery upgrading scheme.

Cracking is not the only form of crude oil conversion; other refinery processes rearrange and combine molecules rather than splitting them to add value. Alkylation is an exothermic process that creates gasoline components by combining the small gaseous byproducts of cracking into higher-carbon-number molecules in the range of the fuel blend stock. This process takes place in a series of large, horizontal vessels and tall, skinny towers. Alkylation creates gasoline components by combining light hydrocarbons in the presence of sulfuric or hydrofluoric acid.

The key cracking and upgrading unit in the refinery is the fluid catalytic cracking (FCC) unit, which breaks long-chain hydrocarbons (e.g., heavy gas oil) into shorter chains such as naphtha. The endothermic cracking reactions occur at temperatures of 480–540°C and are fueled by the heat produced during catalyst regeneration. About 5% of the feed ends up as coke; thus, air is used in the regenerator to burn the coke off. The regenerator exit temperatures are 925–815°C. A fractionator column downstream of the FCC separates and recovers the hydrocarbon vapors. Reforming is an endothermic process that uses heat supplied by fired furnaces, moderate pressure, and catalysts to turn naphtha—a 65–190°C boiling range stream with a low octane—into high-octane gasoline components.

2.2.4 Contaminant Removal

Crude oil contains contaminants (e.g., sulfur and nitrogen) that must be removed from the resultant component streams before they can be blended into the products. Hydrotreating and hydroprocessing are employed for this purpose, raising the streams to a high temperature and pressure in the presence of hydrogen and catalysts. These processes are exothermic and generate excess heat that is either recovered and efficiently used throughout the process or is integrated with other refining units.

2.2.5 Stream Quality Improvement and Blending

Several extraneous refinery streams possess characteristics that prevent them from being blended into fuels. In particular, straight-chain alkane hydrocarbons such as pentanes and hexanes have a low octane value (~60 octane). An isomerization process unit is used to upgrade the octane value of these streams. The octane value is improved by utilizing high temperatures, moderate pressure, and catalysts to reach an adequate octane blending value of 75–92. The component streams from the refining units are combined to meet the specifications for gasoline, diesel, and jet fuel products.

2.2.6 Heat Recovery

Refineries require that high-temperature heat be delivered to various systems. Many refining units operate at high temperatures of 200–800°C. Heat is efficiently recovered from these systems to generate various levels of steam and system preheat. Fired furnaces supply preheat to the hydroprocessing units and reboil heat to the distillation columns.

2.3 U.S. Refining Capacities and Distribution

According to the U.S. EIA report from June 2023 (Q2), [7] the United States has a total of 129 petroleum refineries, of which 125 are currently operable. These refineries are distributed throughout the country, based on their corresponding Petroleum Administration for Defense Districts (PADDs). The PADDs serve as regional groupings of the 50 U.S. states and the District of Columbia, dividing them into five primary districts, as shown in Figure 2. Due to the substantial population it contains, PADD 1 is further divided into three sub-PADDs. Two additional PADDs (i.e., VI and VII) not indicated in the map in Figure 5 are dedicated to U.S. territories. PADDs are essential for enabling nationwide petroleum data users to assess regional petroleum product supplies. Table 2 gives a comprehensive distribution of U.S. petroleum refinery production per both PADD and state. For a more detailed description of each refinery's capacity by company, see Appendix A-2.

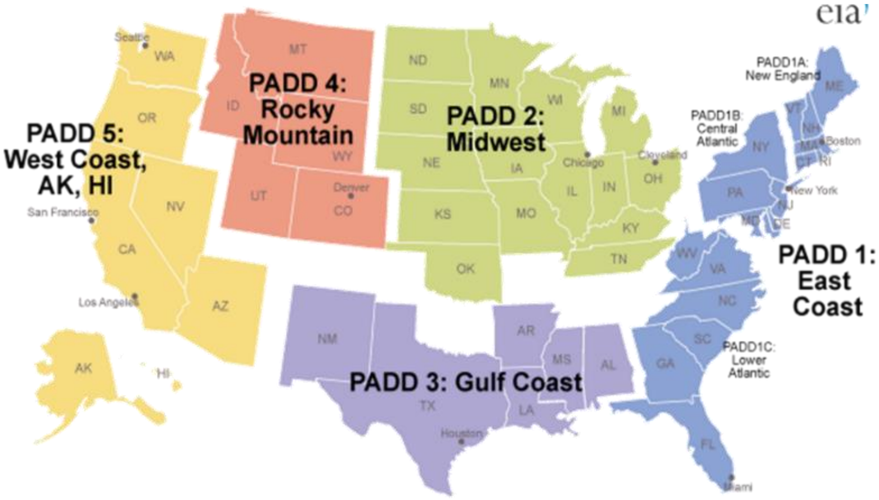


Figure 5. Five primary PADD districts [8].

Table 2. U.S. refinery capacities per both PADD and state [7].

	Operable Refineries (#)			Barrels per Calendar Day			Average Barrels/Refinery		
PADD #	*Total*	*Operating*	*Idle*	*Total*	*Operating*	*Idle*	*Total*	*Operating*	*Idle*
PADD I	7	7	0	877,800	877,800	0	125,400	125,400	0
PADD II	25	22	3	4,206,105	3,948,885	25,7220	168,244	179,495	85,740
PADD III	56	56	0	9,676,729	9,676,729	0	172,799	172,799	0
PADD IV	15	13	2	650,164	537,564	112,600	43,344	41,351	56,300
PADD V	26	26	0	2,649,571	2,643,571	6,000	101,907	101,676	0
U.S. Total	129	124	5	18,060,369	17,684,549	375,820	1,400,023	142,617	75,164

As noted in Table 2, five refineries have shut down in PADDs II and IV. There were two refineries shut down in PADD II: one due to economic reasons and one was converted into a biodiesel plant. In PADD IV, three refineries were shut down: two for economic reasons and one refinery was converted into a biodiesel plant [8].

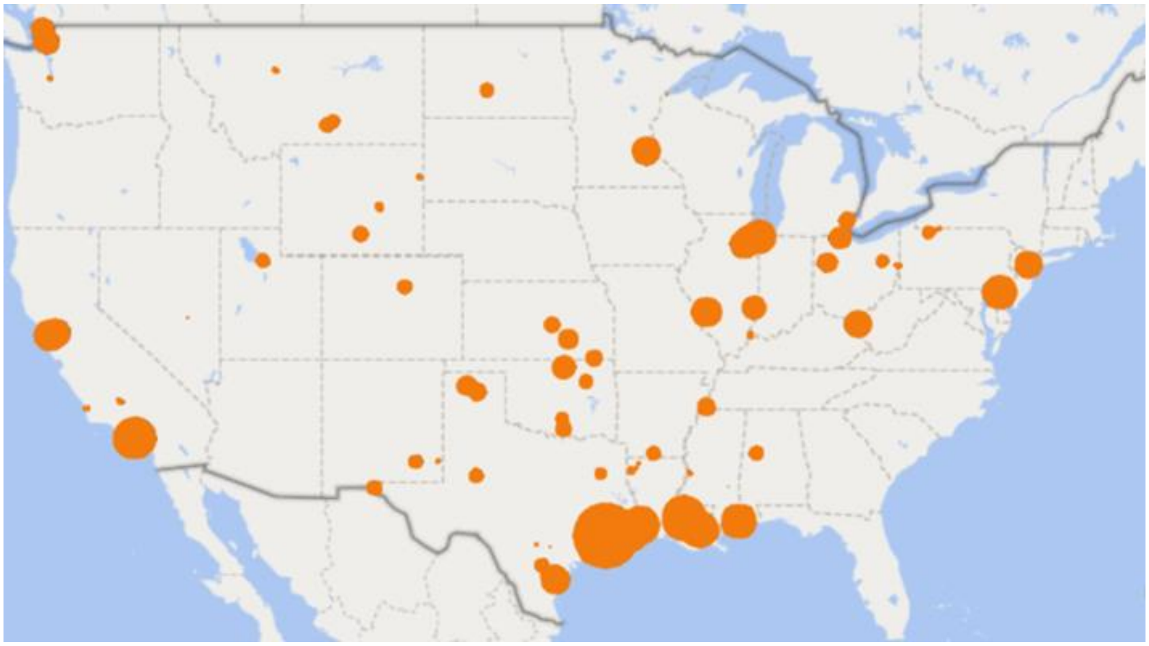


Figure 6. U.S. petroleum refinery distribution.

Figure 6 shows the distribution and relative capacity of the U.S. petroleum refineries, which are critical for meeting the country's energy and petrochemical product needs. They vary in size and capacity, with some being among the largest in the world. The conclusions reached from the refinery distribution data are summarized as:

- Texas has the largest number of refineries in the United States (32), with facilities in cities such as Houston, Beaumont, and Corpus Christi
- Louisiana also hosts a significant number of refineries (15), primarily in areas such as Baton Rouge and Lake Charles
- California has several refineries (14), particularly in the Los Angeles and San Francisco Bay areas
- Pennsylvania has three refineries in the Philadelphia region

- Illinois (specifically the Chicago area and nearby regions) has only four refineries
- Oklahoma has five refineries in cities such as Tulsa and Ponca City.

The capacities of U.S. petroleum refineries vary from 1,700 to 626,000 barrels per day (BPD), or barrels per calendar day. The total U.S. refining capacity exceeds 18 million BPD (18,060,369 BPD). The following are some of the largest refineries in the United States, along with their capacities:

- Motiva Enterprises (Port Arthur, Texas) has a capacity of over 600,000 BPD
- ExxonMobil Baytown (Baytown, Texas) has a capacity of around 564,000 BPD
- Marathon Petroleum (Garyville, Louisiana) has a capacity of approximately 596,000 BPD
- BP p.l.c. (Whiting, Indiana) has a capacity of over 430,000 BPD
- Chevron Pascagoula Refinery (Pascagoula, Mississippi) has a capacity of around 356,440 BPD
- The Phillips 66 Company (Westlake, Louisiana) has a capacity of around 264,000 BPD
- The CITGO Corpus Christi Refinery (Corpus Christi, Texas) has a capacity of roughly 157,000 BPD.

Capacities may change over time due to upgrades, closures, expansions, or changes in ownership. To obtain a reference plant capacity and estimate heat and energy duties, average capacity values are used. The average and mean capacities for the total number of refineries in the United States were calculated at 143,300 and 102,500 BPD, respectively. The median falls above 49% of all the refinery capacities, with 60% of the refinery plants being below the average capacity. Figure 7 shows the refinery capacity distribution. The reference plant capacity selected was 100,000 BPD.

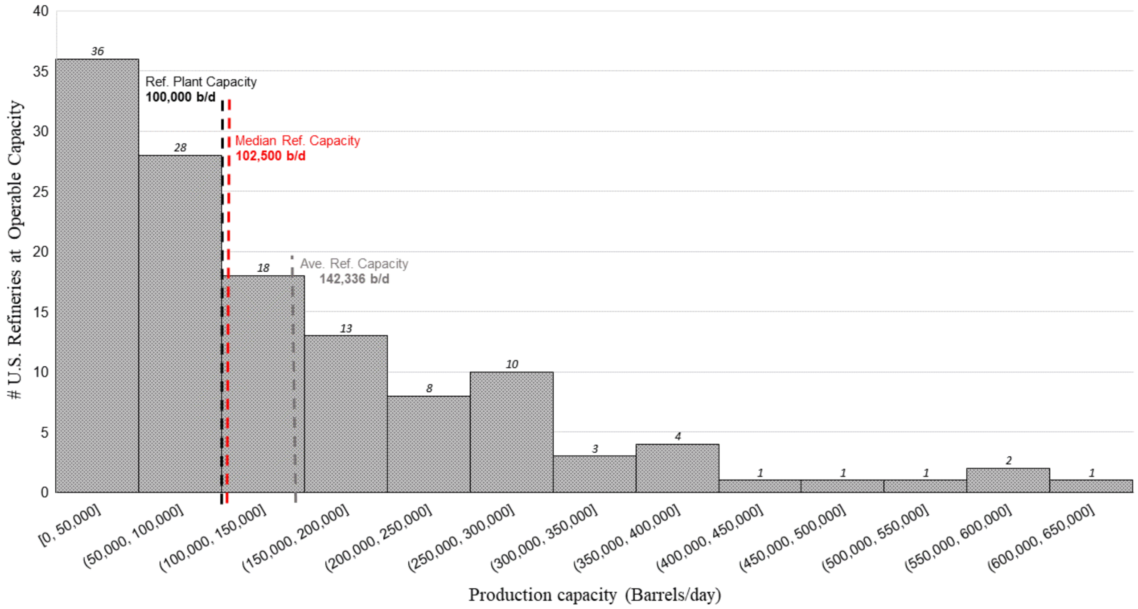


Figure 7. Nationwide refinery capacity distribution chart.

Refineries produce a range of products, and their production mix may be adjusted based on market demand and economic factors. The total production landscape of refineries can change due to market conditions, economic factors, and regulatory changes. The EIA industry report regularly updates information on refinery capacities to reflect the most current data. In 2023, the carbon emission totals for each U.S. refinery were taken from the Environmental Protection Agency (EPA)’s Facility Level Information on Greenhouse Gases Tool (FLIGHT) [1]. The emission totals reported in FLIGHT pertain to direct emissions not lifecycle emissions.

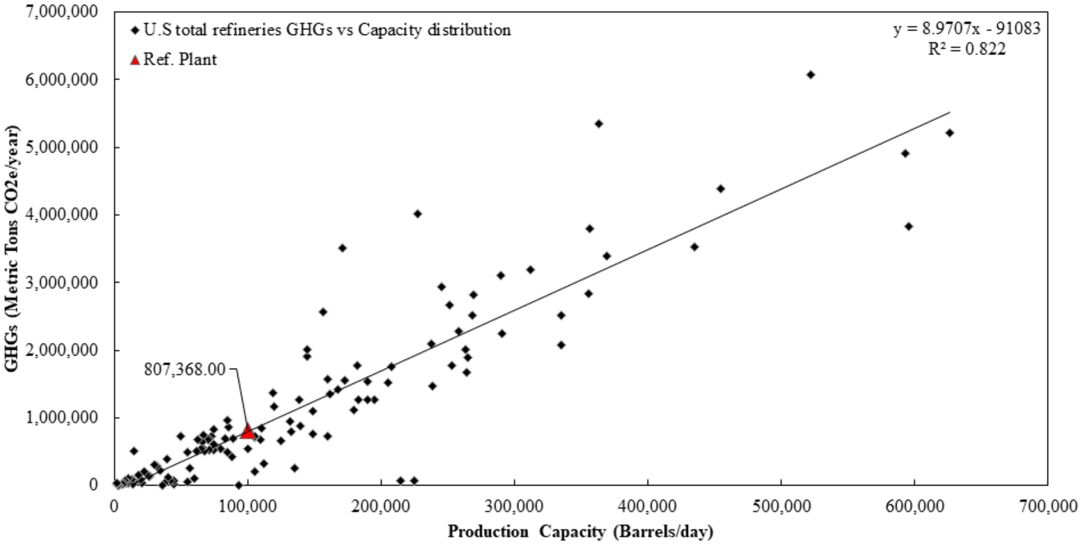


Figure 8. GHGs vs. production capacity correlation for U.S. refineries.

2.4 Reference Refinery Plant

A wide range of refinery configurations exist for fostering crude assay, desired products, utilization of byproducts to satisfy energy demands, and implementation of energy integration systems, such as combined heat and power systems and waste heat recovery.

2.4.1 Configuration, Size, and Assumptions

The reference refinery, a coking-type refinery with deep conversion, uses Arab Medium Stratiev assay crude oil. A reference capacity of 100,000 bbl/day was selected. A specific open-source refinery model, along with expert assessment, were employed to obtain the reference plant results.

2.4.2 Reference Process Details

The refinery's energy and material balance and GHG emissions were assessed using the open-source, Excel-based Petroleum Refinery Life Cycle Inventory Model (PRELIM). PRELIM [9] is a mass- and energy-based process-unit-level tool for estimating the energy use and GHG emissions associated with crude-oil processing in a refinery. It informs decision-making and policy analysis by affording a transparent model that includes data, assumptions, and detailed results. The model draws on an extensive crude oil library and estimates the energy and utility usage and resultant GHG emissions at the refinery process-unit-level and overall refinery level.

Because most U.S. refineries are complex residual upgrading refineries, this configuration was selected as the reference refinery in the PRELIM architecture. PRELIM builds a refinery configuration and chooses a mode of operation based on the selected complexity option and the crude slate. The flowsheet-produced model refinery configuration includes FCC, alkylation, reforming, hydrotreating, and coking process units. The model outcome and results can be used to assess a refinery and its nuclear power plant (NPP) integration. PRELIM's preset reference capacity of 100,000 bbl/day was used for the present work.

The refinery design selected for the reference refinery configuration with the PRELIM model was a complex refinery that includes heavy oil upgrading with FCC, coking, and the associated hydrotreating process units. This configuration emulates the U.S. refinery configuration that has evolved to accommodate the world crude oil supply. The quality of the crude oil has steadily become heavier, which yields higher volumes of high boiling point liquids that require upgrading to gasoline, jet fuel, and diesel fuel from a barrel of crude oil. Over 60% of the U.S. refining capacity currently includes heavy oil upgrading FCC and coking processing capacity [8].

2.4.3 Overall Energy and CO_2 Balance

The energy demands of the reference refinery are based on PRELIM, which provided the power and heating duty demands of the HP, MP, and LP steam; the heating duty from fuel combustion; the hydrogen demand; and the production of refinery fuel gas (RFG). Table 3 summarizes these energy demands—along with the RFG supply—for the reference refinery. A negative steam consumption value denotes that steam is produced in the process unit. The reference refinery is a coking-type refinery with deep conversion, and it operates using Arab Medium Stratiev assay crude oil. Different input conditions would change the results. For example, the high sulfur content of this crude oil carries a significant fuel input demand and increases steam production from the flue gas treatment and sulfur recovery unit. A lower sulfur content would cause both the fuel input and steam production to decrease from over 20 MW to just a few MW.

The electric power requirements of a refinery usually exceed PRELIM's calculations. Power demands of actual refineries of the same size range from 52 to 100 MWe, whereas PRELIM calculates 28 MWe. However, a refinery would likely increase its energy efficiency before integrating with an NPP.

The steam quality listed in Table 3 deviates in some ways from the PRELIM model. First, MP steam was substituted for HP steam. There is only a small demand for MP steam at the refinery (less than 2 MW), and this steam is used for driving compressors. However, HP steam can also be used for this purpose. Second, the coker fractionator requires superheated LP steam, but because the demand is only 1 MW, the superheat was neglected.

Table 3. Overall energy requirements for the reference refinery.

PO #	Process Unit	Power Consumption	Heat from Fuel Combustion	Steam Consumption	Steam Quality	Total Hydrogen Requirement	Total RFG for Onsite Use
		MWe	MWt	MWt		kg/d	MW
A	Desalter	0.1					
B	Atmospheric Tower Furnace		70.1				
C	Atmospheric Tower	3.8		11.3	LP		
D	Vacuum Tower Furnace		19.8				
E	Vacuum Tower	0.6		7.4	LP		
F	Naphtha Hydrotreater	1.9	27.1	2.0	HP	8,396	2
G	Kerosene Hydrotreater	2.5	36.6	2.7	HP	9,647	3
H	Kerosene Merox Unit	0.1					
I	Gas Oil Hydrocracker	5.0	22.9			83,799	30
J	Gas Oil Hydrocracker Fractionator	1.4		19.4	HP		
K	Diesel Hydrotreater	3.1	29.9	2.1	HP	12,464	2
L	Coker Furnace		47.0				
M	Coker	1.7					211
N	Coker Fractionator	0.4		1.0	Superheated, LP		
O	Coker Naphtha Hydrotreater	0.8	11.4	0.8	HP	3,678	1
P	Fluid Catalytic Cracking Feed Hydrotreater	0.0	0.0	0.0	HP	0	
P	Fluid Catalytic Cracking Post Hydrotreater	1.0	14.6	1.2	HP,LP	3,789	2
Q	Fluid Catalytic Cracker	0.5	22.8				40
R	Fluid Catalytic Cracker Main Fractionator	0.4		0.4	HP,LP		
S	Alkylation Unit	0.2		3.8	LP		
T	Catalytic naphtha reformer	1.9	53.8	7.7	HP,LP		95
U	Isomerization Unit	0.2	12.6			7,496	3
V	Fuel gas treatment and sulphur recovery	1.6	22.2	-20.2	HP		
W	Gasoline Blending						
X	Steam Methane Reformer	1.3	16.1	-0.9	HP		
	Total	**28**	**407**	**39**		**129,268**	**388**

Detailed information on the use of each type of steam in the process units was required to determine feedwater temperatures and to model letdown turbines in the final process model. Each process unit's primary type of steam utilized was identified to determine its outlet condition (return inlet). For process units that PRELIM identified as having two different steam pressure levels, a 50:50 distribution of each steam grade was assumed. The outlet stream of the LP steam used for driving ejectors was assumed to be comparable to the outlet stream of the steam for process heating. Process steam is typically produced at saturated conditions, but to match the steam supply header parameters, it was assumed to be superheated from combustion heat. Table 4 summarizes the steam and return water parameters and conditions.

Table 4. Steam and corresponding return water parameters for the reference refinery.

	P (kPa)	T (°C)	h (kJ/kg)	x (-)
HP steam	4,238	300	2,953	superheated
LP steam	377	141	2,735	saturated vapor
HP condensate	4,238	239	1,032	subcooled
LP condensate	377	126	531	subcooled
Turbine condensate	10	46	192	saturated liquid
Make-up	100	20	84	subcooled

Figure 9 shows a simplified process steam model that groups together all steam uses of the same type. This model shows the additional steam flow to the deaerator (at 105°C), which isn't included in the overall energy demand as it is a regenerative heating. For this reference plant, the breakdown of steam demand and internal production and conversion to steam flowrates for each process is shown in Table 5. Note that, due to high internal HP steam production, the net demand from external source for reboilers is zero and return condensate for producing steam is thus at the deaerator's temperature (here 105°C). The HP steam is then directly used only by turbines for the direct drive of equipment as large compressors. Remaining HP steam pressure is reduced to the LP steam level, while producing about 2.7 MWe in a letdown turbine. Since the LP steam is assumed to be saturated, only a portion of the letdown steam passes through a turbine, while a smaller portion goes through a reduction valve to satisfy the desired LP steam header quality. Overall, the demands are distributed into three types: direct drive turbines (only for HP steam, typically a single shaft turbine with large compressors), injection into a process (LP, return as make-up water), and heating of reboilers (HP, LP) with returning condensate subcooled by 15 K.

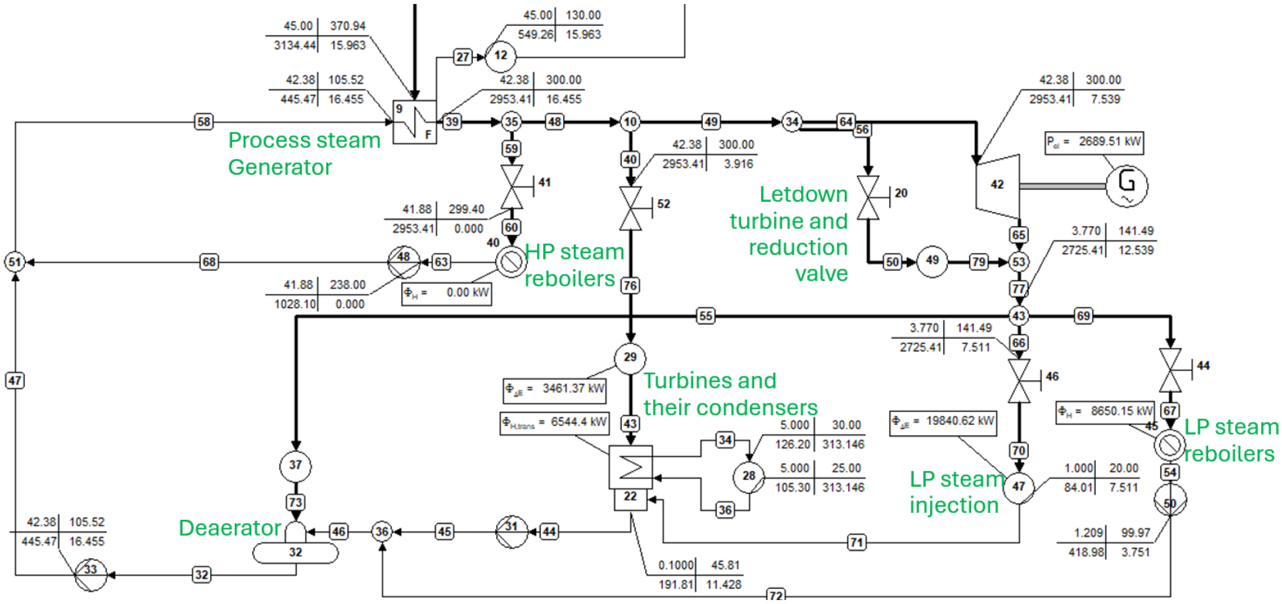


Figure 9. Simplified process steam system for the reference refinery.

Table 5. Process steam parameters for each process in the reference refinery.

Process Unit	Steam HP	Steam LP	Steam use	HP - turb	HP - reb	LP - inj	LP - reb
	MW	MW		kg/s	kg/s	kg/s	kg/s
Atmospheric Tower		11.3	Injection			4.27	
Vacuum Tower		7.4	Injection			2.81	
Naphtha Hydrotreater	2.0		compr. Dr.	0.73			
Kerosene Hydrotreater	2.7		compr. Dr.	0.99			
Gas Oil Hydrocracker Fractionator	19.4		Compr. Dr., reboil.		10.07		
Diesel Hydrotreater	2.1		Compr. Dr.	0.76			
Coker Fractionator		1.0	Injection			0.39	
Coker Naphtha Hydrotreater	0.8		compr. Dr.	0.31			
Fluid Catalytic Cracking Feed Hydrotreater			compr. Dr.				
Fluid Catalytic Cracking Post Hydrotreater	0.6	0.6	HP comp., LP reb.	0.22			0.28
Fluid Catalytic Cracker Main Fractionator	0.2	0.2	HP comp., LP inj.	0.06		0.07	
Alkylation Unit		3.8	Reboilers				1.73
Catalytic naphtha reformer	3.8	3.8	HP compr., LP reboil.	1.39			1.74
Fuel gas treatment and sulphur recovery	-20.2		Produced	-0.55	-9.61		
Steam Methane Reformer	-0.9		Produced		-0.46		
Total	**10.6**	**28.2**		**3.92**	**0.00**	**7.53**	**3.75**

Table 6 summarizes the process lifecycle emissions (in the carbon dioxide equivalent) stemming from the reference refinery, based on PRELIM results. The emissions for each unit include grid emissions for electricity demands, net NG requirements, and RFG. The total national emissions reported are based on EPA FLIGHT and are separated in terms of the percentage of the total emissions for each corresponding process unit in the reference refinery. The emissions from SMR are included for each process unit based on hydrogen consumption.

Table 6. CO_2 emissions from the reference refinery and at the national scale.

PO #	Process Unit	Reference Plant CO2 Emissions	National Scale CO2 Emissions
		Metric Tons per Year	Million Metric Tons per Year
A	Desalter	280	0.03
B	Atmospheric Tower Furnace	138,900	14.30
C	Atmospheric Tower	37,775	3.89
D	Vacuum Tower Furnace	39,237	4.04
E	Vacuum Tower	18,506	1.90
F	Naphtha Hydrotreater	92,945	9.57
G	Kerosene Hydrotreater	119,658	12.32
H	Kerosene Merox Unit	362	0.00
I	Gas Oil Hydrocracker	354,231	36.46
J	Gas Oil Hydrocracker Fractionator	46,716	4.81
K	Diesel Hydrotreater	118,290	12.18
L	Coker Furnace	92,991	9.57
M	Coker	8,182	0.84
N	Coker Fractionator	3,812	0.39
O	Coker Naphtha Hydrotreater	40,184	4.14
P	Fluid Catalytic Cracking Post Hydrotreater	48,591	5.00
Q	Fluid Catalytic Cracker	194,692	20.04
R	Fluid Catalytic Cracker Main Fractionator	2,410	0.25
S	Alkylation Unit	9,230	0.95
T	Catalytic naphtha reformer	118,078	12.15
U	Isomerization Unit	50,957	5.25
V	Fuel gas treatment and sulphur recovery	52,980	5.45
W	Gasoline Blending	0	0.00
X	Steam Methane Reformer	0	0.00
	Total	**1,589,006**	**164**

2.5 Dynamic Operation

Refineries are large manufacturing facilities that operate continuously to refine and convert crude oil into gasoline, diesel, and jet fuel products. They consist of large process units that operate at high temperatures (200–900°C) and pressures. Normally, these process units operate at full capacity unless external factors or constraints (e.g., economic, crude oil supply, and utility constraints) require reduced feed rates.

If required, the refining process units can ramp down by 30–40% yet still maintain safe, stable operation. However, these feed rate reductions and adjustments usually take several hours to accomplish, depending on the magnitude of the change. This ramp time is required to meet unit operating temperatures and pressures, adhere to unit procedure guidelines, and prevent equipment damage.

Refineries schedule shutdowns to inspect, repair, and clean equipment and to replace catalyst systems. These shutdowns occur every 3 years or so and require 30–50 days for the identified work and activities to be accomplished. The shutdown and restart for the process units takes about 1–2 weeks, depending on the unit type. For NPP integration with a refinery, these dynamic operations, variations, and requirements during shutdown and startup will be incorporated into the overall design strategy.

3. METHANOL SYNTHESIS

Methanol is an important chemical commodity widely used in various material and fuel applications. In 2021, the global supply of methanol was estimated at approximately 107 MMT and was mainly used for olefin production, as shown in Figure 10. Since 2017, this essential compound has exhibited an annual growth rate of 5% [9] for several target markets. Common applications for global methanol production include olefins, formaldehyde, gasoline blending, methyl tert-butyl ether, and acetic acid (see Figure 11). Light olefins, specifically ethylene and propylene, are anticipated to be the fastest growing segment in the United States. This distinction is crucial due to the versatile applications of light olefins in areas such as textiles, plastics manufacturing, detergent production, antifreeze formulation, synthetic fibers, coatings, and paints.

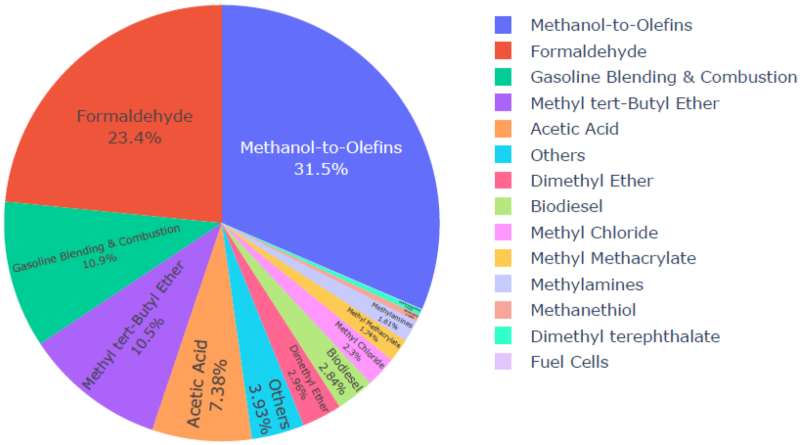


Figure 10. Global methanol demand by application for 2021 [10].

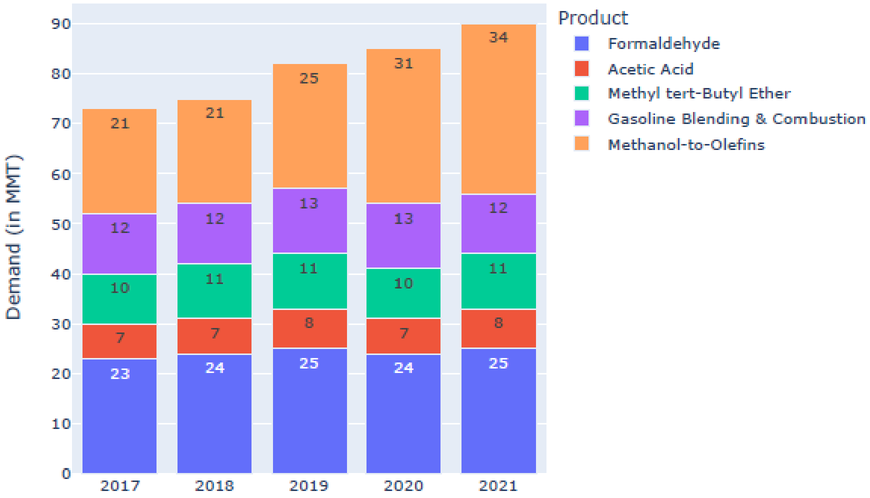


Figure 11. Target markets for methanol, based on end use (in MMT) [10].

The annual market size of methanol in the United States is projected to reach 8.4 MMT by 2027 [10]. This anticipated growth aligns with the global trend, as methanol prices have shot up in the post-pandemic years, coinciding with heightened manufacturing activity, as per the Methanol Institute's 2023 report. China, the largest methanol consumer, contributes approximately 20% of the total global volume—roughly double the amount of methanol consumed in the United States [10].

A reference methanol plant is established in Section 3.3, which has emissions and utility requirements detailed in Figure 12. Light gases are blended with NG to satisfy the combustion heating duty of the plant, while key cooling duties meet process steam and electricity requirements through a robust heat recovery cogeneration system described in Section 3.3.2.2.

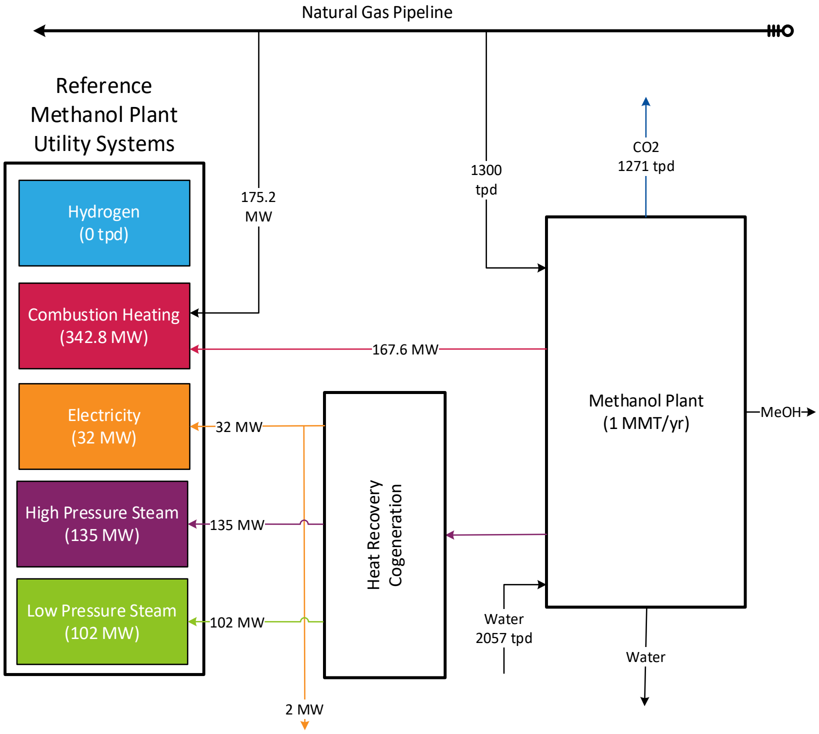


Figure 12. Utility system BFD for reference methanol plant.

3.1 Methanol Production Foundational Processes and Methodologies

Over 85% of global methanol is produced from SMR and coal gasification, though SMR accounts for approximately 65% of that ratio whereas coal gasification accounts for the remaining 35% or so [11]. In either approach, the supply of coal or NG reacts with steam, creating synthetic gas—or syngas—containing carbon monoxide (CO) and hydrogen (H_2). Syngas is catalyzed by a copper/zinc oxide catalyst to synthesize methanol [12].

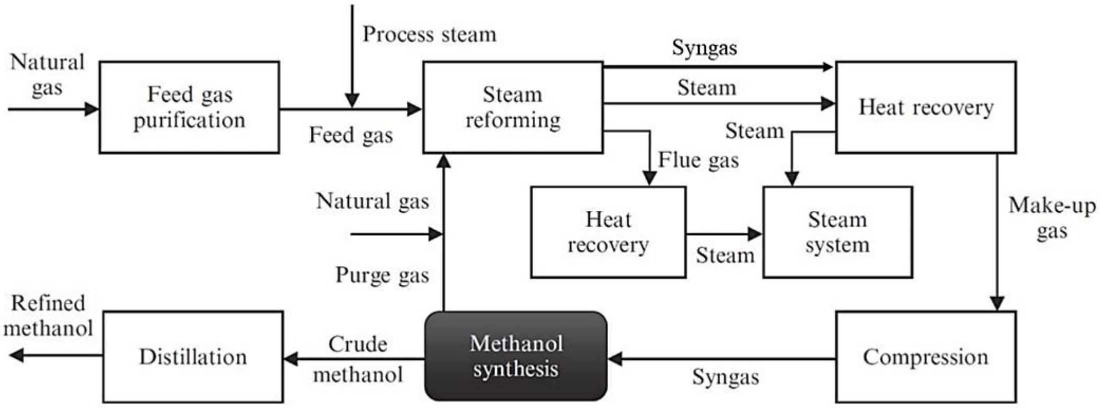


Figure 13. Conventional methane production process [13].

The motivation for an economically viable methanol production process stems from the diverse chemical production processes that rely on methanol as feedstock. Formaldehyde is a major consumer, accounting for approximately 70% of all global methanol usage. The second largest is methyl tert-butyl ether, which constitutes 20% of the annual methanol production. Other significant consumers include acetic acid and dimethyl ether. Beyond these primary applications, methanol plays a crucial role in numerous indirect applications such as paints, resins, silicones, adhesives, antifreeze, and plastics. Additionally, it can serve as a transportation fuel alongside gasoline, highlighting its versatility and importance across various industrial sectors.

3.1.1 Current Methanol Types and Production Methods

Methanol can be produced from various feedstocks, and the different methanol types are classified based on the feedstock used. Among the primary methanol types are:

- NG-Based Methanol: Derived from NG via SMR (or partial oxidation). NG-based methanol is the most common and economically significant methanol type, constituting a large portion of the global methanol production. The methane (CH_4) in NG is converted into syngas and subsequent methanol synthesis.
- Coal-Based Methanol: Produced from coal through gasification, a process that converts solid coal into syngas. Historically, this method has been very important, but today it is less commonly used, due to environmental concerns and the prevalence of NG-based production.
- Biomass-Based Methanol: Obtained from the gasification of biomass feedstocks such as wood, crop residues, and organic waste. It is considered a more environmentally friendly option because it utilizes renewable resources, though it still results in GHG release.
- Carbon Capture and Utilization Methanol: This involves capturing CO_2 emissions from combustion processes, chemical reactions, or directly from the atmosphere and other natural sources. The captured CO_2 is then combined with hydrogen—often derived from renewables—to produce methanol. If hydrogen feedstock is generated via electrolysis, the methanol becomes actual "e-fuel."

3.2 U.S. Methanol Plant Production Capacities: Nationwide Distribution and Locations

The United States has a total of nine operating methanol plants, as per the most recent publicly available information from each company [10]. As Figure 14 shows, methanol production in the United States is concentrated in the Gulf Coast region, due to the logistical benefits of that location such as access to natural gas feedstock, pipelines, and major ports for export. Specifically, Texas and Louisiana account for over 96% of all production (seven out of nine companies), with the remaining 4% or so (two companies) being produced in Tennessee. Table 7 gives a comprehensive distribution of the U.S. methanol production facilities, including their names, locations, and capacities.

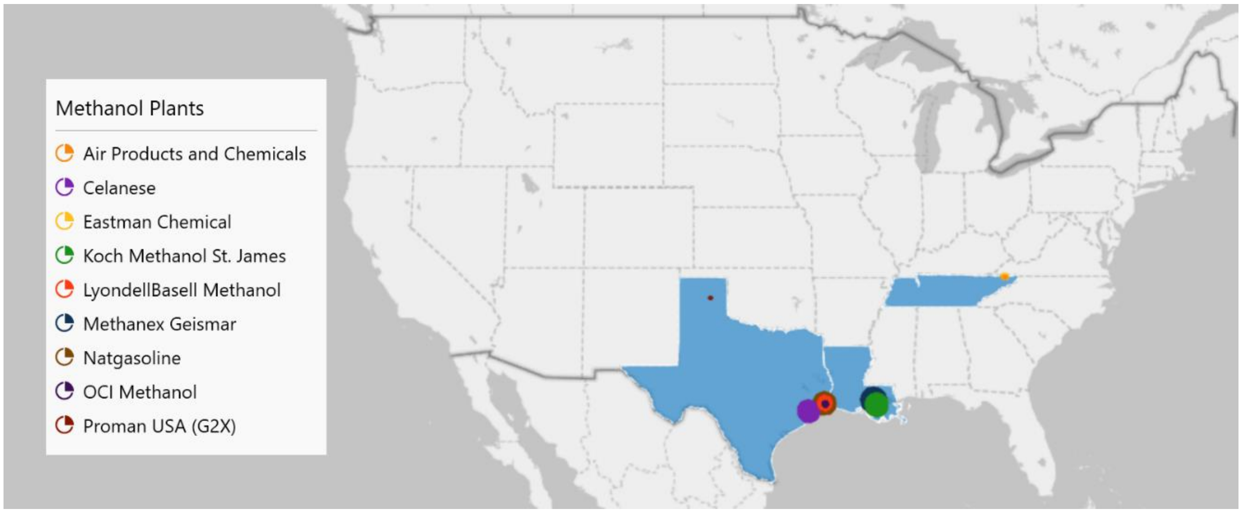


Figure 14. Methanol producers in the United States.

Table 7. Distribution of U.S. methanol production facilities [14–20].

Company	State	Site	Capacity (metric tons per year)
CELANESE [14]	Texas	Pasadena	1,620,000
PROMAN USA (G2X) [15]	Texas	Pampa	78,750
OCI METHANOL [16]	Texas	Beaumont	200,000
LYONDELLBASELL [17]	Texas	Beaumont	800,000
NATGASOLINE [18]	Texas	Beaumont	1,700,000
METHANEX [19]	Louisiana	Geismar	2,200,000
KOCH METHANOL [20]	Louisiana	St. James Parish	1,800,000
EASTMAN CHEMICAL [17]	Tennessee	Kingsport	215,420
AIR PRODUCTS & CHEMICALS [17]	Tennessee	Kingsport	96,000
			Total capacity: 8,710,170

Regarding those states that have a significant concentration of methanol facilities:

- Texas has the majority of methanol production facilities, due to the number of refineries located in the Beaumont area, and it reports about 4.4 MMT of methanol produced yearly—50% of the nation's production.
- Louisiana only has two facilities, but these facilities, Methanex and Koch, located in Donaldsville are the largest in the country, with annual production capacities of 2.2 and 1.8 MMT, respectively.
- Tennessee—specifically, Kingsport—features the nation's smallest production capacity, accounting for an annual production of just 311,420 metric tons.

In the methanol sector, production capacities for the various U.S. facilities vary throughout the years due to upgrades, complete shutdown, expansions, or ownership changes. For reference purposes, and to obtain heat and energy estimations for process modeling using Aspen Plus, the production capacities were simplified to averages. For all methanol facilities in the United States, the yearly production average is 967,800 tons and the capacity average is 800,000 metric tons. For both cases, the median and average capacities fall under 56% of the total capacity for each site and 46% over the nationwide facility capacity. These parameters and distributions are shown in Figure 15. The reference plant's capacity was selected to match the average plant size at approximately 1 MMT per year.

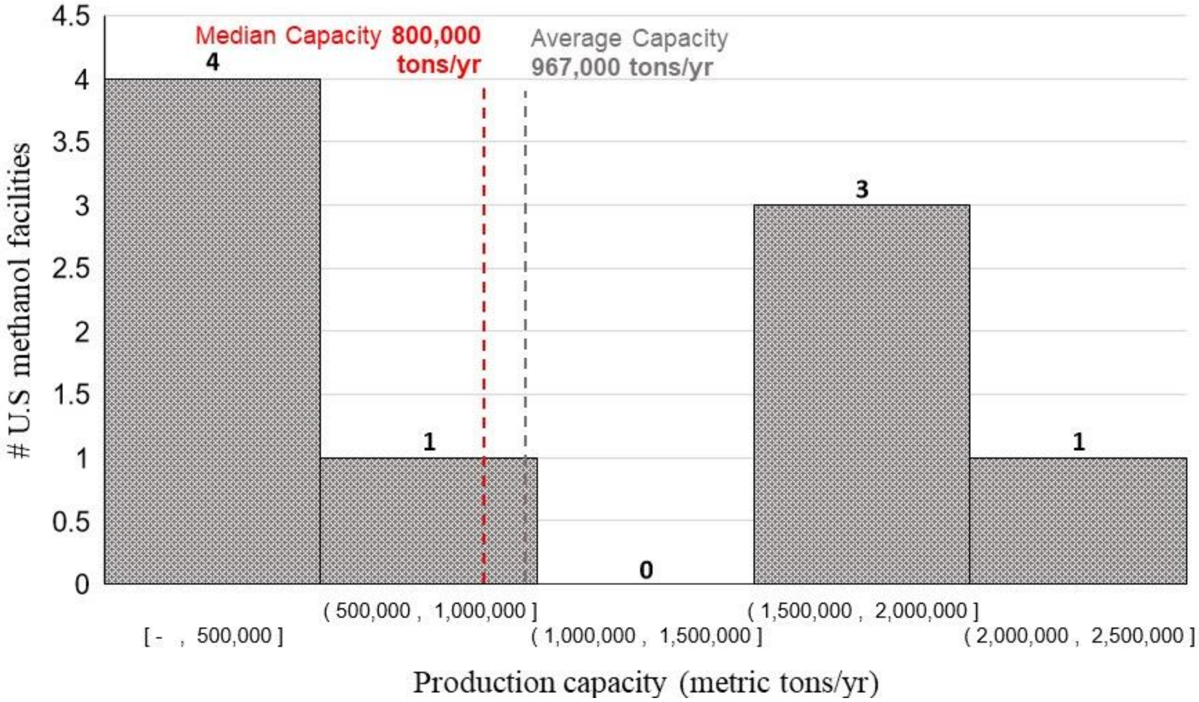


Figure 15. Nationwide methanol plant capacity distribution.

Emissions for six U.S. methanol plants were reported in the EPA FLIGHT [10]. Figure 16 shows the relationship between GHG emissions and production capacity. The chosen reference plant capacity of 1 MMT for the Aspen model corresponds to emissions totaling 464,072 MMT per year, which closely matches the average plant emissions of 476,066 MT per year.

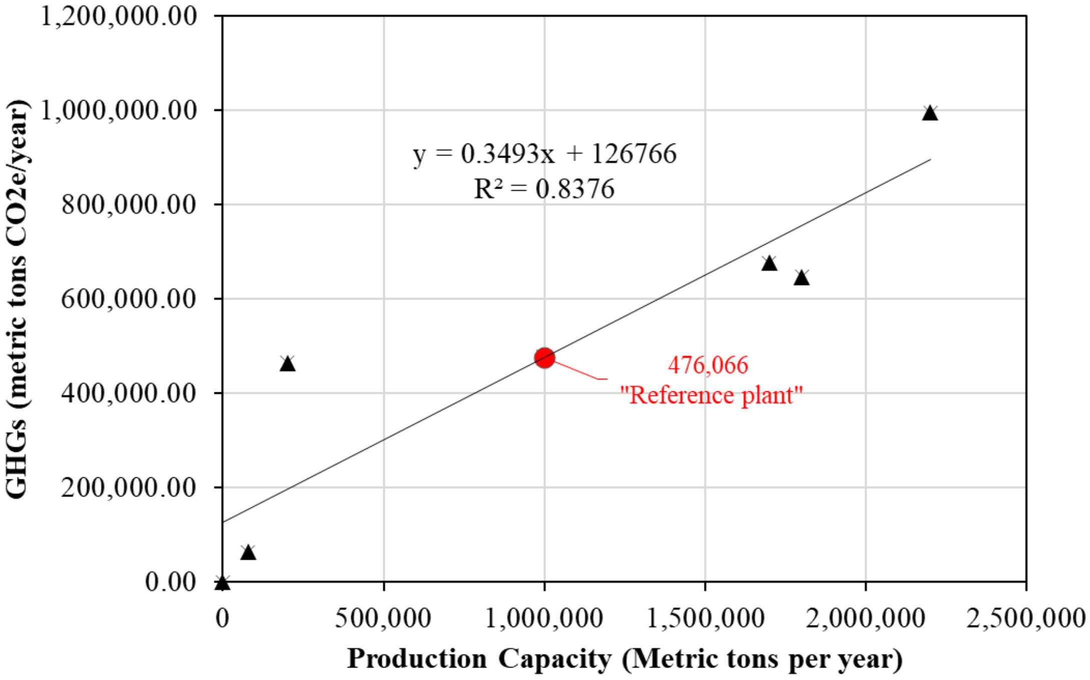


Figure 16. GHGs vs. production capacity correlation for U.S. methanol production.

3.3 Reference Methanol Plant

3.3.1 Configuration, Size, and Assumptions

The chosen reference size of the methanol plant has a production capacity of 2,725 metric ton/day of methanol, or approximately 1 MMT/year. This model was based on the combined SMR plus the autothermal reforming (ATR) process and Lurgi's Mega-Methanol technology. The plant configuration is shown in Figure 17. NG was selected as the feedstock. Two reactors were installed in series: an externally fired furnace for SMR and a reactor for autothermal reforming, with oxygen being supplied from an air separation unit (ASU). In the resulting model, pure oxygen was considered as the output of the ASU; however, for calculating the energy demands, a more reasonable value of 99% oxygen was considered (taken from [21]). Syngas produced from the combined SMR plus ATR process was compressed and fed into methanol synthesis. The first methanol reactor is water cooled to remove the heat produced via the methanol synthesis reactions, and the second reactor is feed cooled, with unheated feed gas being used as the heat transport fluid. The crude methanol stream was condensed (with some heat recovery) and purified in the stripping and distillation columns to obtain industrial-grade purity of the product. Details of the Aspen Plus model for the process depicted in Figure 17, including heat and material balance tables and power and heat integration are in Appendices B-1 through B-3.

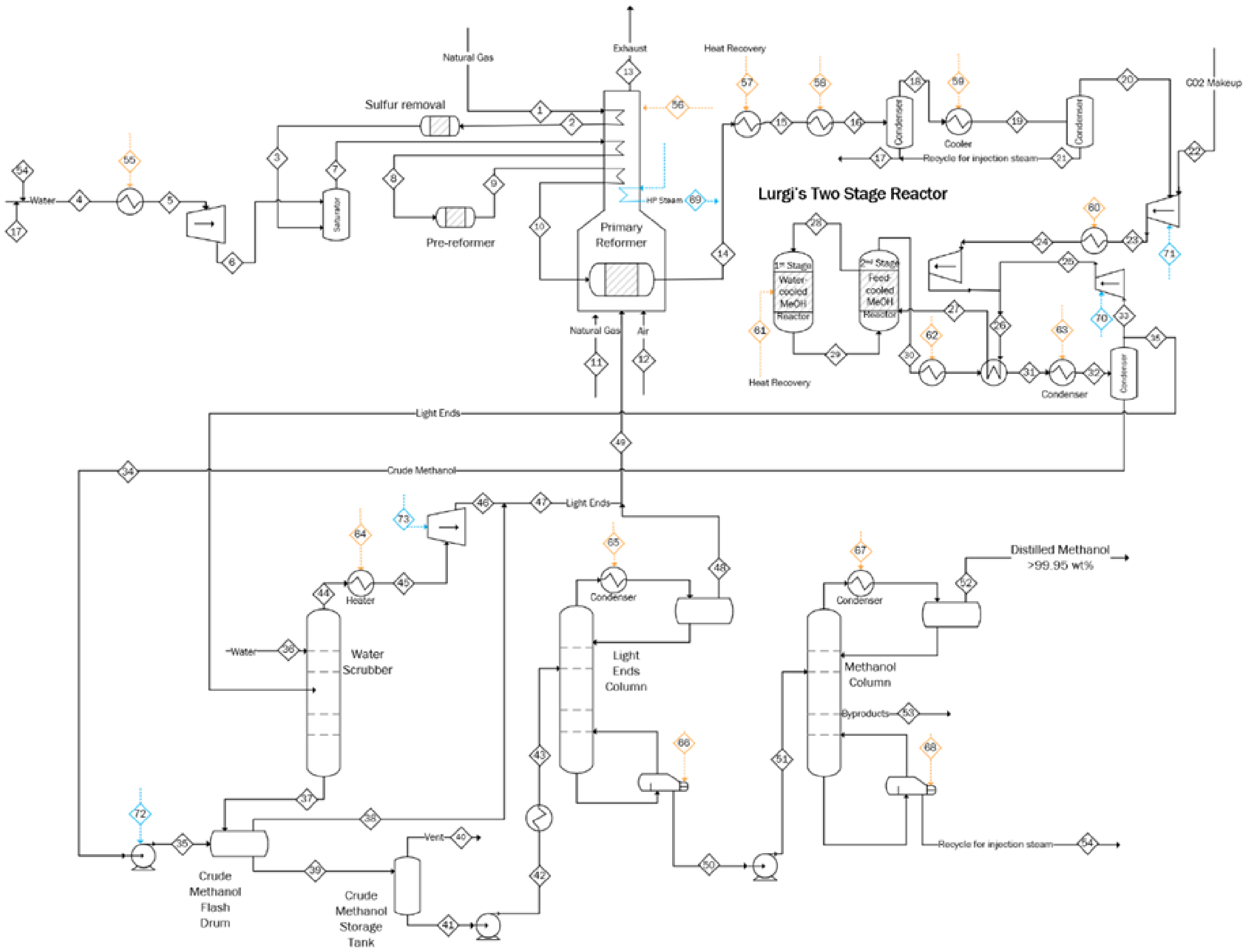


Figure 17. PFD of the reference methanol plant (excluding heat recovery and the auxiliary power production system, which were modeled separately).

The footprint of this production system is relatively compact, consisting of buildings for the steam methane reformer(s), synthesis, compressors, distillation, and process steam systems, which include turbines and pumps. There may be an adjacent water tank for fire safety. A significant footprint results from the cooling system, cooling towers (induced draft) and air coolers. Storage tanks and distribution systems for MeOH, storage for chemical additives, and a wastewater treatment plant are typically standalone systems distanced from the production system.

3.3.2 Reference Process Details

The generic reference plant for the methanol synthesis process via SMR was modeled in Aspen Plus V12, while the heat recovery and energy balance system were in Cycle-Tempo. A BFD of the process is shown in Figure 18. A detailed description of both modeling methods is presented in the following sections.

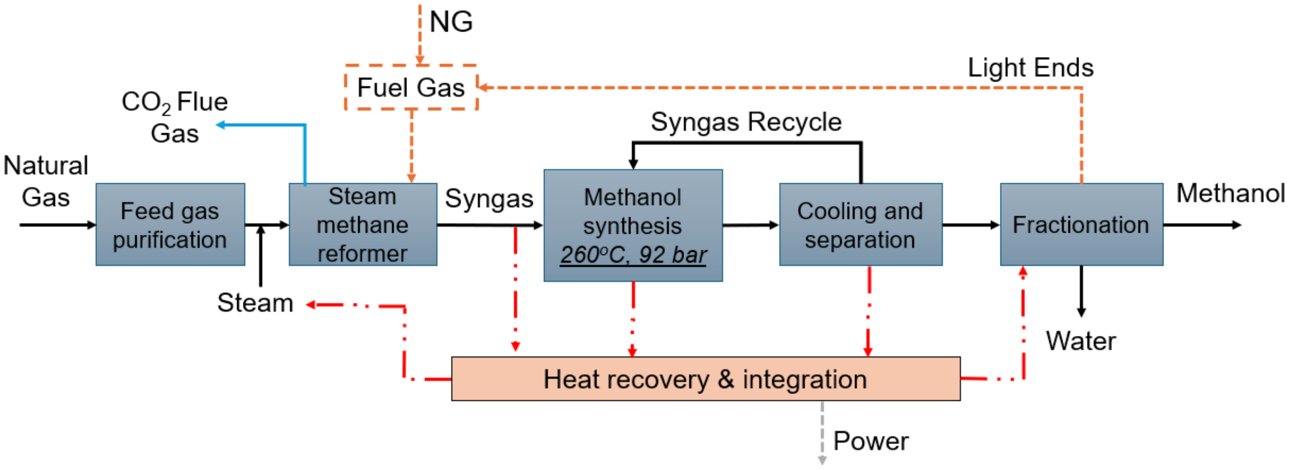


Figure 18. Standard methanol synthesis via SMR.

3.3.2.1 Chemical Process Modeling

This section summarizes the Aspen Plus model built to represent SMR and methanol synthesis. The function of each block and its simulation settings are described in this section. PFDs for syngas production, methanol synthesis, and distillation are presented in Appendix B-2.

NG is mostly comprised of methane, but higher hydrocarbons do account for a non-negligible portion. The NG is assumed to have a molar composition of 95% methane, 3% ethane, and 2% propane. NG used as raw material is introduced into the system at a pressure of 35 bar (500 psi), which is the pressure of gas in NG pipelines.

The energy required for preheating and SMR is supplied from the furnace by burning NG as fuel. NG fuel and ambient air are introduced into the furnace block (FNC-1) along with the light ends from the fractionation section, where complete combustion proceeds. The pressures and temperatures of the fuel and air stream are specified as 1.01 bar and 25°C, respectively. The RGibbs reactor model was employed for the furnace block.

NG contains sulfur in the form of hydrogen sulfide and methyl/ethyl mercaptan, which might damage the reformer catalyst. The mercaptans in NG are additives that give NG its odor. A desulfurization unit (DESULFUR) with a zinc oxide adsorbent was introduced to remove the sulfur compounds. The preferred operating temperature of desulfurization reactions ranges from 350–400°C, so NG is preheated by the heat exchanger connected to the downstream furnace (HX6). The Ryield reactor model was employed for this purpose.

The PRE-RFM block represents the pre-reformer, which converts higher hydrocarbons from the NG into carbon oxides and hydrogen to increase the efficiency of syngas production during SMR. The raw material stream (NG steam mixing gas) is preheated to 600°C by a heat exchanger (HX4) connected to the furnace to meet the operating temperature of the pre-reformer (500~600°C).

The RFM-1 block represents the primary SMR. To fulfill the highly endothermic reaction of SMR, the stream of NG steam mixing gas is preheated to 900°C by a heat exchanger (HX3) connected to the furnace to supply an additional 75 MW of heat from the furnace to the steam methane reformer. An RGibbs model was selected for this reactor.

The syngas stream exiting the SMR is then cooled, and water is removed. Some of the heat from cooling the syngas is used for preheating the steam used in chemical reactions (HX8). A carbon dioxide make-up stream mixes with the syngas to match the stoichiometry of the synthesis reaction and is compressed.

The methanol synthesis process used in this model is based on Lurgi's Mega-Methanol design, as well as the reference methanol synthesis model included in Aspen Plus v11. A series of compressors (K201S1 and K201S2) increases the pressure of the crude syngas stream to 92 bar, which is favorable for the methanol reaction. The modeled process utilizes two methanol reactors in series (R201-A and R201-B) to produce methanol (MeOH) as the target product and ethanol (EtOH) and dimethyl ether (DME) as byproducts. Both reactors were simulated by a kinetic-based reactor model that considers the four reactions listed in Table 8.

Table 8. Methanol synthesis reactions.

No.	Label	Reaction Equations	Ref.
1	Methanol Synthesis	$CO_2 + 3H2 \rightleftharpoons MeOH + H2$	Target Product
2	Reverse Water Gas Shift (RWGS) Reaction	$CO_2 + H2 \rightleftharpoons CO + H2O$	Intermediate
3	Ethanol Formation	$2CO + 4H2 \rightleftharpoons EtOH + H2O$	Byproduct
4	DME Formation	$2CH3OH \rightleftharpoons DME + H2O$	Byproduct

To remove the heat produced by these reactions, the syngas feed cools the second reactor, and boiler feedwater cools the first reactor to generate steam. The configurations of the methanol synthesis reactors simulated for the process are illustrated in Figure 19.

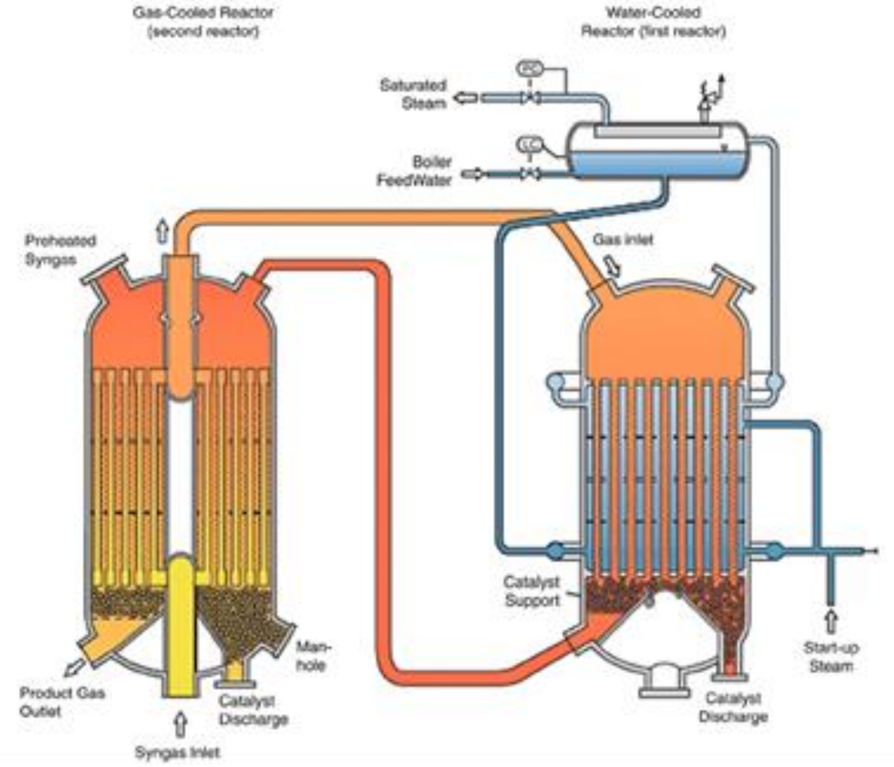


Figure 19. Detailed diagram of Lurgi's two-stage reactor [22].

The stream exiting the synthesis reactors is cooled down using a process-heat exchanger, along with air or water coolers to condense and cool the methanol, water, and byproducts. A flash drum block (V201) separates the unreacted light gases (L13) from the condensed liquid mixture of crude methanol (L14A). Most of the light gases obtained post-condensation are recycled back to the methanol reactor.

The purpose of the water scrubber (C201) is to remove DME from the light gas. Hydrophobic DME, which is not dissolved in the water scrubber, exits the tower with the light gases, whereas hydrophilic methanol and ethanol are captured in the water and sent to the distillation tower along with the crude methanol.

Some light gas remains dissolved in the crude methanol following the water scrubber. The first distillation tower (C301) separates light ends from crude methanol. It has three theoretical stages, and the feed stream is introduced at the third stage. Light gases separated from crude methanol exit the top of the column, whereas the water and methanol mixture exit the bottom and are pumped to the next distillation tower. Here, a RadFrac distillation model, including a partial-vapor condenser and a kettle reboiler, was used. To match the target temperature of first stage of the distillation tower (or condenser) to 45.5℃, the distillate rate was adjusted to 20.249 ton/day. The reboiler duty was 6 MW. After the first distillation tower, the crude methanol had less than a 1×10^{-11} mass fraction of light gas.

The second distillation tower (C302) separates the water and impurities from the methanol to meet Grade A product specifications. Thirty-five stages are required for the distillation, and the feed stream is introduced on the 15th. Methanol exits the top of the distillation tower, a byproduct containing ethanol exits the 30th stage, and water exits the bottom. Here, a RadFrac distillation model, including a partial-vapor condenser and a kettle reboiler, was used. The molar reflux ratio for the separation was 1.5, and the mass flowrate of the liquid side stream at the 30th stage was 27.4 ton/day. To match the target water purity of the bottom stage to 0.999999, the reboiler duty was modified to 93 MW. In the end, methanol purity at a mass fraction of 0.999 at the top of the tower is achievable.

3.3.2.2 Energy Balance and Heat Recovery System

This section describes the auxiliary systems of the model's energy-consuming components and the approach to minimizing external energy demands within the system. The system was optimized for energy use to decrease the size required for an integrated nuclear plant. Decreasing the required capacity of an NPP is expected to also reduce the capital costs involved.

The furnace that supplies heat to the SMR supplies heat for both the reformer and the NG preheat, generating HP steam for power and heat delivery using a backpressure turbine configuration. Steam generated in the furnace could also be used for system startup. The furnace utilizes an air preheater with a flue gas outlet temperature of 117°C, which is reasonably low for decent boiler efficiency but sufficiently higher than the flue gas dew point.

A large amount of high-temperature heat can be recovered from the syngas following the reformer. The reformer outlet gas is 875°C, and it must be cooled to 35°C to condense the water it contains. However, only a portion of this heat can be utilized because a significant amount of the heat, mostly latent heat from water condensation, is removed at relatively low temperatures. The other opportunities for heat recovery are the cooling system for the methanol reactor, which must remain at a constant temperature near 265°C, and the partial cooling of the raw methanol product. Heat in the raw product of the first reactor stage is recuperated in the second reactor stage. Heat from the raw product of the second stage is used to preheat the syngas feed. Recovery of the available heat is illustrated by the PFD in Appendix B-3, Figure B3 and the heat transfer-temperature (Q-T) diagram shown in Figure 20. Starting from the left-hand side at the syngas inlet injection steam superheaters are followed by an injection steam evaporator (illustrated as a drum type) and a preheater. This injection steam, which primarily consists of recycled water effluents, is injected into the NG prior to the reformer. This lower-quality water stream is not introduced into the main process steam system; however, a dedicated backpressure turbine is used to assist with system power demands, the outlet of which is only slightly superheated. Next on the syngas cooling line is an evaporator that generates LP steam for the reboiler duty of column C-301. The last heat recovery taking place is a dirty water preheater, preheating it before being introduced into its own deaerator.

The figure shows that, at the onset of condensation, heat can be supplied for low-temperature waste heat recovery or for boiling water at 1 atm to supply the column C-301 reboiler. While making LP steam only at 2.5 bar to supply the column C302 reboiler would simplify the process steam system, the saturation temperature is too high to generate this steam from the syngas cooling.

The HP steam is generated in the furnace, where the preheater, evaporator and superheater are. It then expands in the first turbine section to MP, where additional steam is added. The MP steam comes from feedwater being preheated in the raw methanol cooler, evaporated while cooling the methanol reactor (part of the steam is then routed for a light end gas heater), and superheated, before being introduced into the turbine. A final turbine section exhausts all the steam at 2.5 bar (backpressure turbine) where all the condensation heat is used to provide a reboiler duty of a C-302 column. Condensate from this system is then split to separate HP and MP streams.

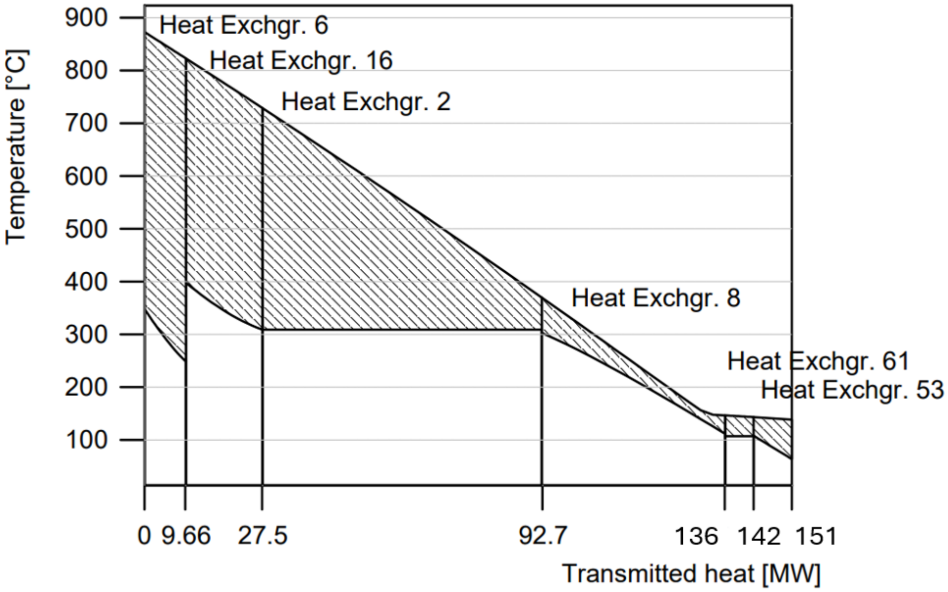


Figure 20. Heat transfer-temperature (Q-T) diagram of waste heat recovery from syngas cooling.

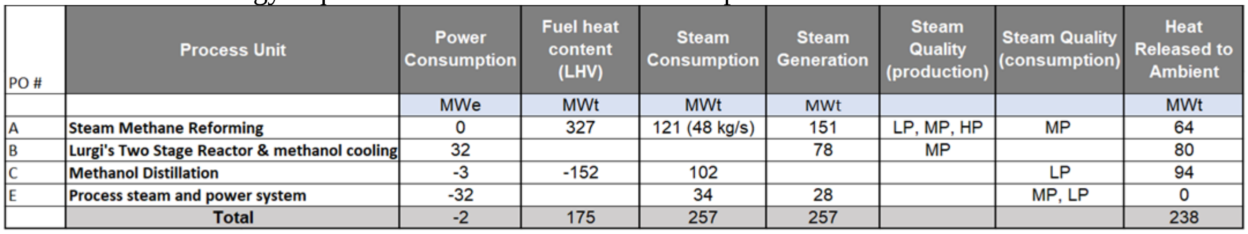
Heat is not recuperated directly from the syngas to final heated processes, due to the complex piping necessary to deliver the syngas to multiple points in the plant. A steam system of this type would require bypasses and an auxiliary heat source to remain flexible in the case of off-design operation. Using process steam as the heating medium rather than the syngas itself solves these issues and is a standard approach used in many industries.

3.3.3 Mass, Energy, and CO_2 Balance

Table 9 summarizes the energy requirements for methanol production. The energy demands of the reference methanol plant are based on the Aspen model with heat integration (see Section 3.3.2). Electricity and steam consumption, heating duty from fuel combustion, steam generation, and heat rejected to the ambient condition are summarized for each process unit. A negative number for power and steam consumption signifies that the steam or power is produced in the specific unit operation.

Steam is generated during SMR, syngas cooling, and methanol cooling. Additional steam is generated in the furnace by combusting additional fuel. Some steam is injected into the SMR process, while the rest is used by the process steam and power system providing power and heat duties to other unit operations. The additional steam from the furnace is generated as implementing only a waste heat recovery system to produce steam is not sufficient to meet all power and heat duty demands internally. Implementation of this additional steam production is sized to meet all heat duty. Power production balance provides a very small excess for export. Water generated as part of a chemical reaction is not considered in the mass or energy balance, as it is not utilized in the energy system streams. The energy demand of the cooling system for rejecting heat to the ambient condition was not included in the model, meaning the actual net electricity export may be smaller by up to several MWe. Considering that the only energy supplied to the system comes from fuel combustion, the energy required for methanol production is 5.535 GJ/tonne MeOH. This value does not include energy generated from the feedstock.

Table 9. Overall energy requirements for reference methanol plant.



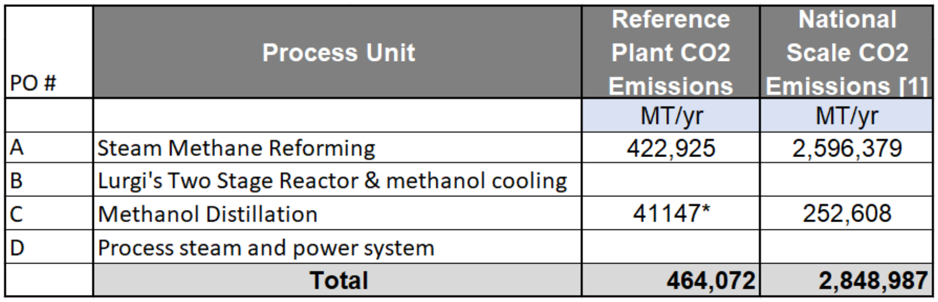
PO #	Process Unit	Power Consumption	Fuel heat content (LHV)	Steam Consumption	Steam Generation	Steam Quality (production)	Steam Quality (consumption)	Heat Released to Ambient
		MWe	MWt	MWt	MWt			MWt
A	**Steam Methane Reforming**	0	327	121 (48 kg/s)	151	LP, MP, HP	MP	64
B	**Lurgi's Two Stage Reactor & methanol cooling**	32			78	MP		80
C	**Methanol Distillation**	-3	-152	102			LP	94
E	**Process steam and power system**	-32		34	28		MP, LP	0
	Total	-2	175	257	257			238

Note: 327 MW is total furnace input consisting of 175 MW in NG and 152 MW in light ends.

NG introduced as feedstock contains 750 MW of energy (23.79 GJ per tonne of methanol), and the product methanol contains MW based on lower heating value (LHV). Thus, the total energy requirement for methanol production, as calculated in the model, is 29.33 GJ per tonne of methanol. This highlights the potential of integrating low-carbon sources when considering an alternative feedstock.

Table 10 summarizes the direct carbon emissions for each process unit, in terms of both the reference plant and the national scale. The total national emissions for 2023 are reported based on the EPA FLIGHT, and the reference plant emissions are from the Aspen model results, including heat integration [2]. Emissions are produced during the combustion of NG in the furnace and are released in the exhaust stream during SMR and in the light ends during distillation. The national-scale emissions are used to calculate the process unit emissions by estimating the relative contribution of emissions from each unit. The linear correlation in Figure 16 estimates carbon emissions generated by the reference plant at 476,066 tonnes per year using national data. This value aligns with the emissions calculated in the Aspen model with heat integration shown in Table 10. The amount of CO_2 emissions associated with methanol production is 0.487 kg CO_2 per kg of methanol.

Table 10. Direct CO_2 emissions for the reference methanol plant and at the national scale.



PO #	Process Unit	Reference Plant CO2 Emissions	National Scale CO2 Emissions [1]
		MT/yr	MT/yr
A	Steam Methane Reforming	422,925	2,596,379
B	Lurgi's Two Stage Reactor & methanol cooling		
C	Methanol Distillation	41147*	252,608
D	Process steam and power system		
	Total	**464,072**	**2,848,987**

* Note: The light gases from the methanol synthesis and distillation sections are recovered and used as fuel in the reforming furnace. Combustion of the light gases reduces the amount of NG required by the reforming furnace.

3.4 Dynamic Operation

Methanol plants are large manufacturing facilities that operate continuously to convert NG into methanol. They consist of large processing units for SMR, methanol synthesis, fractionation, and separation. These units operate at high temperatures (in excess of 300–900°C) and pressures. Multiple steam production systems are incorporated into them for heat recovery. Normally, methanol plants operate at full capacity unless impacted by external factors, such as economics, NG supply, or utility constraints.

If required, the process units can ramp down by 30–40% and still maintain safe, stable operation. However, several hours are usually required to accomplish these feed rate adjustments, depending on the magnitude of the change. This ramp time is necessary to meet unit operating temperatures and pressures, adhere to unit procedure guidelines, and prevent equipment damage.

Methanol plants normally schedule shutdowns to inspect, repair, and clean equipment and to replace catalyst systems. These planned shutdowns occur every 4–5 years and require 30–50 days for the identified work and activities to be accomplished. Approximately 1–2 weeks is needed to shut down and restart the units.

To integrate an NPP with a methanol plant, these dynamic operations, operation variations, and startup and shutdown requirements will be incorporated into the design strategy.

4. PULP AND PAPER

Kraft pulping constitutes 80% of the total chemical pulping industry, making it the predominant method employed worldwide [23]. It involves digesting wood chips at elevated temperatures and pressures in “white liquor” (an aqueous solution composed of sodium sulfide and sodium hydroxide). The chemical dissolution of lignin, which binds cellulose fibers in wood, occurs through this white liquor. The physical pulping of wood chips is conducted in digester systems, via either a batch or a continuous process. Though most kraft pulping occurs in batch digesters, continuous digesters are becoming more prevalent.

After digesting, the contents are transferred to an atmospheric tank commonly known as a blow tank. The entirety of the blow tank contents is sent to pulp washers, where the pulp is separated from the spent cooking liquid. The pulp then undergoes a series of defibrating, washing, and bleaching stages (if needed) [24]. One advantage of the kraft process design is the recovery and recycling of cooking chemicals and heat. Spent cooking liquor and pulp wash water combine to form a weak black liquor that contains about 15% solids and is later concentrated to about 40–55% solids in a multi-effect evaporator system. The liquor is further concentrated to approximately 65–75% solids (strong black liquor) by using flue gases in a direct-contact evaporator or in an indirect-contact condenser. The strong black liquor is fired in a black liquor recovery boiler (BLRB), providing energy for the pulping process and the conversion of sodium sulfate into sodium sulfide [25].

Inorganic chemicals in the black liquor collect as molten smelt at the furnace's base and are later dissolved in water to form green liquor. The green liquor is transferred to a causticizing tank where quicklime (calcium oxide) is added to convert the solution back into white liquor for its return to the digester system. Lime mud precipitates from the tank and undergoes calcination in a lime kiln to regenerate quicklime [25]. This quicklime is added to the green liquor to return it to white liquor for reuse in the digester. A simplified diagram of the kraft pulping process is shown in Figure 21. In 1994, recovery boilers represented 35% of total boiler capacity in the pulp and paper industry. Including other waste fuels, about 50% of boiler capacity is fueled by the chemical recovery process and its byproducts [26]. The recovery furnace provides up to 100% of the total energy requirement for market pulp mills, whereas the percentage for integrated mills widely varies based on the pulp and paper production capacities. The remaining energy requirements are met by conventional fossil fuels, such as NG, fuel oil, or wood boilers.

The wood boiler fuel commonly consists of solid wood waste stemming from log cutting and debarking conducted as part of woodyard processing, enabling a self-sufficient energy cycle at the mill. Additionally, the lime kiln is typically fired with NG or fuel oil because of the process sensitivity to high-temperature levels and contaminants [27].

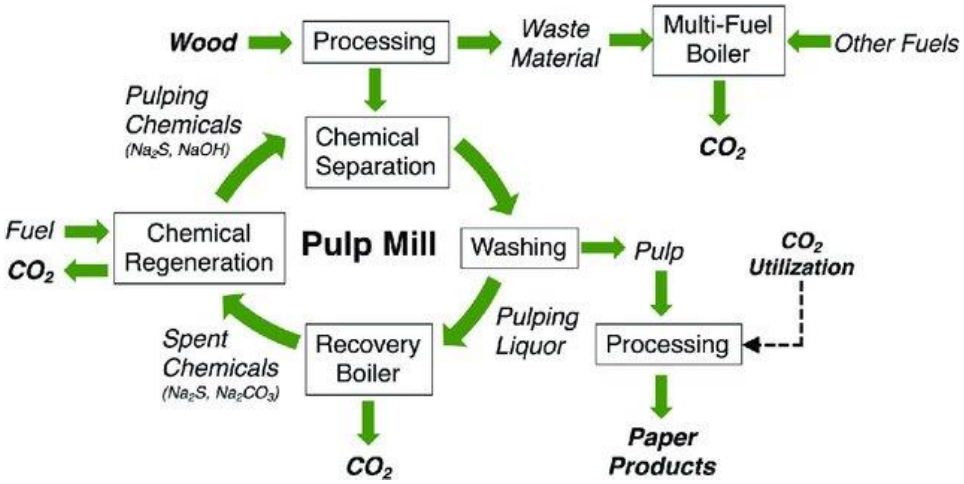


Figure 21. Flow diagram of a kraft pulp mill [28].

A reference kraft pulp mill is established in Section 4.4, which has production and utility requirements detailed in Figure 22. The BLRB, lime kiln, hog boiler, and NG auxiliary boiler are the primary emissions sources. The electricity and process steam demands are satisfied by a CHP system described in Section 4.4.3.2.

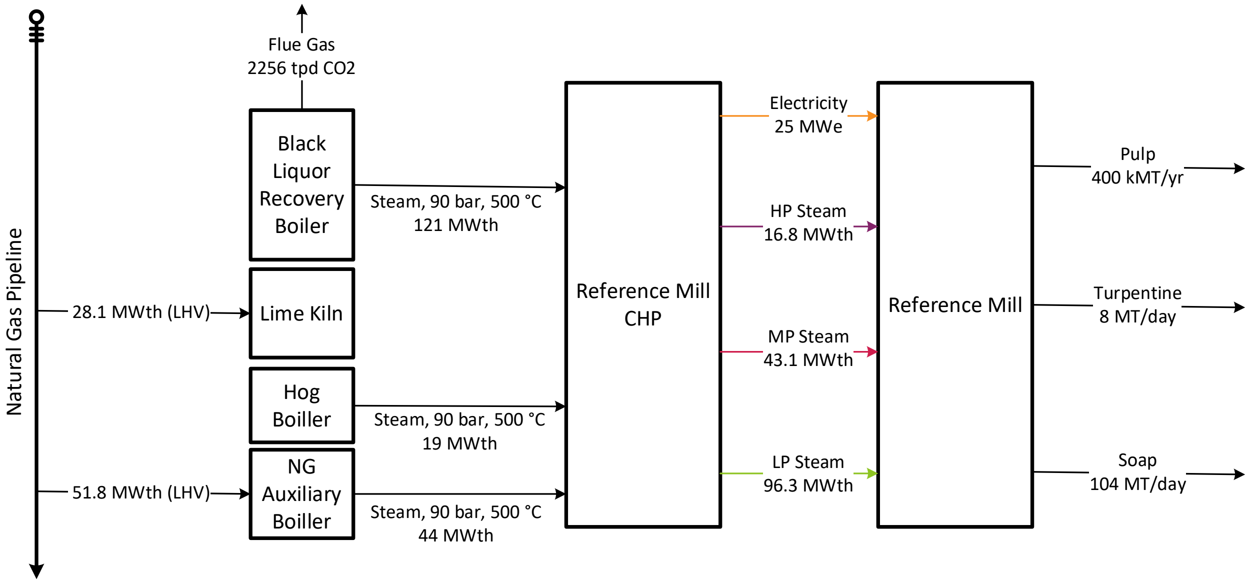


Figure 22. Utility system BFD for reference kraft pulp mill.

4.1 Pollutants and GHG Emissions

Despite the innovations that make it a highly efficient, self-sustaining process, kraft pulping significantly contributes to air and water pollution. CO_2 is emitted from the burning of lignin in the recovery boiler, the combustion of wood waste, the calcination of lime, and the usage of fossil fuels. About 80% of CO_2 emissions reportedly stems from biogenic sources (see Table 11). In addition to these emissions, the entire paper sector releases 0.1 MMT CO_2-eq. of CH_4 and 0.4 MMT CO_2-eq. of N_2O annually [29].

Table 11. Emissions from the pulp and paper industry in MMT CO_2-eq.

Fuel	National CO_2 Emissions from Pulp and Paper	Energy Medium	Average Plant Emissions[b]	Percentage of Total CO_2 Emissions
Spent Liquor	73.5	Steam	0.71	61%
Wood and Other Biomass	26	Steam	0.22	19%
NG (or alternative fossil fuel)	24.2[c]	Steam	0.1	9%
		Combustion Gases[a]	0.13	12%

[a] Combustion of fossil fuels also contributes 0.4 MMT each of CH_4 and N_2O emissions, which are tracked by the EPA [29].

[b] Based on 103 pulp and paper facilities that reported to the EPA in 2021 [29].

[c] Includes the CO_2 released chemically from the calcination of lime. Fossil CO_2 emissions are based on the non-biogenic emissions tracked by the EPA. The percentage of total plant CO_2 emissions allocated to the lime kiln is 12% [30]. The rest of the fossil emissions are allocated to the fuel boiler for steam production.

In addition to tracked GHG emissions, other air emissions from paper mills represent health and environmental hazards. The recovery furnace, lime kiln (calcination), and smelt dissolving tank (recausticizing) primarily contribute to the total emissions resulting from the carryover of solids and the sublimation or condensation of inorganic chemicals. The distinct odor associated with pulp mills is attributed to chemical emissions, especially reduced sulfur compounds such as hydrogen sulfide, methyl mercaptan, dimethyl sulfide, and dimethyl disulfide. These compounds, emitted from various mill points, can cause serious adverse health effects due to their low odor thresholds. The direct-contact evaporator is a major source of hydrogen sulfide emissions, reactions with lignin contribute to methyl mercaptan, and dimethyl sulfide formation arises from digesters and blow tanks. Table 12 lists the quantities of other measured pollutants generated by pulp and paper mills.

Table 12. Measured air pollutant emissions from pulp and paper mills [32–35].

Pollutant	Total Emissions from Pulp and Paper (metric tons in 2021)	Averages of the Reporting Facilities	Source
Pb	0.67	0.0054	Recovery boiler (solids)
CO	8,554	69	Recovery boiler, calcination
NH_3	406	3.3	Recausticizing
NO_2 (eq.)	7,476	60	recovery boiler, calcination
PM10	1,542	12	Recovery boiler, recausticizing, calcination
PM2.5	1,197	2.8	-
SO_2	2,240	0.29	Recovery boiler, recausticizing, calcination,
VOC	2,424	1.8	evaporators, bleaching

4.2 Current Pulping Types and Production Methods

Pulp is a generic term for products that result from the chemical and mechanical treatment of various types of cellulosic material. These pulp products consist of a wide range of technically distinct materials produced via complex manufacturing processes. In the past, only two distinct procedures were available for carrying out the pulping operation at a given location: mechanical or chemical. The choice of procedure depended on the initial mill design, the wood species available, and the end-use application. Today, mills can combine aspects of both processes. Furthermore, the pulping process can be defined per the following categories:

- **Mechanical Pulping:** This is when cellulose fibers are physically separated by a machine. A high percentage (85%–95%) of the original wood components are retained in the final product. The construction costs are lower, but unlike many chemical processes, mechanical pulping requires 100% external energy input. The classifications of mechanical pulping are:
    - **Stone Groundwood:** This involves pressing debarked logs against a huge revolving abrasive stone that shreds fibers from the wood.
    - **Refiner Mechanical Pulp:** This uses wood chips as its main input. These chips are then fed between two large metal disks in a refiner.
    - **Thermomechanical Pulp:** This involves an initial stage in which wood chips, prior to entering a pressurized refiner, are softened by steam, facilitating mechanical separation of the cellulose and reducing the damage to the individual fibers.
- **Combined Pulping Methods:** These involve integrating chemical methods with mechanical methods. Wood chips are first cooked, then ground via a mechanical process. The two main combined methods are:
    - **Chemi-thermomechanical Pulp:** This chemically treats the chips prior to heating and mechanical defibration, leading to pulp that offers additional improvements in physical properties relative to thermomechanical pulp.
    - **Semichemical Pulp:** This type of pulp represents a relatively small subsegment of the total pulp available on the market. Semichemical fibers are usually stiffer and have a higher yield than kraft pulp. The main end uses for semichemical market pulp are containerboard and corrugated mediums.
- **Chemical Pulping:** In this process, chemical solvents separate cellulose fibers from the "wood matrix" to remove as much of the lignin and extractives as possible. This results in a much lower yield (usually 40%–55% of the original wood substance), thereby increasing the unit product cost. The construction costs are higher for this method than for mechanical pulping, but as was mentioned, the chemical recovery process can provide heat and electricity for process operations.
    - **Kraft (sulfate) Pulp:** This involves "digesting" (cooking) wood chips in an alkaline solution for several hours, during which time the chemicals break down the lignin molecules into smaller segments that are then dissolved and removed. A significant feature of kraft pulping technology is the sophisticated recovery system involved, in which chemicals used in the process are captured and extracted for reuse, helping minimize both raw material costs and the environmental impact. Unbleached kraft pulp is dark brown in color, and pulp used for bright or dyed paper must undergo a series of bleaching stages, resulting in both a white product and an additional manufacturing expense. Within the kraft pulp grades are several subgrades used to define specific qualities. Among these subgrades are bleached softwood kraft, northern bleached softwood kraft, southern bleached softwood kraft, radiata pine, bleached hardwood kraft, northern bleached hardwood kraft, southern bleached hardwood kraft, eucalyptus, and tropical hardwood.

- **Sulfite Pulp**: Introduced in 1867, commercial sulfite pulp had, by the early 1900s, surpassed mechanical pulping as the main process employed to produce wood pulp. Kraft pulping was developed only a few years thereafter but would only become popular in the 1940s when development of the recovery boiler enabled kraft pulp mills to recycle nearly all their pulping chemicals. Sulfite market pulp still exists today, but only at a fraction of the size of the kraft pulp market. Sulfite market pulps generally have higher yields and are characterized as being easier to bleach than kraft pulp. However, the fibers are generally weaker.

In the United States, kraft mills are the dominant pulping process. Table 13 shows the number of these types of pulping processes listed above and their share of U.S. pulp production.

Table 13. Summary of mill types, sites and total production. [36]

Mill Type	Sites	Total Production (air-dried-tonnes/yr)	Percentage
Kraft	95	27,577,387	33.8%
Mechanical Recycle	126	21,731,629	26.6%
Kraft/Mechanical Recycle	39	19,582,130	24.0%
Kraft/Other	11	6,466,283	7.9%
Semichemical	7	3,874,800	4.7%
Soda/Sulfite	7	1,029,985	1.3%
Mechanical Pulping	5	1,366,802	1.7%
Total	260	81,629,016	100%

4.3 U.S. Pulp and Paper Kraft Mill Production Capacities: Nationwide Distribution and Locations

A total of 105 pulp and paper facilities exist throughout the United States, 85 of which are kraft pulp mills. Figure 23 discloses the overall state-specific locations of the reportedly operative mills across the country (as of Q2 2023), with most of the kraft mills (69) being clustered in the southeastern United States, due to the resulting biomass feedstock availability and logistical advantages.

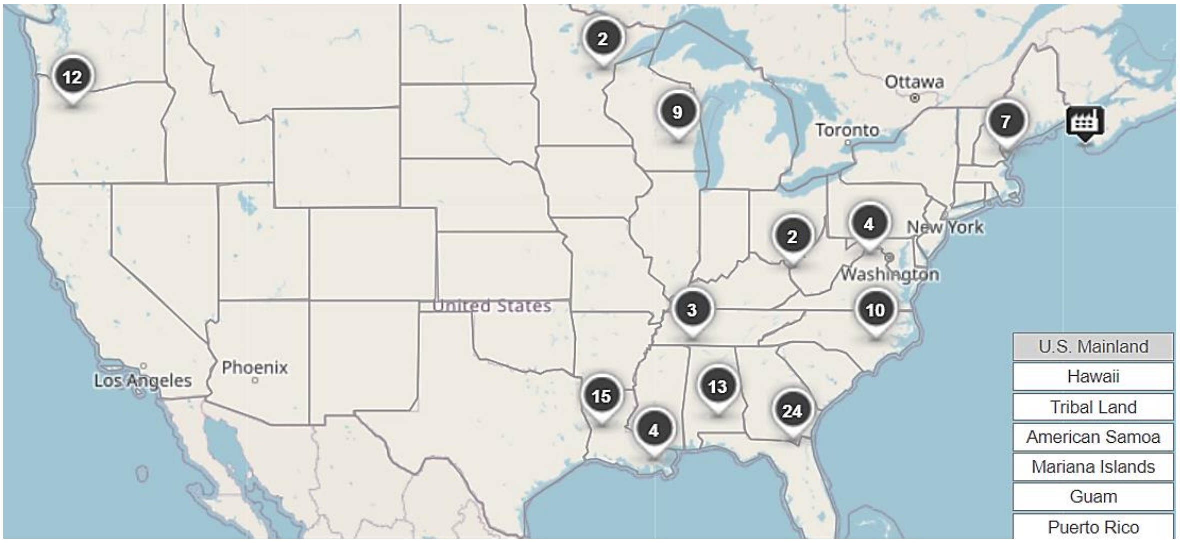


Figure 23. Operative pulp mills in the United States as of 2023 [1].

In particular, Alabama, Georgia, Mississippi, North Carolina, Louisiana, South Carolina, and Florida are the major wood pulp producers in the United States (see Table 14). Appendix A-3 gives a comprehensive distribution of the U.S. pulp mill facilities, along with their names, locations, and capacities. The three states featuring significant concentrations of kraft mill facilities are Georgia, Louisiana, and Alabama, accounting for 18%, 15%, and 14%, respectively, of the nation's annual total kraft pulp production.

Table 14. U.S. pulp production by state [36].

State	Count	Capacity	Share
AL	11	6,362,870	12%
AR	5	2,115,645	4%
FL	6	2,639,369	5%
GA	11	7,675,863	14%
ID	1	731,002	1%
KY	2	1,138,225	2%
LA	7	6,670,045	12%
ME	5	1,900,948	4%
MI	3	1,431,090	3%
MN	2	1,123,385	2%
MS	4	2,809,600	5%
NC	5	1,979,859	4%
NY	2	275,125	1%
OH	1	462,941	1%
OK	1	1,554,350	3%
OR	4	1,909,905	4%
PA	2	673,642	1%
SC	6	3,823,140	7%
TN	2	1,127,835	2%
TX	3	2,064,820	4%
VA	6	2,695,280	5%
WA	6	2,568,179	5%
WI	5	499,757	1%
Total	100	54,232,875	—

4.4 Reference Kraft Pulp Mill

As generally occurs with all diverse production sectors, the kraft pulping capacities for each mill located across the United States will vary over years or months due to upgrades, complete shutdowns, expansions, or ownership changes. This variability is also affected by feedstock quality (heterogeneity) and the challenges associated with woody biomass processing. For reference purposes and determining heat and energy estimations, the yearly average and mean capacities for the total number of kraft mills in the United States were calculated at about 650,000 and 623,000 air-dried-tonnes/year (ADt/yr), respectively. For the process mass balance, industrial use requirements, heat, power, etc., the reference plant capacity was assumed to be 400,000 ADt/yr, which encompasses ~42% of the total number of kraft mills in the nation. These parameters and distributions are reflected in Figure 24. A below-median capacity was chosen as the reference capacity, since about 43% (i.e., 30) of kraft pulp mills operate within this annual capacity and use batch digesters, whereas the remaining mills utilize more than four paper machines to increase their annual production and may use a combination of batch or continuous digesters to process higher quantities of wood chips on a yearly basis.

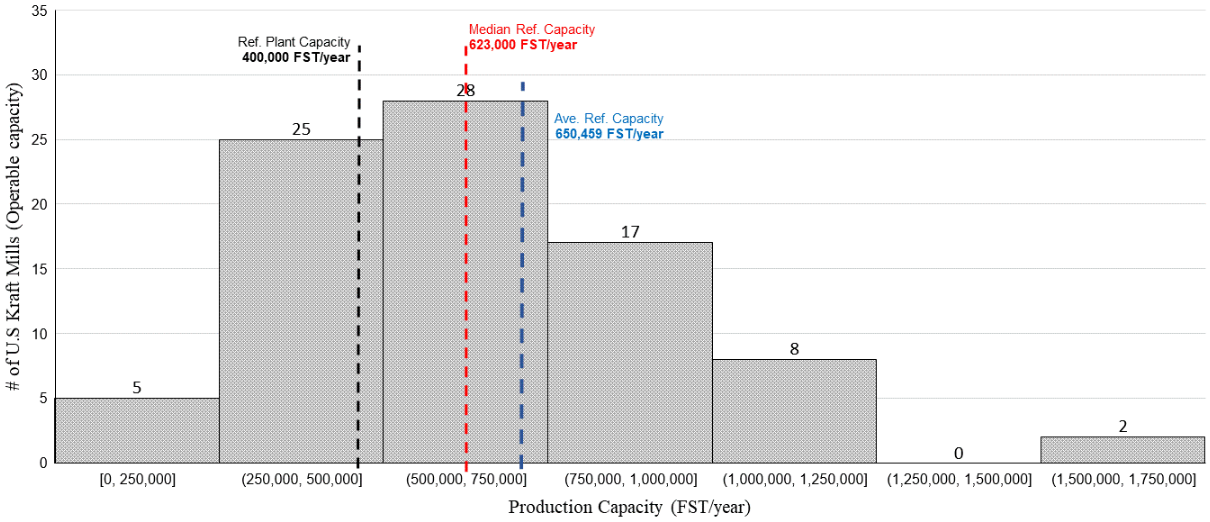


Figure 24. Nationwide kraft pulp mill capacity distribution chart.

4.4.1 Configuration, Size, and Assumptions

The global unbleached softwood kraft pulp market was estimated to be worth U.S. $1,481.5 million in 2023. Future Market Insights projected that the market would expand at a compound annual growth rate of 4.6% over the next 10 years, with a year-on-year growth of 3.5% in 2023. Consequently, the market is anticipated to reach U.S. $2,322.84 billion by 2033 [38]. The future of unbleached softwood kraft pulp in the end-use industries (with advancements) is promising and is expected to remain stable, or even increase, in the future. Thus, the industrial use requirements assessment focused on the unbleached softwood (southern) kraft pulping process (see Figure 25), with a reference capacity of 400,000 ADt/yr. Figure 25 is a block flow diagram of a typical integrated kraft pulp and paper mill. An Excel-sheet-produced model based on real plant process and operative conditions was used to estimate mass and energy balances, along with steam and energy demands. Within the industry, there is no such thing as a "typical" pulp and paper mill, as each mill tends to vary in terms of type and capacity. For example, some integrated mills produce a large portion of market pulp but include small papermaking operations. Others may use a large quantity of recycled materials and only a fraction of virgin pulp. However, the process mechanisms and layouts of all kraft integrated mills are similar and well-documented, enabling energy demands and mass balances to be estimated from reference values. These mass and energy balances will serve as the basis for future work related to the Aspen modeling of the chemical recovery process.

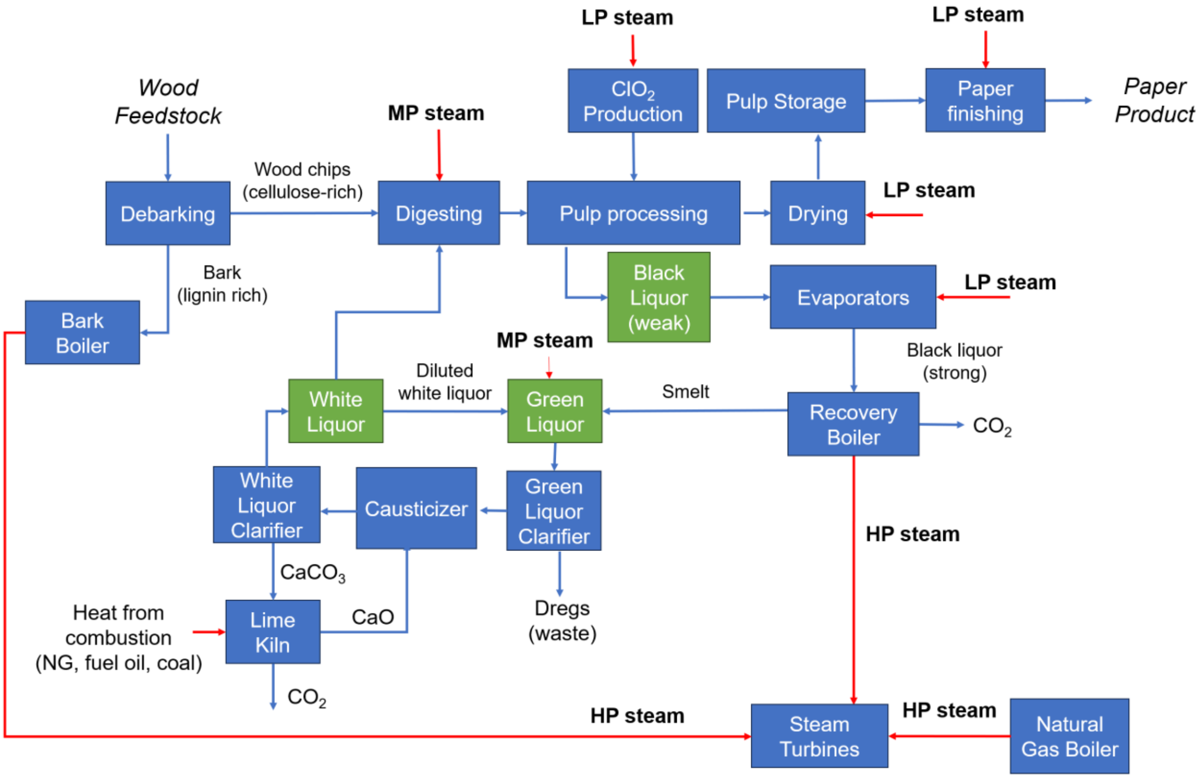


Figure 25. Block flow diagram of a typical kraft pulping process.

4.4.2 Reference Process Details

An annual capacity of 400,000 ADt was assumed for the reference kraft pulp and paper mill, for a final pulping yield of 46 wt%, which falls within the range of industrial operation yields (40–55%). An Excel sheet provided by an industry partner in the pulp and paper industry sector, based on real process operation conditions, chemical dosages, and other parameters; was used to estimate the mass and energy balances, process energy demands and emissions for the reference pulping process. Table 15 summarizes the streams and total flow rates of the process, and a PFD of the reference kraft mill is given in Figure 26.

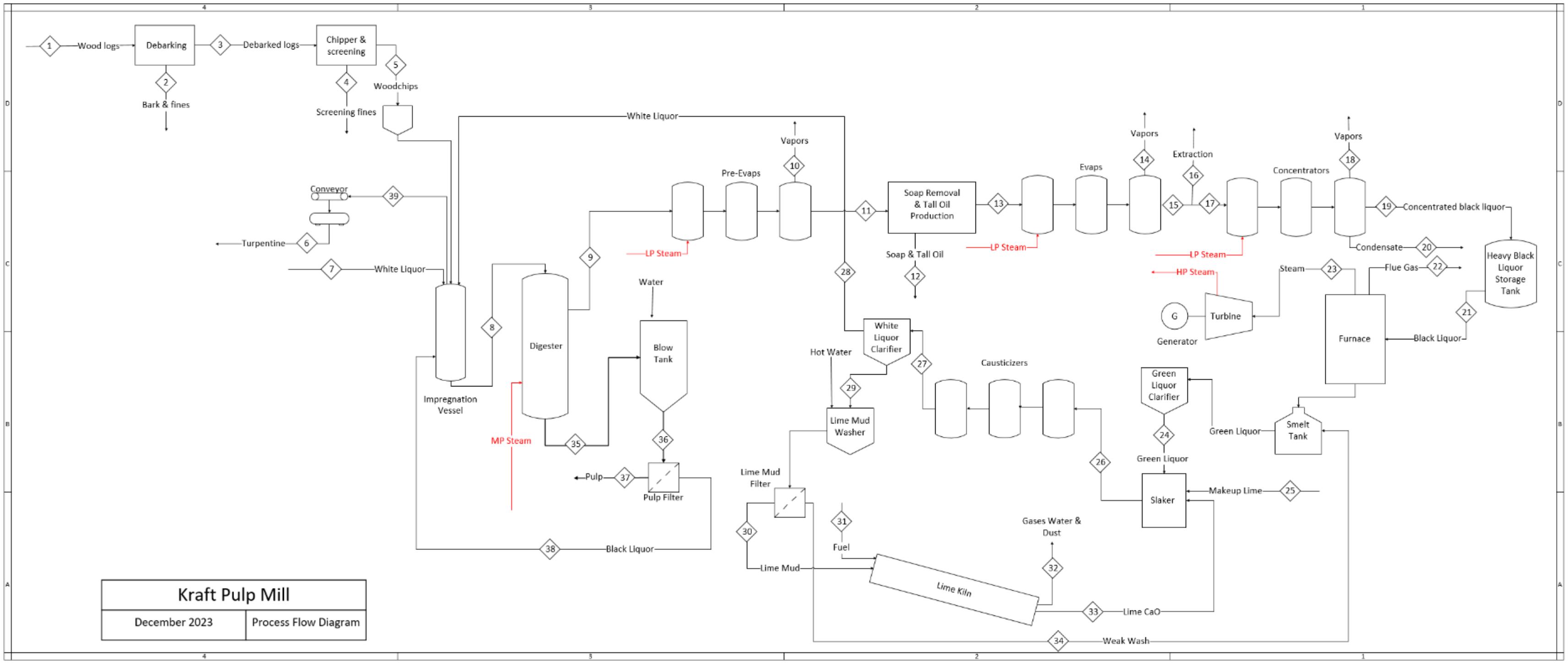


Figure 26. Reference kraft mill PFD.

The reference kraft mill's energy and steam demands are taken from the Excel model, which provides the power and heating duty demands of the HP, MP, and LP steam generated at the recovery boiler, the heating duty from black liquor, and fuel combustion. Table 17 summarizes these energy demands. Note that Table 12 does not show all emissions. Total emissions are shown in Table 18.

Table 15. Pulp mill stream flow rates.

Stream	Components	Total Flow Rate (mt/day)
1	wood logs	2,554
2	bark and fines	230
3	debarked logs	2,324
4	screening fines	76
5	woodchips	2,248
6	turpentine + H_2O	8
7	white liquor	3,929
8	digested woodchips	2,240
9	extracted liquor	11,434
10	black liquor	8,552
11	soaps	104
12	black liquor	8,448
13	black liquor	4,128
14	vapors	2,882
15	black liquor	4,128
16	extracted liquor	537
17	black liquor	3,591
18	vapors	1,383
19	concentrated black liquor	2,208
20	condensate	251
21	black liquor	2,208
22	CO_2 emissions	1,240
23	steam	5,407
24	green liquor	529
25	Make-up lime	329
26	slacked lime	1,103
27	white liquor + lime mud ($CaCO_3$)	1,216
28	white liquor	3,929
29	lime mud	735
30	lime mud	1,320
31	NG, LPG, fuel oil	36
32	CO_2 emissions	131
33	lime CaO	246
34	weak wash	662
35	wet biomass	13,423
36	wet biomass	34,649
37	pulp (bdt)	1,029
38	Make-up black liquor	8,690
39	turpentine and water	8

4.4.3 Energy Integration in the Kraft Pulping Process

4.4.3.1 Process Heating and Heat Recovery

The heat recovery system is designed to capture and reuse the heat generated during various stages of the pulping process—particularly in the recovery of chemicals used in the digestion phase. The following types of heat recovery are typically implemented in a kraft pulping process:

- **Digestion Process:** The steam heat provided during this phase both facilitates the breakdown of lignin and represents a valuable energy resource that enables chemical separation for turpentine recovery.
- **Multi-effect Evaporators:** The multi-effect evaporator system plays a pivotal role in heat recovery. It efficiently concentrates the black liquor by progressively evaporating water from it over several different stages. As the black liquor moves through each evaporator stage, the heat released from the steam condensation in the previous stage is transferred to evaporate water in the current stage. This process significantly minimizes the need for additional external energy input.
- **Recovery Furnace and Steam Generation:** The organic components of the concentrated black liquor are combusted in the recovery furnace. Combustion serves as a means of disposal for the organic material, facilitates chemical conversion of Na_2SO_4 to Na_2S, and generates HP steam.
- **Energy Generation:** The steam produced in the recovery furnace is commonly sent to onsite turbines, generating electricity for the mill and reducing the steam temperature and pressure down to the proper conditions for steam delivery throughout the plant. This process is typically sufficient to meet all the mill’s electricity requirements. Excess electricity can be fed back into the grid.

The integrated nature of the heat recovery system in the kraft pulping process highlights its role in enabling sustainable, closed-loop operation. By efficiently utilizing the heat generated at various stages, industry can maximize resource efficiency, minimize external energy requirements, and actively contribute to the circular economy by recovering and reusing energy and chemicals. This emphasis on sustainability aligns with modern industrial practices that prioritize environmental responsibility and resource conservation.

4.4.3.2 Heat and Power Supply Model

The reference process model provided heat and power requirements of each unit operation and additionally heat content in black liquor and wood residue, that are internally combusted. This information allowed us to develop a model of a typical industrial onsite CHP steam system for power and process-heat production. The steam is generated in a BLRB (assumed fuel input only from black liquor, neglecting the common need for co-firing NG fuel oil[a] [39]), hog boiler, and an NG-fired boiler. Several additional assumptions based on industrial practice were adopted, which are boiler efficiencies [30,40–45] taken as moderate operational (not peak) values. Additionally, an assumption of return condensate temperature is 93°C for each pressure level and after eventual make-up water accounting for losses within the process. The pressure and temperature levels of produced main steam and delivered process steam are then assumed in Table 16. The limits of main steam are given mainly by the BLRB to limit the corrosion from the involved chemicals and precisely manage combustion to perform the chemicals recovery process.

---

[a] BLRBs have auxiliary fuel (gas or oil) burners located near the floor to raise the temperature for initiating combustion of the black liquor and to stabilize combustion if upsets occur. Auxiliary burners may also be provided higher in the furnace to supplement heat input during times of limited liquor availability or other upsets. [39]

Table 16. Boiler efficiency assumptions.

Boiler Efficiency Assumptions:			
BLRB	75%		
Hog Boiler	80%		
NG Boiler	85%		
Steam Parameters Assumptions:			
Stream Name	Pressure (bar)	Temperature (°C)	Notes
Main Steam	90	500	Moderate values, modern boilers up to 110 bar, 515°C
HP	27.6	364	Saturated or slightly superheated
MP	11.0	274	Saturated or slightly superheated
LP	5.2	209	Saturated or slightly superheated
Return condensate	1.2	93	
Other Assumptions:			
Return Condensate Temperature 93 °C			

The steam CHP system model is shown in Figure 27. All the boilers have a common steam bus routing the steam into an extraction backpressure turbine. An auxiliary condenser is used during periods of low demand for LP steam.

The given steam demand and fuel supply from the BLRB and hog boilers was implemented using Table 15. Because of the limits of boiler efficiencies, even in this backpressure configuration there is insufficient heat for all process requirements. An NG boiler was added and sized to provide the remaining heat duty. Since the demand is heat driven, rather than power driven, backpressure turbine is considered and the sufficient turbine isentropic efficiency is only 75%, leading to possibilities of a cheap compact fully impulse turbine. This is also the reason why no feedwater heaters (except for deaerator) are present. The presented system has a net power production of 26.6 MWe, providing a small excess of power required by the pulp mill. Power production efficiency, based on the LHV of the fuels, is 11.3% and total CHP efficiency is over 77%.

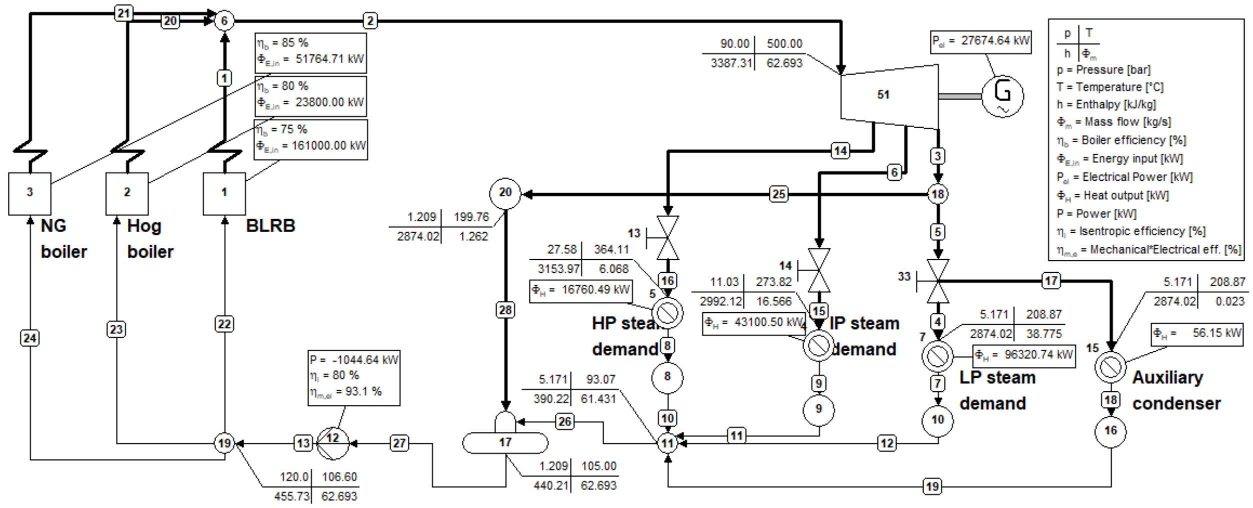


Figure 27. CHP system for reference pulp mill.

4.4.4 Overall Energy and CO_2 Balance

4.4.4.1 Energy Balance

The reference pulp mill's energy and chemical demands were based on plant data pertaining to the wood processing, pulp plant, evaporation plant, lime kiln and causticizing, and recovery boiler process units of an actual operating kraft pulp mill. The electricity and steam demands for the hog fuel boiler and the bleaching, drying, and wastewater treatment were calculated based on data from an American Council for an Energy-Efficient Economy report published in 1995 [39]. Heat recovery in the hog fuel boiler is accomplished via the combustion of byproduct bark and fines from wood processing. The heat of combustion was calculated using an LHV for bark (i.e., 6.72 kJ/kg). The black liquor solids are combusted in a furnace to recover heat as steam used for process heating and electricity generation. The heat of combustion from the black liquor solids was also calculated using the LHV (i.e., 9 MJ/kg) [44]. As Table 17 shows, the resulting electricity and steam demands for the reference plant are 25 MWe and 156 MWt, respectively, before an NG boiler. With the NG boiler sized for the CHP system to meet the steam demand, the net electricity balance results in a surplus of 1 MWe. This confirms the fact that pulp mills are usually energy neutral in terms of process heat and power and can even be net electricity exporters [45].

Table 17. Overall energy and chemical requirements for the reference pulp mill.

PO #	Process Unit	Power Consumption	Heat from Fuel Combustion	Steam Consumption	Steam Quality	Byproduct Fuel Source	Heat of Combustion of Byproducts (LHV)
		MWe	MWt	MWt			MWt
A	Wood Processing	1.4					
B	Hog Fuel Boiler	2.0				bark & fines	23.8
C	Pulp Plant	8.6		29.8	LP, MP		
D	Bleaching*	0.7		22.6	LP, MP		
E	Pulp Drying*	5.2		47.4	LP		
F	Evaporation Plant	1.4		32.5	LP, MP		
G	Lime Kiln + Causticizing	0.9	28.1				
H	Recovery Boiler	3.1		24.1	LP, MP, HP	black liquor solids	161.0
I	Wastewater Plant*	2.0					
J	CHP system	-26.6	45.0	-156.0	LP, MP, HP		-185.0
	Total	**-1**	**73**	**0**			**185**

* Values for bleaching, pulp drying, and wastewater plants came from Reference [39].

In comparison with other systems on heat, power, net, etc. from Reference [46], the yellow star is the reference plant in this work. The red star corresponds to values of a proprietary integrated pulp and paper mill we have available, where the additional NG combustion provides only about 15% of produced steam, another 15% comes from the hog boiler, and the rest comes from recovery boilers. To show how the pulp mills can vary, the red star mill is producing 95% of the electricity demand, also uses a backpressure turbine but about 6% of the steam flow is not used for heat demand and is dumped in an auxiliary condenser.

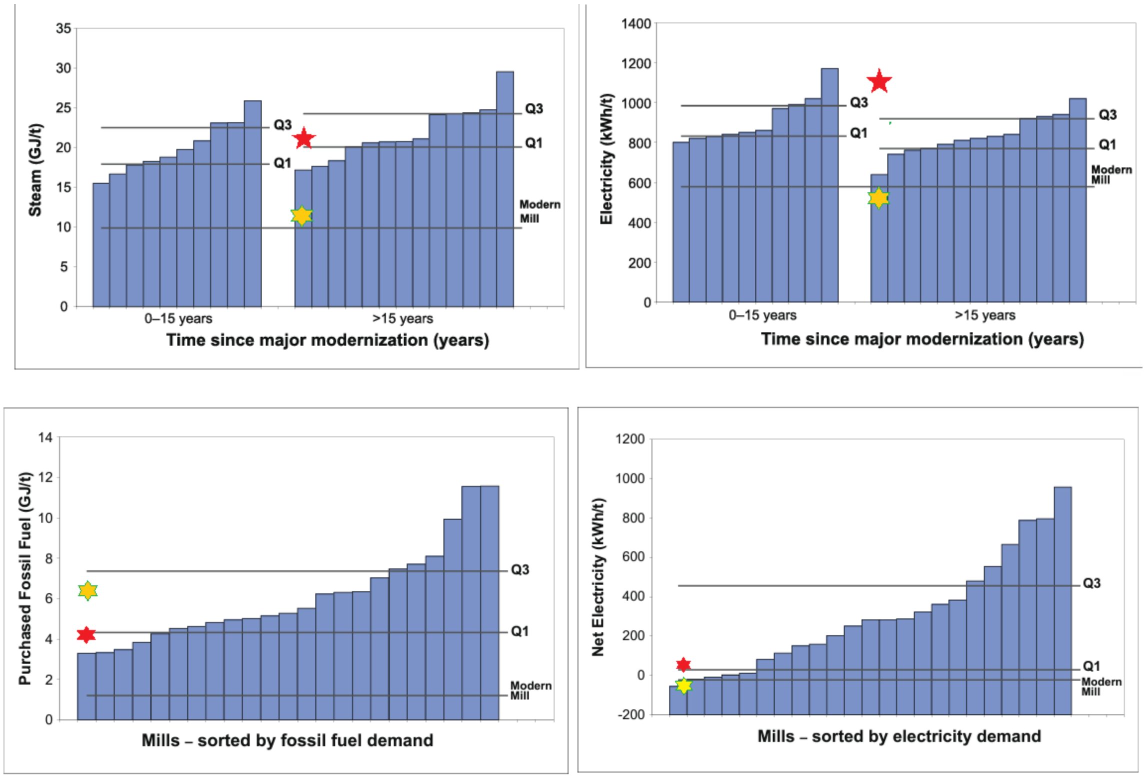


Figure 28. Comparison between a reference kraft mill (yellow star), actual mill with data available at Idaho National Laboratory (INL) (red star), and kraft mills from Reference [46] for (from top left) process steam consumption, (top right) process electricity consumption, (bottom left) purchased fossil fuel demand, and (bottom right) net electricity demand.

4.4.4.2 CO_2 Production

As shown in Figure 29, the recovery boiler is the main source of CO_2 emissions in a pulp and paper process (~90–85% of the total), as a result of black liquor combustion. The remainder of the total emissions (10–15%) stems from other fuel combustion (woody biomass and fossil fuels) activities, along with the CO_2 emitted from converting $CaCO_3$ into CaO in the lime kiln.

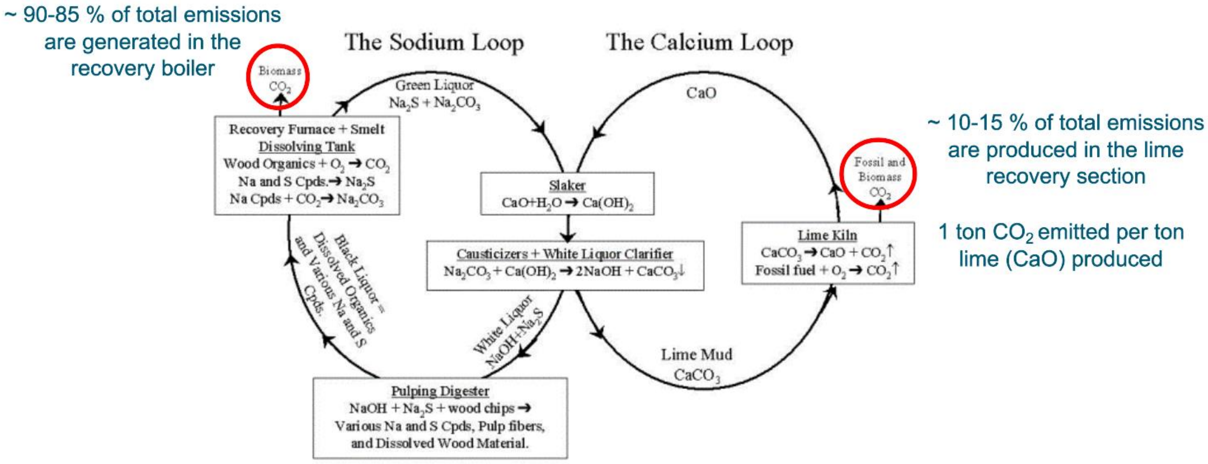


Figure 29. CO_2 sources in kraft mill pulping [46].

An initial correlation for the total CO_2 emissions produced by all operative kraft mills in the United States was conducted by gathering all the plant capacity data and their corresponding GHG emissions (CO_2-eq.) (see Figure 30) [1]. The total emissions generated (biogenic + non-biogenic) were seen to roughly correlate with the plant capacities, though the coefficient of determination (r^2) suggests a small correlation (metric tons CO_2-eq./year). However, a positive relationship is observed between the number of finished tons of dry pulp and the emissions generated by the plant. The variability can be attributed to the variety of kraft pulping equipment and machinery, the age and energy efficiency of the mills, and the carbon capture technologies involved. Initially, for a reference capacity of 400,000 ADt/year, the foreseeable emissions produced under the relevant assumptions seem to be around 748,000 metric tons of CO_2-eq. per year.

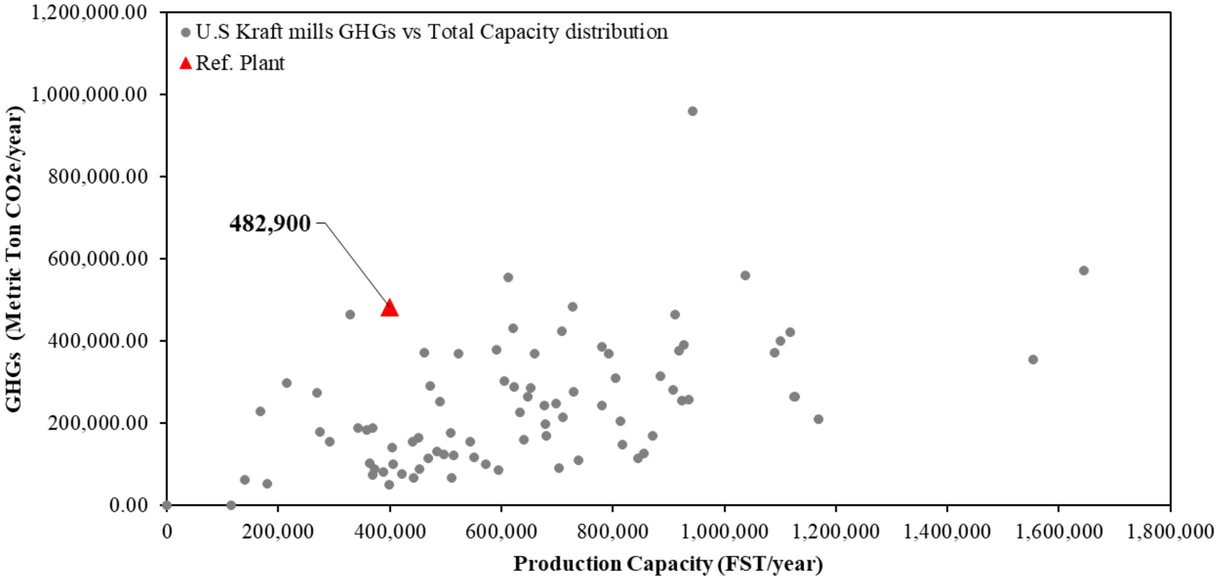


Figure 30. GHGs vs. production capacity correlation for U.S. kraft pulp production.

The CO_2 emissions from the reference plant were derived from Aspen models for the lime kiln and BLRB. The reference plant emissions are shown in Table 18.The total U.S. emissions were obtained from the EPA FLIGHT tool, then divided up by process unit, based on the percentage of carbon emissions from each process unit, as reported by the EPA in 2021 (see Table 18) [29].

Table 18. Direct CO_2 emissions for the reference pulp mill and at the national scale.

PO #	Process Unit	Reference Plant CO2 Emissions	National Scale CO2 Emissions
		Metric Tons per Year	Million Metric Tons per Year
A	Wood processing		
B	Hog Fuel Boiler	3,000	20
C	Pulp Plant		
D	Bleaching		
E	Pulp Drying		
F	Evaporation Plant		
G	Lime Kiln + Causticizing	126,003	19
H	Recovery Boiler	597,360	57
I	Wastewater plant		
J	Natural Gas Boiler	90,859	
	Total	**817,222**	**97**

5. NUCLEAR INTEGRATION

Three separate technoeconomic analyses were performed for a methanol plant [47], petroleum refinery [48] and pulp plant [49] to assess the value of integration with an NPP. For a more thorough description of the technoeconomic analysis results, refer to the cited reports for each sector. The work in this section summarizes the selected relevant results from the three studies, extends the work by further scenarios considered feasible and describes the technical aspects of the thermal integration with a high-temperature gas-cooled reactor (HTGR) model. These studies used the reference plant configurations defined in this report.

All nuclear integration case studies are based on an HTGR system, using parameters of Xe-100 modules (200 MWt/80 MWe) for reference. After developing and tuning a thermodynamic model to the reference plant parameters, the models were extended by steam extraction for CHP operation. Steam is used at multiple pressure levels for process heating, direct drive of equipment and injection into the processes, or as a feedstock for high-temperature steam electrolysis (HTSE), which is a source of hydrogen.

In all cases studied, the power cycle steam (reactor side) is physically separated from the process stream (user side) by isolation heat exchangers. This separation is dictated by safety and regulatory requirements and avoids disrupting the specified chemistry and purity levels of nuclear power cycle steam. Industrial process steam has generally lower purity and chemical treatment requirements and often has a significant make-up rate (losses, injection into the process) and has a higher risk of steam and condensate contamination by leaks in the process steam use from the heated media. Additional isolation heat exchangers are also used for HTSE feedstock steam. Specific parameters of the power cycle and design considerations regarding the separation heat exchangers can be found in Appendix E-1.

5.1 Refinery Nuclear Integration

This section outlines general opportunities for nuclear-sourced energy carriers' integration into the refinery operation and details specific integration scenarios for the reference refinery from Section 2.4. Multiple nuclear integration and decarbonization options are provided, demonstrating the extent of nuclear's potential in the refining industry. The reference refinery designs break down energy demands based on unit operation, allowing for the creation of a process steam model based on demand at each pressure level. This approach allows also for a variety of integration strategies such as increasing (liquefied petroleum gas) LPG production—which lowers the internal byproduct fuel utilization and requires higher external energy input—or meeting only LP steam demands with the NPP while electrifying originally HP steam demand used for equipment direct drives by small turbines. By analyzing each scenario, we identify methods to fully substitute current energy needs at a refinery with nuclear-sourced energy carriers. In the last section, we evaluate the national potential for nuclear energy across the refining sector for supplying all external energy demands.

5.1.1 Nuclear Integration Opportunities for Refineries

Nuclear energy can be integrated within refineries by multiple pathways, as illustrated in a generic BFD in Figure 31. The refinery requires hydrogen supplied typically via an SMR system, high-temperature heating from the combustion of fossil-based fuels, electricity, and process steam. With the proper integration, nuclear power can substitute for all these demands. Integrating nuclear power into the process steam process steam and power demands is relatively straightforward. We will show that, with some refinery modifications, it may be sufficient to supply only LP steam while still meeting all refinery steam and power demands. As an alternative to SMR, hydrogen can be provided via HTSE powered by the NPP. For the refinery's high-temperature heat requirements—beyond what is provided by RFG—two pathways are considered to utilize nuclear energy: producing additional hydrogen that can be combusted as a fuel and direct electrification, which avoids the capital cost and conversion efficiency of electrolysis. Specific scenarios considering these pathways, and their results are discussed in Section 5.2.2.

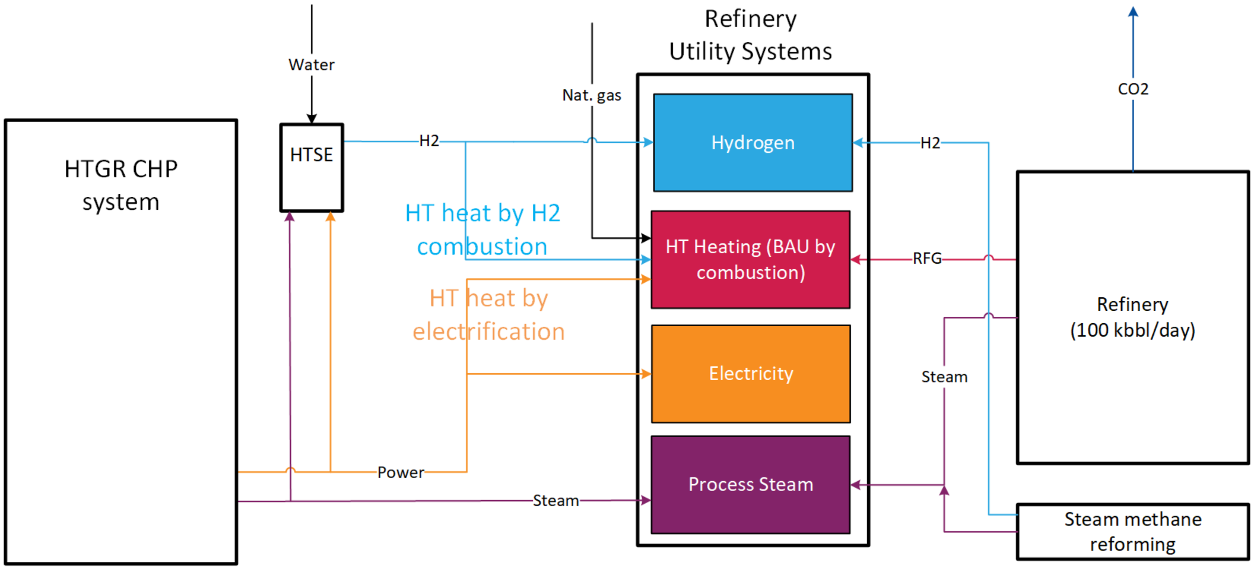


Figure 31. Generic BFD of possible nuclear integration pathways within refinery.

The easiest targets for emissions reductions are the refinery's CHP system, consisting of NG boilers, combustion turbines with heat recovery steam generators. These systems use imported fuel rather than light ends and other waste products stemming from the crude oil processing operations. An advanced nuclear reactor with a CHP system could be incorporated into the plant to generate power and supply HP, MP, and LP steam. Adding a heat source to the steam headers is typically a minor retrofit operation. The feasibility of steam delivery to a refinery from nuclear sources has been proven by demonstration. China recently commissioned a system delivering about 340 MWth of steam from two VVER-1000 reactors to a large petrochemical (835,000 bbl/day) and chemical processing complex, delivering MP steam over distance of 23 km at parameters of 1.8 MPa, superheated by 40 K to 248°C with a system cost of 14,000 /(MWt-km). [50–53]

In addition to steam and power, refineries are large consumers of hydrogen gas, which is used as a reagent in residue cracking and other chemical conversion processes. This hydrogen is derived from fossil fuels via SMR. The associated CO_2 emissions can be avoided by shifting hydrogen production to nuclear-powered HTSE. Meeting the hydrogen demand at the reference refinery via HTSE requires about 184 MWe in direct current (the reported energy demands vary) and 31 MWt of LP steam. When using proton exchange membrane (PEM) electrolyzers instead, oxygen could be utilized and substitute current supply via ASUs. In case of HTSE, only oxygen rich exhaust is available, for which it is harder to find some additional use.

Refineries also require high-temperature heating for a variety of processes. These high temperatures are obtained by combusting RFG and imported NG. The primary use for the RFG is its combustion in furnaces. The ratio between the RFG and required import of NG can vary significantly. The reference refinery requires about 18 MW of external NG (5% of heat duty for furnaces) and when substituting a hydrogen source, only 2 MW (1%) of external NG. If the LPG production is maximized in the reference refinery model, the demand increases to 186 MW (48%). Note that this is a very site-specific parameter since some plants have an excess of RFG production, and in these cases, a significant portion of the RFG remains after providing all high-temperature heating, which is combusted directly in steam boilers.

Theoretically, an HTGR may be able to supply some of this high-temperature duty; however, the implementation challenges are numerous. A new heat exchanging system from the primary loop to intermediate high-temperature heat transfer fluid would need to be designed and consequently approved through regulations and safety procedures, partially removing the advantage of SMNR modularity. Furthermore, replacing a fired heater with a nuclear-derived heat transfer loop represents a large retrofit operation, as it would first require new heat exchanging approach, equipment, and control strategy. It could also upset downstream heat exchanger networks that rely on waste heat from the furnace. This type of integration has a low technological readiness level and high complexity, cost, and risks. Technically feasible options for decarbonizing the refinery's high-temperature heat needs are either blending hydrogen into fuel systems (if burners allow a high hydrogen content or are retrofitted) or electrification.

5.1.2 Reference Refineries Integration Models and Strategies

This section details three main approaches with an alternative for each, totaling six strategies for decarbonizing refineries with nuclear integration. The following strategies were evaluated in this work:

1. The nuclear CHP system satisfies only the steam and power demand of the refinery.
    a. The NPP delivers only LP steam and power to the refinery. HP steam demand is provided by internal steam generation and electric motors instead of turbine direct drives.
2. The NPP delivers HP steam and power. The refinery's H_2 demand is supplied from HTSE fed by the NPP.
    a. The NPP delivers only LP steam and power. The refinery's H_2 demand is supplied from HTSE fed by the NPP.
3. The NPP satisfies all steam and power demands of the refinery. The refinery's H_2 demand is supplied from HTSE fed by the NPP. The high-temperature heat demands that are not satisfied by RFG are provided by electrical heating. This case has a higher high-temperature heat demand because LPG production is maximized.
    a. The NPP satisfies all steam and power demands of the refinery. The refinery's H_2 demand is supplied from HTSE fed by the NPP. The external demand for high-temperature heat is provided by blending additional H_2 from HTSE into the fuel systems. This case has a higher high-temperature heat demand because LPG production is maximized.

Integration Case 1 uses the previously developed process steam model and the reference steam cycle to match the demand of the power and process steam. Looking at the reference refinery steam demand and refinery internal production in greater detail, it was discovered that the external steam demand for HP steam is entirely related to direct drives of selected compressors and pumps. This suggests another nuclear integration path (Case 1a) into the process steam system: delivering only LP saturated steam from the nuclear CHP system over a possibly extended distance between the nuclear and industrial sites. The direct drive turbines can be substituted with electric motors, amounting to an additional electricity demand of 3.5 MWe (for conservative demand results assuming a high turbine efficiency of 90%). For system startup or transient states with insufficient HP steam production, a steam compressor may be added between the LP steam and HP steam line with steam reboilers, or a small backup boiler may be installed. Figure 32 shows a representative BFD for these two cases. A detailed PFD of the process steam and nuclear CHP system can be found in Appendix E-2.1.

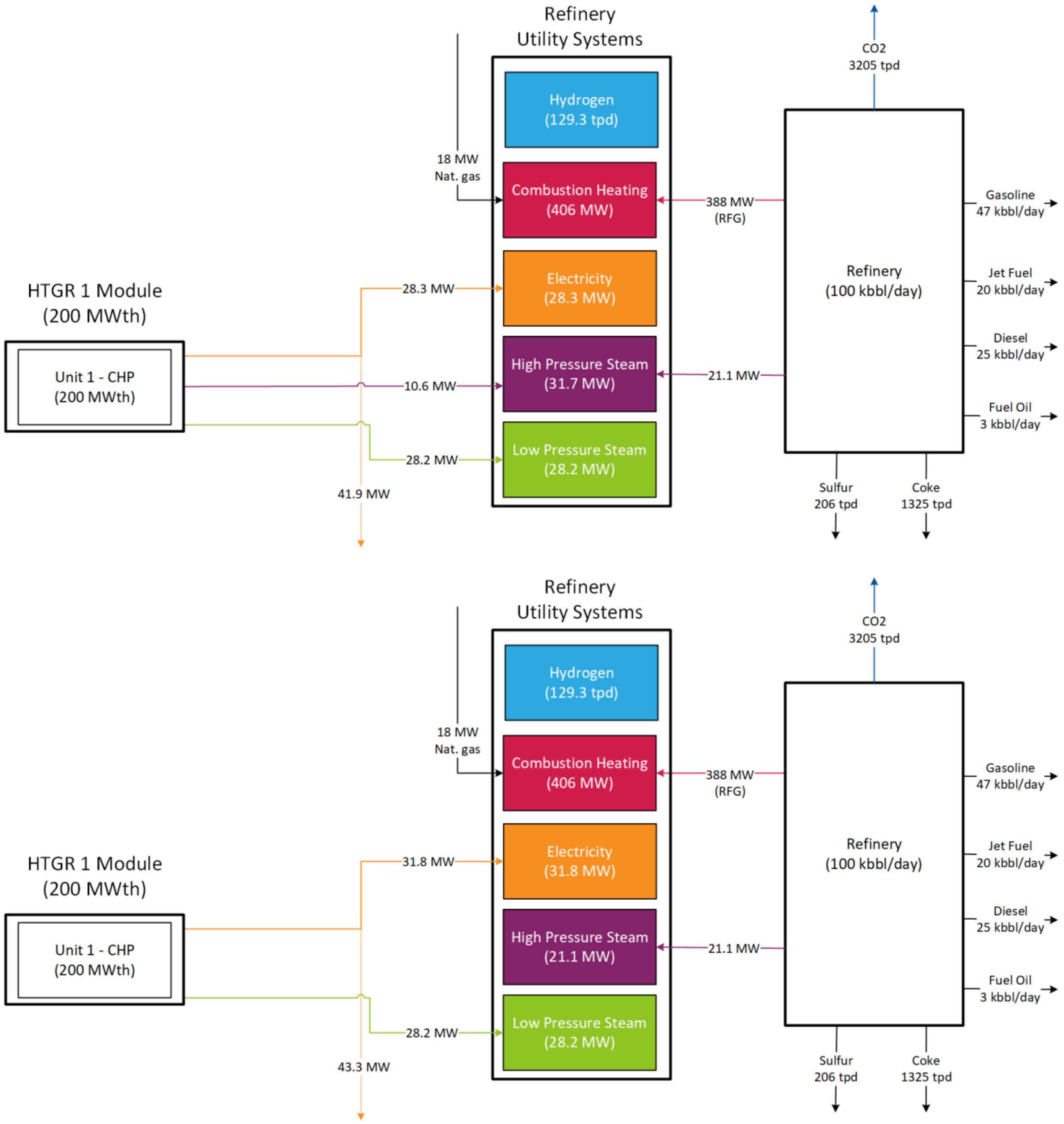


Figure 32. BFD of Refinery Integration Cases 1 (top) and 1a (bottom).

Case 2 substitutes the H_2 production from an SMR with solid oxide electrolyzers, named here as HTSE. The HTSE system increases the overall demands for both steam and power. At the HTSE inlet, feedstock steam is generated by evaporating and slightly superheating deionized water with the process steam system and an additional letdown turbine. The high temperature inside the HTSE stack (750°C) is achieved by internal heat recuperation; therefore, the feedstock steam is sufficient around 150°C, while the HTSE technology can only operate near atmospheric pressure (unlike pressurized PEM electrolyzers). The H_2 product is then dried, compressed, and delivered to the refinery H_2 distribution system operating at around 2 MPa. Without the fuel demand for the steam methane reformer, RFG production can then cover most of the overall fuel demand, reducing the external fuel requirements to only 3 MW. Case 2-a is similar to Case 1-a, in which the LP steam delivery approach is also used while also adding the HTSE system. The only difference in the layout between Case 1-a and Case 2-a is that the stream for superheating the HTSE feedstock does not come from the HP steam line but from a dedicated steam compressor. Figure 33 shows a representative BFD for these two cases. A detailed PFD of the process steam and nuclear CHP system can be found in Appendix E-2.2.

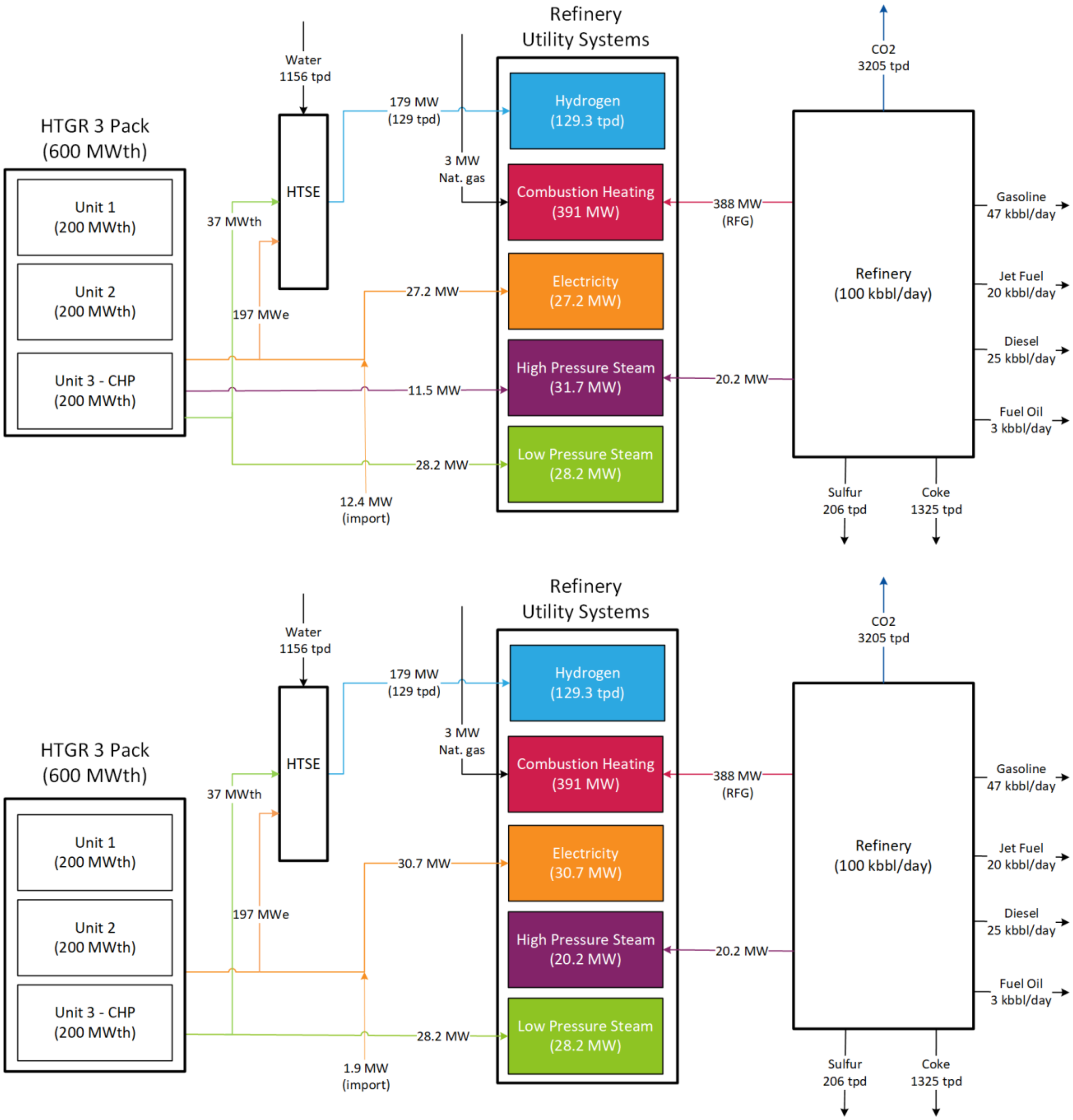


Figure 33. BFD of Refinery Integration Cases 2 (top) and 2a (bottom).

The external HT heat demand at the reference refinery is very low, so it had been modified before illustrating the decarbonization pathways in Case 3 and Case 3a. The reference plant’s external fuel demand is (after the NPP-based H_2 supply) only 3 MW (effectively a rounding error), so it does not make sense to test this integration pathway with the current reference configuration. Adding the LPG production of the reference refinery resulted in less C3 and C4 hydrocarbons to be used as RFG. The resulting decrease in RFG production increases the external fuel demand to 185.6 MW. For the nuclear integration scenarios, electrification (Case 3) is selectively applied to several unit operations while the rest of the system combusts available RFG. Electric heaters substitute 185.6 MW worth of furnaces, and by avoiding stack loss, the plant electricity demand increases by 157.8 MWe. The second approach (Case 3a) uses hydrogen as a substitute fuel for the NG demand. Since the fuel substitute is burned in the same or similar burners and furnaces as in original refinery, the stack losses are maintained and that means the full 185.6 MW in LHV worth of hydrogen is required. As such, the refinery approximately doubles the hydrogen production it requires for refinery operations. Figure 33 shows a representative BFD for these two cases. A detailed summary of the system and comparison with other scenarios can be found in Appendix E-2.3

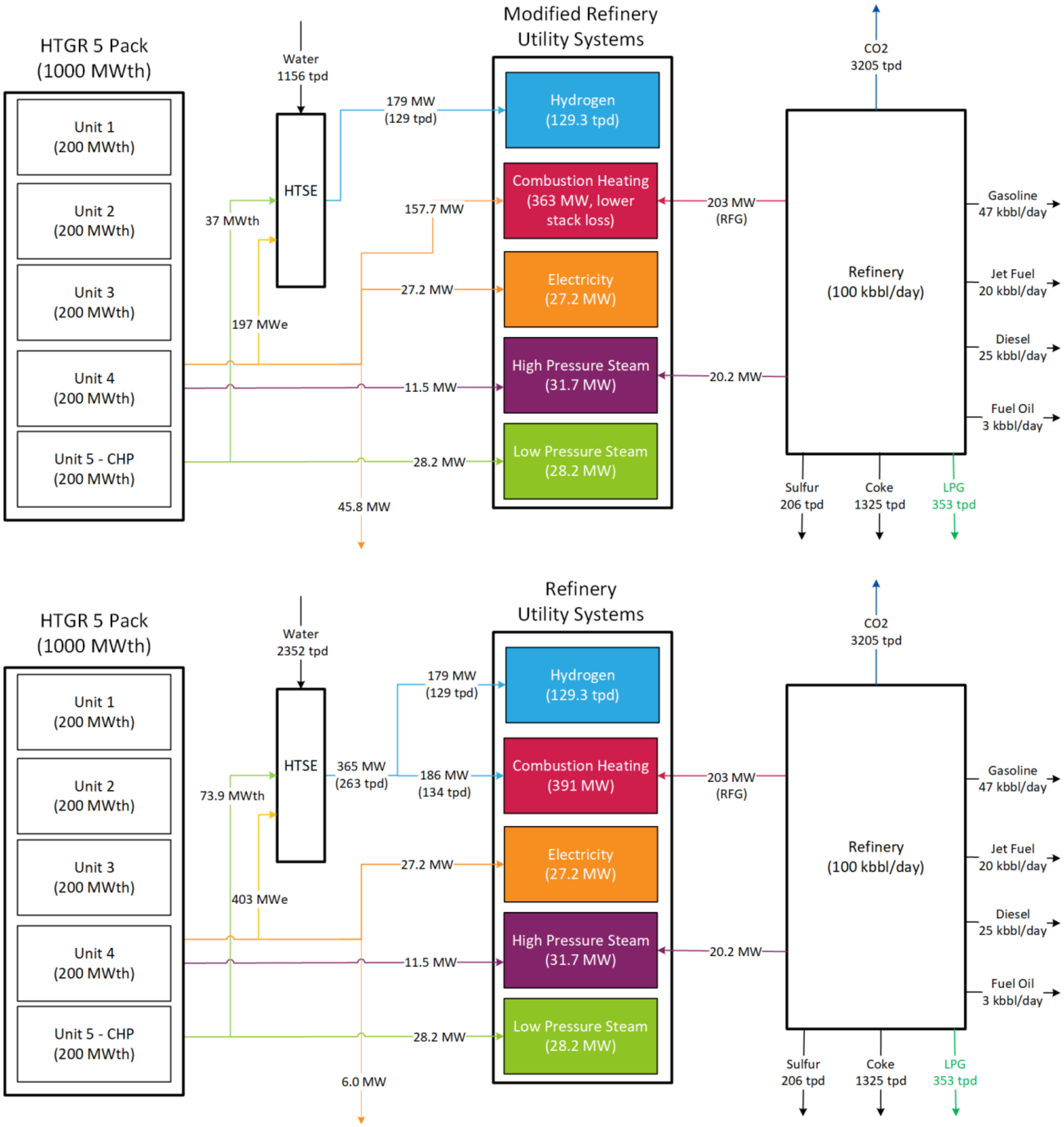


Figure 34. BFD of Refinery Integration Cases 3 (top) and 3a (bottom).

All six scenarios are summarized in the following tables. Table 19 gives the relative energy demands provided by nuclear energy for each case, and Table 20 shows each of these demands by unit operations. In each scenario, the reactor output is quantified as the total demand (i.e., not number of specific modules) to obtain a net-zero balance (no import or export) of all energy supply types.

Table 19. Resulting parameters of HTGR integration into reference refinery—reduction in energy demands.

	Scenario	Electricity Supply	HT Heat Supply	Steam Supply	H_2 Needs Supplied	Total External Energy Demand Reduction	HTGR-Type Reactor Power Required (MWth)	Reactor Power per Supplied Energy (MWth /$MW_{supplied}$)
1	Steam + Power only	100%	0%	100%	0%	25%	98	1.46
1a	LP Steam + Power only	100%	0%	100%	0%	25%	95	1.42
2	Steam + H_2	100%	0%	100%	100%	99%	631	2.64
2a	LP Steam + H_2	100%	0%	100%	100%	99%	605	2.53
3	Steam + H_2 + HT heat electrif.	100%	100%	100%	100%	100%*	1,102	2.55
3a	Steam + H_2 + HT heat via H_2	100%	100%	100%	100%	100%*	1,203	2.79

* Higher basis in HT heating net demand due to lower RFG production.

Table 20. Refinery demands per unit product for considered scenarios.

	Scenario	Electricity demand (MW/100k bbl/day)	NG Demand (MW/100k bbl/day)	Steam Demand (MW/100k bbl/day)	H_2 Demand (MW/100k bbl/day)
0	Business as usual (BAU)	28.3	18.4	38.8	179
1	Steam + Power only	28.3	18.4	38.8	179
1a	LP Steam + Power only	31.8	18.4	28.2	179
2	Steam + H_2	27.2	2.2	39.7	179
2a	LP Steam + H_2	30.7	2.2	28.2	179
3	Steam + H_2 + HT heat electrif.*	186.1	0	39.7	179
3a	Steam + H_2 + HT heat via H_2*	27.2	0	39.7	365

* In these cases, the basis used for the HT heating (186 MW) net demand reduction was higher due to lower RFG production. Electrification demands avoid stack loss.

These integration cases show that an NPP can provide up to 100% of considered external energy inputs. This excludes eventual auxiliary systems that may be added or are specific to each refinery configuration, such as an oxygen or nitrogen supply. The reference plant also represents a refinery with already optimized energy demands, and that may not be the case at existing refineries. We have also shown a progressive approach with minimized demand for an HP steam, such that LP steam delivery may be sufficient in specific cases. In this case, a rather small amount of external demand for MP and HP steam, eventually for short-term situations such as startup operations, etc., can be supplied by steam compression or backup boilers.

The last column of Table 20 is effectively an inverse value of the energy conversion efficiency between nuclear heat and all refinery demands supplied in each scenario. Lower numbers (e.g., Case 1 and Case 1a) show the best efficiency, but efficiency decreases when supplying hydrogen (e.g., Cases 2 and 3a)—the majority of losses comes from the electricity needed for electrolyzers. Electrification (e.g., Case 3) has similar efficiency values to the hydrogen cases, as it is affected by losses in the heat to electricity conversion but takes advantage of stack loss avoidance. If hydrogen was to be used as a fuel (Case 3a), the efficiencies of hydrogen production and consumption would further reduce the overall efficiency of the energy supply and add to the needed nuclear reactor power.

5.1.3 National Potential for Nuclear Integration into Refineries

From Integration Case 3, we quantified the theoretical national capacity potential of nuclear systems to provide refinery energy needs. This case uses the greatest nuclear penetration into refinery demands and has the highest reduction of CO_2 emitted.

The primary purpose is to demonstrate the high-level upper limit of the opportunity for nuclear systems. Footprint size, legal, or geographical limits that would prevent feasible nuclear integration are therefore not considered. The national-level business as usual (BAU) of energy demands is adopted from Reference [54] with several simplifying assumptions. From this reference, all steam is assumed to come from the natural gas combined cycle CHP system following the defined efficiency in the reference, and additional electricity production is assumed to come from a natural gas combined cycle plant with 50% efficiency. This means all heat and power production at the refinery can be combined into a single CHP node. All fuel demand is assumed to go into fired furnaces with 85% efficiency (15% stack loss). Hydrogen production is independent of the refinery (usually provided by an external, though collocated, vendor) and supplied by dedicated SMR systems with 70% efficiency [55]. Steam produced by SMR can also be integrated into the refinery steam system and a steam production rate from the reference plant is adopted.

In the integration case, the NPP is assumed to be an HTGR type. The extraction rate from the CHP systems is adjusted to an average plant derived from the national demand, and the ratio between HP and LP steam demand is taken from the reference plant. H_2 production is assumed to come from a dedicated HTSE system with power and steam provided by the HTGR NPP. While existing furnaces maintain their 85% efficiency, electrified demand for high-temperature heat assumes zero losses when converting power to this heat. The selected integration pathway is substituting all external energy inputs except for feedstock (i.e., crude oil).

Table 21 provides an overview of the demand and nuclear potential with a more illustrative Sankey diagrams for BAU situation in Figure 35 and nuclear integration potential illustrated in Figure 36. Nuclear opportunity is presented in thermal output of the reactors (i.e., primary energy needed for conversion into the refinery demands in the required energy form). While nuclear CHP integration for power and process steam can minimize conversion losses, the electrification of high-temperature heat and hydrogen production substantially contributes to energy conversion losses, as power is converted back to heat. Therefore, primary nuclear (thermal) energy required is almost 60% higher than for BAU with NG fuel. Note that the analysis does not go into the level of detail as for the reference plant and rather illustratively provides a general overview of the current situation and selected nuclear integration pathway.

Table 21. Oil refineries national-level energy demand and potential of nuclear integration for substituting this supply.

No.	Energy Consumption Type	National Scale GW	Nuclear Opportunity GWth (reactors)
1	Steam	27.0	36.8
2	Electricity	7.6	1.2*
3	Fuel (total demand)	71.4	64.26
4	Fuel internal byproduct production (fuel gas and petroleum coke)	-45.7	(reduces fuel substitute)
5	Hydrogen (LHV)	27.4	80.8
	Total	87	183.0

* Most of the electricity demand provided by steam producing CHP systems, here is quantified only additional demand.

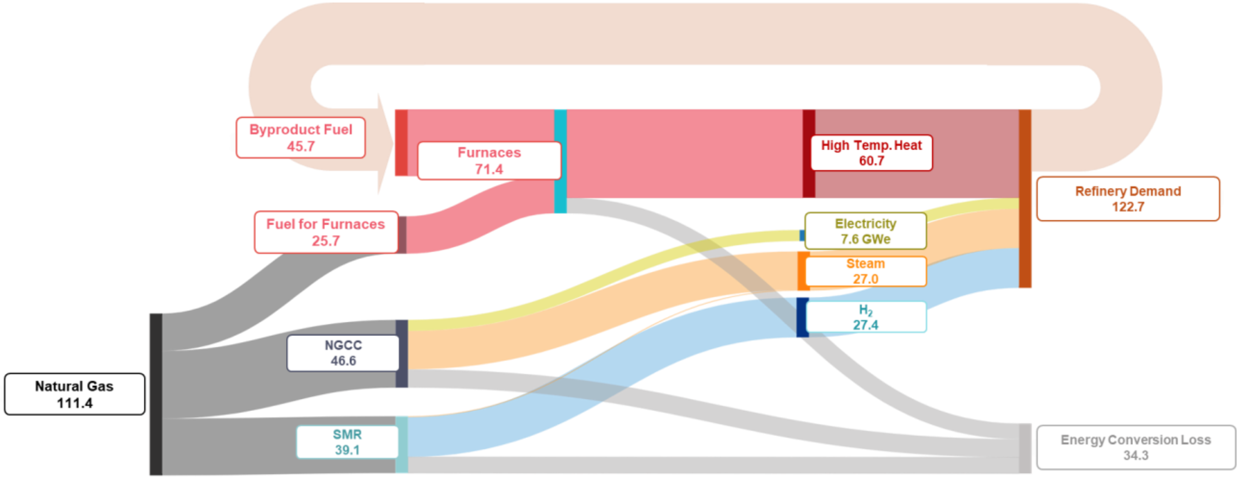


Figure 35. Sankey diagram of national-level energy demand and supply for oil refineries.

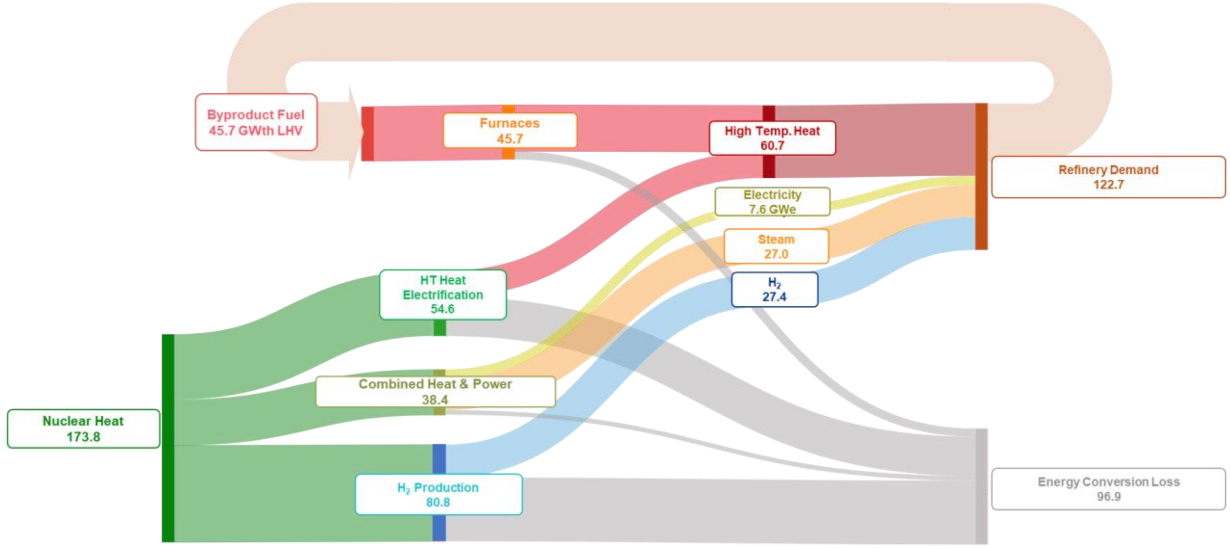


Figure 36. Sankey diagram of national potential of nuclear-integrated scenario for energy demand and supply for oil refineries.

Regarding CO_2 emissions, taken from the same source as energy data [54] and adding the hydrogen production related emissions, decarbonization by the three separate uses of nuclear energy used in the national-level analysis is presented in Figure 37. The remaining emissions are mainly associated with the combustion of the byproduct fuels.

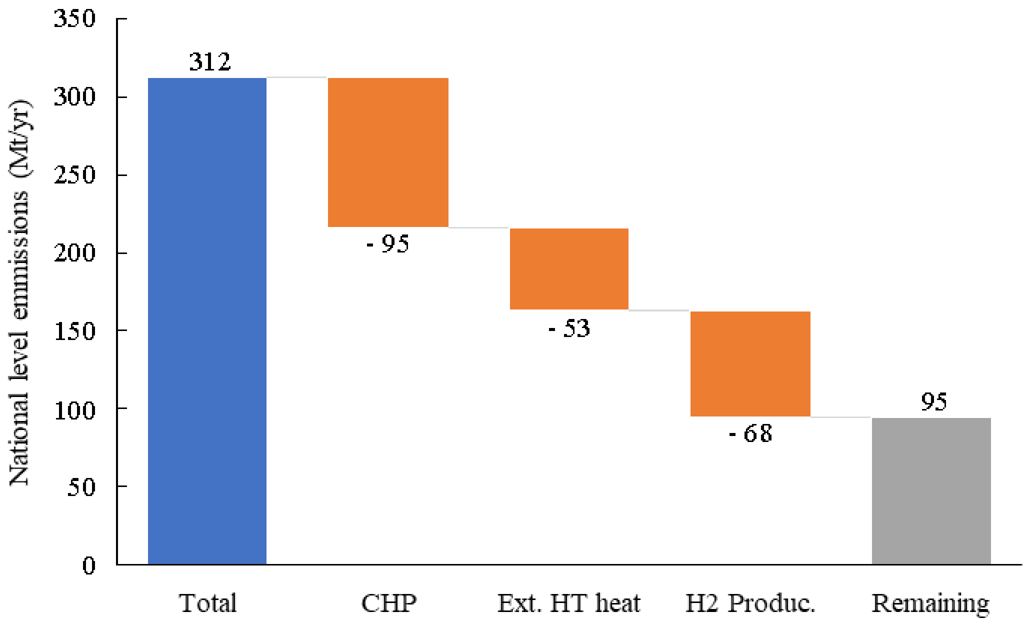


Figure 37. Refineries emission reduction potential from separate uses of nuclear energy used in the national-level analysis.

5.2 Methanol Plant Nuclear Integration

The refinery integration from the previous section was based on providing energy inputs distributed by an internal utility system with no or minimal impact to the unit operations. Methanol production, on the other hand, corresponds to single (or only few) unit operation, and nuclear integration is therefore on a process level. The integration will be shown on two levels. First, the reference plant used here (as described in Section 3.3) requires high-temperature heat for the SMR furnace coming typically from NG, the replacement of which will be demonstrated via hydrogen combustion or electrification. The second option involves effectively an alternative process with feedstock being (nuclear produced) hydrogen and a stream of CO_2. Finally, national potential integration is presented for this deepest decarbonization, producing green methanol.

5.2.1 Nuclear Integration Opportunities

For a conventional methanol plant, chemical and thermodynamic models described in Section 3.3.2.1 and 3.3.2.2 demonstrate that the reference methanol plant only takes in energy in the form of NG (see Figure 12). All electrical and steam requirements of the reference plant can be satisfied by the recuperative cooling of key intermediate and product streams. However, this may differ in the case of a feedstock change, which would be associated with a fundamental change of the process. Nuclear energy has limited pathways for integration in the conventional methanol configuration as illustrated in a generic BFD in Figure 38.

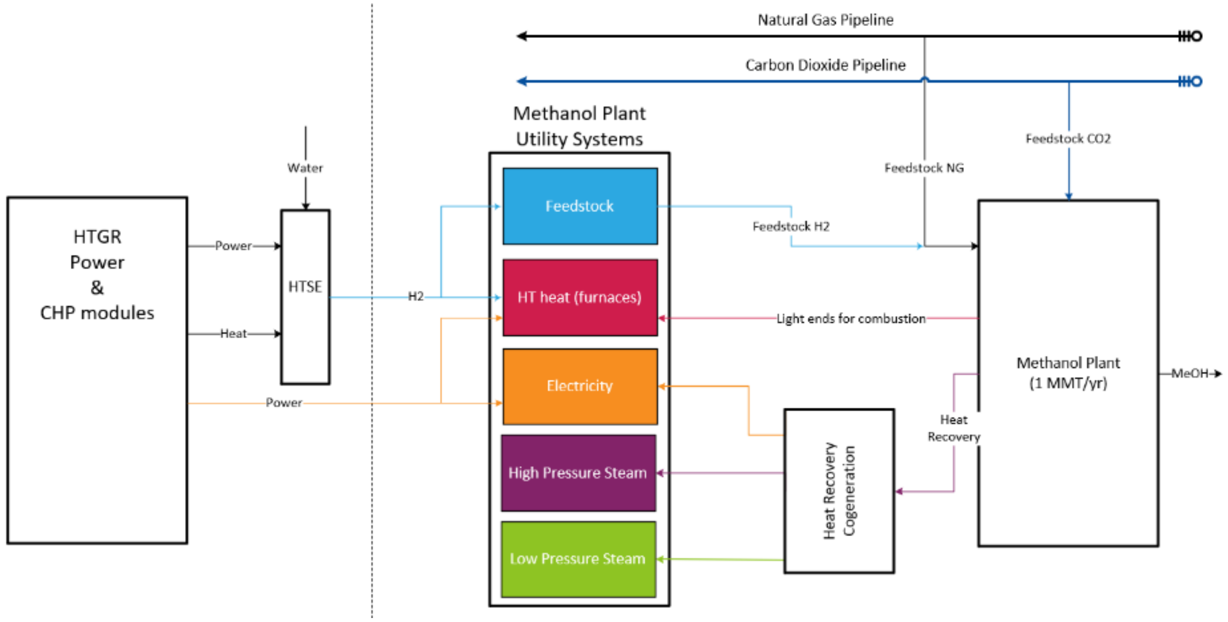


Figure 38. Generic BFD of possible nuclear integration pathways within methanol plants.

The first approach focuses on the furnace, which typically burns NG to provide the high-temperature heat (near 850°C) for the reformer in a near-isothermal heating process. Alternatively, hydrogen can be substituted as the fuel. Fuel switching, however, can impact the burners and change the composition and flowrate of the flue gas, so this is still considered a major retrofit for the plant. As with the refinery, electric heating is another route to replace NG combustion. Using hydrogen, the stack loss remains, but with electrification, the power consumption is significantly decreased by its avoidance. The electrolyzer and hydrogen management system also significantly adds to the system cost.

Electric heating would reduce energy losses by eliminating stack loss, and its feasibility would not rely on the distance between the NPP and the point of use. Electric heaters are increasingly available for industry on a large scale, and except for common resistive heating, new approaches with potential to reduce cost are being developed based on infrared [56] and induction [57] heaters.

An alternative approach to the decarbonization of methanol production is to replace SMR by using CO_2 and H_2 as feedstocks with technologies that do not need the furnace. In this work, this is done by using a reverse water gas shift (RWGS) reactor as an intermediate step for syngas production. This approach provides increased decarbonization, as the CO_2 feedstock is captured from other sources but would require significant energy and hydrogen inputs.

Alternatively, there is existing commercial technology that avoids the intermediate syngas step and thus has wider applications and scalability to utilize sources of cheap renewable electricity and is readily available to source CO_2. Via this method, Carbon Recycling International achieved the world's first industrial-scale commercial production of methanol in 2012, using low-cost baseload renewable electricity and an available CO_2 source. More installations and projects have been announced in the time since, including large-scale systems, [58] so this approach can be considered a proven technology.

Additional options were considered but not pursued. Heat from an HTGR, at an outlet temperature of 750°C, is insufficient for the reformer temperatures, which exceed 850°C near the isothermal heat sink. A specifically designed high-temperature heat transfer system could be used only for partial feedstock preheating, but the cost and losses of such a system are unlikely to outweigh the potential benefit. The fuel input of the reference plant could theoretically be decreased by 27% (88 MWt) if the heating train was converted to accommodate a feedstock gas-steam mixture preheat directly from the HTGR primary or new intermediate loop (with the heat transfer fluid reaching 750°C).This configuration would require less than half of a single Xe-100-type HTGR thermal output, while it would also prompt notable modifications of the furnace itself, primarily increasing required combustion air preheat.

Clean steam and electricity from an SMNR could also be integrated to power the co-electrolysis of CO_2 and water to produce a syngas stream of carbon monoxide and hydrogen. This syngas would be compressed and sent to the synthesis reactor for methanol production. INL is currently conducting research on this co-electrolysis process.

5.2.2 Reference Methanol Plant Integration Models and Strategies

Two methanol refinery configurations were considered in this analysis: conventional methanol production and synthesis from hydrogen and CO_2 with the RWGS step. A total of three integration strategies were selected. Scenarios 1 and 2 were primarily developed in [47].

4. The SMR furnace NG fuel demand is substituted by hydrogen produced via HTSE.
    a. The SMR furnace NG fuel demand provided by electrification.
5. Fuel and feedstock fuel substituted by H_2 produced via HTSE, including RWGS process.

The most direct nuclear integration pathway is to blend nuclear hydrogen, generated by HTSE, into the fuel gas, as depicted by Figure 39 (Case 1). In this scenario, nuclear hydrogen (detailed in Appendix E-3.1) substitutes NG (as the fuel gas) demand in the SMR furnace. Substituting NG for hydrogen in various proportions has been explored in other work [47]. While this pathway appears as a minimal impact to existing plants—the combustors and fuel system would need to be retrofitted—other parts of the furnace may be impacted due to different flame properties and flue gas parameters (volumetric flow, heat capacity, dew point, etc.).

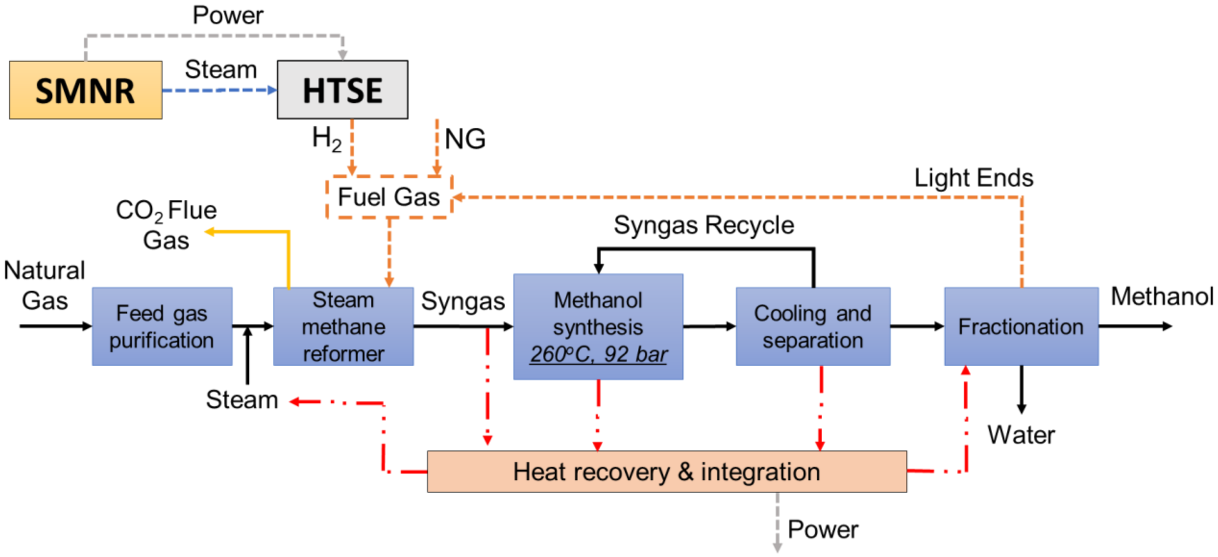


Figure 39. Standard methanol synthesis via SMR with HTSE and H_2 used for fuel substitution. [47]

Since Case 1 would require refurbishing of the furnace, a potentially more feasible solution is to substitute the NG duty completely by electrification (Case 1a). In this case, the retrofitted furnace would be significantly smaller than in BAU or Case 1, combusting only the light ends for feedstock preheating. Electrified heating has no associated conversion losses as with hydrogen production and avoids 15% of the stack loss. With the 2 MWe balance of the waste heat recovery (WHR) system and baseline internal power loads, the net requirement is 147 MWe, which corresponds to only two reactor modules with 13 MWe exported. BFDs for the Case 1 and Case 1a are in Figure 40.

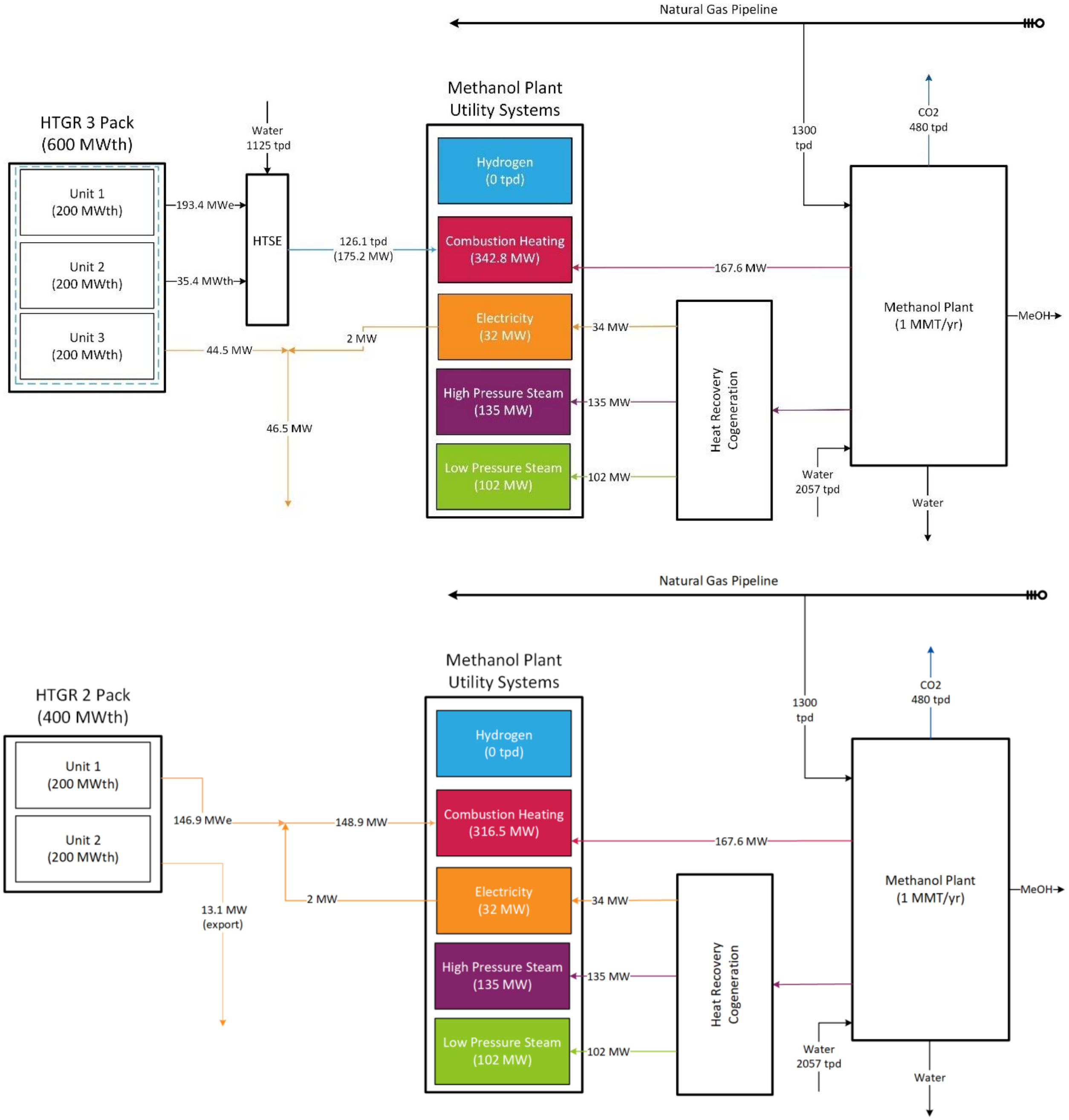


Figure 40. Nuclear integration BFD for methanol plant by hydrogen combustion (top) and partially electrifying (bottom) SMR furnace.

With more significant modifications, syngas can be produced using the RWGS, which utilizes an industrial CO_2 source that would otherwise be emitted from another facility (Case 2). This process would require a large amount of H_2 from an SMNR; a schematic of the RWGS-based methanol synthesis process is included in Figure 41. Through the RWGS pathway, nuclear power is used to reduce CO_2 to methanol through a multistep process. Similarly to the conventional process, light ends are burned to provide heat for the RWGS reactor. There are emissions from the combustion of the light ends, but they are significantly less than the amount of CO_2 getting converted into the product. Depending on the end use of the methanol, the lifecycle emissions for the RWGS synthesis may be close to carbon neutral or net -negative.

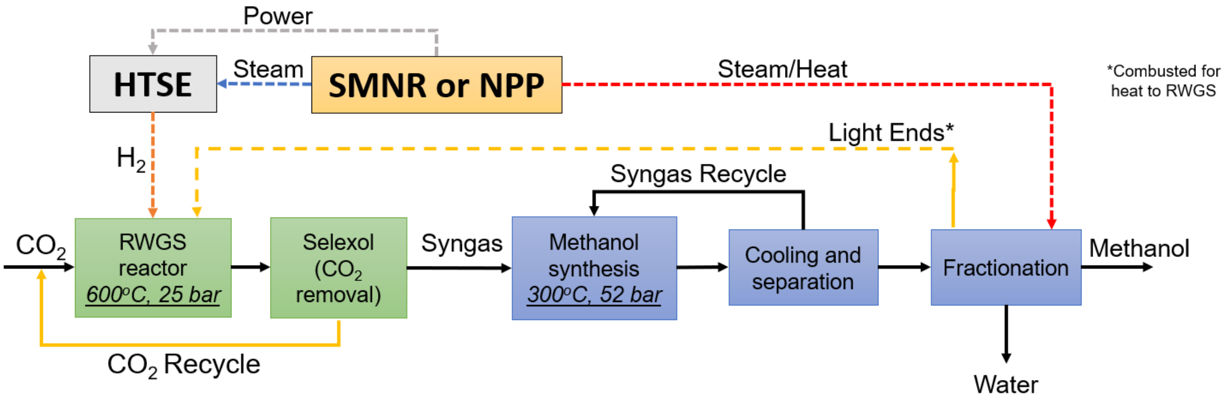


Figure 41. Methanol synthesis via RWGS with SMNR and HTSE. [47]

The chemical conversion process and some direct heat integration for this approach was modeled in Aspen Plus. Indirect heat integration via the steam WHR system was modeled in Cycle-Tempo (see Appendix E-3.2). Since this process is substantially different from the reference plant, a new energy datasheet has been developed, as seen in Table 22. Internal power production, light ends combustion, WHR, and process steam system provide all required heat and steam, but comes 28 MWe short on electricity, that needs to be supplied externally. The remaining energy input is only hydrogen.

Table 22. RWGS methanol plant energy demand datasheet.

PO #	Process Unit	Power Consumption	Heat from Fuel Combustion	Steam Consumption	Other Heat	H2 supply
		MWe	MWt	MWt	MWt	MW(LHV)
A	H2 and CO2 Compression	25			*-21	903
B	RWGS	17	53			
C	Selexol	5			**16	
D	Methanol Synthesis	13				
E	Distillation	1	***-118	81		
F	Furnace					
G	Power & steam system	-32	-65	-84		
	Total	28	0	-3		903

* Interstage cooling, ** chiller duty, ***light ends

Nuclear integration in this case is rather straight forward. Power and hydrogen via HTSE are coupled to the nuclear plant, following the configuration of power conversion cycles dedicated to LP steam and power production for electrolysis. The resulting demands from the HTGR is 2,713 MWth, which translates to 14 units—a massive demand—and 35 MWe export. The resulting BFD is shown in Figure 42.

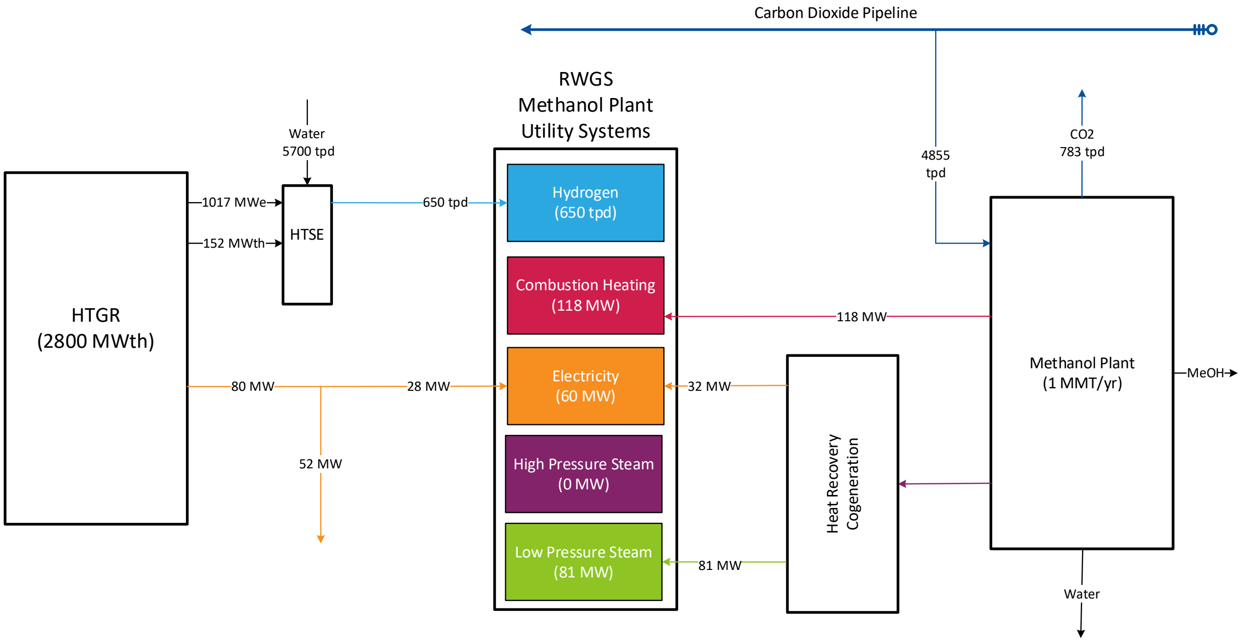


Figure 42. Nuclear integration BFD for RWGS methanol plant.

All three scenarios are summarized in the following tables. Table 23 gives the relative energy demands provided by nuclear energy for each case, and Table 24 shows each of these demands by unit operations. In each scenario, the reactor output is quantified as the total demand (i.e., not number of specific modules) to obtain a net-zero balance (no import or export) of all energy supply types.

These integration cases show that an NPP can provide up to 100% of considered external energy inputs and can substitute feedstock NG with nuclear-generated hydrogen. The last column of Table 20 is effectively an inverse value of the energy conversion efficiency between nuclear heat and all plant energy demands, which are supplied from nuclear sources in each scenario. Lower numbers (e.g., Case 1a) show the best efficiency, but efficiency decreases when supplying hydrogen (e.g., Cases 1 and 2)—the majority of losses comes from the electricity needed for electrolyzers. Electrification (Case 1a) has the best efficiency of the three cases. Case 1and case 2 have similar efficiency as both cases are based on hydrogen, but as a feedstock in Case 2 there are further impacts of conversion process into the methanol product.

Table 23. Resulting parameters of HTGR integration into MeOH plant—reduction in energy demands.

	Scenario	Electricity Supply	HT Heat Supply	Feedstock	Total Net Energy Demand Reduction	HTGR-Type Reactor Power Required (MWth)	Reactor Power per Supplied Energy (MWth/ $MW_{supplied}$)
1	H2 furnace	100% Internally	100%	0%	30%	488	2.78
1a	El. furnace	100% Internally	100%	0%	30%	367	2.09
2	H2 RWGS	100%	100%	100%	100%	2,713	2.93

Table 24. Methanol plant demands per unit product for considered scenarios.

	Scenario	Electricity Supply (MW/1 MMT/yr)	NG Supply (MW/1 MMT/yr)	H_2 Needs Supplied (MW/1 MMT/yr)
0	BAU	-2	776	0
1	H2 furnace	-2	601	175
1a	El. furnace	147	601	0
2	H2 RWGS	28	0	903

5.2.3 National Potential for Nuclear Integration into Methanol Plants

National demand is based on the reference plant model and annual methanol production in the United States. The resulting energy flows of the current approach, using NG both as feedstock and as a fuel (combusted to perform intermediate thermal conversion of the feedstock), is shown in Figure 43. Power production and other thermal requirements (e.g., for distillation) are provided via an internal WHR system.

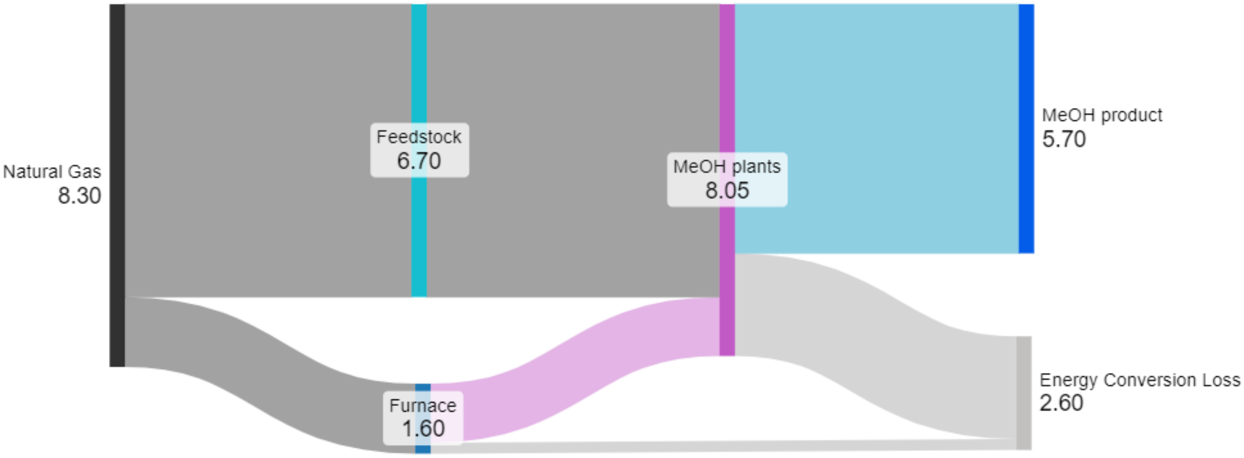


Figure 43. Sankey diagram of national-level energy demand and supply for methanol production.

Nuclear integration national potential is based on Scenario 2, operating with RWGS, which substitutes feedstock NG for hydrogen, see Figure 44. Internal heating demands are met with the combustion of light ends from distillation and an internal WHR system, but a small amount of external power is still needed. Therefore, energy is brought into the system only as hydrogen feedstock and power. As a result of the limited efficiency associated with each conversion, primary energy in nuclear heat is almost 3× higher than primary energy in NG. Conversion to MeOH from H_2 feedstock has very similar efficiency regarding energy inputs as conversion from NG.

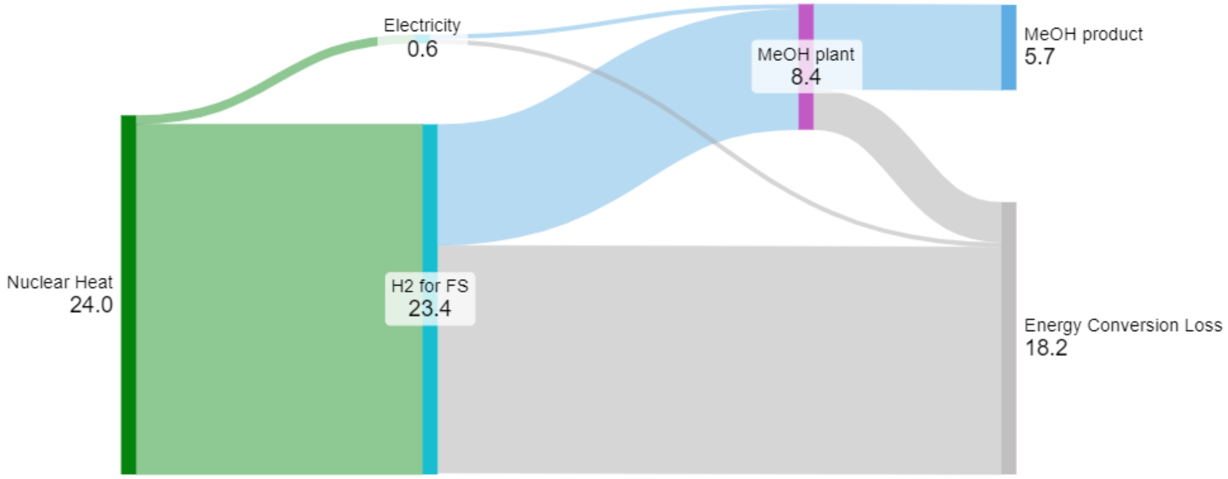


Figure 44. Sankey diagram representing the national potential of selected nuclear-integrated scenario for energy demand and supply for methanol plants.

National-level potential for reduction of CO_2 emissions is presented in Figure 45 for all considered nuclear integration scenarios. Both Cases 1 and 1a are representing the same emission reduction, denoted in the figure as SMR heat. The remaining emissions are associated with the combustion of the byproduct fuels. In the H_2 feedstock pathway (excluding emissions from the product MeOH use), the plant-level emissions are negative as the CO_2 used is recycled from another point source, most of which is converted into the MeOH product.

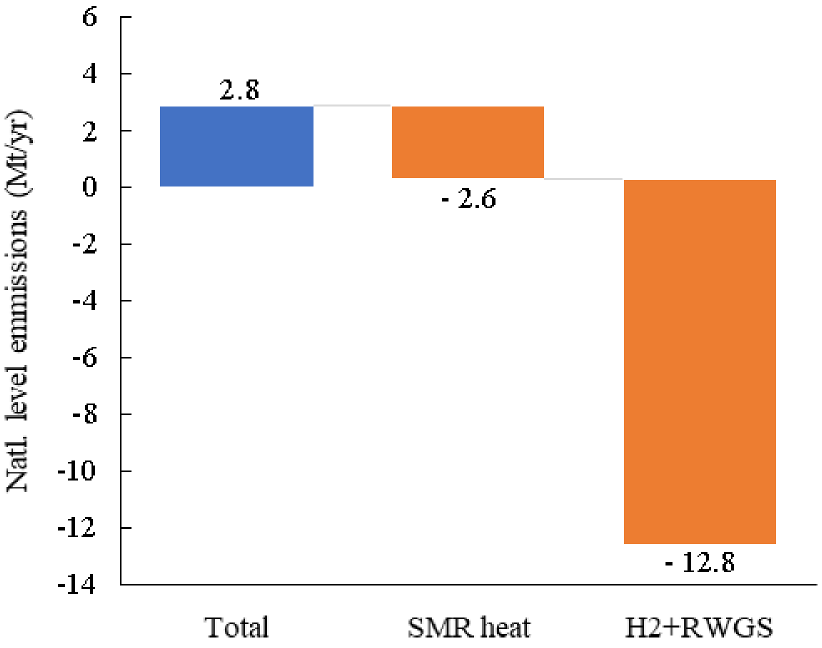


Figure 45. National-level emission reduction potential from two presented scenarios of nuclear energy integration.

5.3 Kraft Pulp Mill Nuclear Integration

5.3.1 Nuclear Integration Opportunities

There are three basic pathways for nuclear integration within the pulp and paper industry: steam and electricity integration to replace fossil fuels, carbon capture, and the production of hydrogen to upgrade biomass to fuels of products. The extent of each integration opportunity is dependent on the mill type, size, and configuration.

Replacement of Fossil Fuels

As described in this report, pulp mills produce much of their steam internally from the black liquor boiler and wood waste. However, some mill configurations require a large portion of steam demands from multifuel or NG boilers. Nearly all the steam is used at low pressures (11 bar or less) and could be easily replaced with steam from a light-water reactor or HTGR. See Appendix E for Figure E12, Figure E13, and Figure E14 showing the steam use for an unbleached kraft softwood pulp mill, bleached kraft softwood pulp mill, and recycled paper mill, respectively. The recovery boiler provides roughly 60% of the steam for both pulp mills, and the recycle mill requires 100% of steam from a dedicated boiler since it doesn’t generate combustible byproducts. Also, roughly 70% of all the steam used in the three configurations is at an LP (4 bar). Most configurations produce electricity through CHP, but nuclear power could also supplement electricity generation that is lost when the boilers are replaced.

Nuclear power could also be integrated through electrification. Because the plant configurations above utilize less than 5% of the steam at pressures supplied by the boilers, electrifying the boilers is not the most efficient use of nuclear integration. However, electrification could replace fossil fuel use in the lime kiln. NG combustion in the lime kiln could also be reduced with clean hydrogen supplied by the NPP.

Carbon Capture

Nuclear power can be integrated to capture carbon from the existing processes at the plant. Commercialized technologies like monoethanolamide (MEA) scrubbing can be retrofitted to the plant and powered with nuclear steam and electricity. Oxyfuel combustion is a low technology readiness level for pulp mills but could lower the energy required for carbon capture. If the boilers and lime kiln are converted to an oxy-fired configuration, the oxygen needs can be supplied with energy from the NPP.

Hydrogen Production

If nuclear steam is integrated into the plant, lignin from the recovery boiler and the wood waste from the pulping process that is typically burned in a hog boiler to power the pulp mill could be processed into other bioproducts or fuels. This waste is rich in carbon, which could be reprocessed into liquid fuels or other bioproducts. Reprocessing wood waste into bioproducts like plastics or food additives would reduce the total carbon output of the mill. Reprocessing wood waste into fuels would not reduce the overall carbon output, but the fuel could be used to displace fossil fuels within the pulp mill or in other applications. Hydrogen, heat, and electricity from the NPP could be integrated into these processes.

In the reference kraft pulp mill studied (see Figure 27), an available drop-in integration opportunity is the external demand from the NG boiler, with the heat content (based on LHV) of 52 MW and a steam duty of 44 MWth. Note that, in source distribution, minor fuel demands, such as the often-necessary co-firing of NG (or other noble fuel) for stabilization in the BLRB operation, have been neglected. If only the NG boiler was replaced, nuclear energy demand would be 44 MWth of 9 MPa 500°C, either injected into the main steam line or designed at any of the process steam parameters (2.8, 1.1, or 0.5 MPa). In the latter configuration, the deficit of power production would be supplied by the nuclear plant. The thermal energy required is less than a quarter of a single Xe-100 reactor unit and 16% of a single NuScale module. Dedicated construction of an SMNR for such a pulp plant integration is therefore highly unlikely. Another drop-in nuclear energy integration can be made for the lime kiln, where electrically heated solutions exist [59–62], though still being significantly less than just a single SMNR module.

Overall, the decarbonization potential of the reference kraft pulp mill is the smallest of any kraft mill or paper production facility, primarily because the reference mill does not include a paper pressing line or bleaching operations, which would increase LP steam utilization. The most impactful, but highly complex, decarbonization opportunity is using an NPP to assist with carbon capture via oxy-fuel combustion.

5.3.2 Reference Kraft Pulp Mill Nuclear Integration

Two integration strategies are presented, both offering a substantially different perspective on the extent of integration work and resulting degree of decarbonization:

6. Drop-in minimally invasive integration by replacing energy coming from fossil fuels—the NPP replaces heat, and electricity demands from the NG steam boiler and provides electric heating to replace NG in the lime kiln.
7. Complex integration by oxyfuel combustion in the BLRB and lime kiln with CO_2 capture. The oxygen demand is fed by CO_2 swept HTSE with the H_2 product exported.

As with the methanol plant, the energy balance of a state-of-the-art kraft mill demonstrates the potential to meet the entire heating demand (excluding the high-temperature heat supplied in the lime kiln) by burning biogenic byproducts of the process. External energy needs depend on BLRB capacity and the mill configuration. The pulp plants can also operate as a net electricity exporter with modern equipment and CHP network. As such, the application of direct nuclear integration into an existing kraft mill is limited.

In the reference plant, 44 MWth of steam is required from an NG-fired boiler. The substitution of this boiler by Xe-100-sourced process steam, delivered at MP, resulting in a lower cost of steam system than for the main 9 MPa steam, required 38.4 MWth as this steam does not perform work in the HP turbine section, is illustrated in Appendix E, Figure E15. As a result of lower total HP steam flow to the plant, the backpressure steam turbine power output is reduced by about 4 MWe, which is made up by the surplus electricity produced from the nuclear plant. The combined power output of the nuclear CHP and pulp mill CHP is 91.3 MWe, with an electricity production efficiency of 23.5% and CHP efficiency of 63.9%.

An electrified lime kiln would demand 28.1 MWe. Combined with the demands for steam, a single HTGR reactor has a net export of 38.2 MWe (Case 1), almost half of the standalone single unit production. See Figure 46 for a BFD with an overview of energy flows. Should the lime kiln NG instead be substituted with HTSE-produced hydrogen, it would consume a larger portion of the HTGR output—about 33 MWe—and a small portion of heat. The impact of vaporization heat on the HTGR output would be very small, meaning that over half the output would still need to be exported and overall results would not differ significantly.

Note that the presented results assume no NG (or other noble fuel) co-firing is required for stable operation of the BLRB. Also, the potential efficiency improvement of an electric kiln over an NG-fired kiln were ignored. Since SMNRs are expected to be installed in bundles of at least four units, it is not feasible to install a full SMNR build to support a paper mill. Alternatively, an SMNR could provide low-carbon steam to a paper mill as part of a larger energy network, as is the case for two cardboard mills connected to the Gösgen Nuclear Power Plant in Switzerland [63]. Microreactors may also better fit the mill demand size.

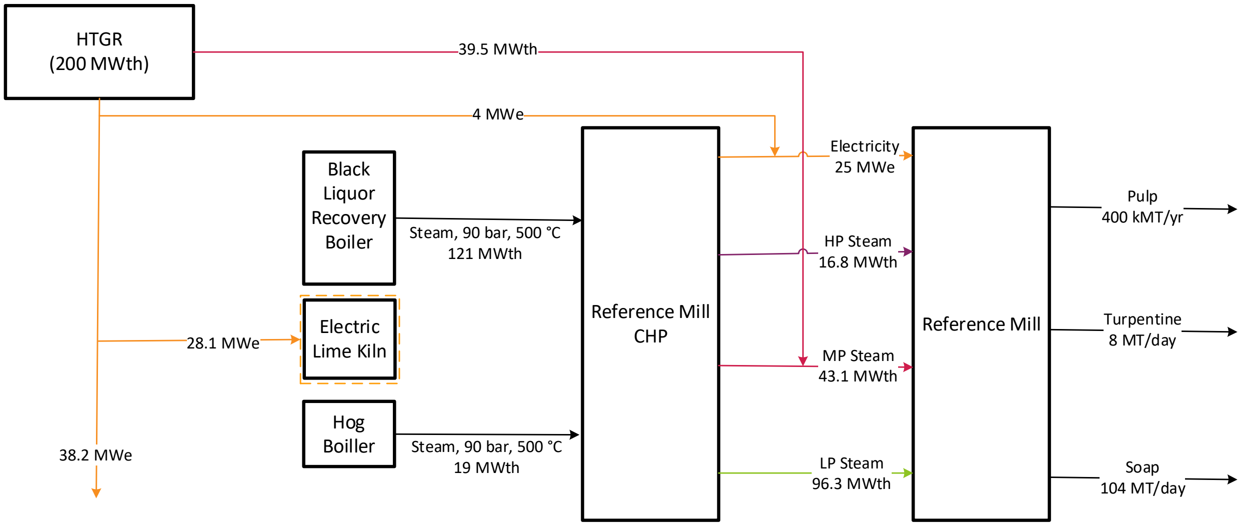


Figure 46. Utility system BFD for simple nuclear integration at reference pulp mill.

Most of the emissions from the plant are biogenic and cannot be avoided without major changes in the mill's operation, making the reference mill a candidate for CO_2 capture utilizing oxyfuel combustion and subsequent post-combustion cleanup. An oxyfuel combustion configuration for the BLRB and lime kiln is presented here in complex integration Case 2. The complex integration schematic, summarized by a BFD in Figure 47, involves replacing the hog boiler and an NG auxiliary boiler with nuclear steam and converting the BLRB and lime kiln to oxyfuel combustion. A CO_2 swept HTSE system provides oxygen mixed with carbon dioxide and no nitrogen, which is used to oxy-fire the BLRB and lime kiln. This produces a flue gas that is very high in CO_2, which can be condensed, cleaned using existing particulate and acid-gas (sulfur) emission controls, and compressed for sequestration. A chlorine deep cleanup would be also necessary for this application but has not been modeled in this work. The NG auxiliary boiler can be removed, as steam is sent from the HTGR to the existing CHP equipment. The steam extracted from the HTGR cycle has the same flowrate and thermodynamic properties as the steam that the decommissioned boilers provided, allowing the existing mill CHP to run on-design and produce the same amount of electricity and HP, MP, and LP steam for use at the mill as in the BAU case. The remaining electricity from the HTGR cycle is used to compress the captured CO_2 with surplus sold to the grid.

The steam cycle model illustrated by Figure E16 in Appendix E shows that a 1,000 MWth HTGR is required to supply the reference mill CHP equipment with 63 MWth of 90 bar, 500°C steam, along with 328.1 MWe and 46.3 MWth to the CO_2 swept HTSE plant. The HTSE plant produces over 1,600 MT/day of oxygen blended to a 40 mol% concentration with CO_2, which is used to fire the BLRB and lime kiln. 203 MT/day of clean hydrogen is produced as a byproduct, while 29.8 MWe of clean electricity may be sold to the grid. The lime kiln requires 28.1 MWth of NG; all other energy inputs are satisfied by the HTGR.

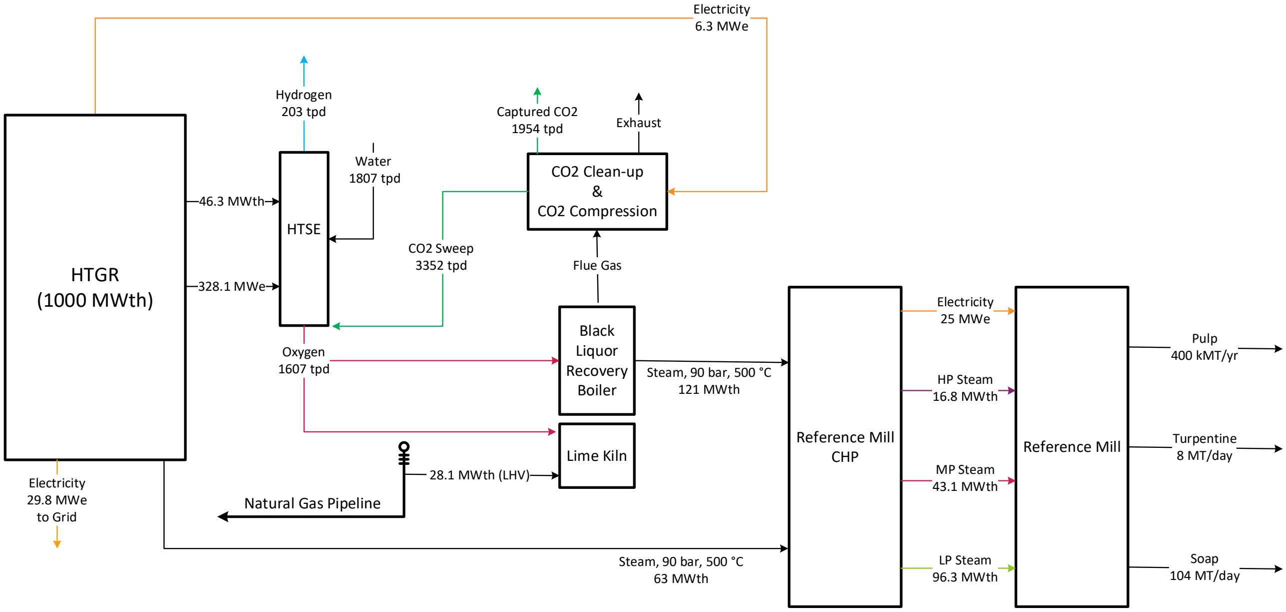


Figure 47. Utility system BFD for nuclear-integrated oxy-fired pulp mill.

A summary of the scenarios, again with modified reactor output for a zero-electricity export, is shown in Table 25 and Table 26. The first scenario shows a simple nuclear integration scenario that replaces all externally needed energy in the reference pulp mill, displacing fossil fuels. Even though the mill is rather large, the demand of nuclear-sourced energy is less than a single nuclear reactor module. On the other hand, the complex integration approach doesn't replace the fuel for the lime kiln but does capture all CO_2 emissions. The overall energy balance results in a large export in Case 2 due to the energy content of the hydrogen byproduct. As the system has carbon capture integration and most of the biogenic emissions are captured along with the lime kiln emissions, the system becomes CO_2 negative.

Table 25. Resulting parameters of HTGR integration into reference paper mill—reduction in energy demands.

	Scenario	Electricity Supply	HT Heat Supply	Steam Supply	H2 Needs Supplied	Total External Energy Demand Reduction	HTGR-Type Reactor Power Required (MWth)	Reactor Power per Supplied Energy (MWth / $MW_{supplied}$)
1	Drop-in fossil fuel replacement	100%	100%	100%	n/a	100%	104.5	1.45
2	Steam + H2	100%	0%	100%	export	61%	925.5	21.0

Table 26. Pulp mill demands per unit product for considered scenarios and resulting energy and CO_2 balance. Integration scenarios source electricity and steam from nuclear CHP.

	Scenario	Electricity Supply (MW/400k ADt/yr)	NG Supply (MW/400k ADt/yr)	Steam Supply (MW/400k ADt/yr)	Energy Exports (MW)	Non-Biogenic CO_2 Balance (t/t_{pulp})
0	BAU	-1	73	0	1	1.21
1	Drop-in fossil fuel replacement	32	0	40	0	0
2	Complex integration	334	28	46	281*	-1.21

* H_2 export

5.3.3 National Potential for Nuclear Integration into Pulp and Paper Plants

National demand has been estimated from the Reference [54] data. Since the data combines the entire forest product industry, estimates for the pulp and paper industry were separated using the breakdown of the demands for subsectors in Reference [64] and fuel demand assessing all the black liquor content of the fuel belonging to the pulp and paper industry. This analysis assumes that all pulp and paper mills are using CHP systems rather than standalone boilers being used in other, more distributed and smaller single-point-demand subsectors. This simplifies energy demands and their supply on a national level, as is shown in Figure 48. External fuels coming into onsite CHP and external power plants is mostly NG within the United States, but other fuels are possible. Conversion efficiencies are adopted from Reference [54]. While the reference plant for nuclear integration only included pulping operations, the estimates for national potential include all pulping and papermaking operations in the United States (paper mills, paperboard mills, newsprint mills, pulp mills, and integrated mills).

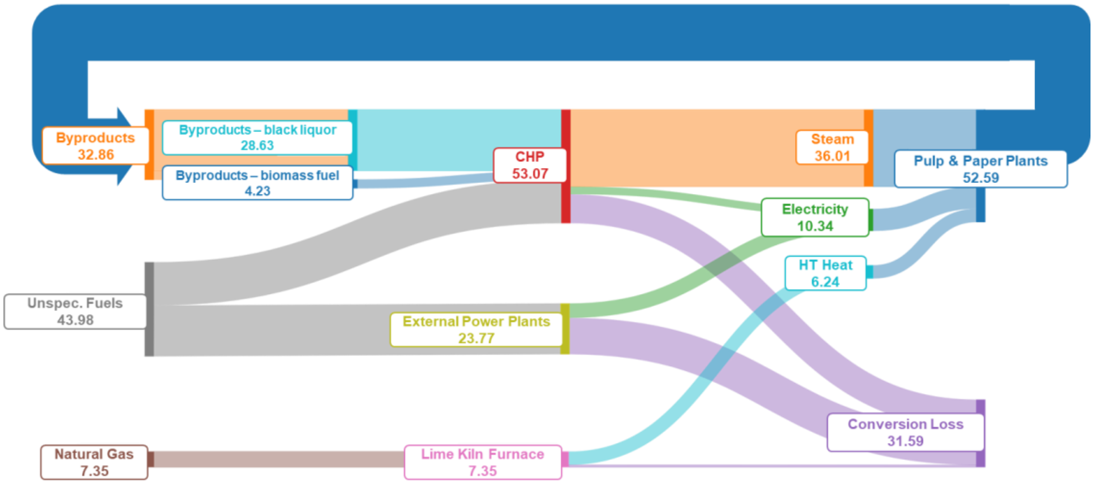


Figure 48. Sankey diagram of national-level energy demand and supply for pulp and paper industry.

The nuclear integration scenario showcased in the national potential is Case 1, where fossil fuels are substituted for clean nuclear energy. The reference integration model was modified for the national average plant with a new ratio between externally required electricity and steam. The reactor output was modified to provide exactly the energy requirements of the plant, for zero electricity exports after meeting the electricity demand for the lime kiln. As for the other industries, electrifying the lime kiln assumes eliminating stack loss. The Sankey diagram of this approach on a national potential level is in Figure 49. Since new nuclear CHP is predicted to have significantly higher efficiency than currently used in aging systems, overall conversion loss is reduced by almost 6 GW of primary energy. A similar, or even higher level, of primary energy saving is also attainable through plant modernization and implementation of energy efficiency solutions and WHR systems, rather than nuclear integration.

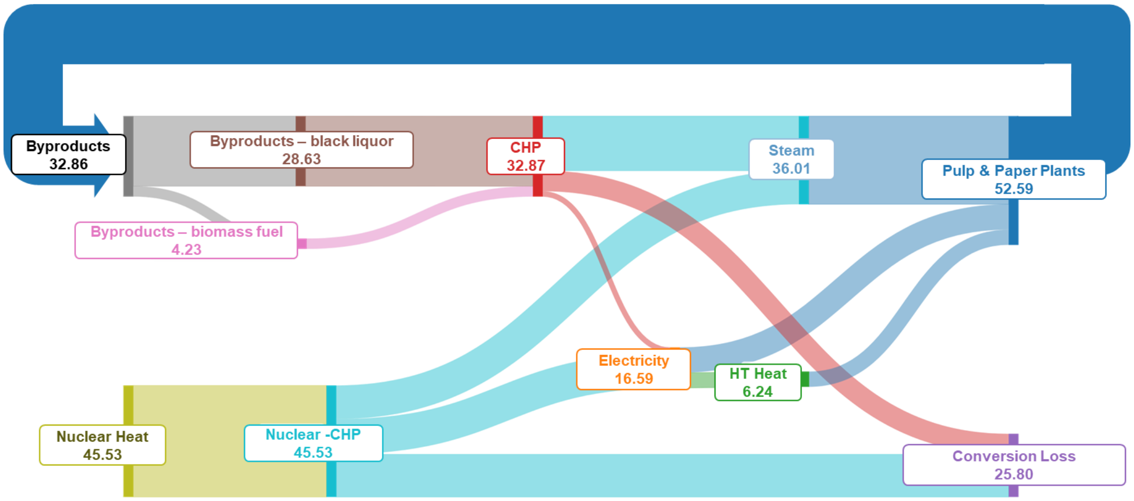


Figure 49. Sankey diagram of national potential of nuclear-integrated scenario for energy demand and supply for pulp and paper plants.

The nuclear integration scenario adopted for the national-level potential effectively reduces all non-biogenic CO_2 emissions. With an assumption that all external fuel is NG and biomass fuel has zero net emissions, proposed nuclear integration has the potential of reducing 73 Mt/yr of CO_2 emissions, with a breakdown for each type of energy supplied in Figure 50.

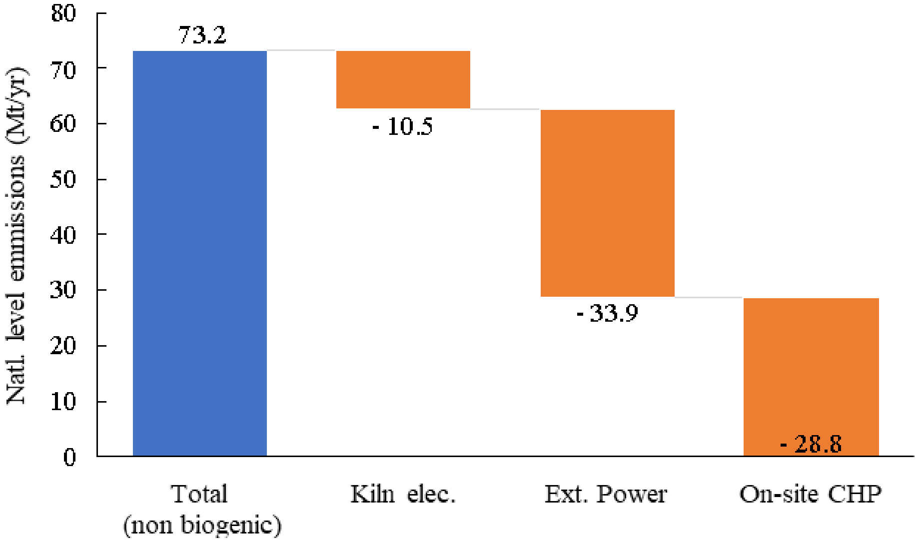


Figure 50. Emission reduction potential from separate uses of nuclear energy used in the national-level analysis.

6. CONCLUSIONS

At INL, the Department of Energy Office of Nuclear Energy's Integrated Energy Systems Program is establishing thermal integration requirements for industrial processes at specific energy requirement levels. At the same time, efforts have been made to develop an understanding of these processes and to engineer industrial decarbonization solutions. The industrial sector could realize decarbonization benefits by integrating manufacturing with existing nuclear reactors or SMNRs for the supplying clean electricity and heat. Understanding industry interest and concerns regarding nuclear integration is critical for identifying opportunities for such integration, along with the degree of SMNR growth and research and development requirements involved. This report analyzed the process efficiency, energy (thermal and electrical) demands, CO_2 emissions, and hazards associated with several key industries. Industrial process improvements were investigated by considering the integration of clean nuclear energy delivery and use to reduce GHG emissions.

Three key industries (i.e., the refining, methanol, pulp and paper) were selected for their high-process-level energy (thermal and electricity) demand and major manufacturing presence in the United States. Per the U.S. Energy Information Administration GHG emissions inventory, these industries account for nearly 27% of U.S. industrial sector GHG inventory emissions, equating to about 5% of the total U.S. emissions in 2023 [65]. Most of the remaining 75% of industrial GHG emissions are tied to smaller facilities that fall under the EPA reporting limits for large GHG emitters.

Table 27 briefly summarizes the various requirements and CO_2 emissions of methanol, oil refining, pulp and paper production. As shown, these industries produce approximately 352 MMT of CO_2 and are significant contributors to both the current and future economies. These systems have been broken down into specific processes to identify unit energy requirements throughout the plant and enable a focused examination of SMNR integration opportunities. Future work will involve a detailed study of these processes and will identify specific integration strategies, develop novel control methods for the complex systems, and conduct technoeconomic assessments of these novel deployments.

Table 27. Summary of identified industry CO_2 emissions and process conditions.

Industrial Product	Annual CO_2 Emissions	Temperature Needs	Approx. Steam Pressure Needs
Oil Refining	164 MMT[b]	300–800°C	0.3, 1.2, 4.2 MPa
Methanol	2.85 MMT[b]	170–1,700°C[a]	0.1–0.3, 3.5 MPa
Pulp and Paper	96.6 MMT[b]	125–325°C	0.45, 1.2, 2.6 MPa

[a] Including potential methanol conversion pathways.

[b] EPA FLIGHT tool data from 2023 [1].

[c] EPA GHG inventory CO_2-eq. data for 2021 [3].

[d] Values are for steam supply to HTSE.

The process and thermodynamic performance, scale, electricity, and heat-use patterns of facilities were evaluated for the refining, methanol, paper and pulp industries. The commonality among these industries is that they convert raw materials into products through physical and chemical changes that require large quantities of thermal electrical energy. Heat demands range from low-temperature steam (0.25 MPa) used for refining and pulp and paper processes to high-temperature process unit operations (up to 950°C) for methanol, refinery operations. Within these industries, fossil-fuel-based furnaces, CHP, and conventional steam boilers account for the majority of the heat loads and resultant CO_2 emissions. These heat supply systems could be switched out with clean nuclear heat from SMNRs to provide electricity, steam, heat, and clean hydrogen.

The practical limit of a clean heat source depends on its temperature output. Current HTGR designs could provide ~565°C, with direct gas heat exchanger future designs potentially providing upward of 750–850°C. And with materials development, 950°C may be attainable with the demonstration of metal alloys' performance at these temperatures. A substitution of a third of the projected 2027 industrial energy demand could be met by about 230 SMNRs featuring a capacity rating of 150 MWt. This study indicates that the scale of the largest industrial energy users is amenable to the scale of SMNR applications.

Emerging SMNR scales and temperatures can be adaptive to individual industry needs. Several industries (e.g., oil refineries and methanol, pulp and paper manufacturing) have high-heat duties in excess of 10 TJ/day (9,500 MMBtu; 120 MWt). SMNR technologies are expected to be well matched to this scale of demand. SMNRs can also power hydrogen generation, which can be used in place of fossil-based hydrogen production and replace or augment conventional combustion gas systems. Hybrid thermal-electricity generation by an SMNR integrated with thermal storage may help balance hourly, daily, and seasonal cycles of industrial applications. To assess integration opportunities, an Aspen HYSYS, Cycle-Tempo and Excel models were developed using an HTGR SMNR parameters as a reference. Evaluations were done for both reference plants and as a national potential.

For a reference refinery with an operating capacity of ~100,000 BPD, the direct CO_2 emissions total approximately 0.8 MMT/year. Most of these emissions stem from the combustion of fossil fuel to supply the thermal heat demands and hydrogen production. The simplest targets for emissions reduction are the refinery's CHP system, which typically relies on natural gas (NG) boilers and combustion turbines to produce steam and electricity. SMNRs can be configured as CHP to supply all electricity and process steam demands. Refineries also consume large quantities of hydrogen for chemical processes like residue cracking. This hydrogen is typically produced via steam methane reforming (SMR), but nuclear HTSE offers a low-emission alternative. For example, at the reference refinery, meeting hydrogen demand through HTSE would require 184 MW of electrical power and 31 MW of LP steam. In addition to steam and hydrogen, refineries require high-temperature heating, which is typically achieved by combusting refinery gas (RFG) or imported NG. While the amount of imported NG varies across refineries configurations, it can be substituted by using clean hydrogen as fuel. An alternative and more efficient substitution (which likely comes at lower cost) is to electrify selected heating duties with resistive furnaces.

Conventional methanol production heavily relies on natural gas (NG), both as a fuel and a feedstock, primarily for the steam methane reforming (SMR) furnace that operates near 850°C. With NG combustion dominating the energy supply in the traditional setup, nuclear integration for methanol plants is more complex compared to refineries. There are two main avenues for replacing NG combustion in methanol production with nuclear-derived energy: hydrogen substitution and electrification. Use of hydrogen as a fuel entails significant retrofitting due to hydrogen's different combustion properties and impacts on flue gas dynamics. Electrification avoids some complexities while significantly reducing stack losses and further avoiding capital costs of electrolyzers. Alternatively, nuclear integration could involve a more transformative process. Rather than using NG and steam, methanol can be synthesized by utilizing CO_2 and hydrogen as feedstocks. This approach involves a reverse water gas shift (RWGS) reactor to produce syngas which is synthesized into methanol in a second reactor. A more direct one-step methanol solution exists commercially but is reserved for future work. From the methanol plant view (emissions scope 1 and 2), the plant becomes CO_2 negative.

Integrating pulp and paper plants with nuclear power is made challenging by the fact that the main heat source (i.e., black liquor combustion) is integral to sustaining the chemical recovery process. Two main pathways are considered, which are drop-in replacement of fossil fuels for steam and electricity generation and second a complex carbon capture technology integration based on oxy-fuel combustion coupled with HTSE hydrogen production. The drop-in, technically feasible, nuclear integration considers substitution of NG boilers by nuclear CHP and changing the NG fired lime kiln to an electrically heated model. The reference mill requires about 105 MWth of nuclear heat for this integration. A more complex integration involves developing an oxy-fuel combustion recovery boiler with CO_2 capture. The oxygen is sourced from HTSE units that are swept by recirculating CO_2. This concept offers greater decarbonization potential, capturing most of the biogenic emissions and creating a CO_2-negative system. This is an energy intensive process, consuming 925 MWth of nuclear heat while also exporting 253 MW of energy in the byproduct hydrogen.

As the United States pursues its decarbonization goals, innovative integration methods will be needed to reduce the environmental impact of these significant industries. Replacing industrial carbon-emitting energy sources with integrated SMNRs will enable direct thermal substitution within processes, “green” electrification, and the production of green chemical processes to produce an identical product. This report will aid researchers and industry in determining how necessary energy streams may be replaced with clean energy sources from SMNRs, giving options for several levels of decarbonization.

7. ACKNOWLEDGEMENTS

This work was supported by the DOE-NE IES Program, with work being conducted at INL under DOE Operations contract no. DEAC0705ID14517.

8. REFERENCES

1. EPA. 2022. “2022 Greenhouse Gas Emissions from Large Facilities.” Environmental Protection Agency. Accessed October 13, 2023. https://ghgdata.epa.gov/ghgp/main.do?site_preference=normal.
2. EPA. 2022. “Facility Level Information on Greenhouse gases Tool (FLIGHT).” Environmental Protection Agency. https://ghgdata.epa.gov/ghgp/main.do.
3. EPA. 2023. “Inventory of U.S. Greenhouse Gas Emissions and Sinks: 1990–2021.” U.S. Environmental Protection Agency, EPA 430-R-23-002. https://www.epa.gov/ghgemissions/inventory-us-greenhouse-gas-emissions-and-sinks-1990-2021.
4. EIA. 2023. “The U.S. energy facts explained.” U.S. Energy Information Administration. https://www.eia.gov/energyexplained/us-energy-facts/.
5. Kozeracki, J., et al. 2023. “Pathways to Commercial Liftoff: Advanced Nuclear.” U.S. Department of Energy: Washington, DC, USA. https://liftoff.energy.gov/wp-content/uploads/2023/03/20230320-Liftoff-Advanced-Nuclear-vPUB.pdf.
6. R. M. Saeed, E. K. Worsham, B-H. Choi, et al. 2023. “Outlook on Industrial Requirements for Incorporating Nuclear Energy into Industrial Processes.” INL/RPT-23-74711, Idaho National Laboratory, Idaho Falls, ID. https://doi.org/10.2172/2293480.
7. EIA. 2023. “Refinery Capacity Report Q2.” U.S. Energy Information Administration. https://www.eia.gov/petroleum/refinerycapacity/.
8. EIA. 2012. “PADD regions enable regional analysis of petroleum product supply and movements.” U.S. Energy Information Administration. https://www.eia.gov/todayinenergy/detail.php?id=4890#:~:text=The%20Petroleum%20Administration%20for%20Defense,PADD%205%20the%20West%20Coast.

9. Abella, J. P., K. Motazedi, J. Guo, K. Cousart, L. L. Jing, and J. A. Bergerson. 2020. "PRELIM: Petroleum Refinery Life Cycle Inventory Model v1.4," v1.4 ed. University of Calgary. https://ucalgary.ca/energy-technology-assessment/open-source-models/prelim.

10. Statista. 2020. "Methanol market volume in the United States from 2016 to 2027 (in million metric tons)." Chart. Accessed May 31, 2023. https://www.statista.com/statistics/1098231/us-methanol-market-size/.

11. Dolan, G. A. 2020. "Methanol: emerging global energy markets." Methanol Institute, presentation for 16th Annual State of the Energy Industry Forum, Washington, DC. https://usea.org/sites/default/files/event-/SOTEI%202020_Greg%20Dolan_MI%20USEA%20Annual%20Forum%2023%20Jan%202020.pdf.

12. Rivarolo, M., et al. 2016. "Feasibility study of methanol production from different renewable sources and thermoeconomic analysis." *Int. J. Hydrog. Energy*, 41(4): 2105–2116. https://doi.org/10.1016/j.ijhydene.2015.12.128.

13. Ghasemzadeh, K., et al. 2018. "Chapter 23 - Economic Assessment of Methanol Production." *Methanol* 613–632. https://doi.org/10.1016/B978-0-444-63903-5.00023-6.

14. Celanese. 2021. "Celanese to Expand Capacity and Utilize Recycled C02 for Methanol Production at Clear Lake, Texas Facility." Celanese. https://www.celanese.com/news-and-media/2021/march/utilizing-recycled-co2-for-methanol-production.

15. Proman. n.d. "Proman USA (Pampa) is a 225 metric ton per day methanol plant in Pampa, Texas which began production in 2015. It is an ISCC certified bio-methanol facility." Proman USA (Pampa). https://www.proman.org/companies/pampa-fuels/#:~:text=Proman%20USA%20(Pampa)%20is%20a,ISCC%20certified%20bio%2Dmethanol%20facility.

16. OCI-Global. 2023. "OCI Global to double its green methanol capacity in the US." OCI Global. https://oci-global.com/news-stories/press-releases/oci-global-to-double-its-green-methanol-capacity-in-the-united-states/#:~:text=OCI%20Global%20(Euronext%3A%20OCI),emissions%20industries%2C%20including%20road%20transport%2C.

17. Clark, B. 2015. "Outlook '16: US Methanol Capacity to stabilize." Independent Commodity Intelligence Services. https://oci-global.com/news-stories/press-releases/oci-global-to-double-its-green-methanol-capacity-in-the-united-states/#:~:text=OCI%20Global%20(Euronext%3A%20OCI),emissions%20industries%2C%20including%20road%20transport%2C.

18. EIA. 2019. "New Methanol Plants Expected to Increase Industrial Natural Gas Use through 2020." U.S. Energy Information Administration. https://www.eia.gov/todayinenergy/detail.php?id=38412.

19. Methanex. n.d. "Geismar." Methanex. https://www.methanex.com/about-us/global-locations/usa-geismar/.

20. Koch Methanol. n.d. "Who We Are." Koch Methanol. https://www.kochmethanol.com/koch-methanol-st-james/who-we-are/#:~:text=Koch%20Methanol%20St.,Association%20(IMPCA)%2Dquality%20methanol.

21. Hu, Y., Li, H., and Yan, J. 2010. "Integration of Evaporative Gas Turbine with Oxy-Fuel Combustion for Carbon Dioxide Capture." *Int. J. of Green Energy*, **7**(6): 615–631. https://doi.org/10.1080/15435075.2010.529405.

22. Bozzano, G. and Manenti, F. 2016. "Efficient methanol synthesis: Perspectives, technologies and optimization strategies." *Progress in Energy and Combustion Science* 56:71–105. https://doi.org/10.1016/j.pecs.2016.06.001.
23. FAO. n.d. "Pulp and Paper Capacities. Survey 2022-2024." Food and Agriculture Organization of the United Nations. https://at.twosides.info/wp-content/uploads/sites/7/2023/09/23-09-FAOForecast.pdf.
24. Oliveira, R. C. P., M. M. Mateus, and D. M. F. Santos. 2019. "Editor's Choice-On the Oxidation of Kraft Black Liquor for Lignin Recovery: A Voltametric Study." *J. Electrochem. Soc*. 166, E547. http://dx.doi.org/10.1149/2.0131916jes.
25. Cheremisinoff, N. P. and P. Rosenfeld. 2010. *Best Practices in the Wood and Paper Industries.* Oxford, Ch. 6: 179-59. ISBN: 9781437778335. https://search.worldcat.org/title/best-practices-in-the-wood-and-paper-industries/oclc/659561107.
26. NREL. n.d "Steam System Opportunity Assessment for the Pulp and Paper, Chemical Manufacturing, and Petroleum Refining Industries: Main Report." National Renewable Energy Laboratory, Golden, CO. https://www.nrel.gov/docs/fy03osti/32806.pdf.
27. Kuparinen, K. and E. Vakkilainen. 2017. "Green pulp Mill: Renewable Alternatives to Fossil Fuels in Lime Kiln Operations." *BioResources* 12(2): 4031–4048. https://bioresources.cnr.ncsu.edu/wp-content/uploads/2017/04/BioRes_12_2_4031_Kuparinen_V_Green_PulP_Mill_Renewable_Alternatives_Fossil_Fuels_Lime_Kiln_Operation_11535.pdf.
28. Sagues, W. J., et al. 2022. "Prospects for Bioenergy with Carbon Capture & Storage (BECCS) in the United States Pulp and Paper Industry." *Energy and Environmental Science* 13(8): 2243–2261. http://dx.doi.org/10.1039/D0EE01107J.
29. EPA. n.d. "2011-2021 Greenhouse Gas Reporting Program Sector Profile: Pulp and Paper." Environmental Protection Agency. https://www.epa.gov/system/files/documents/2023-05/Pulp_and_Paper_Profile_RY2021_05-04-2023%20508.pdf.
30. Yang, F., J. C. Meerman, and A. P. C. Faaij. 2021. "Carbon capture and biomass in industry: A techno-economic analysis and comparison of negative emission options." *Renewable and Sustainable Energy Reviews* 144. https://doi.org/10.1016/j.rser.2021.111028.
31. Vakkilainen, E. K, and P. Ahtila. 2011. "Modern Method to Determine Recovery Boiler Efficiency." *O* PAPEL 72(12): 58–65. http://www.revistaopapel.org.br/noticia-anexos/1359377111_7bc48fa8b7e072c7a6520423b6eeddf1_1138265000.pdf.
32. Maine. 2019. "Air Emissions Inventory." Maine Department of Environmental Protection. https://www.maine.gov/dep/ftp/AIR/DATA/CAP_SUMMARIES/.
33. Dionne, J. and T. R. Walker. 2021. "Air Pollution impacts from a pulp and paper mill facility located in adjacent communities, Edmundston, New Brunswick, Canada and Madawaska, Maine, United States." *Environmental Challenges* 5: 100245. https://doi.org/10.1016/j.envc.2021.100245.
34. Scielo. 2018. "Wastes from pulp and paper mills – a review of generation and recycling alternatives." *Ceramica* 64: 371. https://doi.org/10.1590/0366-69132018643712414.
35. DOE. 2005. "Industrial Technologies Program: Energy and Environmental Profile of the U.S. Pulp and Paper Industry." U.S. Department of Energy. https://www.energy.gov/eere/amo/articles/itp-forest-products-energy-and-environmental-profile-us-pulp-and-paper-industry.
36. Personal communication from Edgar Carrejo and Sunkyu Park at North Carolina State University. July 2024.
37. Fisher International. 2023. "Fishersolve database." Fisher International, Accessed October 27, 2023. https://www.holmen.com/en/paper/sustainability/sustainability-stories/fisher-solve-co2/

38. Future Market Insights. 2022. “Unbleached Softwood Kraft Pulp Market Outlook: Forecast 2023-2033.” Future Market Insights, REP-GB-9149. https://www.futuremarketinsights.com/reports/unbleached-softwood-kraft-pulp-market.
39. Parrish, D. 1998. “Black Liquor Recovery Boilers – An Introduction.” Factory Mutual Research Corp. https://www.nationalboard.org/index.aspx?pageID=164&ID=231.
40. Acker & Associates. n.d. “Kraft Black Liquor Recovery Boiler Efficiency Program.” https://ackerandassociates.com/computer_program_examples/30_kraft_black_liquor_recovery_boiler_analysis/.
41. Vakkilainen, E. K. 2017. “11- Recovery Boiler.” Steam Generation from Biomass, 237–259. https://doi.org/10.1016/B978-0-12-804389-9.00011-3.
42. PS Ravishangar, V. 2022. “Energy Efficient Recovery Boilers – An Overview.” Presented at IPPTA Conference. https://ippta.co/wp-content/uploads/2022/08/P.S.Ravishangar-1.pdf.
43. Orr, A. and Simons, H.A. n.d. “Energy Generation and Use in the Kraft Pulp Industry.” American Council for an Energy-Efficient Economy (ACEEE). https://www.aceee.org/files/proceedings/1995/data/papers/SS95_Panel1_Paper04.pdf.
44. Vakkilainen, E. K. 2017. “Black Liquor.” Steam Generation from BioMass. https://www.sciencedirect.com/topics/engineering/black-liquor.
45. Suhr, M., et al. 2015. “Best Available Techniques (BAT) Reference Document for the Production of Pulp, Paper and Board.” JRS Science and Policy Reports, European Commission. https://eippcb.jrc.ec.europa.eu/sites/default/files/2019-11/PP_revised_BREF_2015.pdf.
46. Francis, D. W., M. T. Towers, and T. C. Browne. 2002. “Energy Cost Reduction in the Pulp and Paper Industry – An Energy Benchmarking Perspective.” Canadian Industry Program for Energy Conservation. https://www.osti.gov/etdeweb/servlets/purl/21552944.
47. Reyes Molina, E.A., K. Sweeney, S. Root, et al. 2024. “Technical and Economic Assessment and Gap analysis of Advanced Nuclear Reactor Integration with a Reference Methanol Synthesis Plant.” Idaho National Laboratory, Idaho Falls, ID. https://doi.org/10.2172/2438487.
48. Reyes Molina, E.A., K. Sweeney, S. Root, et al. 2024. “Technical and Economic Assessment and Gap Analysis of Advanced Nuclear Reactor Integration with a Reference Oil Refinery.” Idaho National Laboratory, Idaho Falls, ID. https://doi.org/10.2172/2438486.
49. Worsham, E., Reyes Molina, E. A., N. Guaita, et al. In Progress. “Technoeconomic Analysis of Kraft Pulp Mill Integration with an Advanced Nuclear Reactor: Decarbonization with Carbon Capture, Oxy-Fuel Combustion, and Thermal Substitution.” M2CT-24IN12030512, Idaho National Laboratory, Idaho Falls, ID.
50. World Nuclear News. 2024. “Start up of Chinese industrial nuclear steam projects begins.” https://www.world-nuclear-news.org/Articles/Start-up-of-Chinese-industrial-nuclear-steam-proje.
51. World Nuclear Association. 2024. “Tianwan Nuclear Power Plant Steam Supply, 2016–2024.” Updated Thursday, 16 May 2024. https://world-nuclear.org/our-association/publications/world-nuclear-performance-report/tianwan-nuclear-power-plant-steam-supply.
52. Traceyhonney. 2024. “China launches nuclear steam project at Tianwan.” Nuclear Engineering International. https://www.neimagazine.com/news/china-launches-nuclear-steam-project-at-tianwan-npp-11576695/.

53. LianYunGang Municipal People's Government. 2023. "Petrochemical Industry."https://www.lyg.gov.cn/lygmhywwz/industrialorientation/content/4639a2af-214d-4468-80e1-1d2bdb5c28da.html#:~:text=Lianyungang%20Petrochemical%20Industry%20Base%20is,6%20million%20tons%20of%20ethylene.

54. Energy.Gov. 2018. "Manufacturing Energy and Carbon Footprint." Sector: Petrochemicals (NAICS 325110). https://www.energy.gov/sites/default/files/2022-10/2018_mecs_petrochemicals_energy_carbon_footprint_2022-10-12.pdf.

55. Hydrogen Newsletter. 2022. "What is Steam Methane Reforming (SMR)?" https://www.hydrogennewsletter.com/what-is-steam-methane-reforming-smr/.

56. DLR. n.d. "High Temperature Laboratory Stuttgart (HITS)." DLR Institute of Solar Research. https://www.dlr.de/sf/en/desktopdefault.aspx/tabid-13657/23719_read-54410/.

57. VGVE. 2023. "Maximizing Efficiency: The Potential of Storage Technologies in Power-to-Heat-to-Power Systems 2023." VGBE. https://events.vgbe.energy/events/maximizing-efficiency--the-potential-of-storage-technologies-in-powertoheattopower-systems-2023/11225/KW3R4/info.

58. CRI. n.d. "Projects." Carbon Recycling International. https://www.carbonrecycling.is/projects.

59. AGICO Cement. n.d. https://www.rotarykilnfactory.com/electric-heating-rotary-kiln/.

60. VTT Research. 2022. "Revolutionary electric rotary kiln ready to be scaled for more environmentally friendly cement production after successful customer testing." https://www.vttresearch.com/en/news-and-ideas/revolutionary-electric-rotary-kiln-ready-be-scaled-more-environmentally-friendly.

61. IBU|tec. n.d. "Rotary Kilns Multiple Options – Flexible Service." https://www.ibu-tec.com/facilities/rotary-kilns/.

62. Noritake. 2024. "Indirect-Heating Continuous Rotary Kiln." https://www.noritake.co.jp/eng/products/eeg/parts/detail/145/.

63. Power. 2022. "District Heating Supply from Nuclear Power Plants." https://www.powermag.com/district-heating-supply-from-nuclear-power-plants/.

64. Energy.gov. 2012. "U.S. Manufacturing Energy Use and Greenhouse Gas Emissions Analysis." https://www.energy.gov/sites/default/files/2013/11/f4/energy_use_and_loss_and_emissions_forest.pdf.

65. U.S. Energy Information Administration, "U.S. Energy-Related Carbon Dioxide Emissions, 2023" April 2024, https://www.eia.gov/environment/emissions/carbon/

66. Taghdisian, H., Farhadi, F., and Pishvaie, M. R. 2012. "AN Optimization-oriented green design for methanol plants." *Journal of Chemical Technology and Biotechnology* 87(8):1111–1120. https://doi.org/10.1002/jctb.3727.

67. Nielsen, S.E. 2010. "Latest Developments in Ammonia Production Technology 1." Semantic Scholar. https://www.semanticscholar.org/paper/Latest-developments-in-ammonia-production-1-Nielsen/6504b1edb419d7c9c1195e480a1d5f70b602e1e7.

68. GHG Data. 2022. "OCI Beaumont LLC." GHG Data. https://ghgdata.epa.gov/ghgp/service/facilityDetail/2022?id=1010636&ds=E&et=&popup=true.

69. Siemens Energy. 2021. "Efficiency: More value to your facility - Industrial steam turbines from 2 to 250 MW." Siemens Energy. https://p3.aprimocdn.net/siemensenergy/79160ff2-731e-4ffe-9ab2-b0360117cdde/SE-Brochure-Industrial-Steam-Turbines-2021-pdf_Original%20file.pdf.

Appendix A

List of Operative Plants in the U.S.

A-1. Methanol Synthesis Operative MeOH Synthesis Plants as of Q2 2023

Company	State	Site	Capacity (metric tons per year)	GHGs Metric Tons CO_2-eq./year
Celanese	Texas	Pasadena	1,620,000	Not specified
Proman USA (G2X)	Texas	Pampa	78,750	63,678
OCI Methanol	Texas	Beaumont	200,000	464,535
LyondellBasell Methanol	Texas	Beaumont	Not specified	—
Natgasoline	Texas	Beaumont	1,700,000	677,351
Methanex Geismar	Louisiana	Geismar	2,200,000	995,769
Koch Methanol St. James	Louisiana	St. James Parish	1,800,000	647,654
Eastman Chemical	Tennessee	Kingsport	Not specified	Not specified
Air Products and Chemicals	Tennessee	Kingsport	96,000	Not specified
Ref. Plant Capacity			1,000,000	476,066

A-2. Petroleum Refining Operative Refineries as of Q2 2023

Corporation	Company	State	Site	Barrels per Calendar Day (EIA report)	GHGs (metric tons CO_2e /year) EPA FLIGHT
Saudi Aramco	Motiva Enterprises LLC	Texas	Port Arthur	626,000	5,213,960
Marathon Petroleum Corp	Marathon Petroleum Co LP	Louisiana	Garyville	596,000	3,837,434
Marathon Petroleum Corp	Marathon Petroleum Co LP	Texas	Galveston Bay	593,000	4,908,983
ExxonMobil Corp	ExxonMobil Refining & Supply Co	Texas	Baytown	564,440	12,560,398
ExxonMobil Corp	ExxonMobil Refining & Supply Co	Louisiana	Baton Rouge	522,500	6,077,597

Corporation	Company	State	Site	Barrels per Calendar Day (EIA report)	GHGs (metric tons CO_2e /year) EPA FLIGHT
PDV America Inc	CITGO Petroleum Corp	Louisiana	Lake Charles	455,000	4,389,676
BP PLC	BP Products North America Inc	Indiana	Whiting	435,000	3,532,434
ExxonMobil Corp	ExxonMobil Refining & Supply Co	Texas	Beaumont	369,024	3,390,268
Marathon Petroleum Corp	Tesoro Refining & Marketing Co	California	Carson	363,000	5,344,156
Chevron Corp	Chevron USA Inc	Mississippi	Pascagoula	356,440	3,796,984
WRB Refining LP	WRB Refining LP	Illinois	Wood River	356,000	2,835,160
Koch Industries Inc	Flint Hills Resources LP	Minnesota	Saint Paul	335,000	2,516,855
Valero Energy Corp	Premcor Refining Group Inc	Texas	Port Arthur	335,000	2,071,772
Pemex	Deer Park Refining Ltd Partnership	Texas	Deer Park	312,500	3,196,503
Marathon Petroleum Corp	Marathon Petroleum Co LP	Kentucky	Catlettsburg	291,000	2,238,609
Valero Energy Corp	Valero Refining Co Texas LP	Texas	Corpus Christi	290,000	3,100,540
Chevron Corp	Chevron USA Inc	California	El Segundo	269,000	2,816,051
Koch Industries Inc	Flint Hills Resources LP	Texas	Corpus Christi, West	268,500	2,521,656
Phillips 66 Company	Phillips 66 Company	Texas	Sweeny	265,000	1,892,093
Phillips 66 Company	Phillips 66 Company	Louisiana	Westlake	264,000	1,667,356
Access Industries	Houston Refining LP	Texas	Houston	263,776	2,012,451
Phillips 66 Company	Phillips 66 Company	New Jersey	Linden	258,500	2,285,121
Marathon Petroleum Corp	Marathon Petroleum Co LP	Illinois	Robinson	253,000	1,781,114

Corporation	Company	State	Site	Barrels per Calendar Day (EIA report)	GHGs (metric tons CO_2e /year) EPA FLIGHT
ExxonMobil Corp	ExxonMobil Refining & Supply Co	Illinois	Joliet	251,800	2,675,486
Chevron Corp	Chevron USA Inc	California	Richmond	245,271	2,934,724
BP PLC	BP Products North America Inc	Washington	Ferndale	238,500	1,476,499
TotalEnergies SE	TotalEnergies Petrochemicals & Refining USA	Texas	Port Arthur	238,000	2,097,521
Shell PLC	Shell Oil Products US	Louisiana	Norco	227,900	4,024,139
Valero Energy Corp	Valero Refining Co Texas LP	Texas	Texas City	225,000	77,590
Valero Energy Corp	Valero Refining New Orleans LLC	Louisiana	Norco	215,000	70,079
Phillips 66 Company	Phillips 66 Company	Oklahoma	Ponca City	208,000	1,760,333
Valero Energy Corp	Valero Refining Co Texas LP	Texas	Houston	205,000	1,524,955
Valero Energy Corp	Diamond Shamrock Refining Co LP	Texas	Sunray	195,000	1,263,515
PBF Energy Co LLC	Chalmette Refining LLC	Louisiana	Chalmette	190,000	1,542,493
Delta Air Lines Inc	Monroe Energy LLC	Pennsylvania	Trainer	190,000	1,270,102
Cenovus Energy Inc	Lima Refining Company	Ohio	Lima	183,000	1,263,350
PDV America Inc	PDV Midwest Refining LLC	Illinois	Lemont	182,685	1,770,675
Valero Energy Corp	Valero Refining Company-Tennessee LLC	Tennessee	Memphis	180,000	1,121,986
PBF Energy Co LLC	Toledo Refining Co LLC	Ohio	Toledo	172,800	1,558,318

Corporation	Company	State	Site	Barrels per Calendar Day (EIA report)	GHGs (metric tons CO_2e /year) EPA FLIGHT
PBF Energy Co LLC	Delaware City Refining Co LLC	Delaware	Delaware City	171,000	3,518,216
PDV America Inc	CITGO Refining & Chemical Inc	Texas	Corpus Christi	167,500	1,423,097
HF Sinclair Corp	HF Sinclair El Dorado Refining LLC	Kansas	El Dorado	162,000	1,359,794
PBF Energy Co LLC	Paulsboro Refining Co LLC	New Jersey	Paulsboro	160,000	724,679
PBF Energy Co LLC	Torrance Refining Co LLC	California	Torrance	160,000	1,579,561
PBF Energy Co LLC	Martinez Refining Co LLC	California	Martinez	156,400	2,571,284
WRB Refining LP	WRB Refining LP	Texas	Borger	149,000	1,101,063
BP-Husky Refining LLC	BP-Husky Refining LLC	Ohio	Toledo	148,920	756,063
HF Sinclair Corp	HF Sinclair Puget Sound Refining LLC	Washington	Anacortes	145,000	1,907,858
Valero Energy Corp	Valero Refining Co California	California	Benicia	145,000	2,016,257
Marathon Petroleum Corp	Marathon Petroleum Co LP	Michigan	Detroit	140,000	881,980
Phillips 66 Company	Phillips 66 Company	California	Wilmington	139,000	1,264,169
Transworld Oil USA Inc	Calcasieu Refining Co	Louisiana	Lake Charles	135,500	253,106
Marathon Petroleum Corp	Western Refining Company LP	Texas	El Paso	133,000	790,786
CVR Energy	Coffeyville Resources Rfg & Mktg	Kansas	Coffeyville	132,000	951,302
Valero Energy Corp	Valero Refining-Meraux LLC	Louisiana	Meraux	125,000	655,088
Phillips 66 Company	Phillips 66 Company	California	Rodeo	120,200	1,171,648

Corporation	Company	State	Site	Barrels per Calendar Day (EIA report)	GHGs (metric tons CO_2e /year) EPA FLIGHT
Marathon Petroleum Corp	Tesoro Refining & Marketing Co	Washington	Anacortes	119,000	1,370,882
Chevron Corp	Pasadena Refining Systems Inc	Texas	Pasadena	112,229	316,742
CHS Inc	CHS McPherson Refinery Inc	Kansas	Mcpherson	110,600	842,297
HF Sinclair Corp	HF Sinclair Navajo Refining LLC	New Mexico	Artesia	110,000	669,694
Kinder Morgan Energy Partners LP	Kinder Morgan Crude & Condensate	Texas	Galena Park	105,000	198,733
Phillips 66 Company	Phillips 66 Company	Washington	Ferndale	105,000	718,028
Marathon Petroleum Corp	St Paul Park Refining Co LLC	Minnesota	Saint Paul	105,000	730,768
Marathon Petroleum Corp	Marathon Petroleum Co LP	Ohio	Canton	100,000	550,349
PAR Pacific Holdings	PAR Hawaii Refining LLC	Hawaii	Ewa Beach	93,500	0
Valero Energy Corp	Diamond Shamrock Refining Co LP	Texas	Three Rivers	89,000	702,322
Vertex Energy LP	Vertex Refining Alabama LLC	Alabama	Saraland	88,000	418,142
Valero Energy Corp	Valero Refining Co Oklahoma	Oklahoma	Ardmore	86,000	858,610
HF Sinclair Corp	HF Sinclair Tulsa Refining LLC	Oklahoma	Tulsa West	85,000	498,897
Valero Energy Corp	Ultramar Inc	California	Wilmington Refinery	85,000	967,407
Delek Group Ltd	Lion Oil Co	Arkansas	El Dorado	83,000	686,296
Delek Group Ltd	Alon Refining Krotz Springs Inc	Louisiana	Krotz Springs	80,000	549,792

Corporation	Company	State	Site	Barrels per Calendar Day (EIA report)	GHGs (metric tons CO_2e /year) EPA FLIGHT
HF Sinclair Corp	HF Sinclair Wyoming Refining C	Wyoming	Sinclair	75,000	827,716
Placid Oil Co	Placid Refining Co	Louisiana	Port Allen	75,000	583,804
Koch Industries Inc	Flint Hills Resources LP	Texas	Corpus Christi, East	74,500	530,245
CVR Energy	Wynnewood Refining Co	Oklahoma	Wynnewood	74,500	607,210
Delek Group Ltd	Alon USA Energy Inc	Texas	Big Spring	73,000	721,257
Delek Group Ltd	Delek Refining Ltd	Texas	Tyler	71,000	530,833
Marathon Petroleum Corp	Tesoro Refining & Marketing Co	North Dakota	Mandan	71,000	695,213
HF Sinclair Corp	HF Sinclair Tulsa Refining LLC	Oklahoma	Tulsa East	70,300	685,854
Marathon Petroleum Corp	Tesoro Alaska Company LLC	Alaska	Kenai	68,000	503,152
Suncor Energy Inc	Suncor Energy (USA) Inc	Colorado	Commerce City West	67,000	741,437
Phillips 66 Company	Phillips 66 Company	Montana	Billings	66,000	559,760
Marathon Petroleum Corp	Tesoro Refining & Marketing Co	Utah	Salt Lake City	66,000	647,519
Red Apple Group Inc	United Refining Co	Pennsylvania	Warren	65,000	545,840
CHS Inc	Cenex Harvest States Coop	Montana	Laurel	62,500	670,674
ExxonMobil Corp	ExxonMobil Refining & Supply Co	Montana	Billings	61,500	517,186
Buckeye Partners LP	Buckeye Texas Processing LLC	Texas	Corpus Christi	60,000	101,407
Calumet Specialty Products Partners, LP	Calumet Shreveport LLC	Louisiana	Shreveport	57,000	260,650
Arctic Slope Regional Corp	Petro Star Inc	Alaska	Valdez	55,000	60,063
Chevron Corp	Chevron USA Inc	Utah	Salt Lake City	54,720	485,756

Corporation	Company	State	Site	Barrels per Calendar Day (EIA report)	GHGs (metric tons CO_2e /year) EPA FLIGHT
Hunt Consolidated Inc	Hunt Refining Co	Alabama	Tuscaloosa	50,000	726,701
Hartree Partners LP	Hartree Channelview LLC	Texas	Channelview	45,000	24,532
Texas International Terminals	Texas International Terminals	Texas	Galveston	45,000	65,358
Magellan Midstream Partners LP	Magellan Terminal Holdings LP	Texas	Corpus Christi	42,500	68,631
PAR Pacific Holdings	U.S. Oil & Refining Co	Washington	Tacoma	40,700	128,320
HF Sinclair Corp	HF Sinclair Woods Cross Refining LLC	Utah	Woods Cross	39,330	395,779
Cenovus Energy Inc	Superior Refining Company LLC	Wisconsin	Superior	38,000	32,040
Suncor Energy Inc	Suncor Energy (USA) Inc	Colorado	Commerce City East	36,000	Not specified
Countrymark Coop Inc	Countrymark Refining & Logistics LLC	Indiana	Mount Vernon	34,500	219,961
FJ Management Inc	Big West Oil Co	Utah	North Salt Lake	31,664	288,509
HF Sinclair Corp	HF Sinclair Casper Refining Co	Wyoming	Evansville	29,850	306,592
Ergon Inc	Ergon Refining Inc	Mississippi	Vicksburg	26,500	138,268
Kern Oil & Refining Co	Kern Oil & Refining Co	California	Bakersfield	26,000	142,322
Petromax Refining Co LLC	Petromax Refining Co LLC	Texas	Houston	25,000	—
Calumet Specialty Products Partners, LP	Calumet Montana Refining LLC	Montana	Great Falls	24,600	170,677
Ergon Inc	Ergon West Virginia Inc	West Virginia	Newell	22,300	199,255
Arctic Slope Regional Corp	Petro Star Inc	Alaska	North Pole	21,000	37,924

Corporation	Company	State	Site	Barrels per Calendar Day (EIA report)	GHGs (metric tons CO_2e /year) EPA FLIGHT
Starlight Relativity Acquisition Co	The San Antonio Refinery	Texas	San Antonio	20,000	83,152
PAR Pacific Holdings	Hermes Consolidated LLC	Wyoming	NewCastle	18,000	160,314
ConocoPhillips	ConocoPhillips Alaska Inc	Alaska	Prudhoe Bay	15,000	504,976
San Joaquin Refining Co Inc	San Joaquin Refining Co Inc	California	Bakersfield	15,000	63,816
Silver Eagle Refining Inc	Silver Eagle Refining	Utah	Woods Cross	15,000	43,429
Blue Dolphin Energy Co	Lazarus Energy LLC	Texas	Nixon	14,000	22,108
Calumet Specialty Products Partners, LP	Calumet Lubricants Co LP	Louisiana	Cotton Valley	13,020	84,708
American Refining Group Inc	American Refining Group Inc	Pennsylvania	Bradford	11,000	111,089
Hunt Consolidated Inc	Hunt Southland Refining Co	Mississippi	Sandersville	11,000	30,582
World Oil Co	Lunday Thagard Co	California	South Gate	8,500	36,986
Calumet Specialty Products Partners, LP	Calumet Lubricants Co LP	Louisiana	Princeton	8,300	65419
Martin Resource Management Grp	Cross Oil Refining & Marketing Inc	Arkansas	Smackover	7,500	34,354
Hilcorp Energy Co	Hilcorp North Slope LLC	Alaska	Prudhoe Bay	6,500	23,096
Valero Energy Corp	Valero Refining Co California	California	Wilmington Asphalt Plant	6,300	14,276
Goodway Refining LLC	Goodway Refining LLC	Alabama	Atmore	4,100	7,917
Silver Eagle Refining Inc	Silver Eagle Refining	Wyoming	Evanston	3,000	11,343
Foreland Refining Corp	Foreland Refining Corp	Nevada	Ely	2,000	16,174
Talley Asphalt Products Inc	Talley Asphalt Products Inc	California	Kern	1,700	34,063
Ref. Plant Capacity				100,000	807,368

A-3. Pulp and Paper Operative Kraft Mills as of Q2 2023

Facility Name	Parent Companies	State	City Name	GHG Quantity (metric tons CO_2e)
Ahlstrom - Munksjo Mosinee LLC	Ahlstrom-Munksjo North America Specialty Solutions LLC (100%)	WI	Mosinee	823,990.8
Alabama River Cellulose, LLC	Koch Industries Inc (100%)	AL	Perdue Hill	1,058,727.25
Billerud Escanaba LLC	Billerud Americas Corp (100%)	MI	Escanaba	2,007,106
Billerud Quinnesec LLC	Billerud Americas Corp (100%)	MI	Quinnesec	373,288.35
Blue Ridge Paper Products LLC	Evergreen Packaging Group Holdings Inc (100%)	NC	Canton	3,616,131.55
Boise Paper	Packaging Corp of America (100%)	WA	Wallula	817,940.1
Boise White Paper LLC	Packaging Corp of America (100%)	MN	International Falls	644,254.3
Boise White Paper Mill	Packaging Corp of America (100%)	AL	Jackson	501,270.2
Brunswick Cellulose, Inc.	Koch Industries Inc (100%)	GA	Brunswick	1,070,890.9
Cascade Pacific Pulp Halsey Pulp Mill	International Grand Investment Corp (100%)	OR	Halsey	218,306.6
Clearwater Paper Corp	Clearwater Paper Corp (100%)	ID	Lewiston	1,143,536.65
Clearwater Paper Corp-Cypress Bend Mill	Clearwater Paper Corp (100%)	AR	Arkansas City	755,424.5
Cosmopolis Pulp Mill	Cosmo Specialty Fibers Inc (100%)	WA	Cosmopolis	219,555.75
Domtar A. W. LLC-Nekoosa	Domtar Corp (100%)	WI	Nekoosa	947,420.1
Domtar Aw LLC Ashdown Mill	Domtar Corp (100%)	AR	Ashdown	1,756,354.7
Domtar Paper Co LLC Hawesville Mill	Domtar Corp (100%)	KY	Hawesville	661,759
Domtar Paper Company LLC - Plymouth	Domtar Corp (100%)	NC	Plymouth	585,457.1
Domtar Paper Company, LLC	Domtar Corp (100%)	PA	Johnsonburg	780,523.7
Evergreen Packaging-Pine Bluff	Reynolds Consumer Products Inc (100%)	AR	Pine Bluff	678,554.05
Finch Paper LLC	Atlas Holdings LLC (100%)	NY	Glens Falls	601,745.85
Foley Cellulose, LLC	Koch Industries Inc (100%)	FL	Perry	1,209,297.55
Georgia Pacific Monticello LLC	Koch Industries Inc (100%)	MS	Monticello	1,745,585.45
Georgia Pacific Palatka LLC	Koch Industries Inc (100%)	FL	Palatka	1,571,115.3

Facility Name	Parent Companies	State	City Name	GHG Quantity (metric tons CO_2e)
Georgia-Pacific Brewton LLC	Koch Industries Inc (100%)	AL	Brewton	1,598,849.75
Georgia-Pacific Cedar Springs LLC	Koch Industries Inc (100%)	GA	Cedar Springs	2,324,871.5
Georgia-Pacific Naheola LLC	Koch Industries Inc (100%)	AL	Pennington	1,011,653.8
Georgia-Pacific Toledo LLC	Koch Industries Inc (100%)	OR	Toledo	1,303,896.8
Gp Big Island, LLC	Koch Industries Inc (100%)	VA	Big Island	700,988.95
Graphic Packaging International	Graphic Packaging Holding Co (100%)	GA	Macon	457,168.15
Graphic Packaging International - Augusta	Graphic Packaging Holding Co (100%)	GA	Augusta	1,255,059.6
Graphic Packaging International - Texarkana Mill	Graphic Packaging Holding Co (100%)	TX	Queen City	1,194,411.5
Graphic Packaging International LLC, West Monroe Mill Plant 31	Graphic Packaging Holding Co (100%)	LA	West Monroe	1,557,851.9
Green Bay Packaging/Ark Kraft	Green Bay Packaging Inc (100%)	AR	Morrilton	363,232.9
Greif Riverville LLC	Greif Inc (100%)	VA	Riverville	557,826.4
Hood Container of Louisiana, LLC	Hood Cos Inc (100%)	LA	Saint Francisville	728,673.6
International Paper - Bogalusa Mill	International Paper Co (100%)	LA	Bogalusa	474,091.85
International Paper - Columbus Mill	International Paper Co (100%)	MS	Columbus	480,449.65
International Paper - Flint River Mill	International Paper Co (100%)	GA	Oglethorpe	279,382.15
International Paper - Georgetown Mill	International Paper Co (100%)	SC	Georgetown	1,181,305.8
International Paper - Mansfield Mill	International Paper Co (100%)	LA	Mansfield	2,369,222.55
International Paper - New Bern Mill	International Paper Co (100%)	NC	Vanceboro	365,378.45
International Paper - Orange Mill	International Paper Co (100%)	TX	Orange	1,007,528.7
International Paper - Pensacola Mill	International Paper Co (100%)	FL	Cantonment	1,098,239.4
International Paper - Pine Hill Mill	International Paper Co (100%)	AL	Pine Hill	523,389.7

Facility Name	Parent Companies	State	City Name	GHG Quantity (metric tons CO_2e)
International Paper - Red River Mill	International Paper Co (100%)	LA	Campti	607,742.6
International Paper - Springfield Mill	International Paper Co (100%)	OR	Springfield	415,141.1
International Paper - Valliant Mill	International Paper Co (100%)	OK	Valliant	1,466,527
International Paper - Vicksburg Mill	International Paper Co (100%)	MS	Redwood	645,312.55
International Paper Rome Linerboard Mill	International Paper Co (100%)	GA	Rome	697,930.4
International Paper Savanna Mill	International Paper Co (100%)	GA	Savannah	1,542,517.65
International Paper-Franklin Mill	International Paper Co (100%)	VA	Franklin	305,224.2
International Paper-Port Wentworth	International Paper Co (100%)	GA	Port Wentworth	417,568.85
International Paper-Prattville Mill	International Paper Co (100%)	AL	Prattville	1,663,394.7
International Paper-Riegelwood	International Paper Co (100%)	NC	Riegelwood	892,440.9
International Paper-Riverdale Mill	International Paper Co (100%)	AL	Selma	1,528,573.65
Interstate Paper LLC Dba Ds Smith Riceboro	Ds Smith Plc (100%)	GA	Riceboro	206,661.7
Leaf River Cellulose LLC	Koch Industries Inc (100%)	MS	New Augusta	350,641.8
Marlboro Paper Mill	Domtar Corp (100%)	SC	Bennettsville	427,595.25
Nd Otm LLC	Nd Paper LLC (100%)	ME	Old Town	184,894.95
Nd Paper Inc - Rumford Division	Nd Paper LLC (100%)	ME	Rumford	1,790,754.05
New-Indy Catawba LLC	New-Indy Jv Corp (100%)	SC	Catawba	819,965.3
Nippon Dynawave Packaging	Nippon Paper Industries Co Ltd (100%)	WA	Longview	1,528,743.8
Packaging Corp. Of America	Packaging Corp of America (100%)	GA	Valdosta	279,129
Packaging Corporation of America	Packaging Corp of America (100%)	TN	Counce	1,091,977.05
Packaging Corporation of America	Packaging Corp of America (100%)	LA	Deridder	1,091,317.2
Packaging Corporation of America	Packaging Corp of America (100%)	MI	Filer City	713,878.85
Packaging Corporation of America-Tomahawk	Packaging Corp of America (100%)	WI	Tomahawk	587,835.05

Facility Name	Parent Companies	State	City Name	GHG Quantity (metric tons CO_2e)
Phoenix Paper Wickliffe LLC	Phoenix Paper Wickliffe LLC (100%)	KY	Wickliffe	515,587.7
Pixelle Androscoggin LLC	Pixelle Specialty Solutions LLC (100%)	ME	Jay	41,682.6
Pixelle Specialty Solutions	Hig Capital (100%)	PA	Spring Grove	1,926,915.55
Pixelle Specialty Solutions LLC	Pixelle Specialty Solutions LLC (100%)	OH	Chillicothe	1,536,425.45
Port Townsend Paper Corp	Crown Paper Group Inc (100%)	WA	Port Townsend	331,041.35
Rayonier Performance Fibers LLC Fernandina Mill	Rayonier Advanced Materials Inc (100%)	FL	Fernandina Beach	365,561.05
Rayonier Performance Fibers LLC, Jesup Mill	Rayonier Advanced Materials Inc (100%)	GA	Jesup	940,373.4
Resolute Forest Product - Coosa Pines Operation	Resolute Fp Us Inc (100%)	AL	Coosa Pines	644,042.65
Resolute Forest Products Calhoun Operation	Resolute Fp Us Inc (100%)	TN	Calhoun	277,896.45
Sappi Cloquet LLC	Sappi North America Inc (100%)	MN	Cloquet	702,653.1
Sappi North America, Inc. - Somerset Operations	Sappi North America Inc (100%)	ME	Skowhegan	1,161,290.35
Sylvamo Eastover Mill	Sylvamo North America LLC (100%)	SC	Eastover	847,015
Sylvamo Ticonderoga Mill	Sylvamo North America LLC (100%)	NY	Ticonderoga	739,065.2
Thilmany Mill	Ahlstrom-Munksjo North America Specialty Solutions LLC (100%)	WI	Kaukauna	1,230,255.05
Twin Rivers Pine Bluff LLC	Twin Rivers Paper Co LLC (100%)	AR	White Hall	255,814.3
Wauna Mill	Koch Industries Inc (100%)	OR	Clatskanie	1,137,930
West Rock Mill Company, LLC	Westrock Co (100%)	AL	Demopolis	310,901.4
Westrock Charleston Kraft, LLC	Westrock Co (100%)	SC	Charleston	2,295,950.15
Westrock Coated Board, LLC	Westrock Co (100%)	AL	Phenix City	1,292,832.9
Westrock Cp, LLC - Fernandina Beach Mill (Formerly Smurfit-Stone)	Westrock Co (100%)	FL	Fernandina Beach	1,622,214.25
Westrock Cp, LLC - Florence Mill	Westrock Co (100%)	SC	Florence	1,030,523.85

Facility Name	Parent Companies	State	City Name	GHG Quantity (metric tons CO_2e)
Westrock Cp, LLC - Hodge Mill	Westrock Co (100%)	LA	Hodge	1,288,500.3
Westrock Cp, LLC - Panama City Mill (Formerly Rocktenn)	Westrock Co (100%)	FL	Panama City	623,525.05
Westrock Cp, LLC - Stevenson Mill	Westrock Co (100%)	AL	Stevenson	402,861.25
Westrock Cp, LLC - West Point Mill	Westrock Co (100%)	VA	West Point	1,928,542.35
Westrock Cp, LLC--Hopewell Mill (Formerly Rocktenn)	Westrock Co (100%)	VA	Hopewell	478,121.5
Westrock Kraft Paper, LLC	Westrock Co (100%)	NC	Roanoke Rapids	1,045,667.2
Westrock Longview LLC	Westrock Co (100%)	WA	Longview	868,574.25
Westrock Tacoma Mill	Westrock Co (100%)	WA	Tacoma	543,355.35
Westrock Texas, L.P.	Westrock Co (100%)	TX	Evadale	1,534,109.75
Westrock Virginia, Covington	Westrock Co (100%)	VA	Covington	3,975,139.75
Wisconsin Rapids Paper Mill	Billerud Americas Corp (100%)	WI	Wisconsin Rapids	12,005.95
Woodland Pulp LLC	International Grand Investment Corp (100%)	ME	Baileyville	782,685.85
Ref. Plant Capacity				748,281

Appendix B

SMR Methanol Flowsheets, Heat, and Mass Balances

B-1. SMR PFD and Detailed Material and Energy Balances

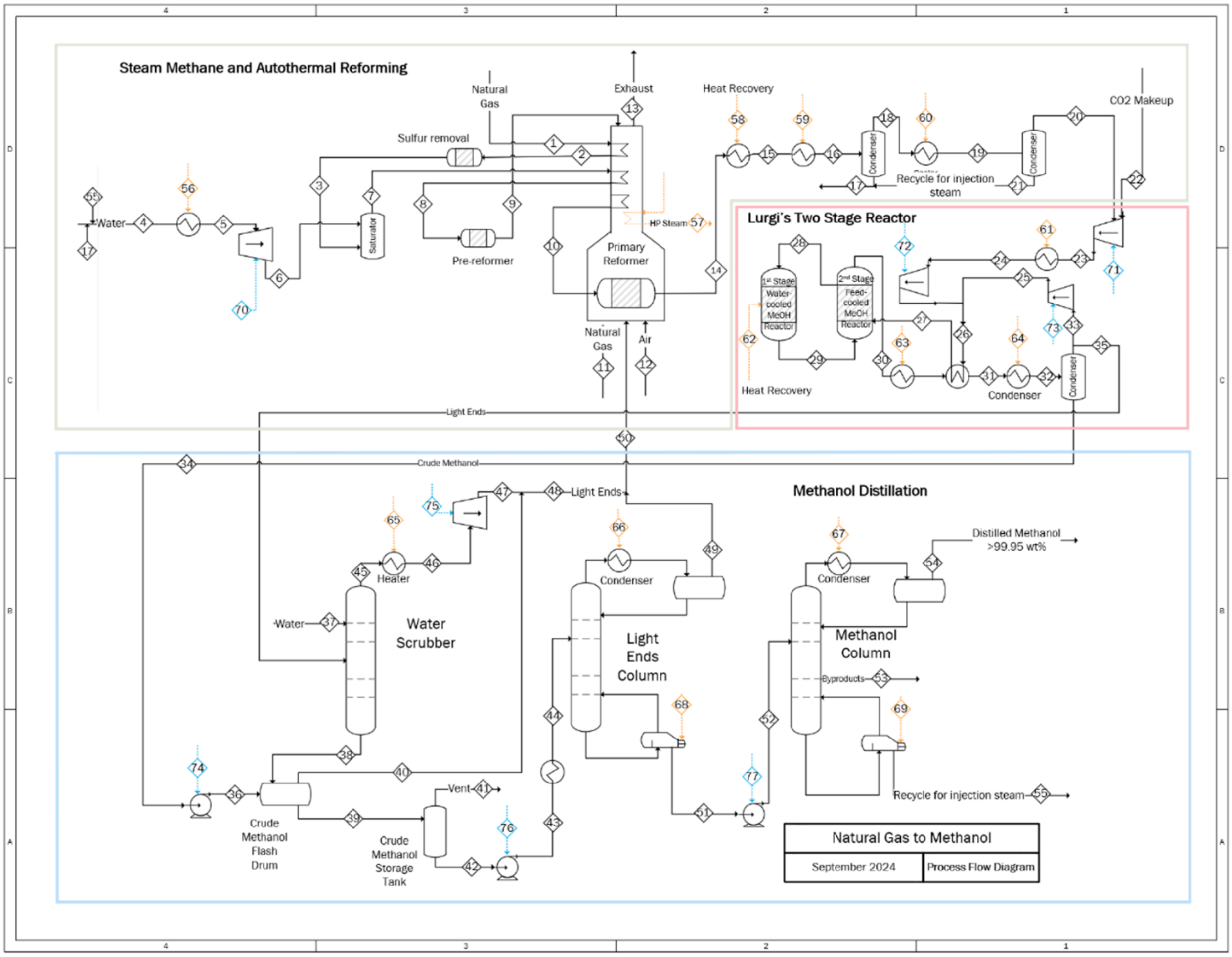


Table B28. SMR material balance.

Stream Number	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22
Temperature (°C)	25.0	381.4	381.4	65.7	400.0	204.9	254.6	381.4	254.6	900.0	25.0	25.0	112.9	875	164.3	80.0	80.0	80.0	35.0	33.9	35.0	25.0
Pressure (bar)	16.5	16.5	16.5	110.0	100.0	16.5	16.5	16.5	16.5	16.5	16.5	1.0	1.0	16.484	16.5	16.5	16.5	16.5	16.5	14.7	16.5	14.7
Enthalpy Flow (MW)	-66.9	-51.8	-51.8	-759.7	-100.7	-139.2	-682.8	-51.8	-682.8	-553.6	-66.9	0.0	-362.3	-399.095	-527.7	-604.3	-407.3	-197.0	-206.4	-248.2	-26.1	-67.9
Mass Flows (tonne/day)	1300.0	1300.0	1300.0	4147.0	4147.0	4147.0	5447.0	1300.0	5447.0	5447.0	1300.0	11750.4	12374.9	5447	5447.0	5447.0	2231.9	3215.1	3215.1	3730.1	141.0	656.0
O2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	2736.9	490.5	4.58E-15	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
H2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	79.1	0.0	0.0	0.0	483.9248	483.9	483.9	0.0	483.9	483.9	484.3	0.0	0.4
N2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	9013.5	9010.1	0	0.0	0.0	0.0	0.0	0.0	0.8	0.0	0.8
CO	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	9.6	0.0	0.0	0.1	1405.956	1406.0	1406.0	0.0	1406.0	1406.0	1406.0	0.0	0.0
CO2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	479.5	0.0	0.0	1317.7	1043.243	1043.2	1043.2	0.6	1042.7	1042.7	1697.4	0.0	654.8
NH3	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
H2O	0.0	0.0	0.0	4147.0	4147.0	4147.0	4147.0	0.0	4147.0	3748.3	0.0	0.0	1547.3	2388.642	2388.6	2388.6	2231.3	157.3	157.3	16.3	141.0	0.0
CH4	1163.8	1163.8	1163.8	0.0	0.0	0.0	1163.8	1163.8	1163.8	1130.5	1163.8	0.0	0.0	125.2286	125.2	125.2	0.0	125.2	125.2	125.2	0.0	0.0
C2H6	68.9	68.9	68.9	0.0	0.0	0.0	68.9	68.9	68.9	0.0	68.9	0.0	0.0	0.004346	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
C3H8	67.3	67.3	67.3	0.0	0.0	0.0	67.3	67.3	67.3	0.0	67.3	0.0	0.0	4.04E-07	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
I-C4	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	1.49E-11	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
N-C4	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	3.39E-11	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
MEOH	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.000714	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ETOH	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	2.74E-07	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
DME	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	2.16E-10	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
AR	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
H2S	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NO	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	9.2	0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NO2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0

Table B29. Material balance for Lurgi’s Two Stage Reactor.

Stream Number	23	24	25	26	27	28	29	30	31	32	33
Temperature (°C)	218.0	37.8	37.8	52.7	190.0	262.3	265.0	208.6	221.0	37.8	37.8
Pressure (bar)	55.0	55.0	89.8	92.0	91.7	91.7	91.3	91.2	90.9	89.8	89.8
Enthalpy Flow (MW)	-225.7	-248.2	-452.3	-689.3	-600.0	-551.8	-621.6	-669.8	-661.9	-831.4	-465.8
Mass Flows (tonne/day)	3730.1	3730.1	8777.0	12497.9	12497.9	12497.9	12497.9	12497.9	12497.9	12497.9	9039.2
O2	0	0	0	0	0	0	0	0	0	0	0
H2	484.3	484.3	2199.8	2684.1	2684.1	2684.1	2303.8	2266.0	2266.0	2266.0	2265.5
N2	0.8	0.8	28.1	28.9	28.9	28.9	28.9	28.9	28.9	28.9	28.9
CO	1406.0	1406.0	319.5	1725.4	1725.4	1725.4	630.5	329.1	329.1	329.1	329.0
CO2	1697.4	1697.4	2099.0	3796.5	3796.5	3796.5	2175.6	2216.7	2216.6	2216.6	2161.7
NH3	0	0	0	0	0	0	0	0	0	0	0
H2O	16.3	16.3	8.9	16.0	16.0	16.0	680.7	664.7	664.7	664.7	9.2
CH4	125.2	125.2	3959.7	4085.0	4085.0	4085.0	4085.0	4085.0	4085.0	4085.0	4078.0
C2H6	0.0	0.0	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1
C3H8	4.04E-07	4.04E-07	9.15E-06	9.55E-06	9.55E-06	9.55E-06	9.55E-06	9.55E-06	9.55E-06	9.55E-06	9.42E-06
I-C4	0	0	0	0	0	0	0	0	0	0	0
N-C4	3.39E-11	3.39E-11	0	3.39E-11	3.39E-11	3.39E-11	3.39E-11	3.39E-11	3.39E-11	3.39E-11	0
MEOH	0.000253	0.000253	156.868	156.8683	156.8683	156.8683	2585.132	2897.145	2897.144	2897.144	161.5531
ETOH	9.78E-08	9.78E-08	0.219273	0.219273	0.219273	0.219273	3.100397	4.84752	4.847757	4.847757	0.225822
DME	2.15E-10	2.15E-10	4.862091	4.862091	4.862091	4.862091	5.112789	5.39864	5.39864	5.39864	5.007303
AR	0	0	0	0	0	0	0	0	0	0	0
H2S	0	0	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0	0	0

Table B30. Material balance for methanol distillation.

Stream Number	35	36	37	38	39	40	41	42	43	44	45	46	47	48	49	50	51	52	53	54	55
Temperature (°C)	37.8	38.2	25.0	39.9	37.9	39.9	37.7	37.7	37.7	54.4	39.3	220.0	12.7	13.7	45.5	14.1	84.1	84.1	107.0	50.4	117.2
Pressure (bar)	89.8	2.4	81.8	81.8	2.4	81.8	1.0	1.0	1.8	1.6	81.8	81.727	2.5	2.4	1.4	1.4	1.7	2.0	1.658529	1.2	1.7
Enthalpy Flow (MW)	-365.6	-365.9	-5.6	-5.9	-367.8	-5.9	-1.0	-366.8	-366.8	-364.0	-13.2	-10.4	-13.6	-17.6	-1.9	-19.5	-357.1	-357.1	-12.1	-236.1	-111.9
Mass Flows (tonne/day)	3458.7	3458.7	30.0	34.1	3449.0	34.1	10.3	3438.6	3438.6	3438.6	258.0	258.0	258.0	301.9	20.3	322.1	3418.4	3418.4	70.0	2728.2	620.1
O2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
H2	0.6	0.6	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	65.7	65.7	65.7	66.2	0.0	66.2	0.0	0.0	0.0	0.0	0.0
N2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.8	0.8	0.8	0.8	0.0	0.8	0.0	0.0	0.0	0.0	0.0
CO	0.1	0.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	9.5	9.5	9.5	9.7	0.0	9.7	0.0	0.0	0.0	0.0	0.0
CO2	54.9	54.9	0.0	0.0	23.2	0.0	8.2	15.0	15.0	15.0	62.7	62.7	62.7	94.4	15.0	109.4	0.0	0.0	0.0	0.0	0.0
NH3	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
H2O	655.5	655.5	30.0	29.6	684.8	29.6	0.1	684.7	684.7	684.7	0.7	0.7	0.7	1.0	0.0	1.0	684.7	684.7	64.5	0.0	620.1
CH4	7.0	7.0	0.0	0.0	0.3	0.0	0.3	0.0	0.0	0.0	118.3	118.3	118.3	124.9	0.0	125.0	0.0	0.0	0.0	0.0	0.0
C2H6	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
C3H8	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
I-C4	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
N-C4	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
MEOH	2735.6	2735.6	0.0	4.5	2735.7	4.5	1.7	2734.0	2734.0	2734.0	0.2	0.2	0.2	4.6	4.9	9.5	2729.1	2729.1	3.4	2725.7	0.0
ETOH	4.6	4.6	0.0	0.0	4.6	0.0	0.0	4.6	4.6	4.6	0.0	0.0	0.0	0.0	0.0	0.0	4.6	4.6	2.1	2.5	0.0
DME	0.4	0.4	0.0	0.0	0.3	0.0	0.0	0.2	0.2	0.2	0.1	0.1	0.1	0.3	0.2	0.5	0.0	0.0	0.0	0.0	0.0
AR	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
H2S	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NO	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NO2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0

Table B31. PFD Detailed energy balance.

Stream	Equipment	Heat Duty (MWt)	Power Consumption (MWe)
*56	heater	135	-
*57	heater	-27.5	-
*58	heat recovery	-151	-
59	cooler	-52.9	-
60	cooler	-9.4	-
61	cooler	-22.4	-
*62	heat released	-69.8	-
*63	heat recovery	-7.9	-
*64	cooler	-80.2	-
*65	heater	2.8	-
66	dist. twr condens	-1.0	-
67	dist. twr condens	-93.3	-
68	reboiler	6	-
69	reboiler	93.1	-
70	turbine	-	-13.8
71	compressor		22.4
72	compressor	-	8.0
73	compressor	-	1.4
74	hydraulic turbine	-	-0.2
75	expander	-	-2.9
76	pump	-	0.0
77	pump	-	0.0

*heat integration

B-2. SMR Aspen Flow Sheets and Material and Energy Balances

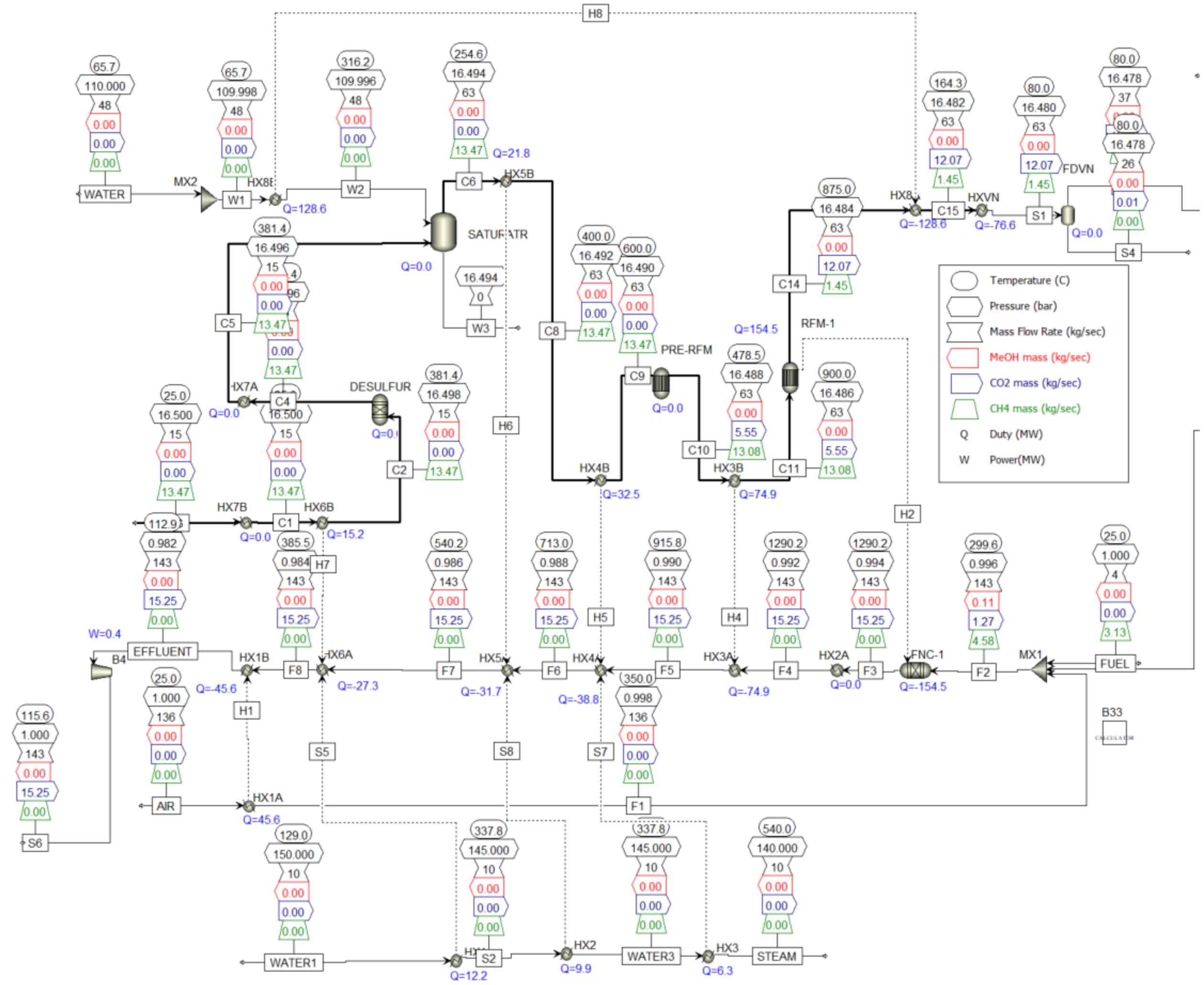


Figure B1. Aspen Plus SMR model reforming section and furnace.

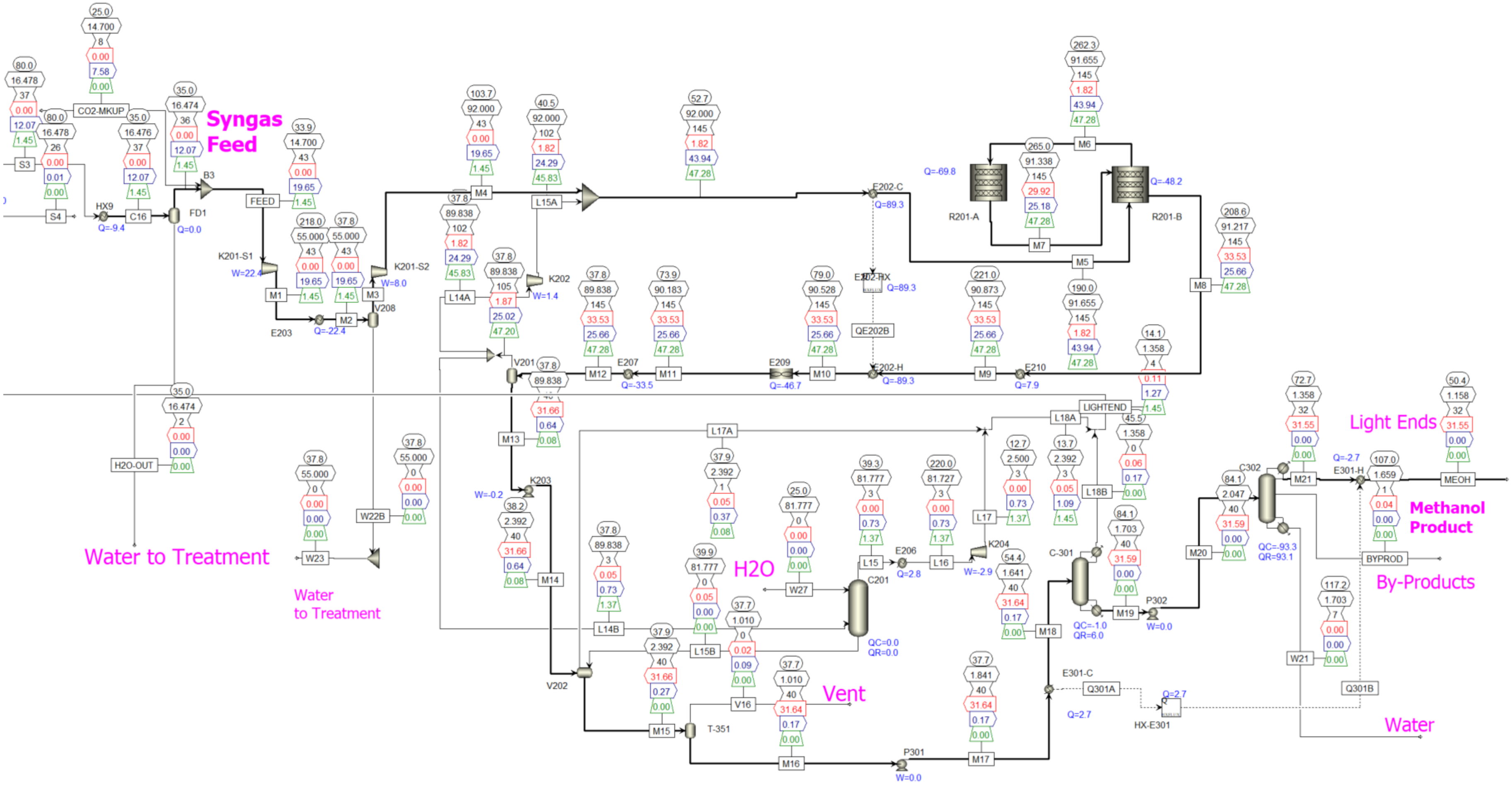


Figure B2. Aspen Plus SMR model methanol synthesis and distillation sections.

B-2.1   SMR Model Material and Heat Balance

Table B32. Material balance for SMR methanol model.

Stream Name	**Units**	**AIR**	**BYPROD**	**C1**	**C2**	**C4**	**C5**	**C6**	**C8**	**C9**	**C10**	**C11**	**C14**	**C15**	**C16**	**C17**	**CO2-MKUP**	**EFFLUENT**	**F1**	**F2**	**F3**	**F4**	**F5**	**F6**	**F7**	**F8**	**FEED**	**FUEL**
From			C302	HX7B	HX6B	DESULFUR	HX7A	SATURATR	HX5B	HX4B	PRE-RFM	HX3B	RFM-1	HX8A	HX9	FD1		HX1B	HX1A	MX1	FNC-1	HX2A	HX3A	HX4A	HX5A	HX6A	B3	
To		HX1A		HX6B	DESULFUR	HX7A	SATURATR	HX5B	HX4B	PRE-RFM	HX3B	RFM-1	HX8A	HXVN	FD1	B3	B3	B4	MX1	FNC-1	HX2A	HX3A	HX4A	HX5A	HX6A	HX1B	K201-S1	MX1
Temperature	C	25	107.0222996	25	381.4	381.3999181	381.3999181	254.5561719	400	600	478.481007	900	875	164.2829814	35	34.99959303	25	112.9435245	350	299.6301267	1290.180115	1290.180257	915.8169462	713.0021209	540.2492838	385.4764242	33.85632655	25
Pressure	bar	1	1.658528568	16.5	16.498	16.496	16.496	16.494	16.492	16.49	16.488	16.486	16.484	16.482	16.476	16.474	14.7	0.982	0.998	0.996	0.994	0.992	0.99	0.988	0.986	0.984	14.7	1
Mass Vapor Fraction		1	0	1	1	1	1	1	1	1	1	1	1	1	0.956135188	1	1	1	1	1	1	1	1	1	1	1	1	1
Mass Flows	kg/hr	489600.00	2916.67	54166.67	54166.67	54166.67	54166.67	226958.33	226958.33	226958.33	226958.33	226958.33	226958.33	226958.33	133963.07	128086.87	27333.33	515621.05	489600.00	515621.05	515621.05	515621.05	515621.05	515621.05	515621.05	515621.05	155420.20	12600.00
O2	kg/hr	114036	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	20437	114036	114036	20437	20437	20437	20437	20437	20437	0	0
H2	kg/hr	0	0	0	0	0	0	0	0	0	3298	3298	20164	20164	20163	20163	16	0	0	2760	0	0	0	0	0	0	20180	0
N2	kg/hr	375564	0	0	0	0	0	0	0	0	0	0	0	0	0	0	35	375420	375564	375599	375420	375420	375420	375420	375420	375420	35	0
CO	kg/hr	0	0	0	0	0	0	0	0	0	399	399	58581	58581	58581	58581	0	2	0	403	2	2	2	2	2	2	58581	0
CO2	kg/hr	0	0	0	0	0	0	0	0	0	19979	19979	43468	43468	43445	43444	27282	54906	0	4558	54906	54906	54906	54906	54906	54906	70726	0
NH3	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2O	kg/hr	0	2689	0	0	0	0	172792	172792	172792	156179	156179	99527	99527	6555	680	0	64472	0	41	64472	64472	64472	64472	64472	64472	680	0
CH4	kg/hr	0	0	48491	48491	48491	48491	48491	48491	48491	47103	47103	5218	5218	5218	5218	0	0	0	16486	0	0	0	0	0	0	5218	11280
C2H6	kg/hr	0	0	2870	2870	2870	2870	2870	2870	2870	2	2	0	0	0	0	0	0	0	668	0	0	0	0	0	0	0	668
C3H8	kg/hr	0	0	2806	2806	2806	2806	2806	2806	2806	0	0	0	0	0	0	0	0	0	653	0	0	0	0	0	0	0	653
I-C4	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
N-C4	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MEOH	kg/hr	0	141	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	396	0	0	0	0	0	0	0	0
ETOH	kg/hr	0	87	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
DME	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	21	0	0	0	0	0	0	0	0
AR	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2S	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	382	0	0	382	382	382	382	382	382	0	0
NO2	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	1	0	0	1	1	1	1	1	1	0	0
Mole Flows	kmol/hr	16970.30	155.54	3181.66	3181.66	3181.66	3181.66	12773.06	12773.06	12773.06	13709.44	13709.44	18931.30	18931.30	13769.96	13443.82	629.34	18879.34	16970.30	19538.58	18879.34	18879.34	18879.34	18879.34	18879.34	18879.34	14073.15	740.10
O2	kmol/hr	3564	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	639	3564	3564	639	639	639	639	639	639	0	0
H2	kmol/hr	0	0	0	0	0	0	0	0	0	1636	1636	10002	10002	10002	10002	8	0	0	1369	0	0	0	0	0	0	10010	0
N2	kmol/hr	13407	0	0	0	0	0	0	0	0	0	0	0	0	0	0	1	13401	13407	13408	13401	13401	13401	13401	13401	13401	1	0
CO	kmol/hr	0	0	0	0	0	0	0	0	0	14	14	2091	2091	2091	2091	0	0	0	14	0	0	0	0	0	0	2091	0
CO2	kmol/hr	0	0	0	0	0	0	0	0	0	454	454	988	988	987	987	620	1248	0	104	1248	1248	1248	1248	1248	1248	1607	0
NH3	kmol/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2O	kmol/hr	0	149	0	0	0	0	9591	9591	9591	8669	8669	5525	5525	364	38	0	3579	0	2	3579	3579	3579	3579	3579	3579	38	0
CH4	kmol/hr	0	0	3023	3023	3023	3023	3023	3023	3023	2936	2936	325	325	325	325	0	0	0	1028	0	0	0	0	0	0	325	703
C2H6	kmol/hr	0	0	95	95	95	95	95	95	95	0	0	0	0	0	0	0	0	0	22	0	0	0	0	0	0	0	22
C3H8	kmol/hr	0	0	64	64	64	64	64	64	64	0	0	0	0	0	0	0	0	0	15	0	0	0	0	0	0	0	15
I-C4	kmol/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
N-C4	kmol/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MEOH	kmol/hr	0	4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	12	0	0	0	0	0	0	0	0
ETOH	kmol/hr	0	2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
DME	kmol/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AR	kmol/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2S	kmol/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO	kmol/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	13	0	0	13	13	13	13	13	13	0	0
NO2	kmol/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Mass fraction																												
O2		0.232917001	0	0	0	0	0	0	0	0	7.14381E-29	7.14381E-29	8.40567E-19	8.40567E-19	0	0	0	0.039635781	0.232917001	0.221162739	0.039635781	0.039635781	0.039635781	0.039635781	0.039635781	0.039635781	0	0
H2		0	0	0	0	0	0	0	0	0	0.014530024	0.014530024	0.088842445	0.088842445	0.150514158	0.157419226	0.000603392	3.27305E-07	0	0.005353563	3.27305E-07	3.27305E-07	3.27305E-07	3.27305E-07	3.27305E-07	3.27305E-07	0.129840447	0
N2		0.767082999	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0.001289996	0.728093528	0.767082999	0.7284403	0.728093528	0.728093528	0.728093528	0.728093528	0.728093528	0.728093528	0.000226868	0
CO		0	0	0	0	0	0	0	0	0	0.00175679	0.00175679	0.258115616	0.258115616	0.437294376	0.457355983	0	4.46755E-06	0	0.000780708	4.46755E-06	4.46755E-06	4.46755E-06	4.46755E-06	4.46755E-06	4.46755E-06	0.376922015	0
CO2		0	0	0	0	0	0	0	0	0	0.088027257	0.088027257	0.191526237	0.191526237	0.324306315	0.3391757	0.998106612	0.106485534	0	0.008840003	0.106485534	0.106485534	0.106485534	0.106485534	0.106485534	0.106485534	0.455060108	0
NH3		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	2.55585E-13	0	0	2.55585E-13	2.55585E-13	2.55585E-13	2.55585E-13	2.55585E-13	2.55585E-13	0	0
H2O		0	0.921978299	0	0	0	0	0.761336515	0.761336515	0.761336515	0.688139214	0.688139214	0.438524394	0.438524394	0.048934471	0.005311532	0	0.125036818	0	7.96284E-05	0.125036818	0.125036818	0.125036818	0.125036818	0.125036818	0.125036818	0.004377407	0
CH4		0	0	0.895209565	0.895209565	0.895209565	0.895209565	0.213653834	0.213653834	0.213653834	0.20753955	0.20753955	0.02299038	0.02299038	0.038949215	0.040736063	0	2.40964E-27	0	0.031973324	2.40964E-27	2.40964E-27	2.40964E-27	2.40964E-27	2.40964E-27	2.40964E-27	0.033571921	0.895209565
C2H6		0	0	0.052987265	0.052987265	0.052987265	0.052987265	0.012646125	0.012646125	0.012646125	7.14826E-06	7.14826E-06	7.9778E-07	7.9778E-07	1.35155E-06	1.41356E-06	0	0	0	0.001295167	0	0	0	0	0	0	1.16496E-06	0.052987265
C3H8		0	0	0.05180317	0.05180317	0.05180317	0.05180317	0.012363525	0.012363525	0.012363525	1.1301E-09	1.1301E-09	7.42526E-11	7.42526E-11	1.25797E-10	1.31568E-10	0	0	0	0.001265891	0	0	0	0	0	0	1.0843E-10	0.05180317
I-C4		0	0	0	0	0	0	0	0	0	1.10655E-13	1.10655E-13	2.72885E-15	2.72885E-15	0	0	0	0	0	0	0	0	0	0	0	0	0	0
N-C4		0	0	0	0	0	0	0	0	0	1.55434E-13	1.55434E-13	6.2294E-15	6.2294E-15	1.05537E-14	1.10379E-14	0	0	0	0	0	0	0	0	0	0	9.0967E-15	0
MEOH		0	0.048222477	0	0	0	0	0	0	0	1.51985E-08	1.51985E-08	1.31149E-07	1.31149E-07	1.13045E-07	8.21788E-08	0	3.1339E-26	0	0.000767772	3.1339E-26	3.1339E-26	3.1339E-26	3.1339E-26	3.1339E-26	3.1339E-26	6.77262E-08	0
ETOH		0	0.029799224	0	0	0	0	0	0	0	1.95511E-11	1.95511E-11	5.03431E-11	5.03431E-11	4.39652E-11	3.18048E-11	0	0	0	6.33374E-07	0	0	0	0	0	0	2.62114E-11	0
DME		0	0	0	0	0	0	0	0	0	9.80406E-16	9.80406E-16	3.95649E-14	3.95649E-14	6.70192E-14	7.00936E-14	0	0	0	4.02705E-05	0	0	0	0	0	0	5.77664E-14	0
AR		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2S		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0.000741628	0	0	0.000741628	0.000741628	0.000741628	0.000741628	0.000741628	0.000741628	0	0
NO2		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	1.91509E-06	0	0	1.91509E-06	1.91509E-06	1.91509E-06	1.91509E-06	1.91509E-06	1.91509E-06	0	0
Mole fraction																												
O2		0.21	0	0	0	0	0	0	0	0	3.69592E-29	3.69592E-29	3.14923E-19	3.14923E-19	0	0	0	0.033829656	0.21	0.182396279	0.033829656	0.033829656	0.033829656	0.033829656	0.033829656	0.033829656	0	0
H2		0	0	0	0	0	0	0	0	0	0.11932408	0.11932408	0.52834988	0.52834988	0.726382046	0.744003768	0.013	4.43437E-06	0	0.070083529	4.43437E-06	4.43437E-06	4.43437E-06	4.43437E-06	4.43437E-06	4.43437E-06	0.711313983	0
N2		0.79	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0.002	0.709845703	0.79	0.686221956	0.709845703	0.709845703	0.709845703	0.709845703	0.709845703	0.709845703	8.94381E-05	0
CO		0	0	0	0	0	0	0	0	0	0.00103831	0.00103831	0.110474172	0.110474172	0.151882244	0.155566862	0	4.35606E-06	0	0.000735541	4.35606E-06	4.35606E-06	4.35606E-06	4.35606E-06	4.35606E-06	4.35606E-06	0.148610062	0
CO2		0	0	0	0	0	0	0	0	0	0.033112671	0.033112671	0.052172863	0.052172863	0.071689976	0.073427269	0.985	0.066082288	0	0.005300792	0.066082288	0.066082288	0.066082288	0.066082288	0.066082288	0.066082288	0.114191924	0
NH3		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	4.09874E-13	0	0	4.09874E-13	4.09874E-13	4.09874E-13	4.09874E-13	4.09874E-13	4.09874E-13	0	0
H2O		0	0.959650609	0	0	0	0	0.750908298	0.750908298	0.750908298	0.632356129	0.632356129	0.291822335	0.291822335	0.026425688	0.002809059	0	0.189557401	0	0.000116645	0.189557401	0.189557401	0.189557401	0.189557401	0.189557401	0.189557401	0.00268344	0
CH4		0	0	0.95	0.95	0.95	0.95	0.236637116	0.236637116	0.236637116	0.214164866	0.214164866	0.017180383	0.017180383	0.023619573	0.02419257	0	4.10221E-27	0	0.052595237	4.10221E-27	4.10221E-27	4.10221E-27	4.10221E-27	4.10221E-27	4.10221E-27	0.023110702	0.95
C2H6		0	0	0.03	0.03	0.03	0.03	0.007472751	0.007472751	0.007472751	3.93549E-06	3.93549E-06	3.18068E-07	3.18068E-07	4.37277E-07	4.47885E-07	0	0	0	0.001136673	0	0	0	0	0	0	4.27856E-07	0.03
C3H8		0	0	0.02	0.02	0.02	0.02	0.004981834	0.004981834	0.004981834	4.24266E-10	4.24266E-10	2.01871E-11	2.01871E-11	2.77535E-11	2.84268E-11	0	0	0	0.000757582	0	0	0	0	0	0	2.71556E-11	0.02
I-C4		0	0	0	0	0	0	0	0	0	3.15172E-14	3.15172E-14	5.62852E-16	5.62852E-16	0	0	0	0	0	0	0	0	0	0	0	0	0	0
N-C4		0	0	0	0	0	0	0	0	0	4.42711E-14	4.42711E-14	1.28487E-15	1.28487E-15	1.76648E-15	1.80933E-15	0	0	0	0	0	0	0	0	0	0	1.72842E-15	0
MEOH		0	0.028220272	0	0	0	0	0	0	0	7.85246E-09	7.85246E-09	4.90693E-08	4.90693E-08	3.43227E-08	2.44354E-08	0	2.67121E-26	0	0.000632336	2.67121E-26	2.67121E-26	2.67121E-26	2.67121E-26	2.67121E-26	2.67121E-26	2.33427E-08	0
ETOH		0	0.012129119	0	0	0	0	0	0	0	7.0257E-12	7.0257E-12	1.31008E-11	1.31008E-11	9.28436E-12	6.57757E-12	0	0	0	3.62818E-07	0	0	0	0	0	0	6.28342E-12	0
DME		0	0	0	0	0	0	0	0	0	3.52308E-16	3.52308E-16	1.02959E-14	1.02959E-14	1.41528E-14	1.44961E-14	0	0	0	2.30683E-05	0	0	0	0	0	0	1.38478E-14	0
AR		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2S		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0.000675025	0	0	0.000675025	0.000675025	0.000675025	0.000675025	0.000675025	0.000675025	0	0
NO2		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	1.1369E-06	0	0	1.1369E-06	1.1369E-06	1.1369E-06	1.1369E-06	1.1369E-06	1.1369E-06	0	0
Enthalpy, mixture	MJ/kmol	-0.01	-280.08	-75.71	-58.56	-58.56	-58.56	-192.46	-186.30	-177.15	-165.05	-145.38	-75.89	-100.35	-53.96	-48.29	-388.21	-69.09	9.68	1.95	-27.45	-27.45	-41.74	-49.13	-55.18	-60.39	-63.49	-75.42

Stream Name	**Units**	**H2O-OUT**	**L13**	**L14A**	**L14B**	**L15**	**L15A**	**L15B**	**L16**	**L17**	**L17A**	**L18A**	**L18B**	**LIGHTEND**	**M1**	**M2**	**M3**	**M4**	**M5**	**M6**	**M7**	**M8**	**M9**	**M10**	**M11**	**M12**	**M13**	**M14**	**M15**	**M16**	**M17**
From		FD1	V201	B2	B2	C201	K202	C201	E206	K204	V202	B7	C-301	B10	K201-S1	E203	V208	K201-S2	E202-C	R201-B	R201-A	R201-B	E210	E202-H	E209	E207	V201	K203	V202	T-351	P301
To		HX1A	B2	K202	C201	E206	B5	V202	K204	B7	B7	B10	B10	MX1	E203	V208	K201-S2	B5	R201-B	R201-A	R201-B	E210	E202-H	E209	E207	V201	K203	V202	T-351	P301	E301-C
Temperature	C	34.99959303	37.77777778	37.77777778	37.77777778	39.33688568	40.52087267	39.92198778	220	12.67042988	37.9139599	13.71248062	45.5	14.1155872	217.9765015	37.77777778	37.77777778	103.6504294	190	262.3085583	265.0163593	208.6285246	221	78.99017066	73.88888889	37.77777778	37.77777778	38.19589938	37.9139599	37.68329998	37.69791516
Pressure	bar	16.474	89.83843645	89.83843645	89.83843645	81.777	92	81.777	81.727	2.5	2.392201459	2.392201459	1.357987865	1.357987865	55	55	55	92	91.65526214	91.65526214	91.33759609	91.21738607	90.87265004	90.52791218	90.18317431	89.83843645	89.83843645	2.392201459	2.392201459	1.01	1.840620875
Mass Vapor Fraction		0	1	1	1	1	1	0	1	1	1	1	1	1	1	0.997525027	1	1	1	1	1	1	1.000003548	1	0.772764513	0.723257983	0	0	0	0	0
Mass Flows	kg/hr	5876.20	376632.23	365709.90	10922.33	10750.32	365709.90	1422.02	10750.32	10750.32	1826.93	12577.25	843.80	13421.05	155420.20	155420.20	155035.54	155035.54	520745.44	520745.44	520745.44	520745.44	520743.97	520743.97	520743.97	520743.97	144111.74	144111.74	143706.82	143276.60	143276.60
O2	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2	kg/hr	0	94395	91657	2737	2737	91657	0	2737	2737	23	2760	0	2760	20180	20180	20180	20180	111837	111837	95990	94417	94418	94418	94418	94418	23	23	0	0	0
N2	kg/hr	0	1204	1169	35	35	1169	0	35	35	0	35	0	35	35	35	35	35	1204	1204	1204	1204	1204	1204	1204	1204	0	0	0	0	0
CO	kg/hr	0	13709	13311	398	398	13311	0	398	398	5	403	0	403	58581	58581	58581	58581	71892	71892	26271	13714	13714	13714	13714	13714	5	5	0	0	0
CO2	kg/hr	1	90072	87460	2612	2611	87460	1	2611	2611	1321	3931	627	4558	70726	70726	70725	70725	158185	158185	90649	92361	92359	92359	92359	92359	2287	2287	967	627	627
NH3	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2O	kg/hr	5875	382	371	11	29	371	1232	29	29	12	40	1	41	680	680	296	296	667	667	28364	27696	27696	27696	27696	27696	27314	27314	28535	28530	28530
CH4	kg/hr	0	169916	164989	4928	4928	164989	0	4928	4928	278	5205	1	5206	5218	5218	5218	5218	170207	170207	170207	170207	170207	170207	170207	170207	290	290	13	1	1
C2H6	kg/hr	0	5	5	0	0	5	0	0	0	0	0	0	0	0	0	0	0	5	5	5	5	5	5	5	5	0	0	0	0	0
C3H8	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
I-C4	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
N-C4	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MEOH	kg/hr	0	6731	6536	195	7	6536	188	7	7	184	191	205	396	0	0	0	0	6536	6536	107714	120714	120714	120714	120714	120714	113983	113983	113987	113916	113916
ETOH	kg/hr	0	9	9	0	0	9	0	0	0	0	0	0	0	0	0	0	0	9	9	129	202	202	202	202	202	193	193	193	193	193
DME	kg/hr	0	209	203	6	6	203	0	6	6	5	11	10	21	0	0	0	0	203	203	213	225	225	225	225	225	16	16	12	10	10
AR	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2S	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO2	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Mole Flows	kmol/hr	326.14	60232.15	58485.41	1746.73	1741.80	58485.41	74.32	1741.80	1741.80	65.39	1807.19	20.98	1828.17	14073.15	14073.15	14051.81	14051.81	72537.23	72537.23	66210.61	65391.80	65391.84	65391.84	65391.84	65391.84	5159.70	5159.70	5168.62	5157.58	5157.58
O2	kmol/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2	kmol/hr	0	46825	45468	1358	1358	45468	0	1358	1358	11	1369	0	1369	10010	10010	10010	10010	55478	55478	47617	46837	46837	46837	46837	46837	11	11	0	0	0
N2	kmol/hr	0	43	42	1	1	42	0	1	1	0	1	0	1	1	1	1	1	43	43	43	43	43	43	43	43	0	0	0	0	0
CO	kmol/hr	0	489	475	14	14	475	0	14	14	0	14	0	14	2091	2091	2091	2091	2567	2567	938	490	490	490	490	490	0	0	0	0	0
CO2	kmol/hr	0	2047	1987	59	59	1987	0	59	59	30	89	14	104	1607	1607	1607	1607	3594	3594	2060	2099	2099	2099	2099	2099	52	52	22	14	14
NH3	kmol/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2O	kmol/hr	326	21	21	1	2	21	68	2	2	1	2	0	2	38	38	16	16	37	37	1574	1537	1537	1537	1537	1537	1516	1516	1584	1584	1584
CH4	kmol/hr	0	10591	10284	307	307	10284	0	307	307	17	324	0	325	325	325	325	325	10610	10610	10610	10610	10610	10610	10610	10610	18	18	1	0	0
C2H6	kmol/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C3H8	kmol/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
I-C4	kmol/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
N-C4	kmol/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MEOH	kmol/hr	0	210	204	6	0	204	6	0	0	6	6	6	12	0	0	0	0	204	204	3362	3767	3767	3767	3767	3767	3557	3557	3557	3555	3555
ETOH	kmol/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	3	4	4	4	4	4	4	4	4	4	4
DME	kmol/hr	0	5	4	0	0	4	0	0	0	0	0	0	0	0	0	0	0	4	4	5	5	5	5	5	5	0	0	0	0	0
AR	kmol/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2S	kmol/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO	kmol/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO2	kmol/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Mass fraction																															
O2		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2		5.52163E-07	0.250627814	0.250627814	0.250627814	0.254637167	0.250627814	7.10305E-06	0.254637167	0.254637167	0.012576888	0.219476245	2.31083E-06	0.205677615	0.129840447	0.129840447	0.130162591	0.130162591	0.21476309	0.21476309	0.184331263	0.181312107	0.181312932	0.181312932	0.181312932	0.181312932	0.000160311	0.000160311	9.43836E-07	1.36092E-08	1.36092E-08
N2		0	0.003196584	0.003196584	0.003196584	0.003247728	0.003196584	3.14867E-08	0.003247728	0.003247728	0.000220712	0.002808033	8.46334E-08	0.002631493	0.000226868	0.000226868	0.000227431	0.000227431	0.002312612	0.002312612	0.002312612	0.002312612	0.002312735	0.002312735	0.002312735	0.002312735	2.82162E-06	2.82162E-06	2.39846E-08	4.98432E-10	4.98432E-10
CO		4.03019E-07	0.036398214	0.036398214	0.036398214	0.036980572	0.036398214	3.75452E-07	0.036980572	0.036980572	0.002734244	0.032006057	1.25174E-06	0.02999387	0.376922015	0.376922015	0.377857198	0.377857198	0.138056865	0.138056865	0.050449514	0.026335054	0.02633498	0.02633498	0.02633498	0.02633498	3.49829E-05	3.49829E-05	3.24924E-07	7.37186E-09	7.37186E-09
CO2		0.000190082	0.239151952	0.239151952	0.239151952	0.242866113	0.239151952	0.000850546	0.242866113	0.242866113	0.72285583	0.312587923	0.742586987	0.339622561	0.455060108	0.455060108	0.456186911	0.456186911	0.303767271	0.303767271	0.174074698	0.177362751	0.177360017	0.177360017	0.177360017	0.177360017	0.015868418	0.015868418	0.006731943	0.004373324	0.004373324
NH3		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2O		0.99980789	0.001014321	0.001014321	0.001014321	0.002662345	0.001014321	0.86669765	0.002662345	0.002662345	0.006375567	0.003201716	0.000935377	0.003059228	0.004377407	0.004377407	0.001909422	0.001909422	0.001280809	0.001280809	0.054468068	0.053185724	0.053185812	0.053185812	0.053185812	0.053185812	0.189534634	0.189534634	0.198563821	0.199128433	0.199128433
CH4		2.86327E-07	0.451146798	0.451146798	0.451146798	0.458361363	0.451146798	3.19691E-05	0.458361363	0.458361363	0.151975401	0.413856693	0.001539424	0.387933735	0.033571921	0.033571921	0.033655215	0.033655215	0.326851837	0.326851837	0.326851837	0.326851837	0.326853019	0.326853019	0.326853019	0.326853019	0.002014498	0.002014498	8.84407E-05	9.06614E-06	9.06614E-06
C2H6		1.19421E-11	1.28296E-05	1.28296E-05	1.28296E-05	1.30347E-05	1.28296E-05	1.43114E-09	1.30347E-05	1.30347E-05	1.79458E-05	1.3748E-05	3.70337E-06	1.31165E-05	1.16496E-06	1.16496E-06	1.16785E-06	1.16785E-06	9.35767E-06	9.35767E-06	9.35767E-06	9.35767E-06	9.3577E-06	9.3577E-06	9.3577E-06	9.3577E-06	2.84053E-07	2.84053E-07	5.67241E-08	2.18103E-08	2.18103E-08
C3H8		1.55796E-16	1.04238E-09	1.04238E-09	1.04238E-09	1.05905E-09	1.04238E-09	2.79295E-14	1.05905E-09	1.05905E-09	2.08978E-09	1.20878E-09	1.01617E-09	1.19667E-09	1.0843E-10	1.0843E-10	1.08699E-10	1.08699E-10	7.64405E-10	7.64405E-10	7.64405E-10	7.64405E-10	7.64409E-10	7.64409E-10	7.64409E-10	7.64409E-10	3.79399E-11	3.79399E-11	1.14798E-11	5.98457E-12	5.98457E-12
I-C4		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
N-C4		1.60119E-21	0	0	0	0	0	0	0	0	0	0	0	0	9.0967E-15	9.0967E-15	9.11927E-15	9.11927E-15	2.71498E-15	2.71498E-15	2.71498E-15	2.71498E-15	2.71498E-15	2.71498E-15	2.71498E-15	2.71498E-15	0	0	0	0	0
MEOH		7.85852E-07	0.017872549	0.017872549	0.017872549	0.000667837	0.017872549	0.132228146	0.000667837	0.000667837	0.100569417	0.015179229	0.242908803	0.029496903	6.77262E-08	6.77262E-08	6.38376E-08	6.38376E-08	0.01255158	0.01255158	0.206845483	0.231810726	0.231811293	0.231811293	0.231811293	0.231811293	0.790934576	0.790934576	0.793193059	0.795075246	0.795075246
ETOH		3.09032E-10	2.49825E-05	2.49825E-05	2.49825E-05	1.09197E-06	2.49825E-05	0.000183633	1.09197E-06	1.09197E-06	0.000132684	2.02067E-05	8.58453E-05	2.43335E-05	2.62114E-11	2.62114E-11	2.47503E-11	2.47503E-11	1.75448E-05	1.75448E-05	0.000248074	0.000387867	0.000387887	0.000387887	0.000387887	0.000387887	0.001336329	0.001336329	0.001340224	0.00134359	0.00134359
DME		5.61037E-18	0.000553956	0.000553956	0.000553956	0.000562748	0.000553956	5.44097E-07	0.000562748	0.000562748	0.002541308	0.000850147	0.011936213	0.001547143	5.77664E-14	5.77664E-14	5.79097E-14	5.79097E-14	0.000389033	0.000389033	0.000409092	0.000431964	0.000431965	0.000431965	0.000431965	0.000431965	0.000113146	0.000113146	8.1163E-05	7.02973E-05	7.02973E-05
AR		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2S		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO2		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Mole fraction																															
O2		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2		4.93504E-06	0.777416455	0.777416455	0.777416455	0.779616444	0.777416455	6.74175E-05	0.779616444	0.779616444	0.174302175	0.757713352	4.61029E-05	0.749018193	0.711313983	0.711313983	0.712394243	0.712394243	0.76482044	0.76482044	0.719171021	0.716249518	0.716250235	0.716250235	0.716250235	0.716250235	0.002221126	0.002221126	1.30177E-05	1.87542E-07	1.87542E-07
N2		0	0.000713523	0.000713523	0.000713523	0.000715544	0.000713523	2.15057E-08	0.000715544	0.000715544	0.000220117	0.000697617	1.21506E-07	0.000689613	8.94381E-05	8.94381E-05	8.95739E-05	8.95739E-05	0.000592653	0.000592653	0.000649282	0.000657412	0.000657445	0.000657445	0.000657445	0.000657445	2.81324E-06	2.81324E-06	2.3805E-08	4.94275E-10	4.94275E-10
CO		2.59236E-07	0.008125496	0.008125496	0.008125496	0.00814851	0.008125496	2.56465E-07	0.00814851	0.00814851	0.002727171	0.00795234	1.79729E-06	0.007861098	0.148610062	0.148610062	0.148835757	0.148835757	0.035383697	0.035383697	0.014165618	0.00748716	0.007487112	0.007487112	0.007487112	0.007487112	3.48828E-05	3.48828E-05	3.22527E-07	7.31118E-09	7.31118E-09
CO2		7.78179E-05	0.033979241	0.033979241	0.033979241	0.034059746	0.033979241	0.000369778	0.034059746	0.034059746	0.458877313	0.049431626	0.678612683	0.056652249	0.114191924	0.114191924	0.114364785	0.114364785	0.04955142	0.04955142	0.031108861	0.032093363	0.032092755	0.032092755	0.032092755	0.032092755	0.010070697	0.010070697	0.004252981	0.002760525	0.002760525
NH3		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2O		0.999916224	0.000352065	0.000352065	0.000352065	0.00091211	0.000352065	0.920490171	0.00091211	0.00091211	0.009887172	0.00123687	0.002088188	0.00124664	0.00268344	0.00268344	0.001169391	0.001169391	0.000510397	0.000510397	0.023779286	0.023510197	0.023510153	0.023510153	0.023510153	0.023510153	0.29384807	0.29384807	0.306451351	0.307058907	0.307058907
CH4		3.21567E-07	0.175844153	0.175844153	0.175844153	0.176340793	0.175844153	3.8128E-05	0.176340793	0.176340793	0.26466012	0.179536598	0.003859252	0.177520485	0.023110702	0.023110702	0.023145799	0.023145799	0.146263634	0.146263634	0.160239543	0.162246022	0.162246033	0.162246033	0.162246033	0.162246033	0.003507221	0.003507221	0.000153276	1.5699E-05	1.5699E-05
C2H6		7.15553E-12	2.66792E-06	2.66792E-06	2.66792E-06	2.67544E-06	2.66792E-06	9.10641E-10	2.67544E-06	2.67544E-06	1.66736E-05	3.18196E-06	4.95328E-06	3.20229E-06	4.27856E-07	4.27856E-07	4.28506E-07	4.28506E-07	2.23411E-06	2.23411E-06	2.44758E-06	2.47823E-06	2.47823E-06	2.47823E-06	2.47823E-06	2.47823E-06	2.63843E-07	2.63843E-07	5.24495E-08	2.01494E-08	2.01494E-08
C3H8		6.36561E-17	1.47812E-10	1.47812E-10	1.47812E-10	1.48231E-10	1.47812E-10	1.21186E-14	1.48231E-10	1.48231E-10	1.32401E-09	1.90776E-10	9.26805E-10	1.99223E-10	2.71556E-11	2.71556E-11	2.71968E-11	2.71968E-11	1.24447E-10	1.24447E-10	1.36338E-10	1.38045E-10	1.38045E-10	1.38045E-10	1.38045E-10	1.38045E-10	2.40307E-11	2.40307E-11	7.23825E-12	3.77014E-12	3.77014E-12
I-C4		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
N-C4		4.96342E-22	0	0	0	0	0	0	0	0	0	0	0	0	1.72842E-15	1.72842E-15	1.73104E-15	1.73104E-15	3.35335E-16	3.35335E-16	3.67378E-16	3.71978E-16	3.71978E-16	3.71978E-16	3.71978E-16	3.71978E-16	0	0	0	0	0
MEOH		4.41883E-07	0.003487818	0.003487818	0.003487818	0.000128639	0.003487818	0.078957734	0.000128639	0.000128639	0.087687652	0.003296932	0.304891629	0.006758101	2.33427E-08	2.33427E-08	2.19813E-08	2.19813E-08	0.002812167	0.002812167	0.050771748	0.057612133	0.05761207	0.05761207	0.05761207	0.05761207	0.689436154	0.689436154	0.688271136	0.689312074	0.689312074
ETOH		1.2086E-10	3.39091E-06	3.39091E-06	3.39091E-06	1.46294E-07	3.39091E-06	7.62665E-05	1.46294E-07	1.46294E-07	8.04647E-05	3.05259E-06	7.49431E-05	3.87762E-06	6.28342E-12	6.28342E-12	5.92749E-12	5.92749E-12	2.73403E-06	2.73403E-06	4.23515E-05	6.70465E-05	6.70497E-05	6.70497E-05	6.70497E-05	6.70497E-05	0.000810176	0.000810176	0.000808855	0.00081019	0.00081019
DME		2.19417E-18	7.51891E-05	7.51891E-05	7.51891E-05	7.53925E-05	7.51891E-05	2.25975E-07	7.53925E-05	7.53925E-05	0.001541141	0.00012843	0.010420329	0.000246542	1.38478E-14	1.38478E-14	1.38689E-14	1.38689E-14	6.06236E-05	6.06236E-05	6.98409E-05	7.46691E-05	7.4669E-05	7.4669E-05	7.4669E-05	7.4669E-05	6.85972E-05	6.85972E-05	4.89837E-05	4.23896E-05	4.23896E-05
AR		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2S		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO2		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Enthalpy, mixture	MJ/kmol	-287.97	-27.84	-27.84	-27.84	-27.31	-27.75	-283.77	-21.47	-28.08	-220.51	-35.04	-330.30	-38.43	-57.75	-63.49	-63.15	-61.09	-29.78	-27.39	-33.80	-36.88	-36.44	-41.36	-43.93	-45.77	-255.10	-255.31	-256.16	-256.01	-256.01

Stream Name	**Units**	**M18**	**M19**	**M20**	**M21**	**MEOH**	**NG**	**S1**	**S2**	**S3**	**S4**	**S6**	**S24**	**STEAM**	**V16**	**W1**	**W2**	**W3**	**W21**	**W22B**	**W23**	**W27**	**WATER**	**WATER1**	**WATER3**
From		E301-C	C-301	P302	C302	E301-H		HXVN	HX1	FDVN	FDVN	B4	B5	HX3	T-351	MX2	HX8B	SATURATR	C302	V208	B9				HX2
To		C-301	P302	C302	E301-H		HX7B	FDVN	HX2	HX9			E202-C			HX8B	SATURATR			B9		C201	MX2	HX1	HX3
Temperature	C	54.44444444	84.11266765	84.12165238	72.66345351	50.35115058	25	80	337.8247279	79.99878607	79.99878607	115.6389551	52.69559345	540	37.68329998	65.7000412	316.1649631		117.1766204	37.77777778	37.77777778	25	65.7	129	337.8247279
Pressure	bar	1.640620875	1.70272573	2.047463594	1.357987865	1.157987865	16.5	16.48	145	16.478	16.478	1	92	140	1.01	109.998	109.996		1.70272573	55	55	81.777	110	150	145
Mass Vapor Fraction		0	0	0	0	0	1	0.590252113	0.0001	1	0	1	1	1	1	0	1		0	0	0	0	0	0	1
Mass Flows	kg/hr	143276.60	142432.80	142432.80	113676.98	113676.98	54166.67	226958.33	34398.00	133963.07	92995.26	515621.05	520745.44	34398.00	430.22	172791.67	172791.67		25839.15	384.66	384.66	1250.00	172791.67	34398.00	34398.00
O2	kg/hr	0	0	0	0	0	0	0	0	0	0	20437	0	0	0	0	0		0	0	0	0	0	0	0
H2	kg/hr	0	0	0	0	0	0	20164	0	20163	0	0	111837	0	0	0	0		0	0	0	0	0	0	0
N2	kg/hr	0	0	0	0	0	0	0	0	0	0	375420	1204	0	0	0	0		0	0	0	0	0	0	0
CO	kg/hr	0	0	0	0	0	0	58581	0	58581	0	2	71892	0	0	0	0		0	0	0	0	0	0	0
CO2	kg/hr	627	0	0	0	0	0	43468	0	43445	23	54906	158185	0	341	0	0		0	0	0	0	0	0	0
NH3	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
H2O	kg/hr	28530	28530	28530	1	1	0	99527	34398	6555	92971	64472	667	34398	5	172792	172792		25839	384	384	1250	172792	34398	34398
CH4	kg/hr	1	0	0	0	0	48491	5218	0	5218	0	0	170207	0	11	0	0		0	0	0	0	0	0	0
C2H6	kg/hr	0	0	0	0	0	2870	0	0	0	0	0	5	0	0	0	0		0	0	0	0	0	0	0
C3H8	kg/hr	0	0	0	0	0	2806	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
I-C4	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
N-C4	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
MEOH	kg/hr	113916	113711	113711	113570	113570	0	0	0	0	0	0	6536	0	72	0	0		0	0	0	0	0	0	0
ETOH	kg/hr	193	192	192	106	106	0	0	0	0	0	0	9	0	0	0	0		0	0	0	0	0	0	0
DME	kg/hr	10	0	0	0	0	0	0	0	0	0	0	203	0	2	0	0		0	0	0	0	0	0	0
AR	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
H2S	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
NO	kg/hr	0	0	0	0	0	0	0	0	0	0	382	0	0	0	0	0		0	0	0	0	0	0	0
NO2	kg/hr	0	0	0	0	0	0	0	0	0	0	1	0	0	0	0	0		0	0	0	0	0	0	0
Mole Flows	kmol/hr	5157.58	5136.60	5136.60	3546.76	3546.76	3181.66	18931.30	1909.38	13769.96	5161.34	18879.34	72537.23	1909.38	11.05	9591.40	9591.40	0.00	1434.29	21.34	21.34	69.39	9591.40	1909.38	1909.38
O2	kmol/hr	0	0	0	0	0	0	0	0	0	0	639	0	0	0	0	0	0	0	0	0	0	0	0	0
H2	kmol/hr	0	0	0	0	0	0	10002	0	10002	0	0	55478	0	0	0	0	0	0	0	0	0	0	0	0
N2	kmol/hr	0	0	0	0	0	0	0	0	0	0	13401	43	0	0	0	0	0	0	0	0	0	0	0	0
CO	kmol/hr	0	0	0	0	0	0	2091	0	2091	0	0	2567	0	0	0	0	0	0	0	0	0	0	0	0
CO2	kmol/hr	14	0	0	0	0	0	988	0	987	1	1248	3594	0	8	0	0	0	0	0	0	0	0	0	0
NH3	kmol/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2O	kmol/hr	1584	1584	1584	0	0	0	5525	1909	364	5161	3579	37	1909	0	9591	9591	0	1434	21	21	69	9591	1909	1909
CH4	kmol/hr	0	0	0	0	0	3023	325	0	325	0	0	10610	0	1	0	0	0	0	0	0	0	0	0	0
C2H6	kmol/hr	0	0	0	0	0	95	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C3H8	kmol/hr	0	0	0	0	0	64	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
I-C4	kmol/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
N-C4	kmol/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MEOH	kmol/hr	3555	3549	3549	3544	3544	0	0	0	0	0	0	204	0	2	0	0	0	0	0	0	0	0	0	0
ETOH	kmol/hr	4	4	4	2	2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
DME	kmol/hr	0	0	0	0	0	0	0	0	0	0	0	4	0	0	0	0	0	0	0	0	0	0	0	0
AR	kmol/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2S	kmol/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO	kmol/hr	0	0	0	0	0	0	0	0	0	0	13	0	0	0	0	0	0	0	0	0	0	0	0	0
NO2	kmol/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Mass fraction																									
O2		0	0	0	0	0	0	8.40567E-19	0	0	0	0.039635781	0	0	0	0	0	0	0	0	0	0	0	0	0
H2		1.36092E-08	1.79695E-28	1.79695E-28	0	0	0	0.088842445	0	0.150514158	2.08666E-06	3.27305E-07	0.21476309	0	0.000310738	0	0	0	0	1.93327E-06	1.93327E-06	0	0	0	0
N2		4.98432E-10	6.5745E-29	6.5745E-29	0	0	0	0	0	0	0	0.728093528	0.002312612	0	7.84561E-06	0	0	0	0	8.0303E-10	8.0303E-10	0	0	0	0
CO		7.37186E-09	1.59076E-27	1.59076E-27	0	0	0	0.258115616	0	0.437294376	2.05726E-06	4.46755E-06	0.138056865	0	0.00010608	0	0	0	0	1.38244E-06	1.38244E-06	0	0	0	0
CO2		0.004373324	1.58422E-11	1.58422E-11	1.98496E-11	1.98496E-11	0	0.191526237	0	0.324306315	0.000251677	0.106485534	0.303767271	0	0.792223753	0	0	0	0	0.000908406	0.000908406	0	0	0	0
NH3		0	0	0	0	0	0	0	0	0	0	2.55585E-13	0	0	0	0	0	0	0	0	0	0	0	0	0
H2O		0.199128433	0.200302568	0.200302568	1.25416E-05	1.25416E-05	0	0.438524394	1	0.048934471	0.999742928	0.125036818	0.001280809	1	0.010530546	1	1	0	0.999999	0.999085704	0.999085704	1	1	1	1
CH4		9.06614E-06	1.98595E-20	1.98595E-20	0	0	0.895209565	0.02299038	0	0.038949215	1.09362E-06	2.40964E-27	0.326851837	0	0.026522588	0	0	0	0	9.37936E-07	9.37936E-07	0	0	0	0
C2H6		2.18103E-08	1.14791E-18	1.14791E-18	0	0	0.052987265	7.9778E-07	0	1.35155E-06	5.1339E-11	0	9.35767E-06	0	1.16841E-05	0	0	0	0	3.71322E-11	3.71322E-11	0	0	0	0
C3H8		5.98457E-12	1.83399E-20	1.83399E-20	0	0	0.05180317	7.42526E-11	0	1.25797E-10	1.20395E-15	0	7.64405E-10	0	1.84157E-09	0	0	0	0	4.80678E-16	4.80678E-16	0	0	0	0
I-C4		0	0	0	0	0	0	2.72885E-15	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
N-C4		0	0	0	0	0	0	6.2294E-15	0	1.05537E-14	2.29561E-20	0	2.71498E-15	0	0	0	0	0	0	4.97688E-21	4.97688E-21	0	0	0	0
MEOH		0.795075246	0.798346389	0.798346389	0.999059308	0.999059308	0	1.31149E-07	0	1.13045E-07	1.57229E-07	3.1339E-26	0.01255158	0	0.166367661	0	0	0	6.51656E-07	1.63501E-06	1.63501E-06	0	0	0	0
ETOH		0.00134359	0.001351041	0.001351041	0.000928149	0.000928149	0	5.03431E-11	0	4.39652E-11	5.95307E-11	0	1.75448E-05	0	0.000219337	0	0	0	3.48344E-07	6.15067E-10	6.15067E-10	0	0	0	0
DME		7.02973E-05	1.27052E-09	1.27052E-09	1.59191E-09	1.59191E-09	0	3.95649E-14	0	6.70192E-14	1.58551E-17	0	0.000389033	0	0.003699767	0	0	0	0	1.60772E-17	1.60772E-17	0	0	0	0
AR		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2S		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO		0	0	0	0	0	0	0	0	0	0	0.000741628	0	0	0	0	0	0	0	0	0	0	0	0	0
NO2		0	0	0	0	0	0	0	0	0	0	1.91509E-06	0	0	0	0	0	0	0	0	0	0	0	0	0
Mole fraction																									
O2		0	0	0	0	0	0	3.14923E-19	0	0	0	0.033829656	0	0	0	0	0	0	0	0	0	0	0	0	0
H2		1.87542E-07	2.47176E-27	2.47176E-27	0	0	0	0.52834988	0	0.726382046	1.86503E-05	4.43437E-06	0.76482044	0	0.006003776	0	0	0	0	1.7286E-05	1.7286E-05	0	0	0	0
N2		4.94275E-10	6.50774E-29	6.50774E-29	0	0	0	0	0	0	0	0.709845703	0.000592653	0	1.09082E-05	0	0	0	0	5.16693E-10	5.16693E-10	0	0	0	0
CO		7.31118E-09	1.57478E-27	1.57478E-27	0	0	0	0.110474172	0	0.151882244	1.32333E-06	4.35606E-06	0.035383697	0	0.000147505	0	0	0	0	8.89601E-07	8.89601E-07	0	0	0	0
CO2		0.002760525	9.98159E-12	9.98159E-12	1.44558E-11	1.44558E-11	0	0.052172863	0	0.071689976	0.000103037	0.066082288	0.04955142	0	0.701121909	0	0	0	0	0.000372047	0.000372047	0	0	0	0
NH3		0	0	0	0	0	0	0	0	0	0	4.09874E-13	0	0	0	0	0	0	0	0	0	0	0	0	0
H2O		0.307058907	0.308304563	0.308304563	2.23127E-05	2.23127E-05	0	0.291822335	1	0.026425688	0.999875673	0.189557401	0.000510397	1	0.022766954	1	1	0	0.999999497	0.999607803	0.999607803	1	1	1	1
CH4		1.5699E-05	3.4326E-20	3.4326E-20	0	0	0.95	0.017180383	0	0.023619573	1.22825E-06	4.10221E-27	0.146263634	0	0.064391995	0	0	0	0	1.05381E-06	1.05381E-06	0	0	0	0
C2H6		2.01494E-08	1.05855E-18	1.05855E-18	0	0	0.03	3.18068E-07	0	4.37277E-07	3.07622E-11	0	2.23411E-06	0	1.51343E-05	0	0	0	0	2.22582E-11	2.22582E-11	0	0	0	0
C3H8		3.77014E-12	1.15326E-20	1.15326E-20	0	0	0.02	2.01871E-11	0	2.77535E-11	4.9193E-16	0	1.24447E-10	0	1.62659E-09	0	0	0	0	1.9648E-16	1.9648E-16	0	0	0	0
I-C4		0	0	0	0	0	0	5.62852E-16	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
N-C4		0	0	0	0	0	0	1.28487E-15	0	1.76648E-15	7.11616E-21	0	3.35335E-16	0	0	0	0	0	0	1.54338E-21	1.54338E-21	0	0	0	0
MEOH		0.689312074	0.690882243	0.690882243	0.99933196	0.99933196	0	4.90693E-08	0	3.43227E-08	8.84118E-08	2.67121E-26	0.002812167	0	0.202228425	0	0	0	3.66385E-07	9.19743E-07	9.19743E-07	0	0	0	0
ETOH		0.00081019	0.000813193	0.000813193	0.000645727	0.000645727	0	1.31008E-11	0	9.28436E-12	2.32826E-11	0	2.73403E-06	0	0.000185438	0	0	0	1.3622E-07	2.40647E-10	2.40647E-10	0	0	0	0
DME		4.23896E-05	7.64727E-10	7.64727E-10	1.10751E-09	1.10751E-09	0	1.02959E-14	0	1.41528E-14	6.20094E-18	0	6.06236E-05	0	0.003127954	0	0	0	0	6.29028E-18	6.29028E-18	0	0	0	0
AR		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2S		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO		0	0	0	0	0	0	0	0	0	0	0.000675025	0	0	0	0	0	0	0	0	0	0	0	0	0
NO2		0	0	0	0	0	0	0	0	0	0	1.1369E-06	0	0	0	0	0	0	0	0	0	0	0	0	0
Enthalpy, mixture	MJ/kmol	-254.10	-250.27	-250.27	-236.87	-239.64	-75.71	-114.91	-256.64	-51.51	-284.07	-69.01	-34.21	-226.09	-327.04	-285.14	-236.87		-280.83	-287.68	-287.68	-288.71	-285.14	-279.56	-237.98

Table B33. Heat balance for SMR methanol model.

Stream Name	H1	H2	H4	H5	H6	H7	H8	Q301A	Q301B	QE202A	QE202B	S5	S7	S8
QCALC MW	-45.6428709	-154.532206	-74.9268135	-32.456924	-21.8372328	-15.1594531	-128.595832	-2.73051039	-2.73051035	-89.2843715	-89.2843715	-12.1579325	-6.30451515	-9.89725984
TBEGIN C	25		478.481007	400	254.556172	25	65.7000412	37.6979152		52.6955935		129	337.824728	337.824728
TEND C	350		900	600	400	381.4	316.164963	54.4444444		190		337.824728	540	337.824728

B-3. Energy Integration and Heat Recovery System

Recovery of the available heat is illustrated by the PFD in Figure B3. Starting from the left-hand side at the syngas inlet, we see MP steam superheater followed by superheater, evaporator and preheater of a steam for injection into the NG for reforming. Between two sections of preheater is a LP process steam loop for heating reboiler C-301. This injection steam uses recycled water of limited purity and is therefore kept separate from the rest of the system. To assist with power production, it is generated at higher pressure and superheated, letdown turbine then produces additional 13 MWe of electricity. Methanol reactor cooling generates saturated MP steam, part of which is used for heater E206, and the rest is after superheating admitted to the backpressure turbine and afterwards used for heating reboiler C-302. Finally, to provide all necessary heat for C-302 reboiler and get balanced power production, HP steam is produced in the SMR furnace using heat exchanger placed at suitable locations besides the methane conversion heat exchangers, requiring additional 27.5 MWth from natural gas combustion.

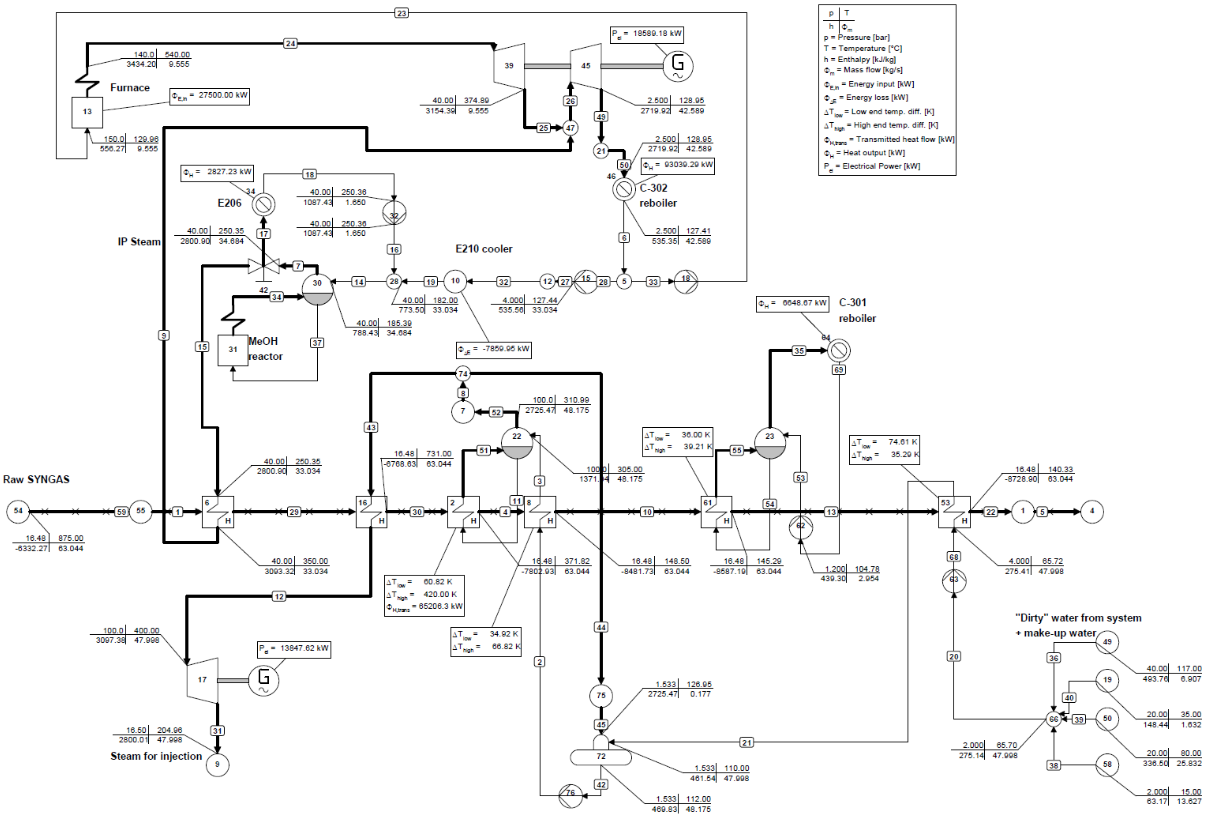
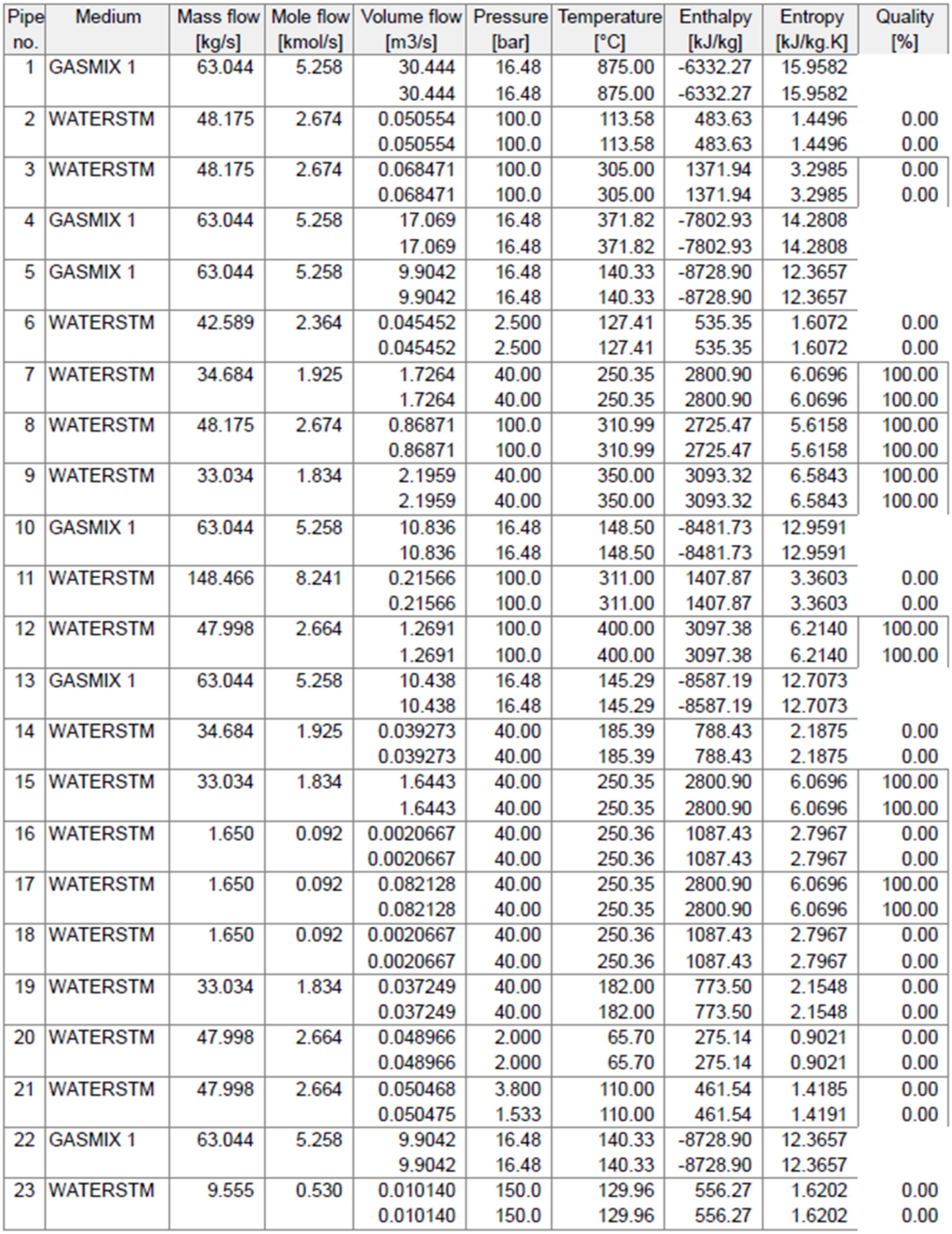


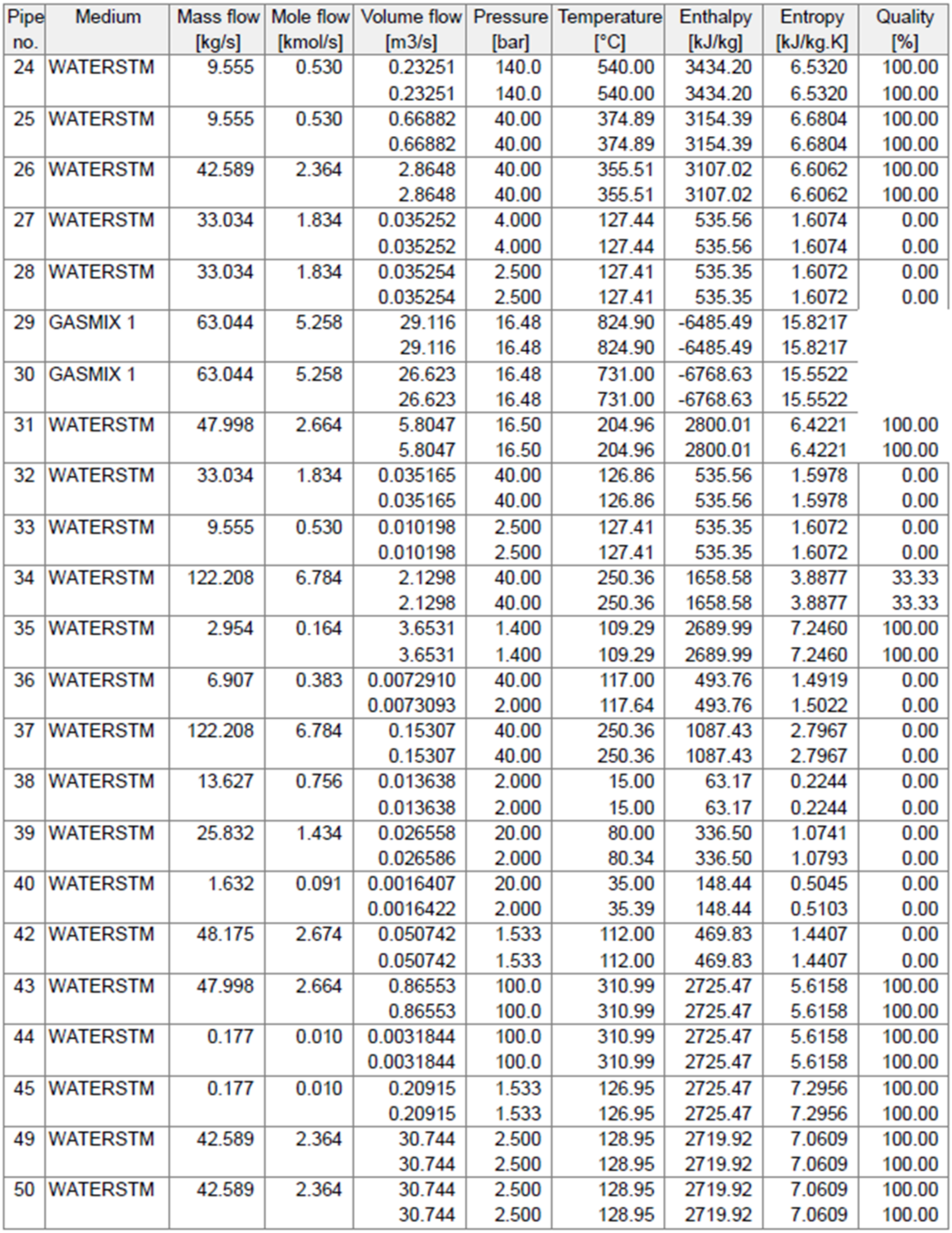
Figure B3. PFD of heat integration and power production system.

B-3.1 Material Balance for Heat and Power Integration System

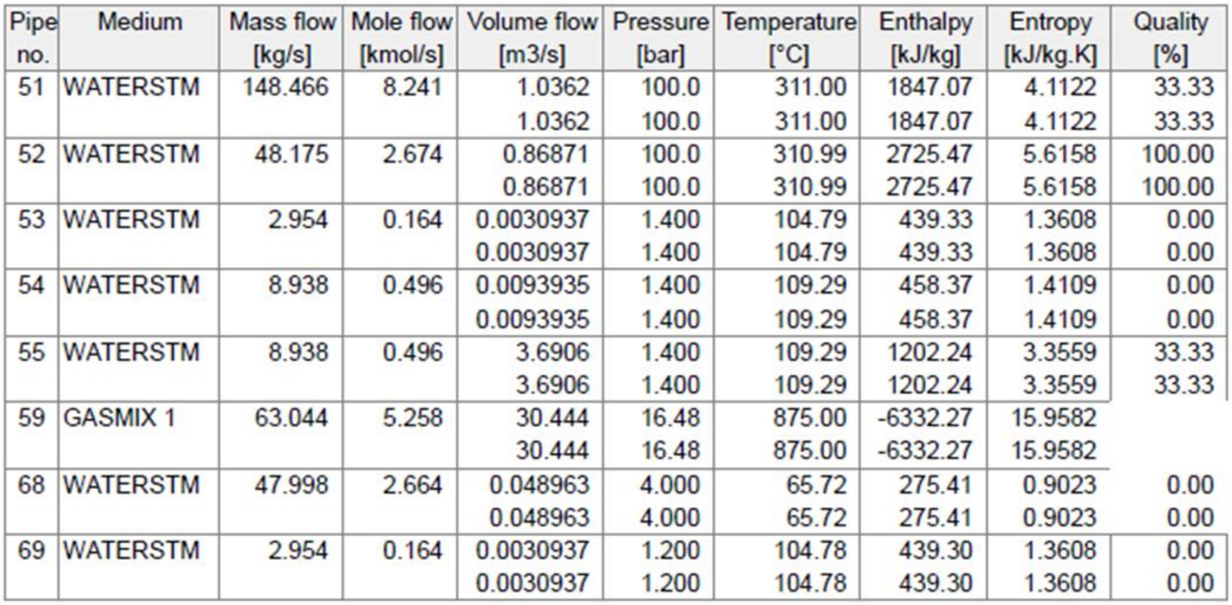
Table B34. Material balance for heat and power system.



Pipe no.	Medium	Mass flow [kg/s]	Mole flow [kmol/s]	Volume flow [m3/s]	Pressure [bar]	Temperature [°C]	Enthalpy [kJ/kg]	Entropy [kJ/kg.K]	Quality [%]
1	GASMIX 1	63.044	5.258	30.444	16.48	875.00	-6332.27	15.9582	
				30.444	16.48	875.00	-6332.27	15.9582	
2	WATERSTM	48.175	2.674	0.050554	100.0	113.58	483.63	1.4496	0.00
				0.050554	100.0	113.58	483.63	1.4496	0.00
3	WATERSTM	48.175	2.674	0.068471	100.0	305.00	1371.94	3.2985	0.00
				0.068471	100.0	305.00	1371.94	3.2985	0.00
4	GASMIX 1	63.044	5.258	17.069	16.48	371.82	-7802.93	14.2808	
				17.069	16.48	371.82	-7802.93	14.2808	
5	GASMIX 1	63.044	5.258	9.9042	16.48	140.33	-8728.90	12.3657	
				9.9042	16.48	140.33	-8728.90	12.3657	
6	WATERSTM	42.589	2.364	0.045452	2.500	127.41	535.35	1.6072	0.00
				0.045452	2.500	127.41	535.35	1.6072	0.00
7	WATERSTM	34.684	1.925	1.7264	40.00	250.35	2800.90	6.0696	100.00
				1.7264	40.00	250.35	2800.90	6.0696	100.00
8	WATERSTM	48.175	2.674	0.86871	100.0	310.99	2725.47	5.6158	100.00
				0.86871	100.0	310.99	2725.47	5.6158	100.00
9	WATERSTM	33.034	1.834	2.1959	40.00	350.00	3093.32	6.5843	100.00
				2.1959	40.00	350.00	3093.32	6.5843	100.00
10	GASMIX 1	63.044	5.258	10.836	16.48	148.50	-8481.73	12.9591	
				10.836	16.48	148.50	-8481.73	12.9591	
11	WATERSTM	148.466	8.241	0.21566	100.0	311.00	1407.87	3.3603	0.00
				0.21566	100.0	311.00	1407.87	3.3603	0.00
12	WATERSTM	47.998	2.664	1.2691	100.0	400.00	3097.38	6.2140	100.00
				1.2691	100.0	400.00	3097.38	6.2140	100.00
13	GASMIX 1	63.044	5.258	10.438	16.48	145.29	-8587.19	12.7073	
				10.438	16.48	145.29	-8587.19	12.7073	
14	WATERSTM	34.684	1.925	0.039273	40.00	185.39	788.43	2.1875	0.00
				0.039273	40.00	185.39	788.43	2.1875	0.00
15	WATERSTM	33.034	1.834	1.6443	40.00	250.35	2800.90	6.0696	100.00
				1.6443	40.00	250.35	2800.90	6.0696	100.00
16	WATERSTM	1.650	0.092	0.0020667	40.00	250.36	1087.43	2.7967	0.00
				0.0020667	40.00	250.36	1087.43	2.7967	0.00
17	WATERSTM	1.650	0.092	0.082128	40.00	250.35	2800.90	6.0696	100.00
				0.082128	40.00	250.35	2800.90	6.0696	100.00
18	WATERSTM	1.650	0.092	0.0020667	40.00	250.36	1087.43	2.7967	0.00
				0.0020667	40.00	250.36	1087.43	2.7967	0.00
19	WATERSTM	33.034	1.834	0.037249	40.00	182.00	773.50	2.1548	0.00
				0.037249	40.00	182.00	773.50	2.1548	0.00
20	WATERSTM	47.998	2.664	0.048966	2.000	65.70	275.14	0.9021	0.00
				0.048966	2.000	65.70	275.14	0.9021	0.00
21	WATERSTM	47.998	2.664	0.050468	3.800	110.00	461.54	1.4185	0.00
				0.050475	1.533	110.00	461.54	1.4191	0.00
22	GASMIX 1	63.044	5.258	9.9042	16.48	140.33	-8728.90	12.3657	
				9.9042	16.48	140.33	-8728.90	12.3657	
23	WATERSTM	9.555	0.530	0.010140	150.0	129.96	556.27	1.6202	0.00
				0.010140	150.0	129.96	556.27	1.6202	0.00



Pipe no.	Medium	Mass flow [kg/s]	Mole flow [kmol/s]	Volume flow [m3/s]	Pressure [bar]	Temperature [°C]	Enthalpy [kJ/kg]	Entropy [kJ/kg.K]	Quality [%]
24	WATERSTM	9.555	0.530	0.23251	140.0	540.00	3434.20	6.5320	100.00
				0.23251	140.0	540.00	3434.20	6.5320	100.00
25	WATERSTM	9.555	0.530	0.66882	40.00	374.89	3154.39	6.6804	100.00
				0.66882	40.00	374.89	3154.39	6.6804	100.00
26	WATERSTM	42.589	2.364	2.8648	40.00	355.51	3107.02	6.6062	100.00
				2.8648	40.00	355.51	3107.02	6.6062	100.00
27	WATERSTM	33.034	1.834	0.035252	4.000	127.44	535.56	1.6074	0.00
				0.035252	4.000	127.44	535.56	1.6074	0.00
28	WATERSTM	33.034	1.834	0.035254	2.500	127.41	535.35	1.6072	0.00
				0.035254	2.500	127.41	535.35	1.6072	0.00
29	GASMIX 1	63.044	5.258	29.116	16.48	824.90	-6485.49	15.8217	
				29.116	16.48	824.90	-6485.49	15.8217	
30	GASMIX 1	63.044	5.258	26.623	16.48	731.00	-6768.63	15.5522	
				26.623	16.48	731.00	-6768.63	15.5522	
31	WATERSTM	47.998	2.664	5.8047	16.50	204.96	2800.01	6.4221	100.00
				5.8047	16.50	204.96	2800.01	6.4221	100.00
32	WATERSTM	33.034	1.834	0.035165	40.00	126.86	535.56	1.5978	0.00
				0.035165	40.00	126.86	535.56	1.5978	0.00
33	WATERSTM	9.555	0.530	0.010198	2.500	127.41	535.35	1.6072	0.00
				0.010198	2.500	127.41	535.35	1.6072	0.00
34	WATERSTM	122.208	6.784	2.1298	40.00	250.36	1658.58	3.8877	33.33
				2.1298	40.00	250.36	1658.58	3.8877	33.33
35	WATERSTM	2.954	0.164	3.6531	1.400	109.29	2689.99	7.2460	100.00
				3.6531	1.400	109.29	2689.99	7.2460	100.00
36	WATERSTM	6.907	0.383	0.0072910	40.00	117.00	493.76	1.4919	0.00
				0.0073093	2.000	117.64	493.76	1.5022	0.00
37	WATERSTM	122.208	6.784	0.15307	40.00	250.36	1087.43	2.7967	0.00
				0.15307	40.00	250.36	1087.43	2.7967	0.00
38	WATERSTM	13.627	0.756	0.013638	2.000	15.00	63.17	0.2244	0.00
				0.013638	2.000	15.00	63.17	0.2244	0.00
39	WATERSTM	25.832	1.434	0.026558	20.00	80.00	336.50	1.0741	0.00
				0.026586	2.000	80.34	336.50	1.0793	0.00
40	WATERSTM	1.632	0.091	0.0016407	20.00	35.00	148.44	0.5045	0.00
				0.0016422	2.000	35.39	148.44	0.5103	0.00
42	WATERSTM	48.175	2.674	0.050742	1.533	112.00	469.83	1.4407	0.00
				0.050742	1.533	112.00	469.83	1.4407	0.00
43	WATERSTM	47.998	2.664	0.86553	100.0	310.99	2725.47	5.6158	100.00
				0.86553	100.0	310.99	2725.47	5.6158	100.00
44	WATERSTM	0.177	0.010	0.0031844	100.0	310.99	2725.47	5.6158	100.00
				0.0031844	100.0	310.99	2725.47	5.6158	100.00
45	WATERSTM	0.177	0.010	0.20915	1.533	126.95	2725.47	7.2956	100.00
				0.20915	1.533	126.95	2725.47	7.2956	100.00
49	WATERSTM	42.589	2.364	30.744	2.500	128.95	2719.92	7.0609	100.00
				30.744	2.500	128.95	2719.92	7.0609	100.00
50	WATERSTM	42.589	2.364	30.744	2.500	128.95	2719.92	7.0609	100.00
				30.744	2.500	128.95	2719.92	7.0609	100.00



Pipe no.	Medium	Mass flow [kg/s]	Mole flow [kmol/s]	Volume flow [m3/s]	Pressure [bar]	Temperature [°C]	Enthalpy [kJ/kg]	Entropy [kJ/kg.K]	Quality [%]
51	WATERSTM	148.466	8.241	1.0362	100.0	311.00	1847.07	4.1122	33.33
				1.0362	100.0	311.00	1847.07	4.1122	33.33
52	WATERSTM	48.175	2.674	0.86871	100.0	310.99	2725.47	5.6158	100.00
				0.86871	100.0	310.99	2725.47	5.6158	100.00
53	WATERSTM	2.954	0.164	0.0030937	1.400	104.79	439.33	1.3608	0.00
				0.0030937	1.400	104.79	439.33	1.3608	0.00
54	WATERSTM	8.938	0.496	0.0093935	1.400	109.29	458.37	1.4109	0.00
				0.0093935	1.400	109.29	458.37	1.4109	0.00
55	WATERSTM	8.938	0.496	3.6906	1.400	109.29	1202.24	3.3559	33.33
				3.6906	1.400	109.29	1202.24	3.3559	33.33
59	GASMIX 1	63.044	5.258	30.444	16.48	875.00	-6332.27	15.9582	
				30.444	16.48	875.00	-6332.27	15.9582	
68	WATERSTM	47.998	2.664	0.048963	4.000	65.72	275.41	0.9023	0.00
				0.048963	4.000	65.72	275.41	0.9023	0.00
69	WATERSTM	2.954	0.164	0.0030937	1.200	104.78	439.30	1.3608	0.00
				0.0030937	1.200	104.78	439.30	1.3608	0.00

Appendix C

SMR/ATR Methanol Flowsheets, Heat, and Mass Balances

C-1. SMR/ATR Reference Plant PFD

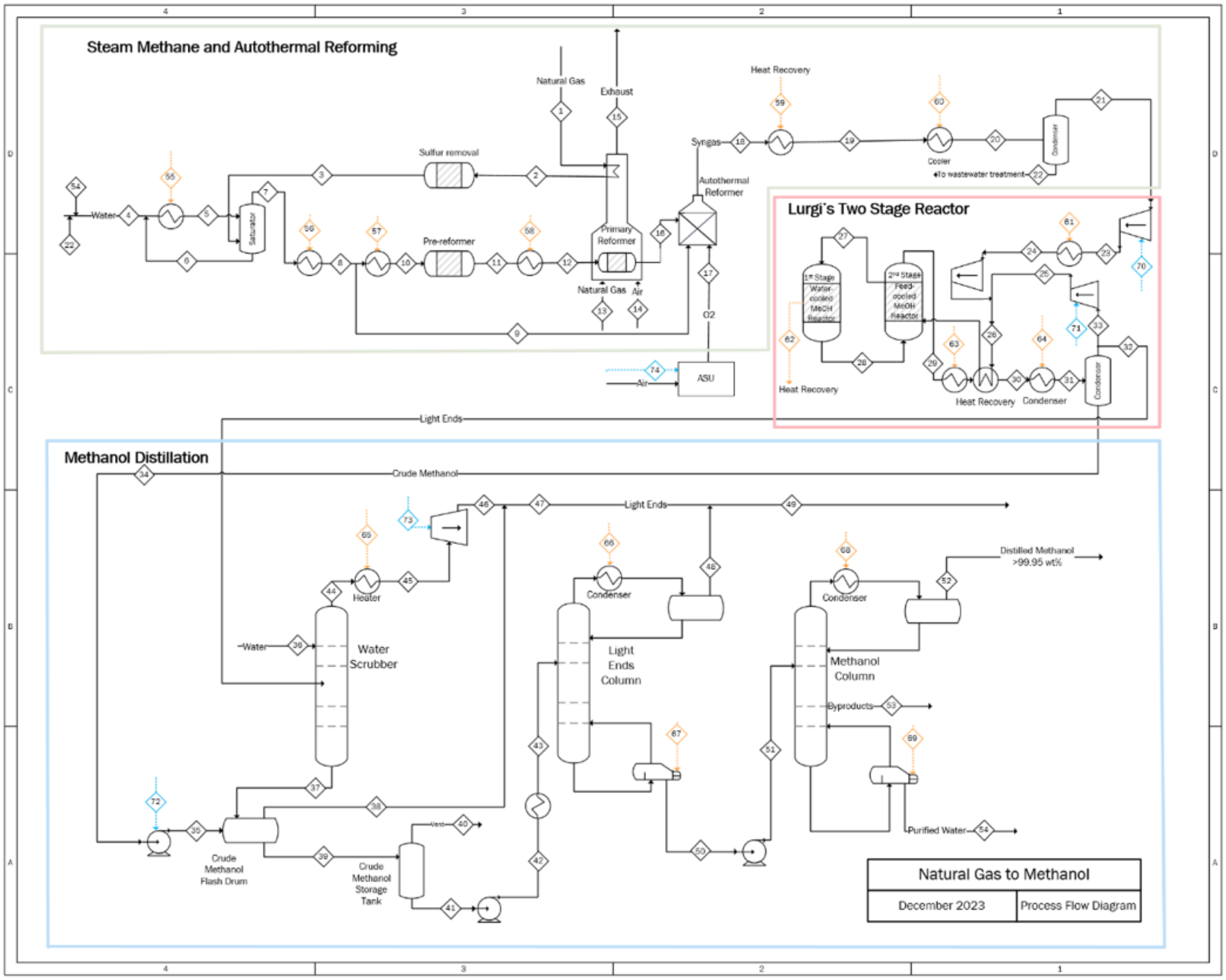


Figure C1. Methanol plant overall energy PFD. (The numbers correspond to the energy input from electricity, heat in fuel, and steam, as listed in Appendix B.)

C-2. Chemical Process Model

This section explains the Aspen Plus model built to represent steam methane reforming with autothermal reforming. PFDs of the steam methane reforming process are presented in Figure C2 and Figure C3, representing syngas production and the methanol synthesis process, respectively. The same assumptions used in the steam methane reforming model were used for NG and air composition, pressures and temperatures as well as the assumptions for the performer and steam methane reformer blocks.

The product stream from the SMR (C12), the bypass stream from the splitter (C8A), and extra oxygen (O_2) for autothermal reforming are mixed at the mixer block (MX3) and introduced to the ATR. An air separation unit supplies oxygen to the ATR. ATR occurs in the block (ATR) simulated by an adiabatic RGibbs reactor model. Most of the energy required for steam methane reforming is self-supplied from the heat produced via the partial combustion of methane. The configurations of the steam methane reformer and the autothermal reformer are illustrated in Figure C2.

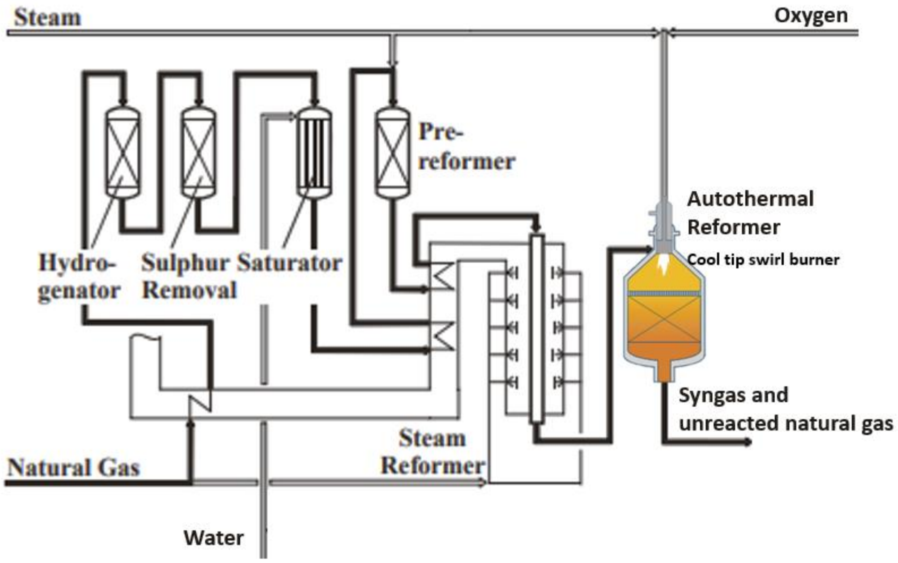


Figure C2. Configuration diagram of the SMR and the ATR in series [66–67].

The product stream exiting the ATR is high temperature (>980°C). Heat is recovered from this stream via two heat exchangers (HX8 and HX9). Some heat is used to preheat water used for chemical reactions (HX8), and some can be used to produce medium pressure (MP) steam from the boiler feedwater (HX9). In this model, only the amount of heat released is calculated for HX8 and HX9.

The methanol synthesis process used in this model is based on Lurgi's Mega-Methanol design, as well as the reference methanol synthesis model included in Aspen Plus v11. A series of compressors (K201-S1 and K201-S2) increases the pressure of the crude syngas stream to 92 bar, which is favorable for the methanol reaction. The modeled process utilizes two methanol reactors in series (R201-A and R201-B) to produce methanol (MeOH) as the target product and ethanol (EtOH) and dimethyl ether (DME) as byproducts. Both reactors were simulated by a kinetic-based reactor model.

Some light gas remains dissolved in the crude methanol following the water scrubber. The first distillation tower (C301) separates light ends from crude methanol. It has 27 stages, and the feed stream is introduced at the 15th. Light gases separated from crude methanol at the top of the distillation tower go to the outlet (light ends), whereas water and methanol mixture exit at the bottom and are pumped to the next distillation tower. Here, a RadFrac distillation model, including a partial-vapor condenser and a kettle reboiler, was used. To match the target temperature of first stage of the distillation tower (or condenser) to 45.5℃, the distillate rate was adjusted to 56.34463 ton/day. The reboiler duty was 52.75 MW. After the first distillation tower, the crude methanol had less than a 1 × 10-7 mass fraction of light gas.

The second distillation tower (C302) separates the water and impurities from the methanol to meet Grade A product specifications. Fifty stages are required for the distillation, and the feed stream is introduced at the 30th. Methanol exits the top of the distillation tower, a byproduct containing ethanol exits the 45th stage, and water exits the bottom. Here, a RadFrac distillation model, including a partial-vapor condenser and a kettle reboiler, was used. The molar reflux ratio for the separation was 1.5, and the mass flowrate of the liquid side stream at the 45th stage was 96 ton/day. To match the target water purities of the bottom stage to 0.999999, the reboiler duty was modified to 372 MW. In the end, methanol purity at the top of the tower is achievable at a mass fraction of 0.9995.

C-2.1 Energy Balance and Heat Recovery System

Recovery of the available heat is illustrated by the PFD in Figure C3. Starting from the left-hand side at the syngas inlet, we see high HP and LP steam superheaters followed by a HP evaporator (either a once-through or drum type). Next is an evaporator that generates MP steam for injection into natural gas prior to the reformer. This steam, which primarily consists of recycled water effluents with limited purity, is generated in a physically separated system and is not introduced into the main process steam system.

After the deaerator, this "dirty" water is preheated with heat from cooler E210 and evaporated using heat from the syngas. A small portion of the steam generated is then extracted for the deaerator. MP feedwater is partially preheated by heat from the syngas before heading to the drum boiler, which cools the MeOH reactor. High-temperature syngas is then used to superheat this steam and is routed to the first turbine section to lower its pressure for mixing with the MP steam. Expansion then continues to the LP level of 2.5 bar, where 85% of the steam mass flow rate is routed to the reboiler of column C302. The remaining steam can be expanded in an LP turbine section or in a small separate turbine to provide nearly 10 MWe of additional power. MP saturated process steam is generated in MeOH reactor.

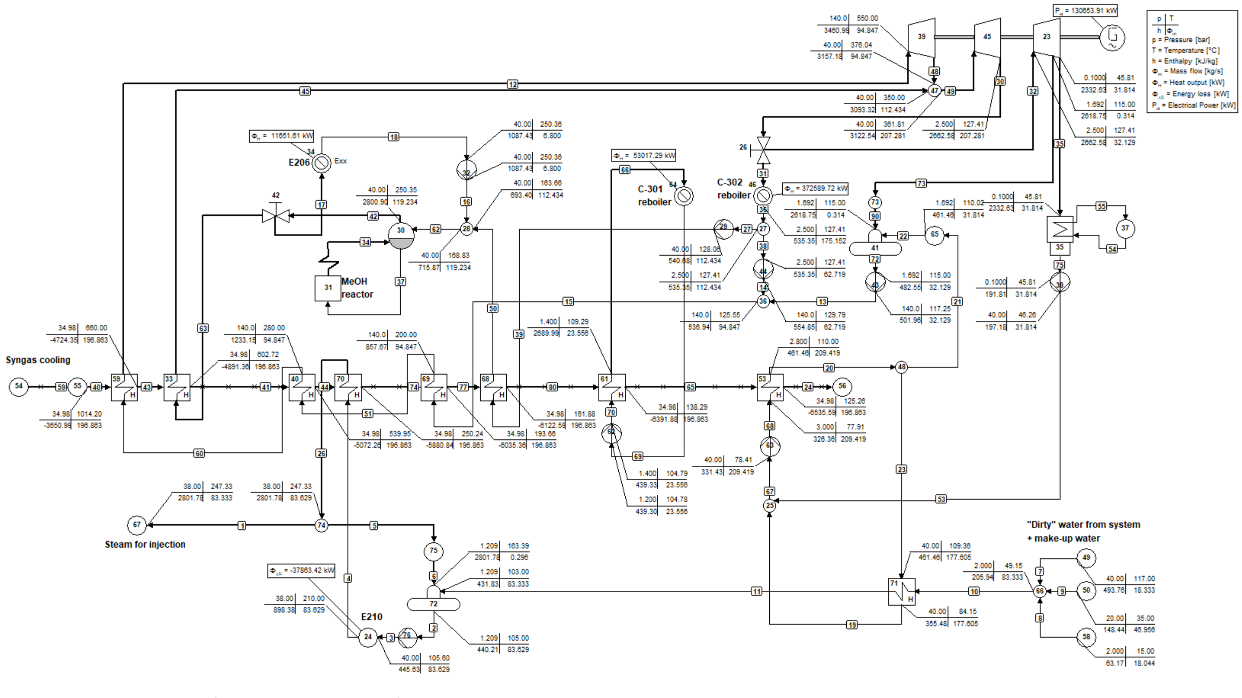


Figure C3. PFD of heat recovery from syngas cooling.

The heat transfer-temperature (*Q*-*T*) diagram in Figure C4 illustrates the heat exchange that occurs during the syngas cooling. The figure shows that, at the onset of condensation, heat can be supplied for low-temperature waste heat recovery or for boiling water at 1 atm to supply the column C-301 reboiler. Making steam only 2.5 bar to supply the column C-302 reboiler would simplify the process steam system, but its saturation temperature is too high for this abundant low-temperature heat to be utilized.

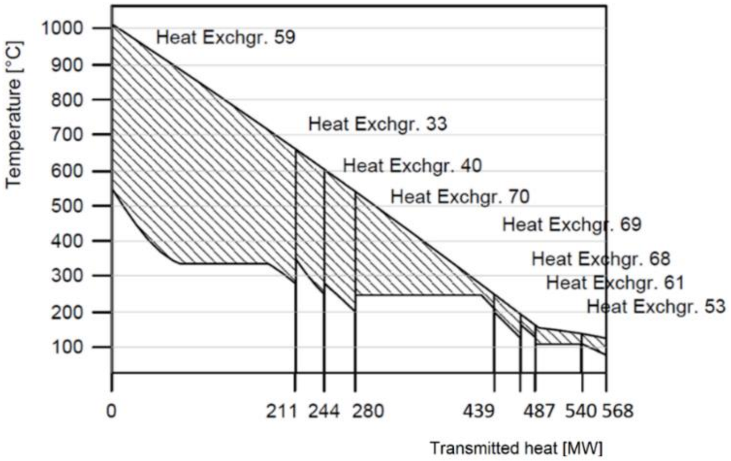


Figure C4. Heat transfer-temperature (*Q*-*T*) diagram of waste heat recovery from syngas cooling.

C-2.2 Overall Energy and CO_2 Balance

Table C summarizes the energy requirements for methanol production. The energy demands of the reference methanol plant are based on the Aspen model with heat integration.

Similar to the SMR ATR process, steam is generated during steam methane reforming, syngas cooling, and methanol cooling. Implementing a waste heat recovery system to produce steam makes it possible to meet all heat duty demands (apart from the heat supplied by NG combustion) while producing more than the required power. Water generated as part of a chemical reaction is not considered in the mass or energy balance, as it is not utilized in the energy system streams. The energy demand of the cooling system for rejecting heat to the ambient condition was not included in the model, meaning the actual net electricity export may be smaller by up to several MWe. Considering that the only energy supplied to the system comes from fuel combustion, the energy required for methanol production is 5.519 GJ/metric ton MeOH. This value does not include energy generated from the feedstock.

Table C35. Overall energy requirements for the reference methanol plant.

PO#	Process Unit	Power Consumption	Fuel heat content (LHV)	Steam Consumption	Steam Generation	Steam Quality (production)	Steam Quality (consumption)	Heat Released to Ambient
		MWe	MWt	MWt	MWt			MWt
A	**Steam Methane and Autothermal Reforming + cooling**	1	698	216* (83 kg/s)	568	LP, MP, HP	MP	105
B	**Lurgi's Two Stage Reactor & methanol cooling**	79			287	MP		313
C	**Methanol Distillation**	-11		437			LP	410
D	**Air separation unit**	42						n.a.
E	**Process steam and power system**	-129		-653			MP, LP	68
	Total	-16	698	0	855			827

*Steam for injection into the process

NG introduced as feedstock contains 3,385 MW of energy (26.8 GJ per metric ton of methanol), and the product methanol contains 2,517 MW (based on a lower heating value [LHV]). Thus, the total energy requirement for methanol production, as calculated in the model, is 32.32 GJ per metric ton of methanol. This highlights the potential of integrating low-carbon sources when considering an alternative feedstock.

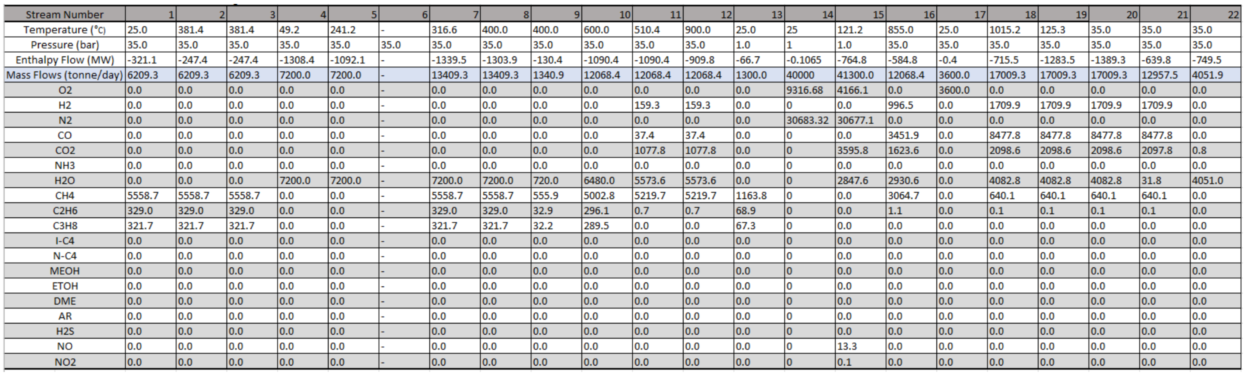
Table C36 summarizes the direct carbon emissions for each process unit, in terms of both the reference plant and the national scale. The total national emissions for 2023 are reported based on the EPA FLIGHT, and the reference plant emissions are from the Aspen model results, including heat integration [68]. Emissions are produced during NG combustion in the furnace and are released in the exhaust stream during steam methane reforming and in the light ends during distillation. The national-scale emissions are used to calculate the process unit emissions by estimating the percent contribution of emissions from each unit. The linear correlation in Figure 16 predicts the carbon emissions generated by approximately 1.3 metric tons per year for the reference plant, using national data. This value aligns with the 1.2 metric tons of CO_2 predicted per year by the Aspen model with heat integration for the reference plant. The amount of CO_2 emissions associated with methanol production is 0.348 kg CO_2 per kg of methanol.

Table C36. Direct CO_2 emissions for the reference methanol plant at the national scale.

PO #	Process Unit	Reference Plant CO2 Emissions	National Scale CO2 Emissions
		Metric Tons per Year	Metric Tons per Year
A	**Steam Methane and Autothermal Reforming + cooling**	1,150,672	2,688,416
B	**Lurgi's Two Stage Reactor & methanol cooling**		
C	**Methanol Distillation**	68,726	160,571
D	**Air separation unit**		
E	**Process steam and power system**		
	Total	**1,219,398**	**2,848,987**

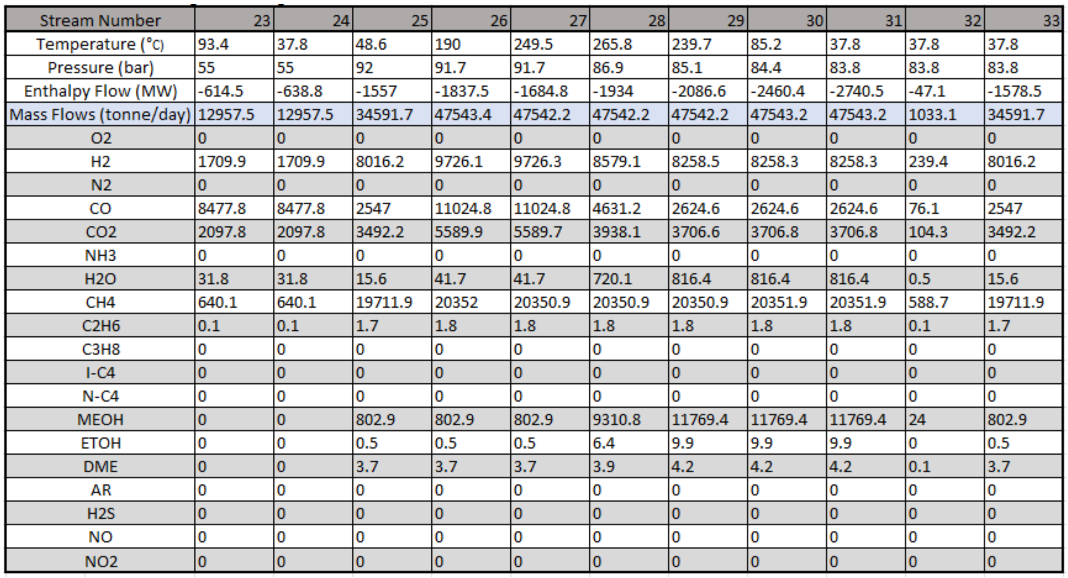
* Note: The light gases from the methanol synthesis and distillation sections are recovered and used as fuel in the reforming furnace. Combustion of the light gases reduces the amount of NG required by the reforming furnace.

Table C37. Methanol material balance for steam methane and autothermal reforming.



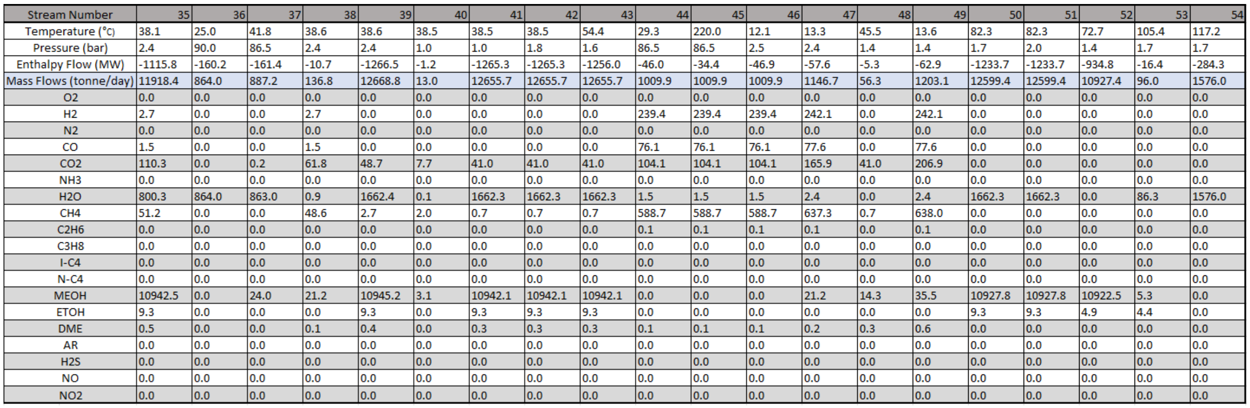
Stream Number	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22
Temperature (°C)	25.0	381.4	381.4	49.2	241.2	-	316.6	400.0	400.0	600.0	510.4	900.0	25.0	25	121.2	855.0	25.0	1015.2	125.3	35.0	35.0	35.0
Pressure (bar)	35.0	35.0	35.0	35.0	35.0	35.0	35.0	35.0	35.0	35.0	35.0	35.0	1.0	1	1.0	35.0	35.0	35.0	35.0	35.0	35.0	35.0
Enthalpy Flow (MW)	-321.1	-247.4	-247.4	-1308.4	-1092.1	-	-1339.5	-1303.9	-130.4	-1090.4	-1090.4	-909.8	-66.7	-0.1065	-764.8	-584.8	-0.4	-715.5	-1283.5	-1389.3	-639.8	-749.5
Mass Flows (tonne/day)	6209.3	6209.3	6209.3	7200.0	7200.0	-	13409.3	13409.3	1340.9	12068.4	12068.4	12068.4	1300.0	40000	41300.0	12068.4	3600.0	17009.3	17009.3	17009.3	12957.5	4051.9
O2	0.0	0.0	0.0	0.0	0.0	-	0.0	0.0	0.0	0.0	0.0	0.0	0.0	9316.68	4166.1	0.0	3600.0	0.0	0.0	0.0	0.0	0.0
H2	0.0	0.0	0.0	0.0	0.0	-	0.0	0.0	0.0	0.0	159.3	159.3	0.0	0	0.0	996.5	0.0	1709.9	1709.9	1709.9	1709.9	0.0
N2	0.0	0.0	0.0	0.0	0.0	-	0.0	0.0	0.0	0.0	0.0	0.0	0.0	30683.32	30677.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CO	0.0	0.0	0.0	0.0	0.0	-	0.0	0.0	0.0	0.0	37.4	37.4	0.0	0	0.0	3451.9	0.0	8477.8	8477.8	8477.8	8477.8	0.0
CO2	0.0	0.0	0.0	0.0	0.0	-	0.0	0.0	0.0	0.0	1077.8	1077.8	0.0	0	3595.8	1623.6	0.0	2098.6	2098.6	2098.6	2097.8	0.8
NH3	0.0	0.0	0.0	0.0	0.0	-	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
H2O	0.0	0.0	0.0	7200.0	7200.0	-	7200.0	7200.0	720.0	6480.0	5573.6	5573.6	0.0	0	2847.6	2930.6	0.0	4082.8	4082.8	4082.8	31.8	4051.0
CH4	5558.7	5558.7	5558.7	0.0	0.0	-	5558.7	5558.7	555.9	5002.8	5219.7	5219.7	1163.8	0	0.0	3064.7	0.0	640.1	640.1	640.1	640.1	0.0
C2H6	329.0	329.0	329.0	0.0	0.0	-	329.0	329.0	32.9	296.1	0.7	0.7	68.9	0	0.0	1.1	0.0	0.1	0.1	0.1	0.1	0.0
C3H8	321.7	321.7	321.7	0.0	0.0	-	321.7	321.7	32.2	289.5	0.0	0.0	67.3	0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
I-C4	0.0	0.0	0.0	0.0	0.0	-	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
N-C4	0.0	0.0	0.0	0.0	0.0	-	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
MEOH	0.0	0.0	0.0	0.0	0.0	-	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ETOH	0.0	0.0	0.0	0.0	0.0	-	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
DME	0.0	0.0	0.0	0.0	0.0	-	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
AR	0.0	0.0	0.0	0.0	0.0	-	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
H2S	0.0	0.0	0.0	0.0	0.0	-	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NO	0.0	0.0	0.0	0.0	0.0	-	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0	13.3	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NO2	0.0	0.0	0.0	0.0	0.0	-	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0	0.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0

Table C38. Methanol material balance for Lurgi's two-stage reactor and methanol cooling.



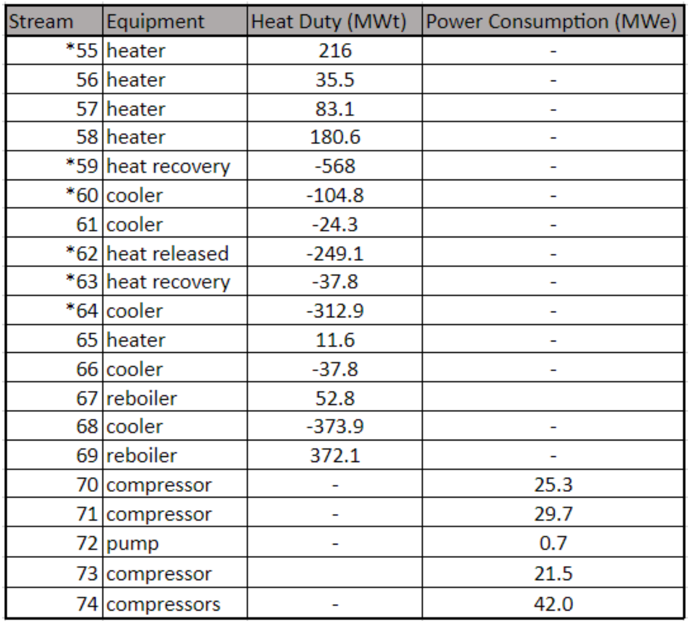
Stream Number	23	24	25	26	27	28	29	30	31	32	33
Temperature (°C)	93.4	37.8	48.6	190	249.5	265.8	239.7	85.2	37.8	37.8	37.8
Pressure (bar)	55	55	92	91.7	91.7	86.9	85.1	84.4	83.8	83.8	83.8
Enthalpy Flow (MW)	-614.5	-638.8	-1557	-1837.5	-1684.8	-1934	-2086.6	-2460.4	-2740.5	-47.1	-1578.5
Mass Flows (tonne/day)	12957.5	12957.5	34591.7	47543.4	47542.2	47542.2	47542.2	47543.2	47543.2	1033.1	34591.7
O2	0	0	0	0	0	0	0	0	0	0	0
H2	1709.9	1709.9	8016.2	9726.1	9726.3	8579.1	8258.5	8258.3	8258.3	239.4	8016.2
N2	0	0	0	0	0	0	0	0	0	0	0
CO	8477.8	8477.8	2547	11024.8	11024.8	4631.2	2624.6	2624.6	2624.6	76.1	2547
CO2	2097.8	2097.8	3492.2	5589.9	5589.7	3938.1	3706.6	3706.8	3706.8	104.3	3492.2
NH3	0	0	0	0	0	0	0	0	0	0	0
H2O	31.8	31.8	15.6	41.7	41.7	720.1	816.4	816.4	816.4	0.5	15.6
CH4	640.1	640.1	19711.9	20352	20350.9	20350.9	20350.9	20351.9	20351.9	588.7	19711.9
C2H6	0.1	0.1	1.7	1.8	1.8	1.8	1.8	1.8	1.8	0.1	1.7
C3H8	0	0	0	0	0	0	0	0	0	0	0
I-C4	0	0	0	0	0	0	0	0	0	0	0
N-C4	0	0	0	0	0	0	0	0	0	0	0
MEOH	0	0	802.9	802.9	802.9	9310.8	11769.4	11769.4	11769.4	24	802.9
ETOH	0	0	0.5	0.5	0.5	6.4	9.9	9.9	9.9	0	0.5
DME	0	0	3.7	3.7	3.7	3.9	4.2	4.2	4.2	0.1	3.7
AR	0	0	0	0	0	0	0	0	0	0	0
H2S	0	0	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0	0	0

Table C39. Methanol material balance for methanol distillation.



Stream Number	35	36	37	38	39	40	41	42	43	44	45	46	47	48	49	50	51	52	53	54
Temperature (°C)	38.1	25.0	41.8	38.6	38.6	38.5	38.5	38.5	54.4	29.3	220.0	12.1	13.3	45.5	13.6	82.3	82.3	72.7	105.4	117.2
Pressure (bar)	2.4	90.0	86.5	2.4	2.4	1.0	1.0	1.8	1.6	86.5	86.5	2.5	2.4	1.4	1.4	1.7	2.0	1.4	1.7	1.7
Enthalpy Flow (MW)	-1115.8	-160.2	-161.4	-10.7	-1266.5	-1.2	-1265.3	-1265.3	-1256.0	-46.0	-34.4	-46.9	-57.6	-5.3	-62.9	-1233.7	-1233.7	-934.8	-16.4	-284.3
Mass Flows (tonne/day)	11918.4	864.0	887.2	136.8	12668.8	13.0	12655.7	12655.7	12655.7	1009.9	1009.9	1009.9	1146.7	56.3	1203.1	12599.4	12599.4	10927.4	96.0	1576.0
O2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
H2	2.7	0.0	0.0	2.7	0.0	0.0	0.0	0.0	0.0	239.4	239.4	239.4	242.1	0.0	242.1	0.0	0.0	0.0	0.0	0.0
N2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CO	1.5	0.0	0.0	1.5	0.0	0.0	0.0	0.0	0.0	76.1	76.1	76.1	77.6	0.0	77.6	0.0	0.0	0.0	0.0	0.0
CO2	110.3	0.0	0.2	61.8	48.7	7.7	41.0	41.0	41.0	104.1	104.1	104.1	165.9	41.0	206.9	0.0	0.0	0.0	0.0	0.0
NH3	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
H2O	800.3	864.0	863.0	0.9	1662.4	0.1	1662.3	1662.3	1662.3	1.5	1.5	1.5	2.4	0.0	2.4	1662.3	1662.3	0.0	86.3	1576.0
CH4	51.2	0.0	0.0	48.6	2.7	2.0	0.7	0.7	0.7	588.7	588.7	588.7	637.3	0.7	638.0	0.0	0.0	0.0	0.0	0.0
C2H6	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.1	0.1	0.1	0.1	0.0	0.1	0.0	0.0	0.0	0.0	0.0
C3H8	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
I-C4	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
N-C4	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
MEOH	10942.5	0.0	24.0	21.2	10945.2	3.1	10942.1	10942.1	10942.1	0.0	0.0	0.0	21.2	14.3	35.5	10927.8	10927.8	10922.5	5.3	0.0
ETOH	9.3	0.0	0.0	0.0	9.3	0.0	9.3	9.3	9.3	0.0	0.0	0.0	0.0	0.0	0.0	9.3	9.3	4.9	4.4	0.0
DME	0.5	0.0	0.0	0.1	0.4	0.0	0.3	0.3	0.3	0.1	0.1	0.1	0.2	0.3	0.6	0.0	0.0	0.0	0.0	0.0
AR	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
H2S	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NO	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NO2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0

Table C40. Methanol overall energy demands.



Stream	Equipment	Heat Duty (MWt)	Power Consumption (MWe)
*55	heater	216	-
56	heater	35.5	-
57	heater	83.1	-
58	heater	180.6	-
*59	heat recovery	-568	-
*60	cooler	-104.8	-
61	cooler	-24.3	-
*62	heat released	-249.1	-
*63	heat recovery	-37.8	-
*64	cooler	-312.9	-
65	heater	11.6	-
66	cooler	-37.8	-
67	reboiler	52.8	
68	cooler	-373.9	-
69	reboiler	372.1	-
70	compressor	-	25.3
71	compressor	-	29.7
72	pump	-	0.7
73	compressor		21.5
74	compressors	-	42.0

* Streams include heat integration results.

C-3. SMR with ATR Aspen Flowsheets

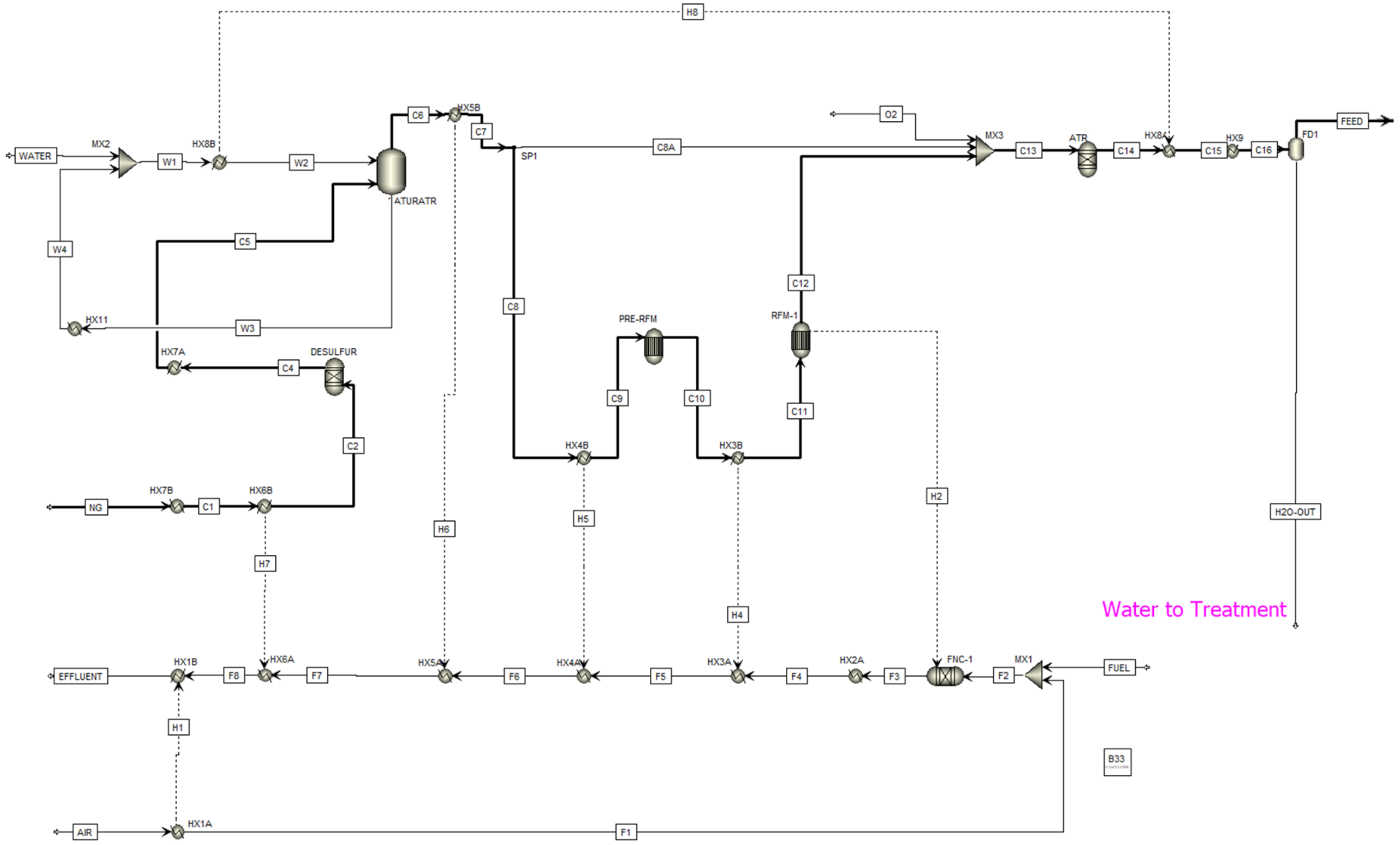


Figure C5. Process diagram of the Aspen Plus model for steam methane reforming.

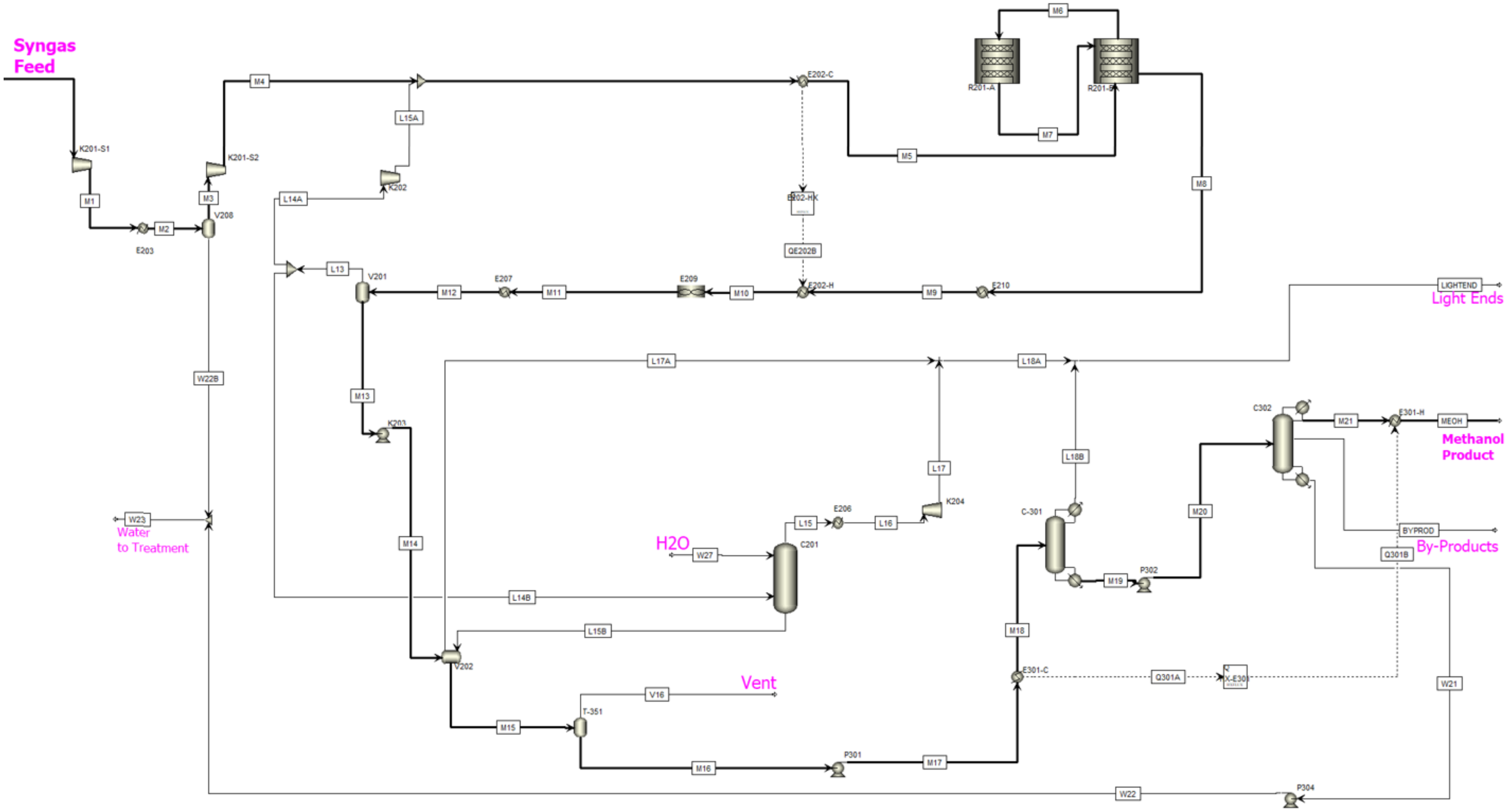


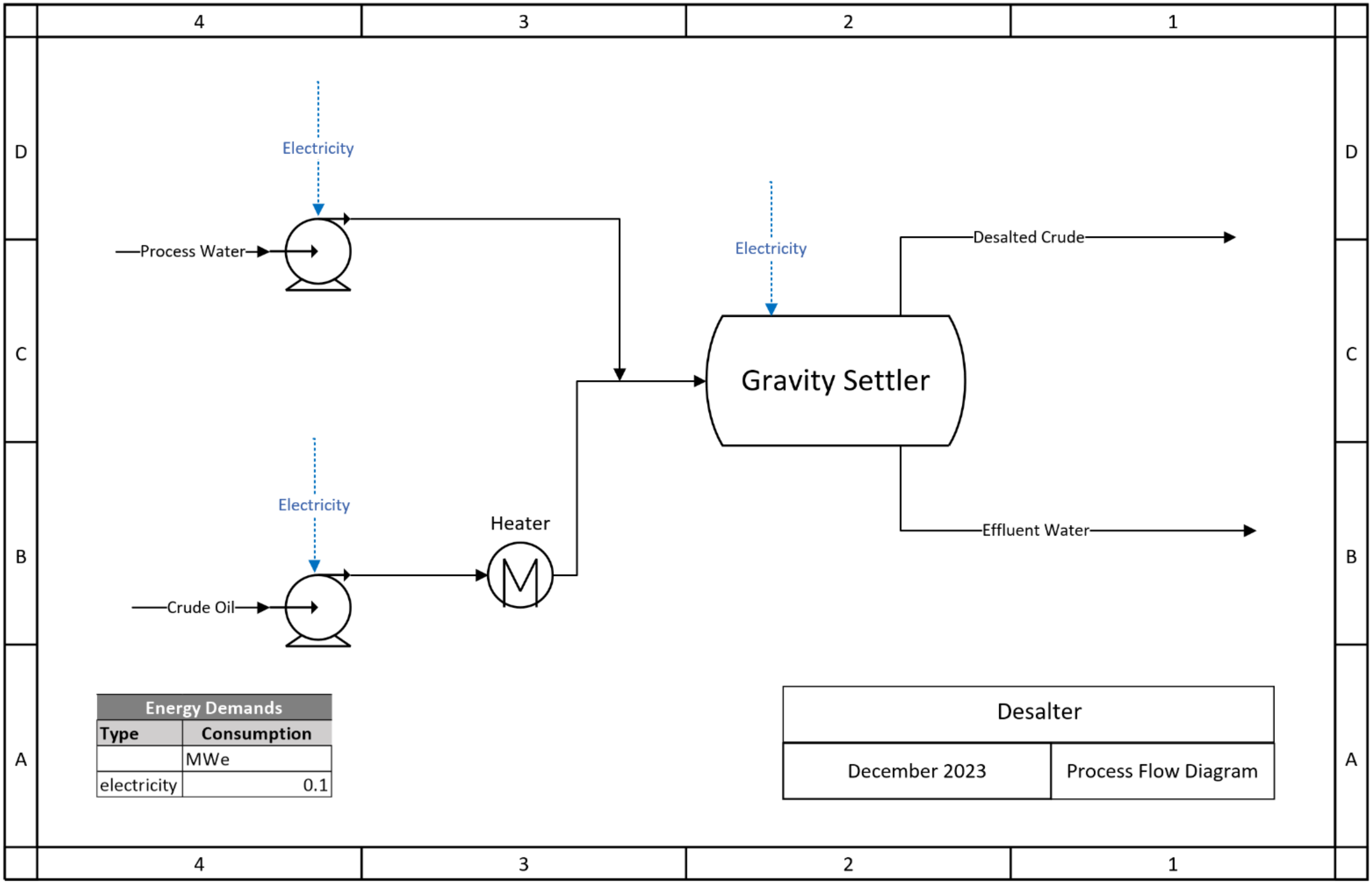
Figure C6. Process diagram of the Aspen Plus model for methanol synthesis and distillation.

Appendix D

Refinery Detailed PFDs, Material, and Energy Balances

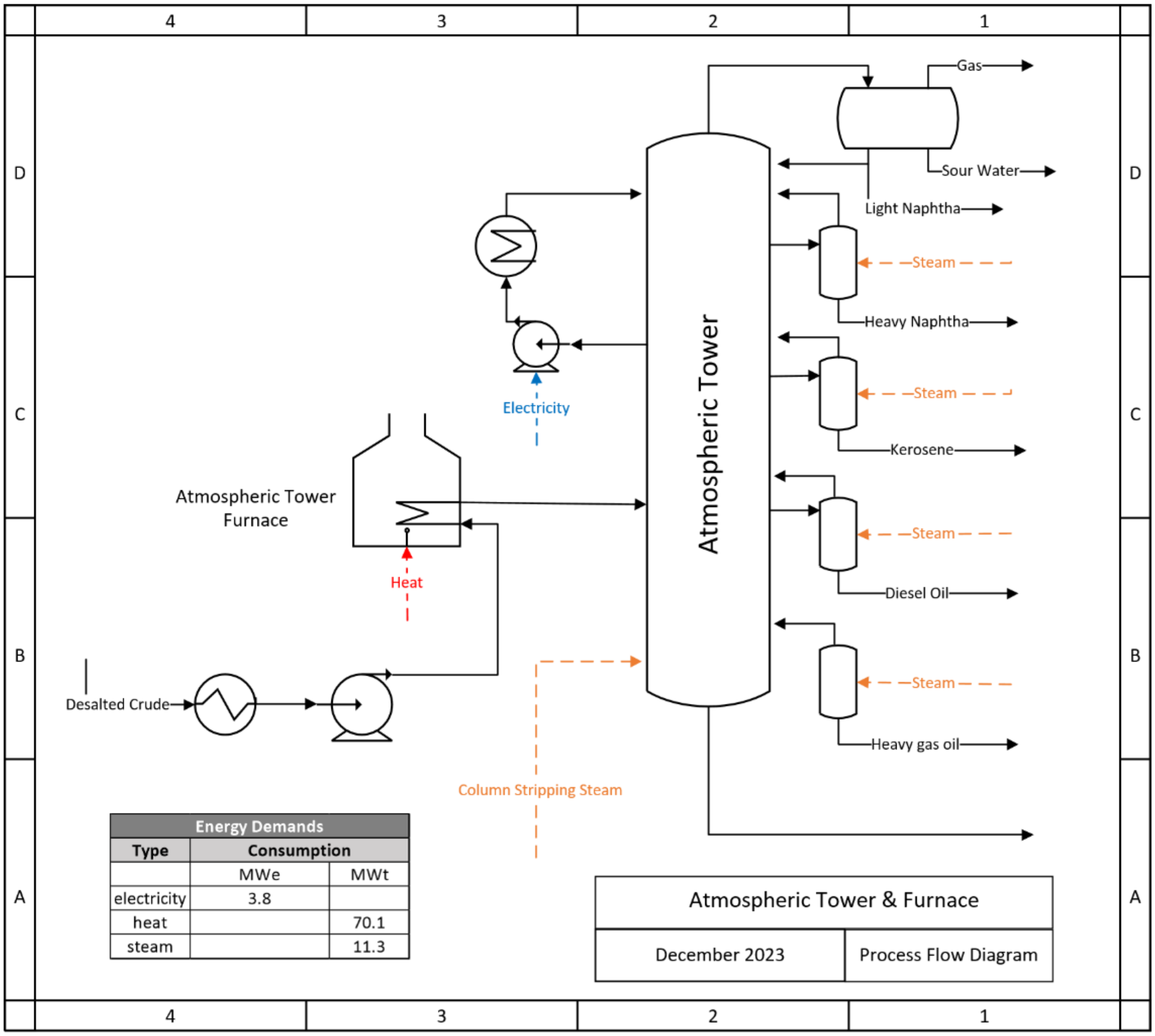
Table D41. Refinery overall energy requirements.

Energy Demands			
Stream	**Type**	**Consumption**	
		MWe	MWt
1	electricity	0.1	
2	heat		70.1
3	electricity	3.8	
4	steam		11.3
5	electricity	1.9	
6	steam		2.0
7	heat		27.1
8	electricity	0.2	
9	heat		12.6
10	electricity	1.9	
11	steam		7.7
12	heat		53.8
13	electricity	2.5	
14	steam		2.7
15	heat		36.6
16	electricity	0.1	
17	electricity	3.1	
18	steam		2.1
19	heat		29.9
20	electricity	5.0	
21	heat		22.9
22	electricity	1.4	
23	steam		19.4
24	electricity	0.6	
25	steam		7.4
26	heat		19.8
27	heat		47.0
28	electricity	1.7	
29	electricity	0.4	
30	steam		1.0
31	electricity	0.8	
32	steam		0.8
33	heat		11.4
34	electricity	0.5	
35	heat		22.8
36	electricity	0.4	
37	steam		0.4
38	steam		3.8
39	electricity	0.2	
40	electricity	1.0	
41	steam		1.2
42	heat		14.6



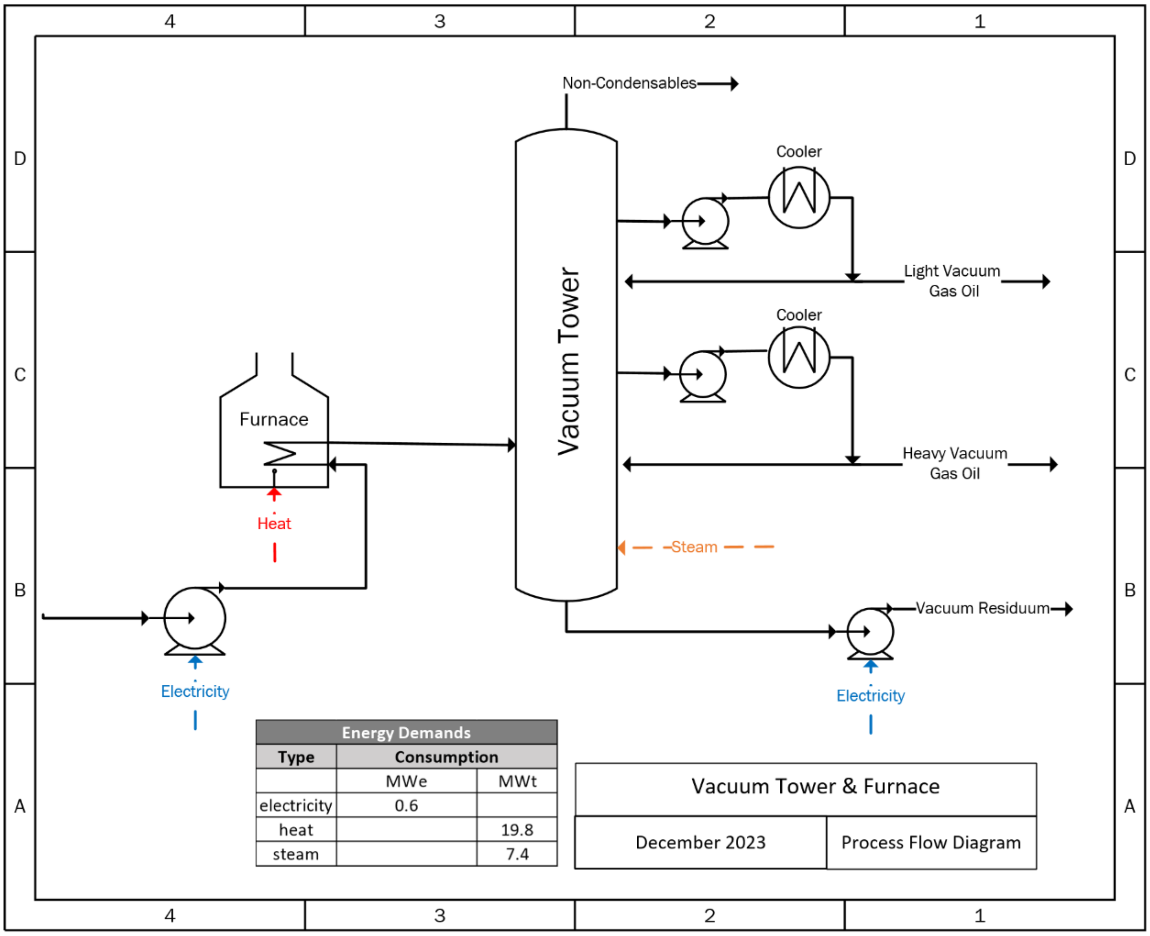
Energy Demands	
Type	**Consumption**
	MWe
electricity	0.1

Figure D7. Refinery process flow diagrams with energy demands for the primary refinery processes.



Energy Demands		
Type	Consumption	
	MWe	MWt
electricity	3.8	
heat		70.1
steam		11.3

Figure D7. Continued.



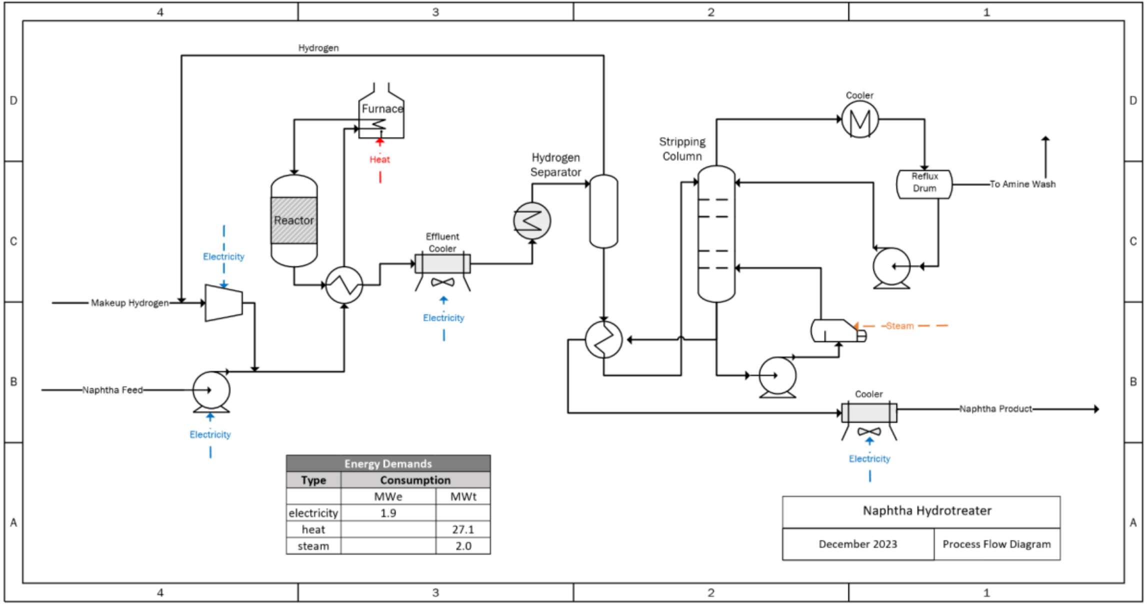
Energy Demands		
Type	Consumption	
	MWe	MWt
electricity	0.6	
heat		19.8
steam		7.4

Vacuum Tower & Furnace

December 2023

Process Flow Diagram

Figure D7. Continued.



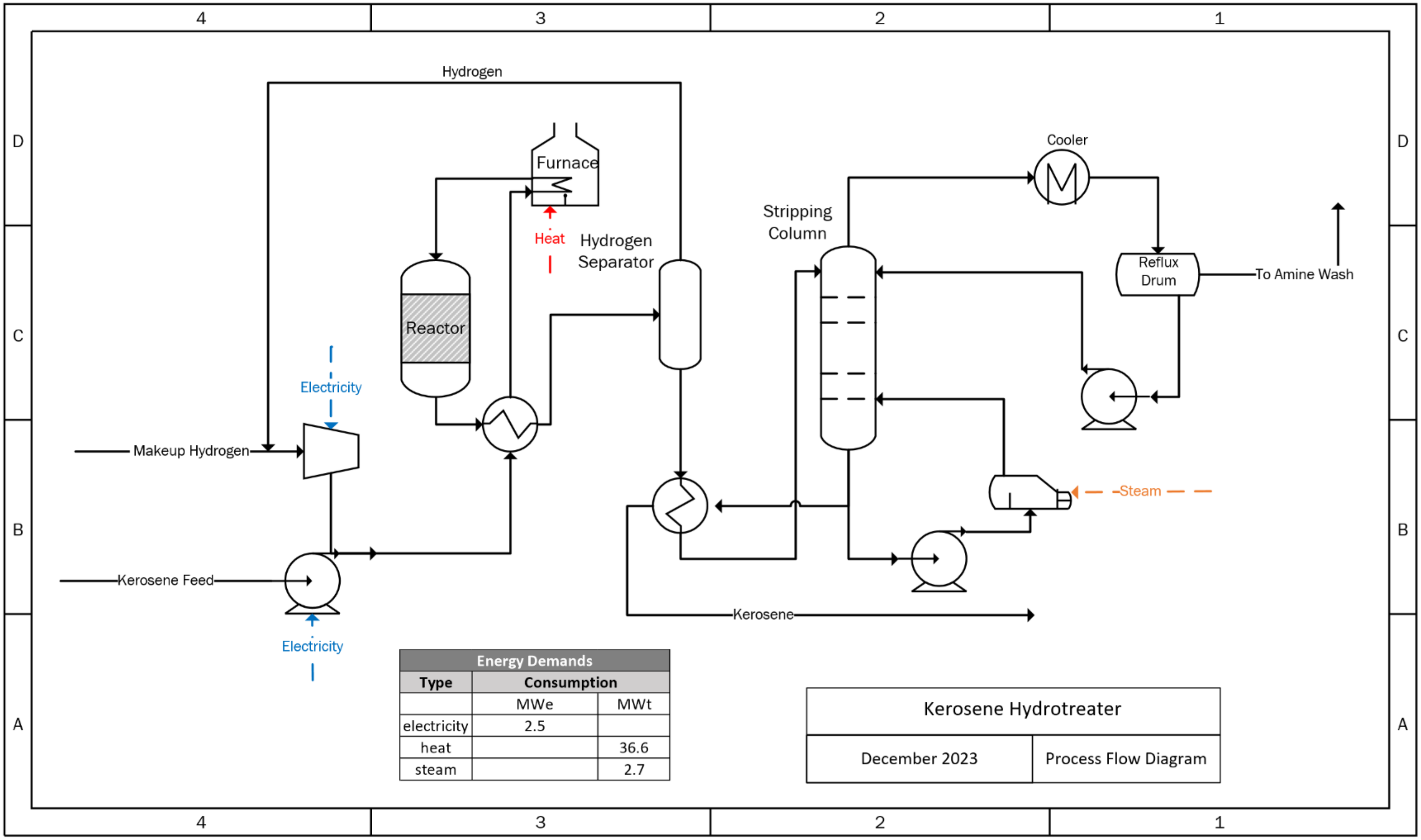
Energy Demands		
Type	Consumption	
	MWe	MWt
electricity	1.9	
heat		27.1
steam		2.0

Naphtha Hydrotreater

December 2023

Process Flow Diagram

Figure D7. Continued.



Energy Demands

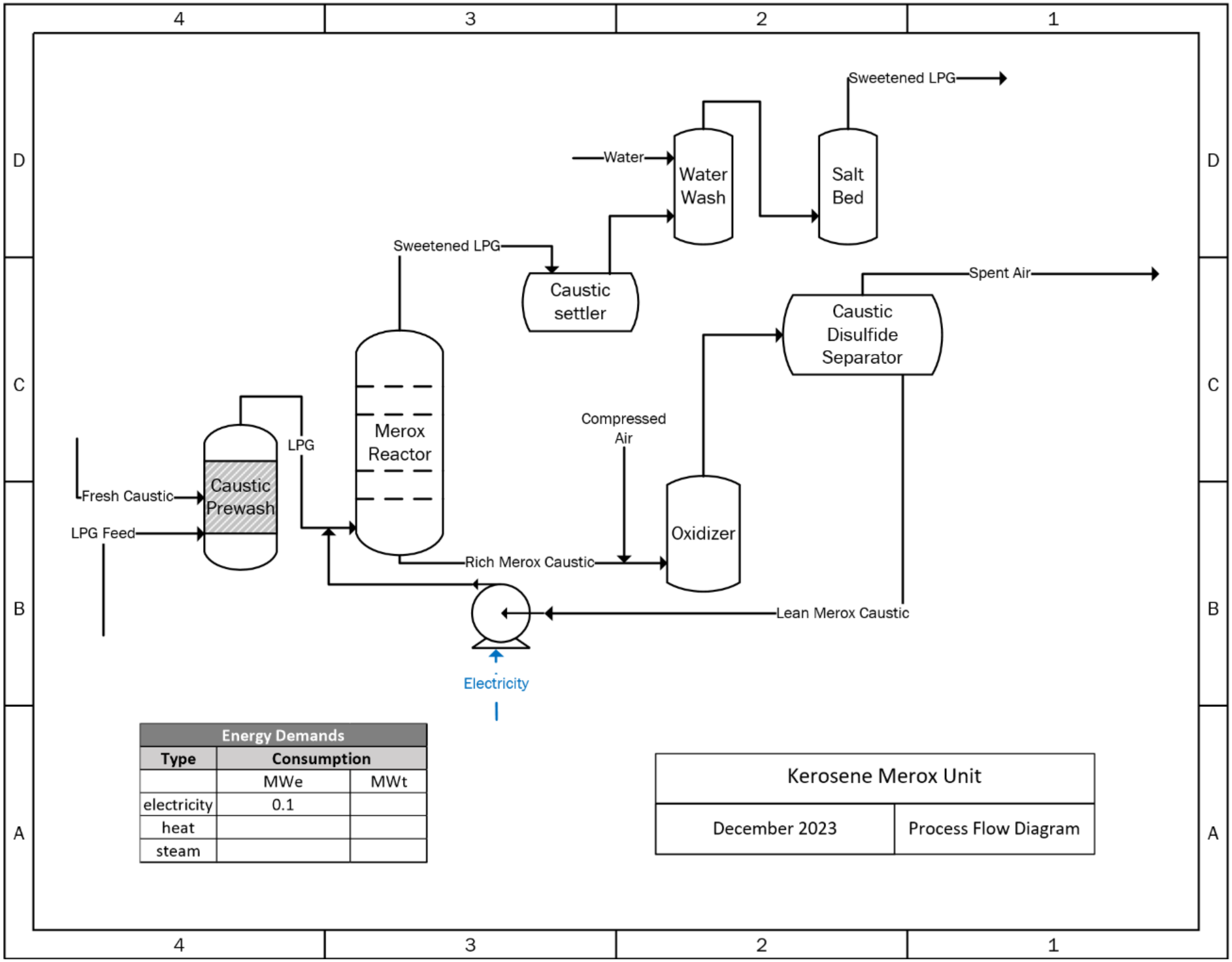
Type	Consumption	
	MWe	MWt
electricity	2.5	
heat		36.6
steam		2.7

Kerosene Hydrotreater

December 2023

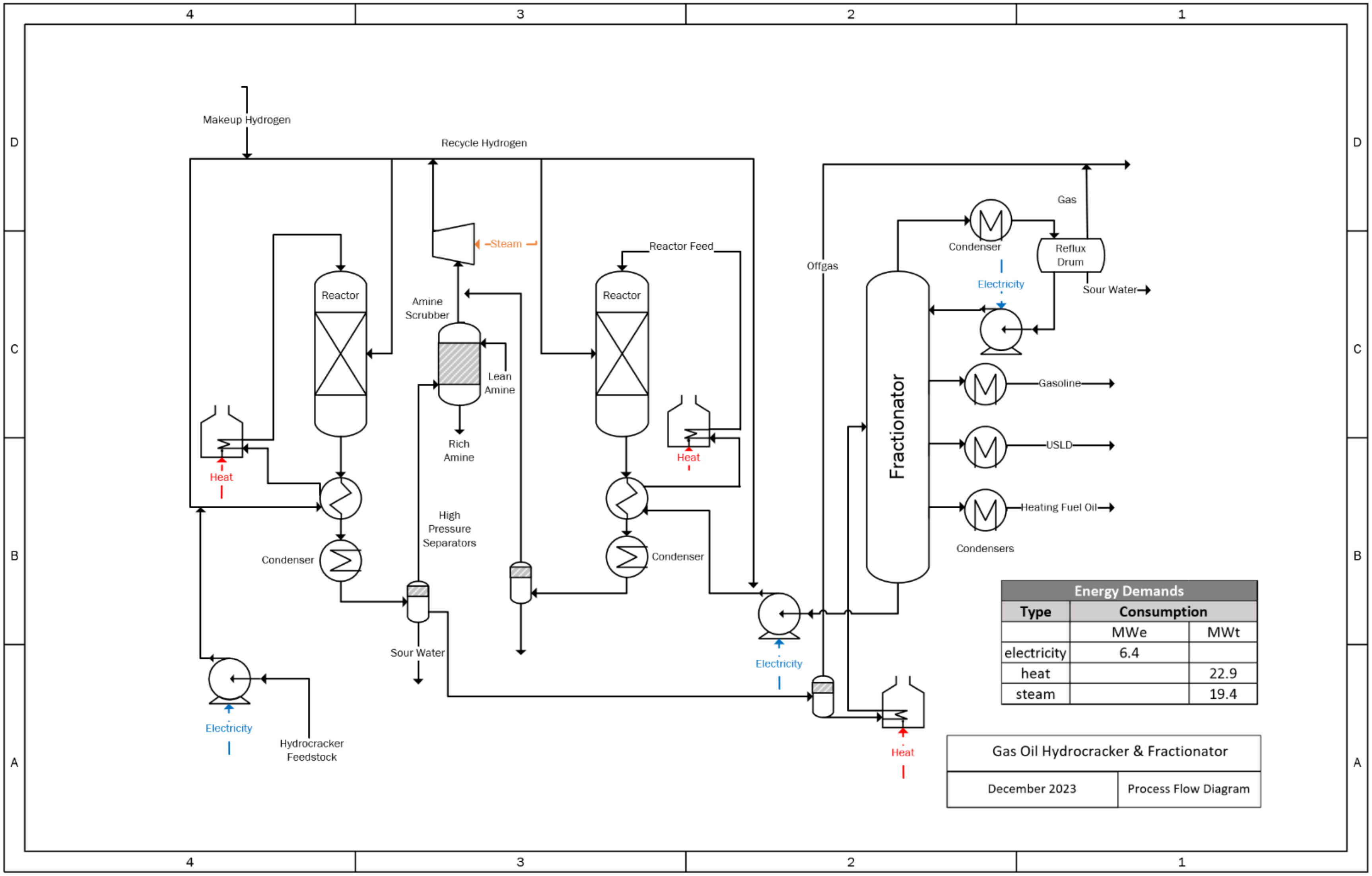
Process Flow Diagram

Figure D7. Continued.



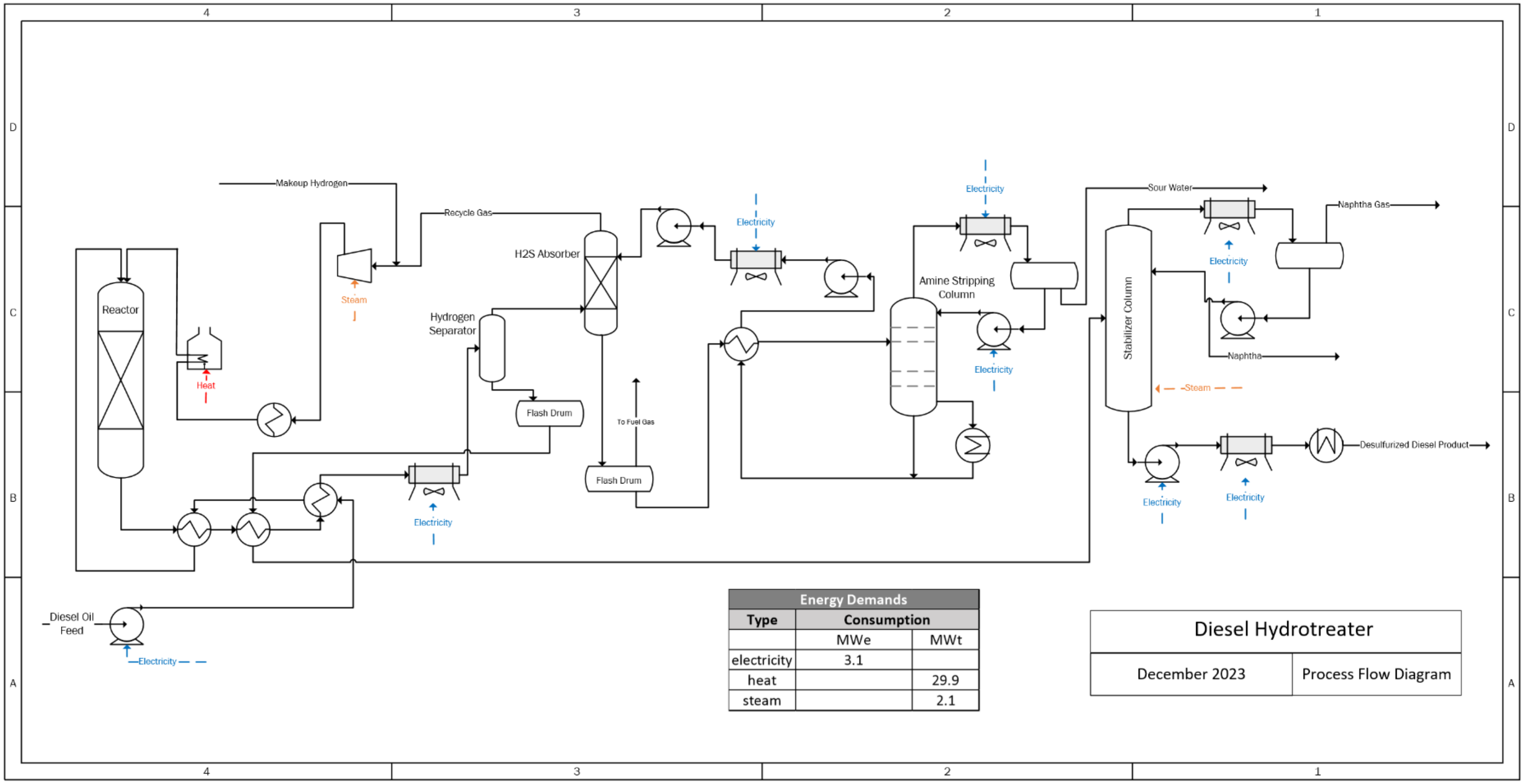
Energy Demands		
Type	Consumption	
	MWe	MWt
electricity	0.1	
heat		
steam		

Figure D7. Continued.



Energy Demands		
Type	Consumption	
	MWe	MWt
electricity	6.4	
heat		22.9
steam		19.4

Figure D7. Continued.



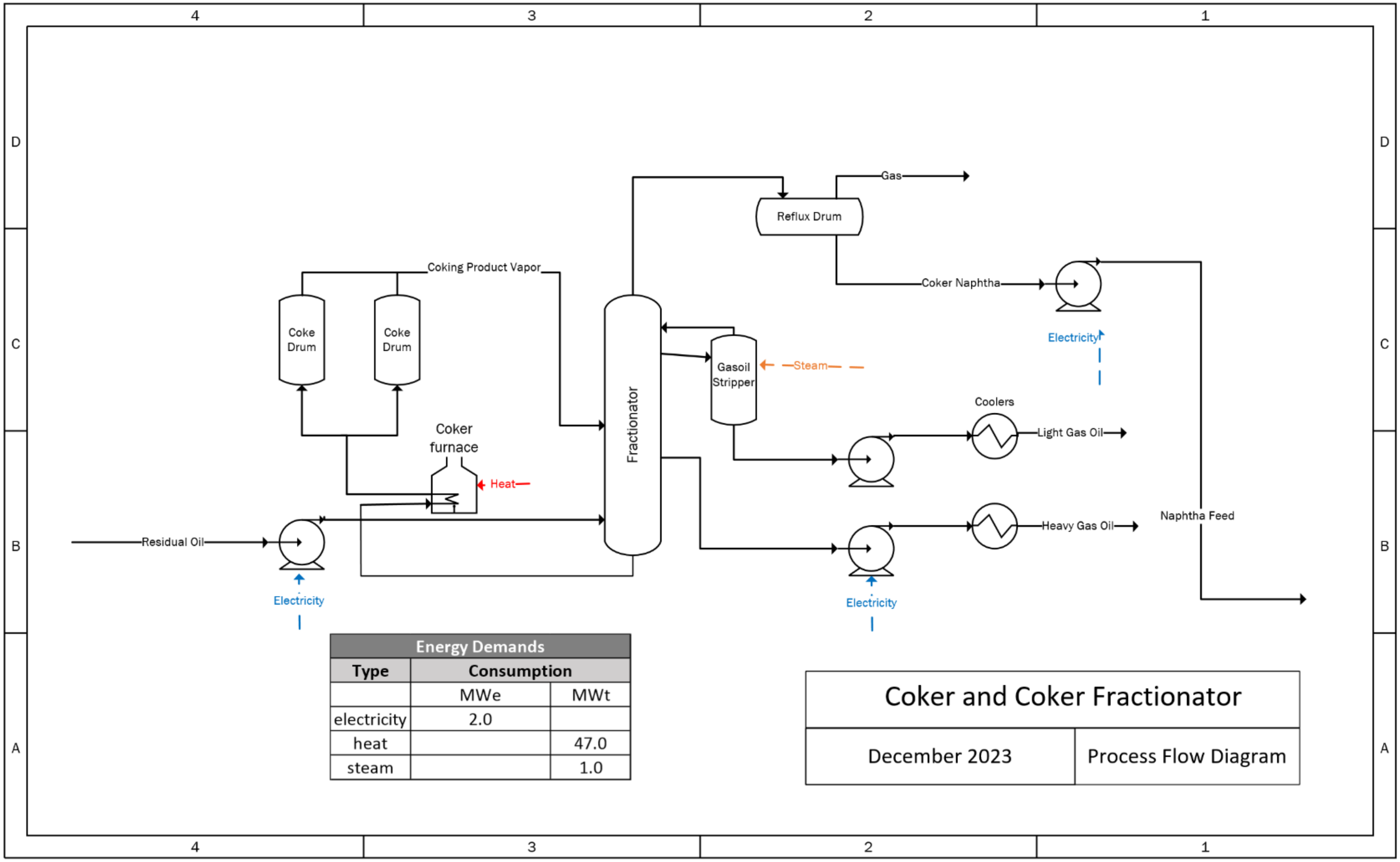
Energy Demands		
Type	Consumption	
	MWe	MWt
electricity	3.1	
heat		29.9
steam		2.1

Diesel Hydrotreater

December 2023

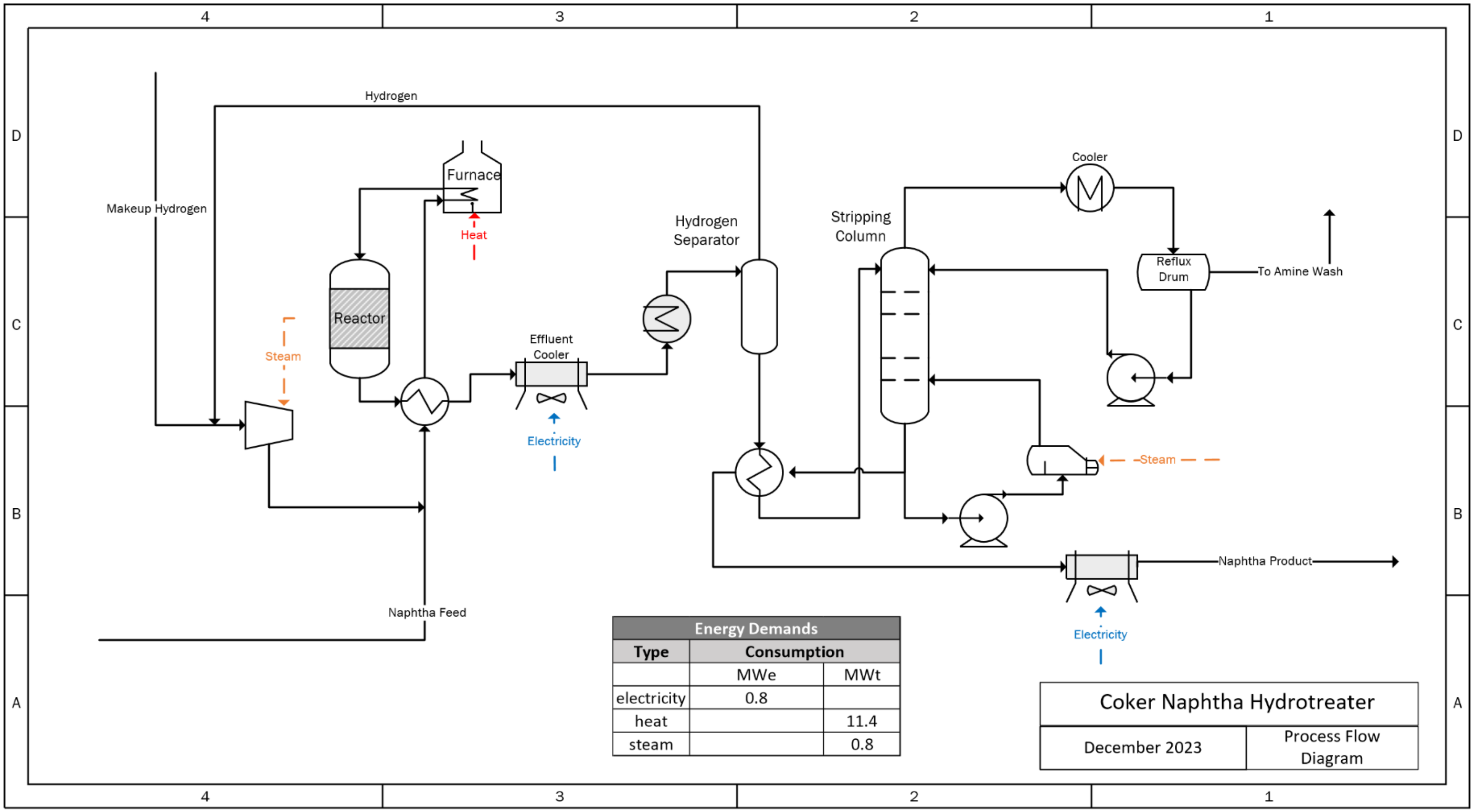
Process Flow Diagram

Figure D7. Continued.



Energy Demands		
Type	Consumption	
	MWe	MWt
electricity	2.0	
heat		47.0
steam		1.0

Figure D7. Continued.



Energy Demands

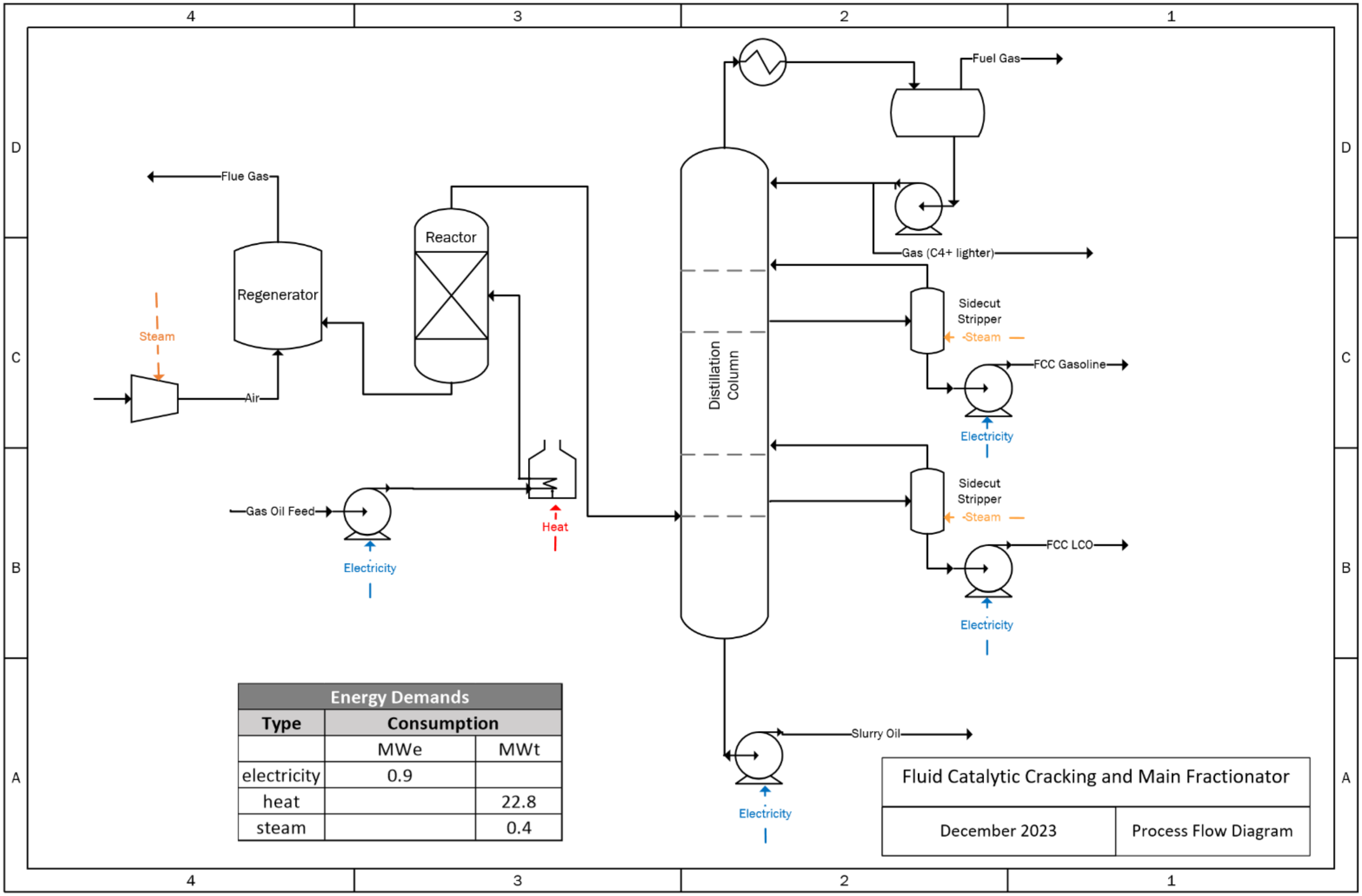
Type	Consumption	
	MWe	MWt
electricity	0.8	
heat		11.4
steam		0.8

Coker Naphtha Hydrotreater

December 2023

Process Flow Diagram

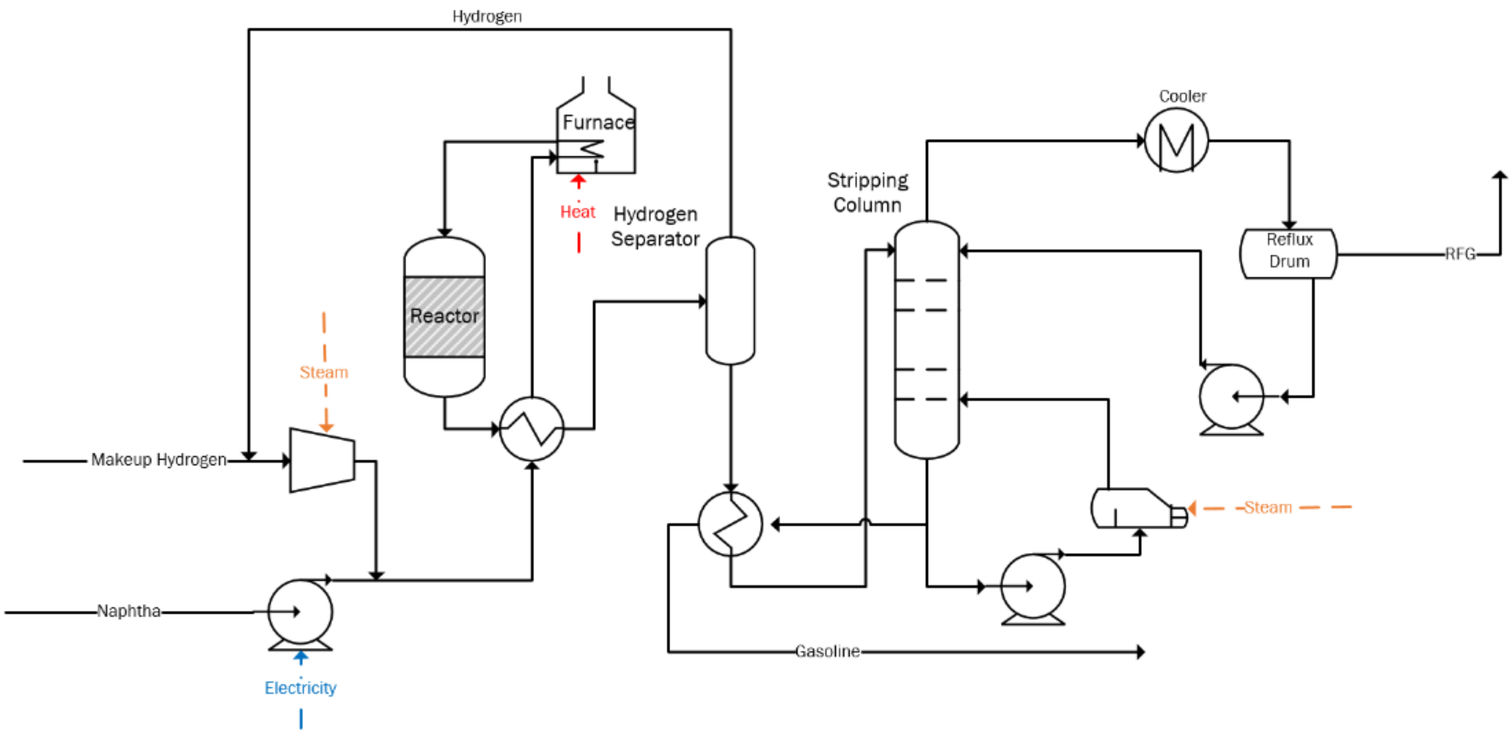
Figure D7. Continued.



Energy Demands		
Type	Consumption	
	MWe	MWt
electricity	0.9	
heat		22.8
steam		0.4

Fluid Catalytic Cracking and Main Fractionator	
December 2023	Process Flow Diagram

Figure D7. Continued.



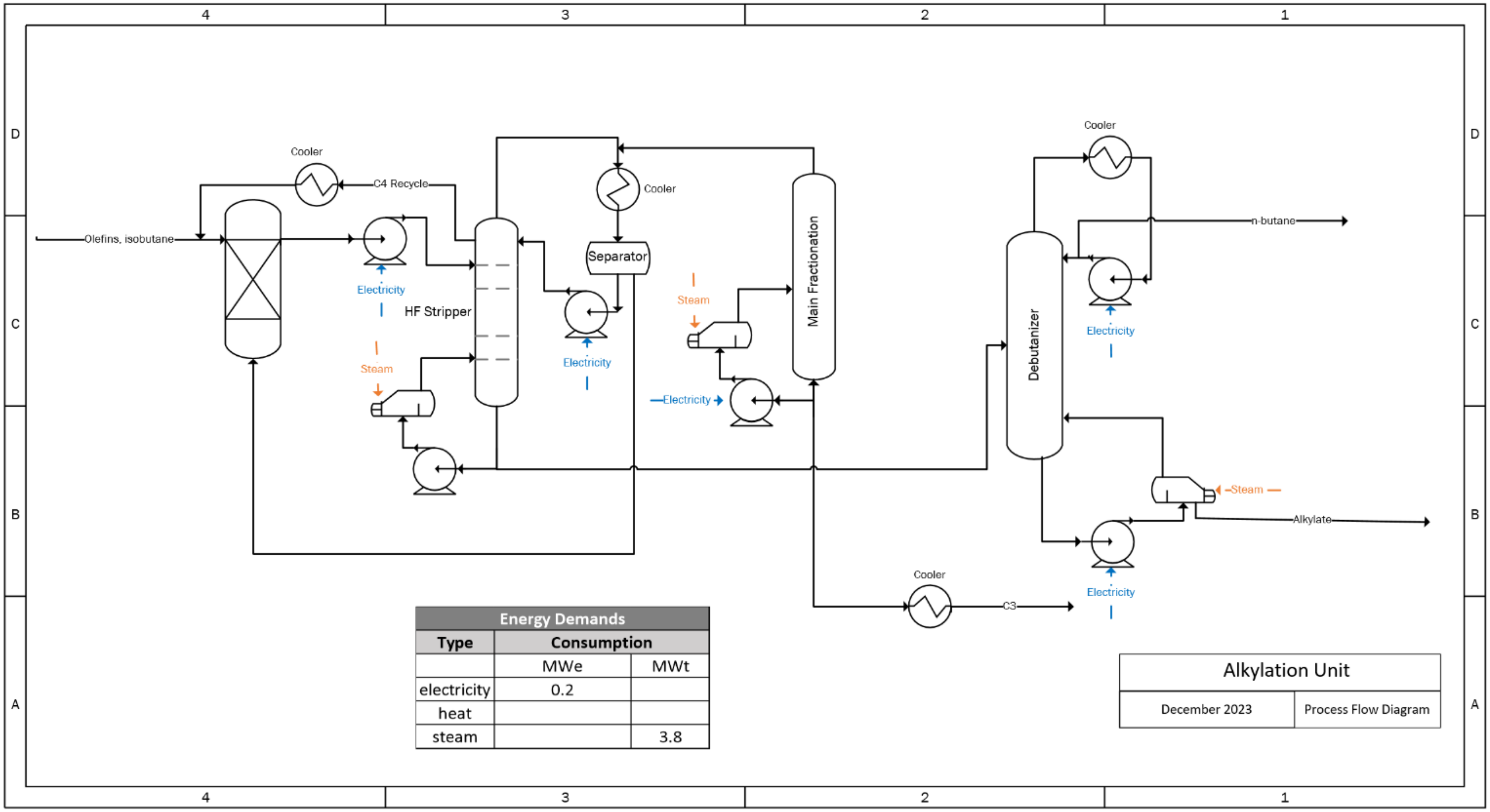
Energy Demands		
Type	Consumption	
	MWe	MWt
electricity	1.0	
heat		14.6
steam		1.2

Fluid Catalytic Cracking Post Hydrotreater

December 2023

Process Flow Diagram

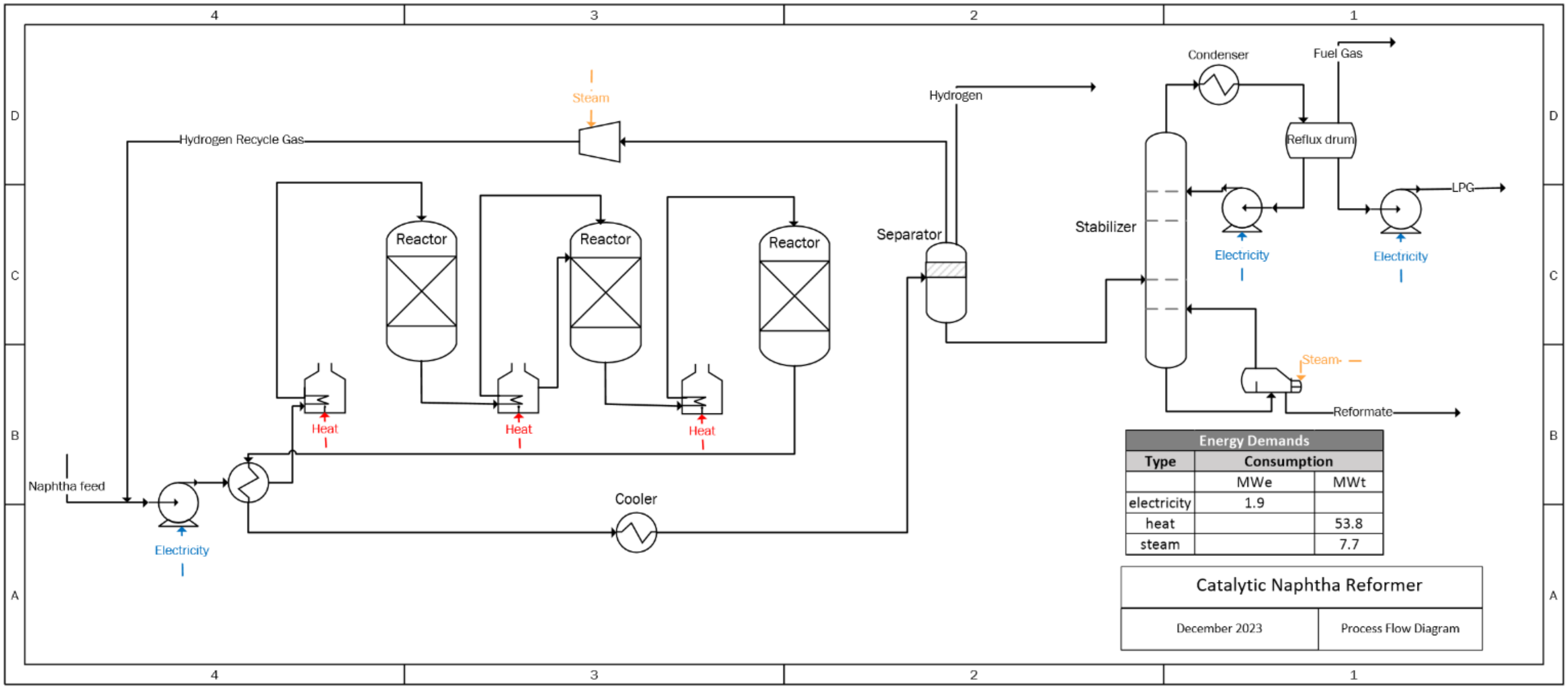
Figure D7. Continued.



Energy Demands		
Type	Consumption	
	MWe	MWt
electricity	0.2	
heat		
steam		3.8

Alkylation Unit	
December 2023	Process Flow Diagram

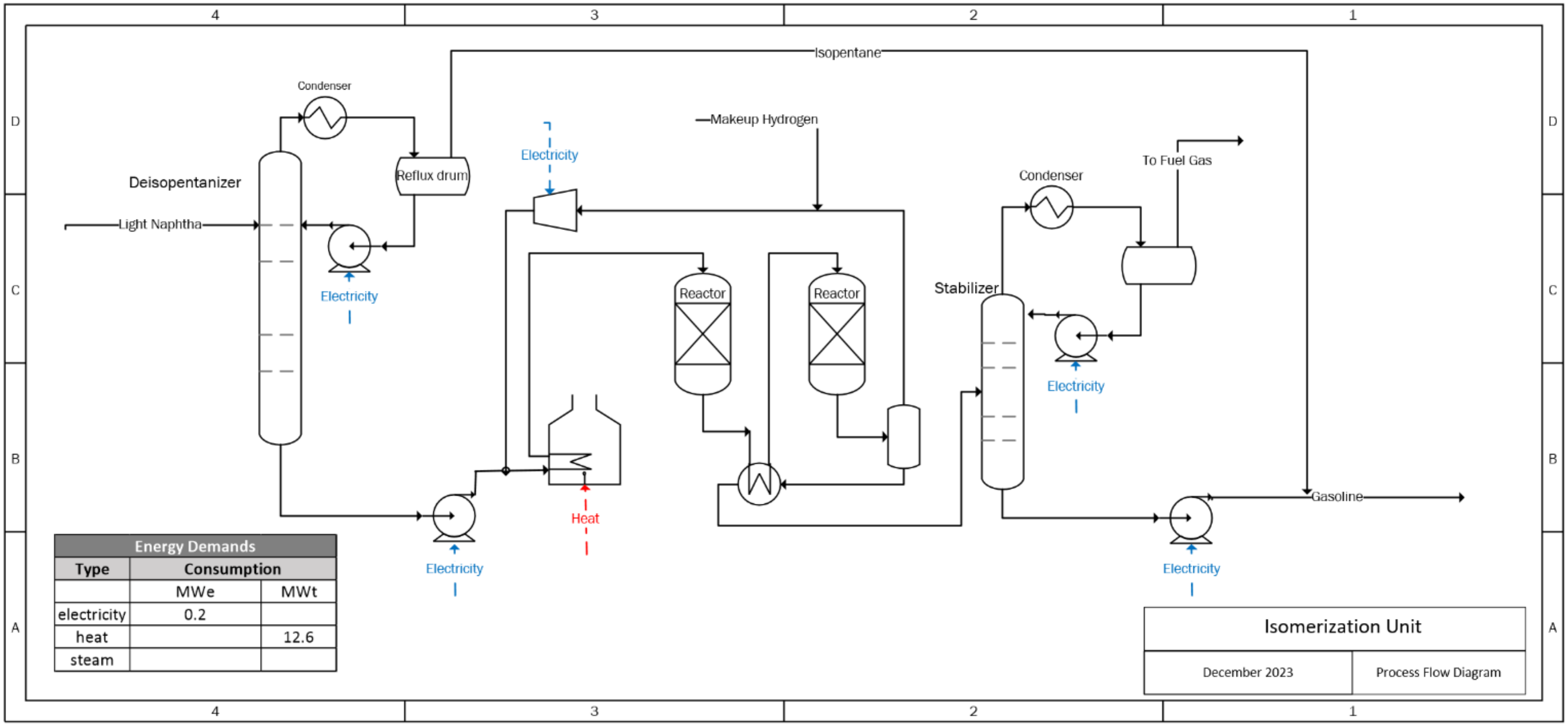
Figure D7. Continued.



Energy Demands

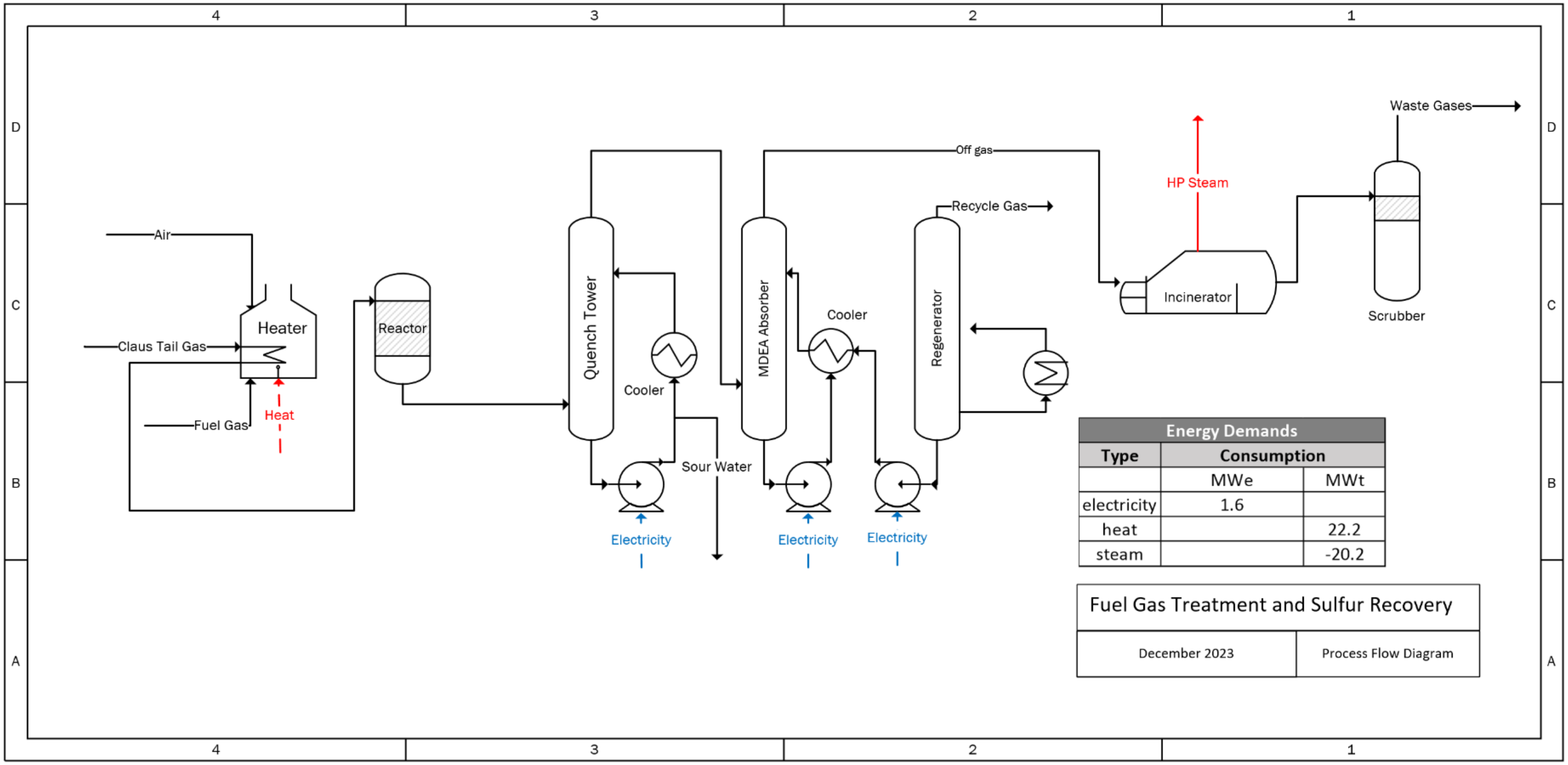
Type	Consumption	
	MWe	MWt
electricity	1.9	
heat		53.8
steam		7.7

Figure D7. Continued.



Energy Demands		
Type	Consumption	
	MWe	MWt
electricity	0.2	
heat		12.6
steam		

Figure D7. Continued.



Energy Demands

Type	Consumption	
	MWe	MWt
electricity	1.6	
heat		22.2
steam		-20.2

Fuel Gas Treatment and Sulfur Recovery

December 2023 | Process Flow Diagram

Figure D7. Continued.

Appendix E

Details of Nuclear Integration Scenarios

E-1. HTGR Power Conversion and CHP Systems

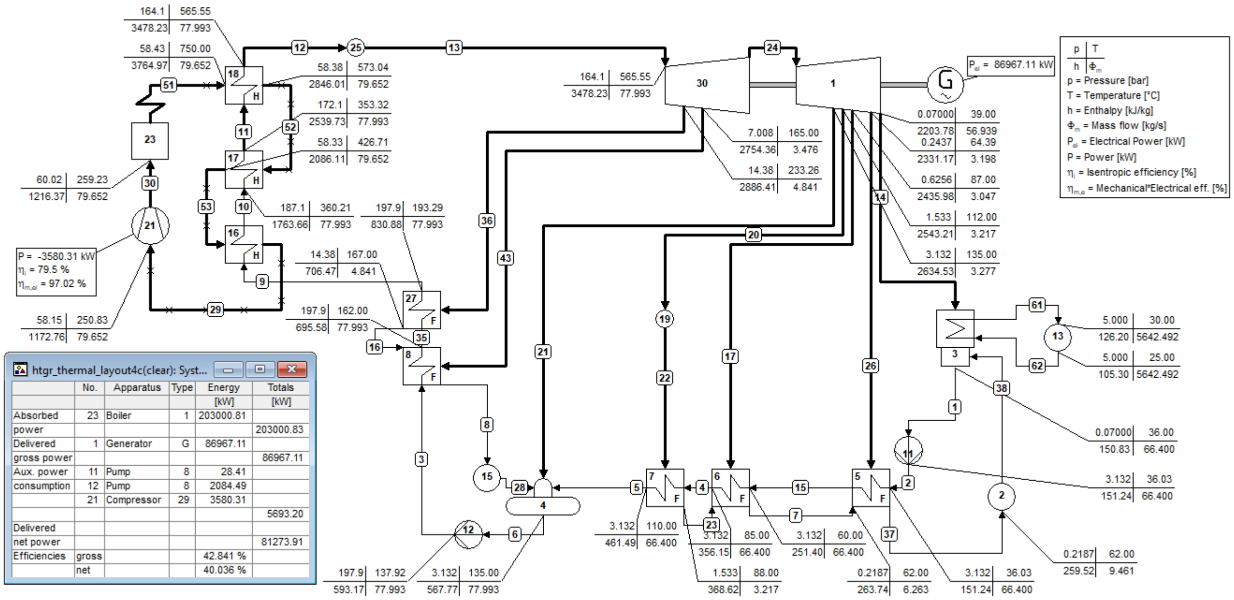
A base-case model was developed in Aspen HYSYS and Cycle-Tempo, based on a generic high-temperature gas-cooled reactor (HTGR). The primary coolant is helium, which circulates between the reactor core and the steam generator. The thermodynamic properties of the helium and steam are summarized in Table E.

Table E42. Thermodynamic properties of HTGR coolants.

Coolant		Temperature (°C/°F)	Pressure (bar/psia)	Phase
Primary Helium	Core Inlet	260/500	60/870	Supercritical
	Core Outlet	750/1382		
Secondary Steam	Steam Generator Inlet	220/428	165/2393	Subcooled Liquid
	Steam Generator Outlet	565/1049		Gas

The base case model was sized with a 200 MWt core to represent a single HTGR unit but is modular and can be scaled to represent a four pack (800 MWt), depending on the needs of each industry. For “power only” operation, this corresponds to 80 MWe (or 320 MWe for a four pack). This study requires individualized CHP systems for each industry; any process-heat usage that cannot be covered by waste heat utilization will result in reduced electricity generation.

As part of another work package, an HTGR steam cycle model was developed and then tuned for the parameters of the generic HTGR. The design parameters for the steam cycle correspond with an existing industrial steam turbine offer. These turbines are offered as modular frames (thus the lower cost), with the customer-specified flow rate, inlet and outlet conditions, and controlled/uncontrolled extractions all within the limits of the specific frames. Considering, for example, a Siemens brochure [69], in which the industrial turbine offer is well described, frame SST-600 fits exactly within the inlet parameter limits and offers up to two controlled and six uncontrolled extractions. Figure E1 shows a reference PFD of the HTGR power conversion system.



	No.	Apparatus	Type	Energy [kW]	Totals [kW]
Absorbed	23	Boiler	1	203000.81	
power					203000.83
Delivered	1	Generator	G	86967.11	
gross power					86967.11
Aux. power	11	Pump	8	28.41	
consumption	12	Pump	8	2084.49	
	21	Compressor	29	3580.31	
					5693.20
Delivered					
net power					81273.91
Efficiencies	gross			42.841 %	
	net			40.036 %	

Figure E1. HTGR reference plant.

For multiple safety and regulatory reasons as well as to avoid using a nuclear power cycle steam with specified chemistry and purity levels for industrial process, isolation exchangers are expected to be between the nuclear plant and industrial process steam. Industrial process steam has generally lower purity and chemical treatment requirements, often has a significant make-up rate (losses, injection into process) and there is high risk of steam and condensate contamination from the industrial processes.

As condensation on the primary side and boiling on the secondary side have high heat transfer coefficients, the size of the exchanger would be relatively moderate. A drum type of heat exchanger is the highly recommended configuration from the control and physical operation aspect, since nuclear-sourced steam is condensing while process steam is boiling. For enhanced effectiveness, a heat exchange network that preheats and superheats process steam (through desuperheating and subcooling of the extraction steam) may be added.

The nuclear integration scenarios are expecting a significant difference between process steam source and refinery headers, and thus a different approach is taken for deaerator placement than might be expected. Rather than a deaerator on the industrial site, preferably cold condensate is pumped to the nuclear site (lower losses, possibility of atmospheric buffer tanks) and the deaerator is placed within the set of isolation heat exchangers. A schematic configuration of this system of components is provided in Figure E2. Note that all these components together will be represented in the following PFDs as a single heat exchanger.

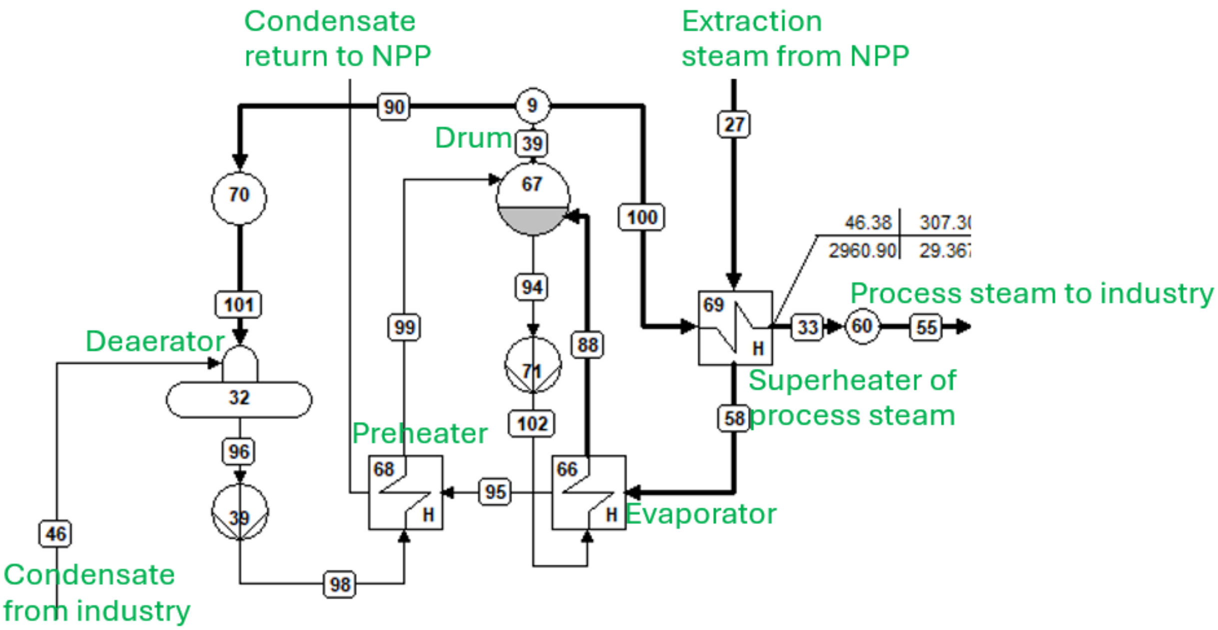


Figure E2. Steam generation system at a nuclear power plant site with a deaerator, preheater, drum boiler and superheater.

E-2. Refinery Integration

E-2.1 Case 1 and 1a

Using the previously developed process steam model and reference steam cycle, a steam extraction has been added to the steam cycle to supply the process heat. Figure E3 illustrates the integrated system for the supply of steam only. Overall energy balance is then presented along with other scenarios at the end of this section.

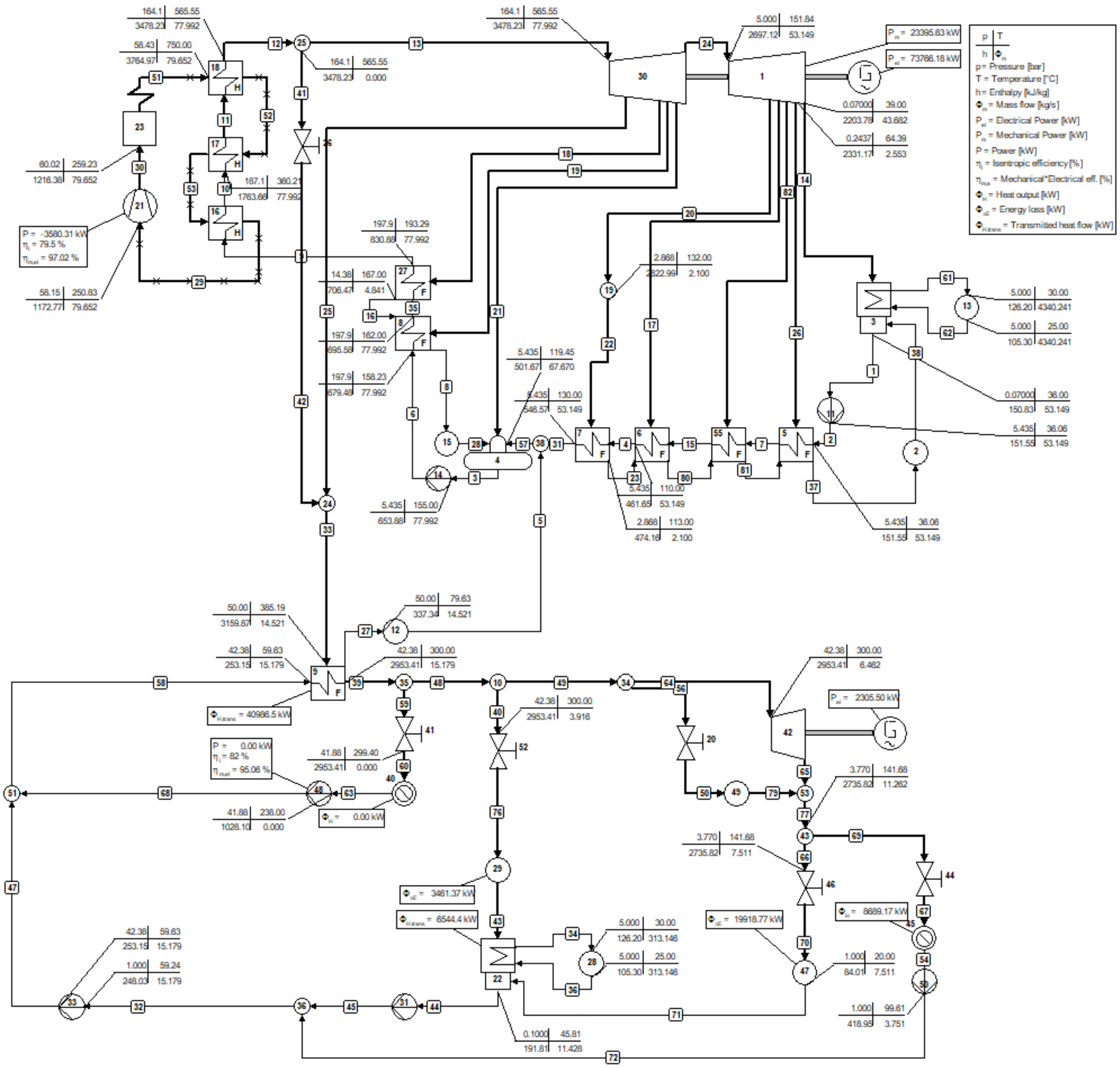


Figure E3. PFD of a nuclear-integrated CHP system in a refinery—for process steam and power supply only (Case 1).

When looking at the net demands of the reference refinery process steam system in Figure E4, note that the net flow to the HP steam reboilers is zero. In our reference refinery system, this is caused by internal steam production, primarily from the sulfur treatment system. HP steam is then used only for turbines used for direct drives of selected compressors and pumps. This suggest another nuclear integration path into the process steam system, as is suggested in the PFD, delivering only LP saturated steam from the nuclear CHP system over an extended distance, which is likely to be present between the nuclear and industrial sites. The direct drive turbines are then substituted with electric motors, in the reference refinery amounting for 3.7 MWe (for conservative demand results assuming high turbine efficiency of 90%). For system startup or transient states with insufficient HP steam production, a steam compressor is added between the LP steam and HP steam line with steam reboilers. For a steam compressor application with usually no HP steam demand, a more in-depth analysis should be done on feasibility against a small backup fossil fuel boiler. Also, note that the 2 MWe originally produced by the letdown turbine are not available, but more power is produced in the main turbine and with better efficiency.

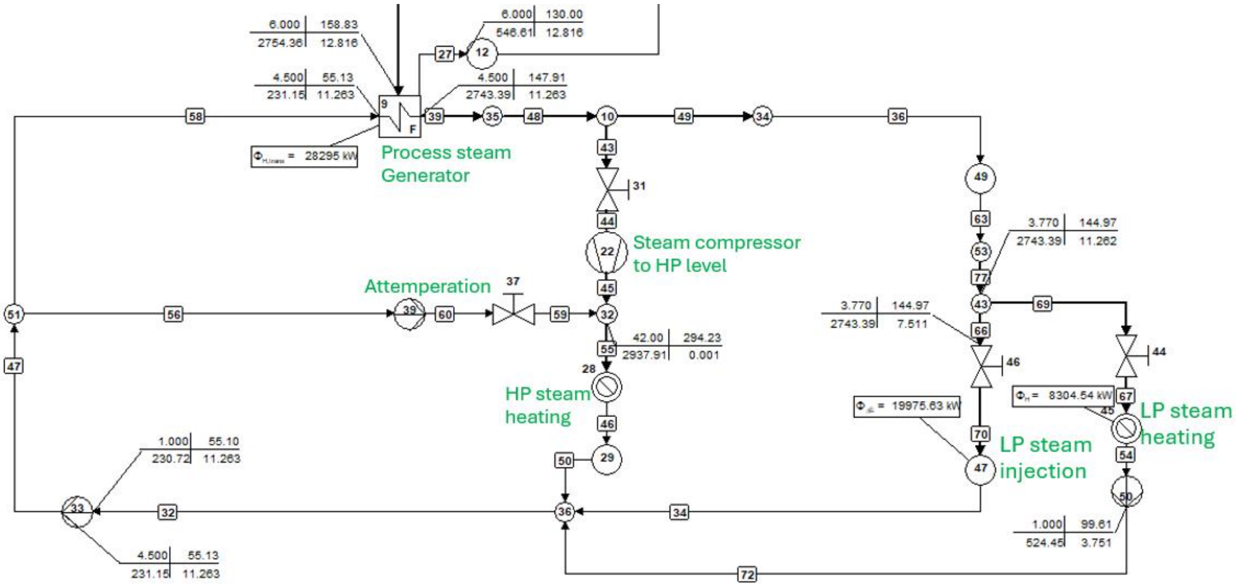


Figure E4. Reference refinery process steam PFD after supplying only LP steam.

Overall implementation into the reference HTGR plant follows same CHP strategy as for the previous system, only with steam extraction from the NPP at lower pressure, as shown Figure E5. By delivering saturated, rather than superheated steam, the heat transfer station between the nuclear plant steam cycle and process steam is further simplified. The LP of the components then has the potential for cost reduction. Applicability of this concept to specific refineries then depends on actual net HP steam demand for heating purposes and actual system costs. The overall energy balance is again provided in a summary at the end of this section.

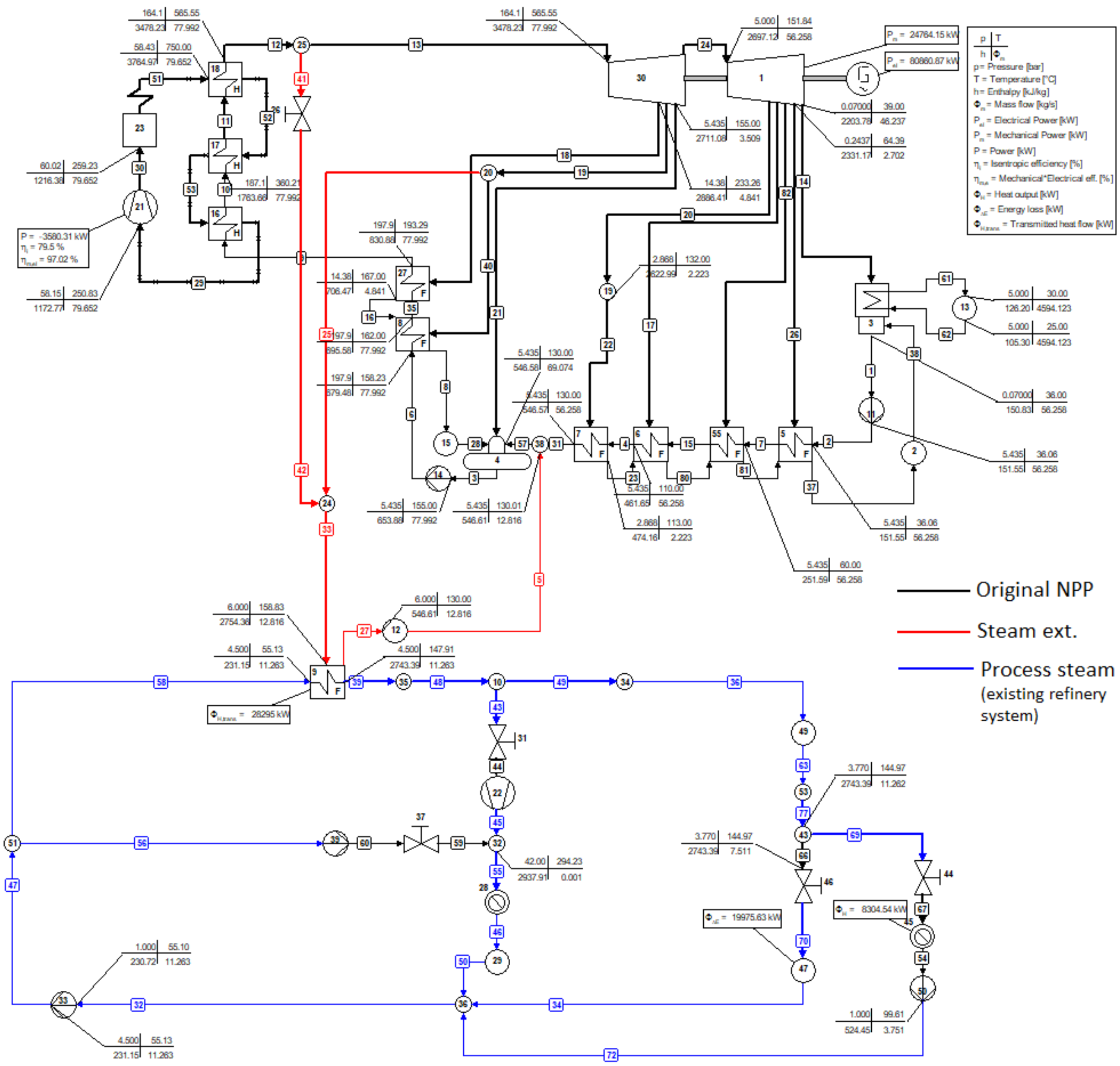


Figure E5. Overall integration PFD for Case 1a.

E-2.2 Case 2 and 2a

For the following work of nuclear integration with H_2 production by steam methane reforming substituted by solid oxide electrolyzers, named here as high-temperature steam electrolysis (HTSE), was developed a modified process steam system, which is shown in Figure E6. In this scenario, steam production from the steam methane reformed becomes zero, and instead, a new demand appears for LP steam (1.4 bar with assumptions taken here). This steam, drawn from the LP header via a small letdown turbine is required to evaporate the deionized water. A small superheat is then provided by HP steam. The high temperature inside the HTSE stack around 750°C is then achieved by internal heat recuperation and this temperature provided to the HTSE is sufficient. The current refinery H_2 distribution system operates at around 2 MPa, and HTSE provides H_2 at near atmospheric condition, so additional power is included for an H_2 compressor. Note that without the fuel demand for the steam methane reformer, refinery fuel gas production can then cover most of the overall fuel demand, with a net requirement of only 3 MW in external fuel.

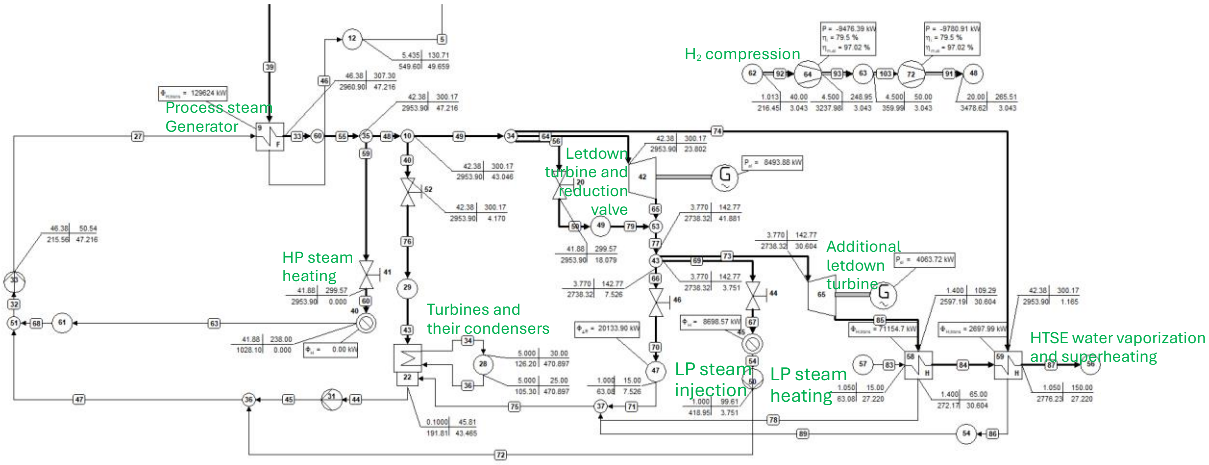


Figure E6. Simplified process steam system for the refinery, with additional components for H_2 production.

A PFD of the entire system in Figure E15 shows the same steam integration point by implementing a turbine extraction upstream of the feedwater heaters as in Case 1, only with a higher extraction flowrate.

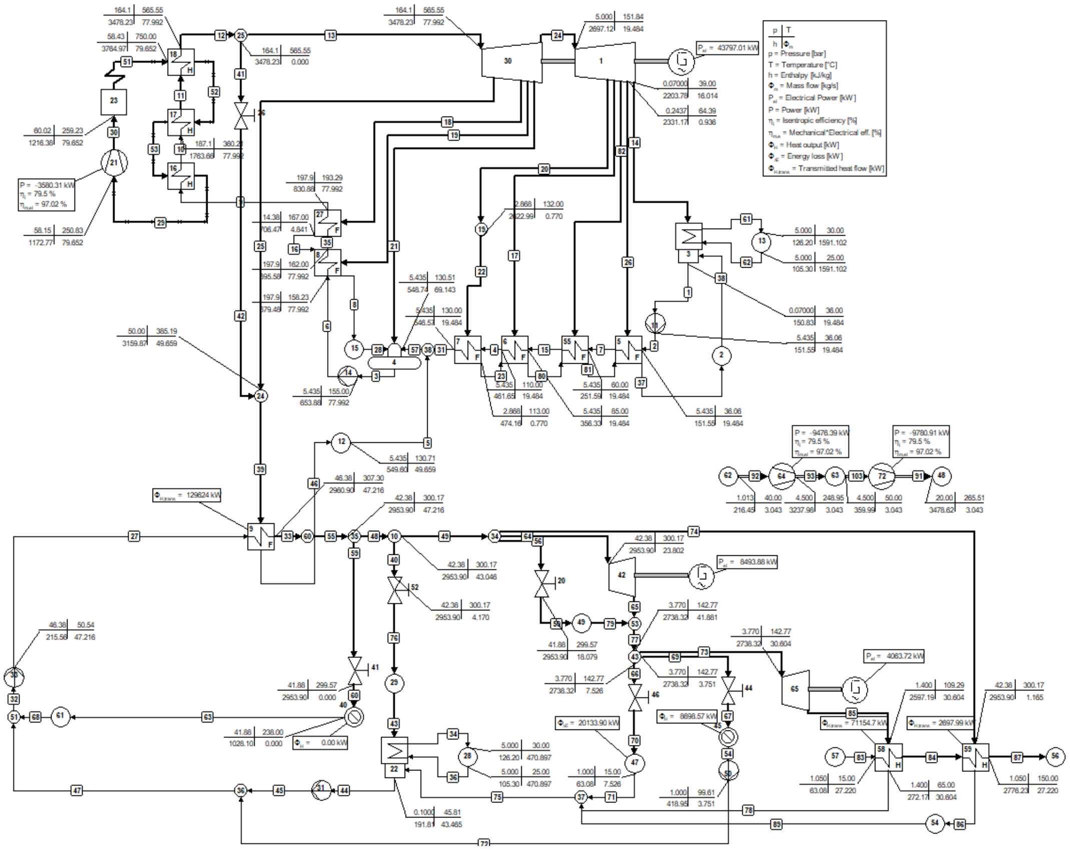


Figure E7. PFD of a nuclear-integrated CHP system in a refinery—for process steam and the H_2 supply generated via HTSE.

Similarly to Case 1a, the LP steam delivery approach can also be used when the HTSE system is employed. The layout of the process steam system is in Figure E8. A steam compression system allowing HP steam production in periods of insufficient internal supply, such as start-ups and possibly transient states, is included in the layout but with zero flow.

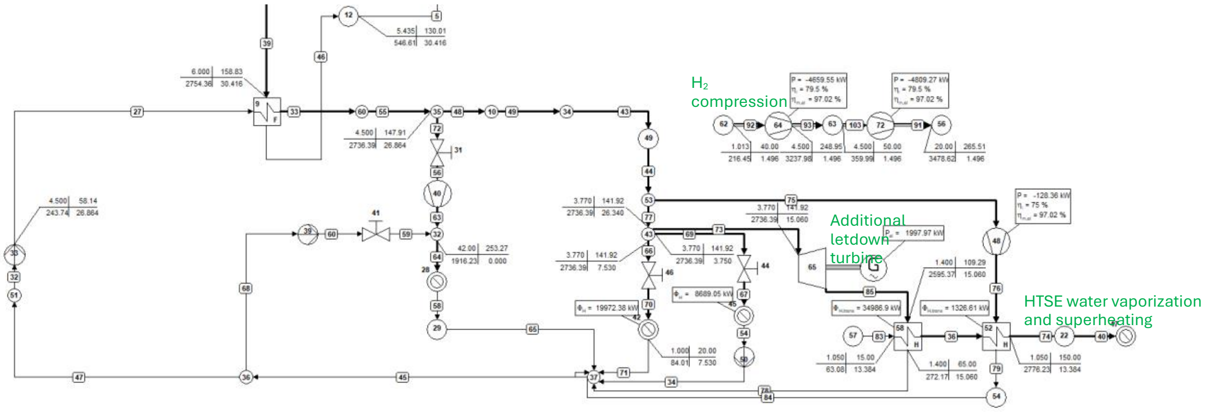


Figure E8. Simplified process steam system for the refinery with external supply of only LP steam, with additional components for H_2 production.

E-2.3 Summary with 200 MWth HTGR Modules

In all cases, a single reactor module with a CHP cycle is sufficient to provide all the process steam. For resilience purposes, steam extraction may be, however, split between different units or multiple units may be configured for steam extraction and serve as a backup. Table E43 summarizes the steam cycle parameters. Table E44 gives integration scenarios corresponding to the number of reactor systems, the steam cycles, and the overall electricity balance.

Table E43. Parameters of HTGR systems with different CHP configurations.

	W_{net} (MW)	Q_{steam} (MW)	Electrical Efficiency	Heating Efficiency	CHP Efficiency
Baseline cycle	81.3	0.0	40.0%	0.0%	40.0%
CHP cycle, HP steam only	70.3	38.8	34.6%	19.1%	53.7%
CHP cycle, HP steam plus HTSE	50.3	76.0	24.8%	37.4%	62.2%
CHP cycle, HP steam plus HTSE+H_2 HT heat	31.1	113.5	15.3%	55.9%	71.3%
CHP cycle, LP steam only	75.2	28.2	37.0%	13.9%	50.9%
CHP cycle, LP steam plus HTSE	59.1	64.5	29.1%	31.8%	60.9%

Note that in the overall system, a net export affords advantages for cases involving maintenance on the nuclear system of one of the reactor systems. It is also suggested that, for contingency purposes, there would be auxiliary interconnections between reactor units and steam cycles, enabling the CHP cycle to be operated even when its dedicated reactor is undergoing maintenance. For similar reasons, when the CHP cycle is undergoing maintenance, the reduction and desuperheater station from the steam between the steam generator and the turbine ensures an uninterrupted steam supply.

Table E44. Resulting parameters of HTGR integration scenarios involving a reference refinery.

	Scenario	CHP Module Count	Baseline Module Count	Electricity Demand (MW)	Steam Supply (MW)	CHP System Power Supply (MW)	Baseline System Power Supply (MW)	Net Electricity Balance (MW)
1	Steam + Power only	1	0	28.3	38.8	70.3	0.0	41.9
1a	LP Steam + Power only	1	0	31.8	28.2	75.2	0.0	43.3
2	Steam + H_2, net export	1	3	225.3	76.0	50.3	243.8	68.9
2	Steam + H_2, net import	1	2	225.3	76.0	50.3	162.5	-12.4
2a	LP Steam + H_2, net export	1	3	228.9	64.5	64.5	243.8	79.4
2a	LP Steam + H_2, net import	1	2	228.9	64.5	64.5	162.5	-1.9
3	Steam + H_2 + HT heat electrif.	1	5	410.9	76.0	50.3	406.4	45.8

3a	Steam + H_2 + HT heat via H_2	1	5	431.5	113.5	31.1	406.4	6.0

E-3. Methanol Plant Integration

E-3.1 Case 1

The production of H_2 for Case 1 is assumed to come out of the nuclear-integrated HTSE system. The extracted heat is then used to evaporate and slightly superheat LP demineralized water serving as an HTSE feedstock, while the HTSE also consumes the produced electricity for electrolysis. A dedicated H_2 production unit using the 200 MWth HTGR, as shown in the Figure E9, produces 0.598 kg/s of H_2 (71.7 MW LHV). Unlike H_2 production for the refinery, since it only serves as a fuel for combustion, the compression system and its power consumption are omitted here. The MeOH furnace requires 175.2 MW (LHV) of hydrogen (already including stack loss), resulting in a need of 489 MWth in reactors, or having three units and exporting 44 MWe of excess electricity.

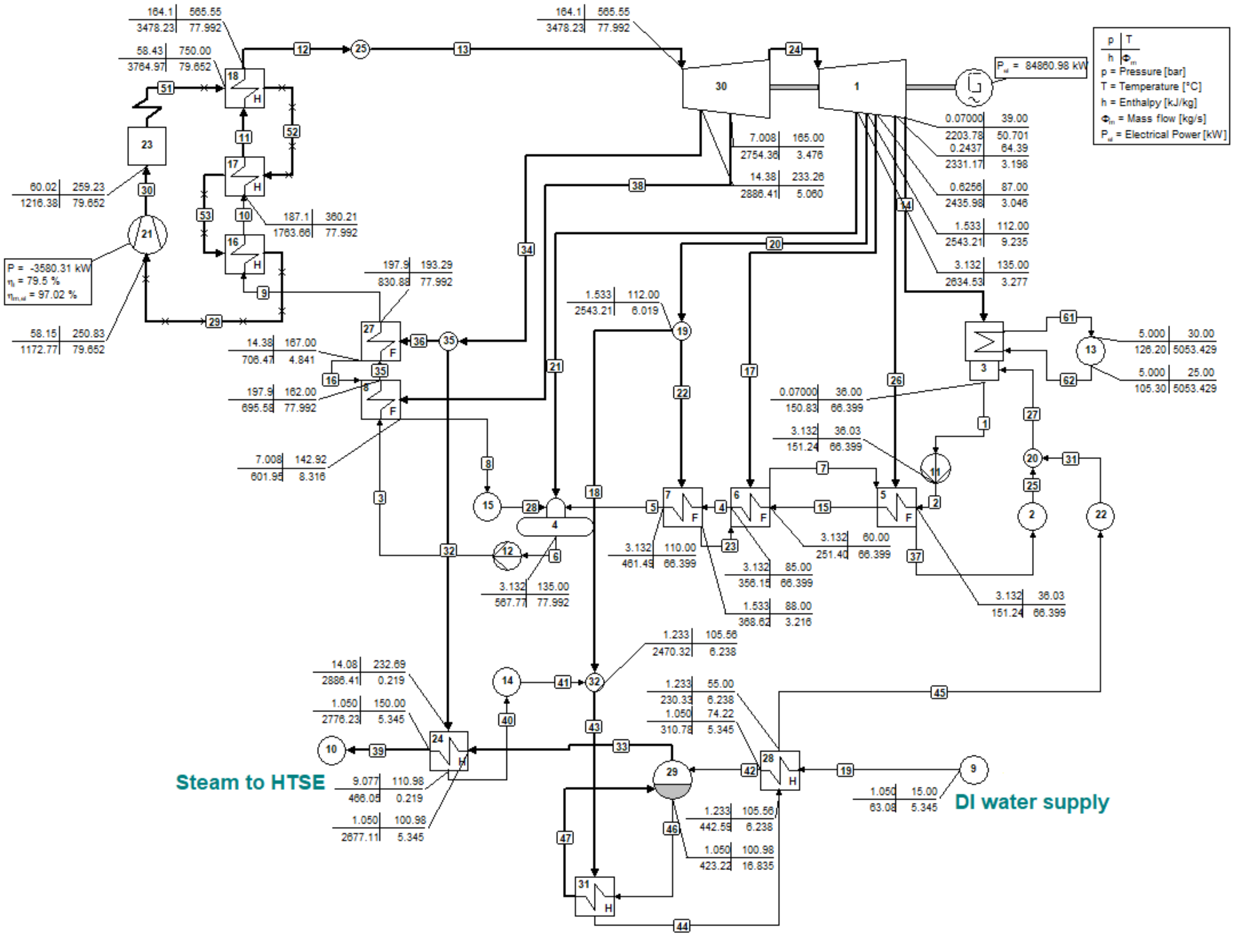


Figure E9. 200 MWth HTGR with heat extraction for HTSE and tuned as dedicated to H_2 production with net-zero export or import.

E-3.2 Case 2

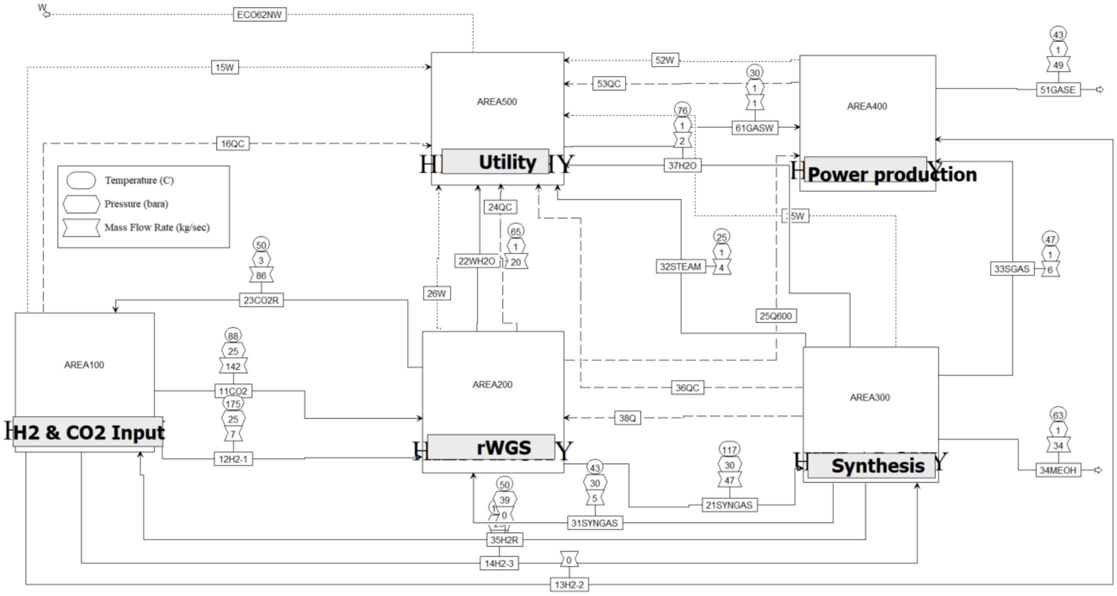


Figure E10. RWGS methanol model blocks (1 MMT/yr).

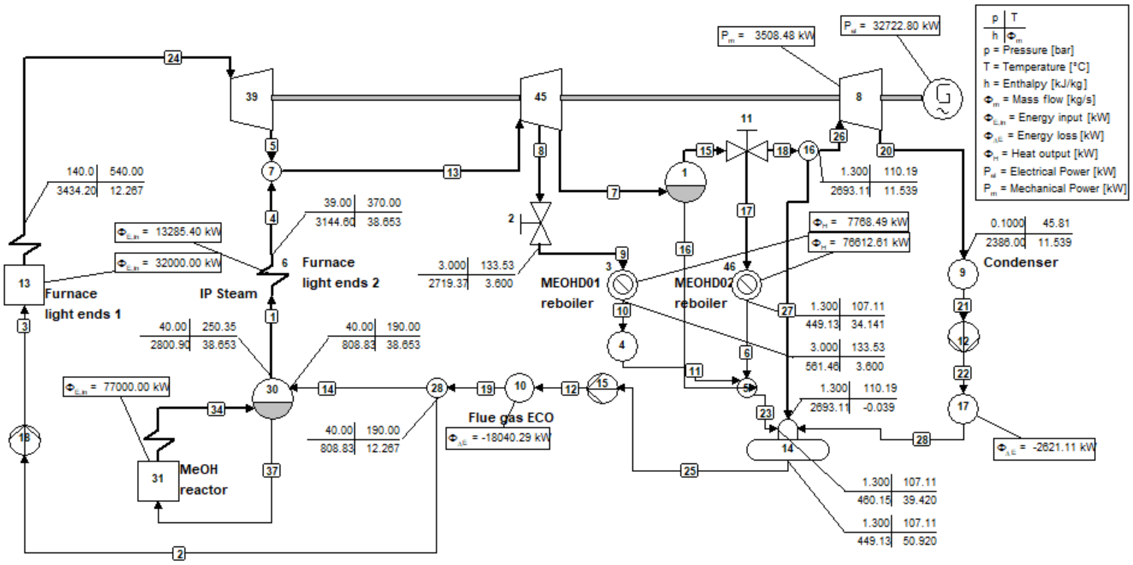


Figure E11. PFD of heat recovery for the RWGS methanol plant.

E-4. Pulp Mill Integration

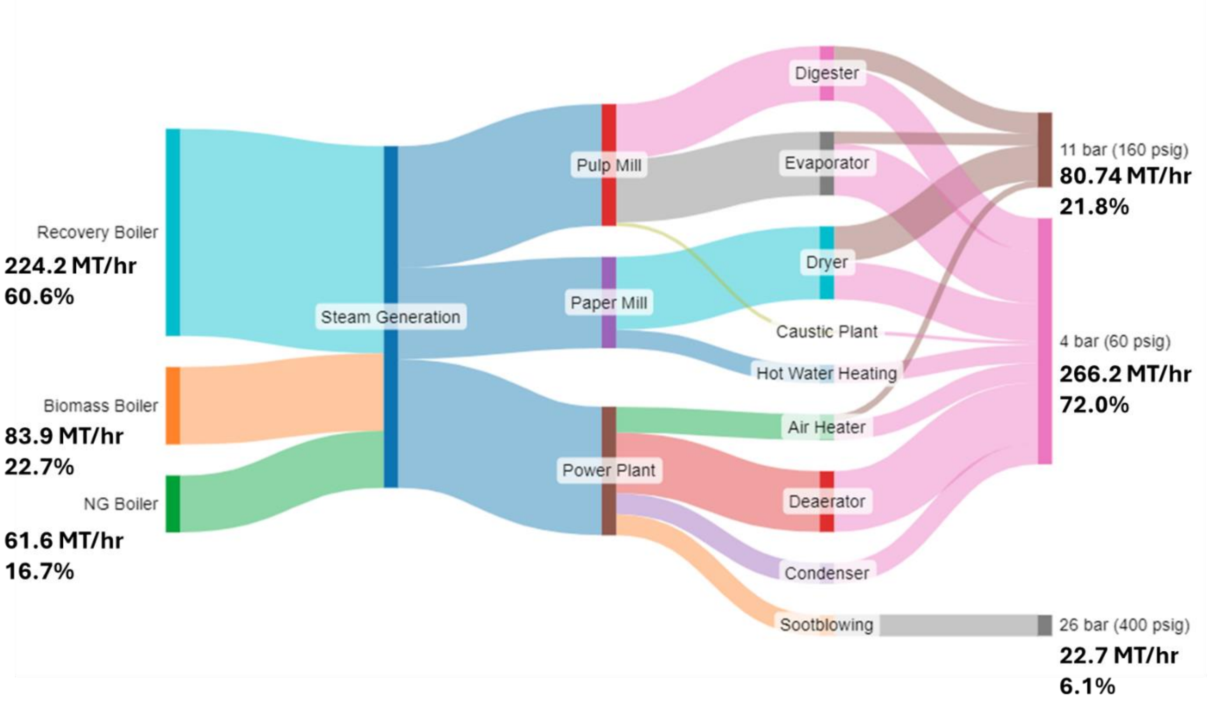


Figure E12. Steam production and use at a 500,877 air-dried metric tons unbleached softwood kraft pulp mill.[b]

---

[b] Based on communications with Dr. Sunkyu Park and Edgar Carrejo at North Carolina State University, July 2024.

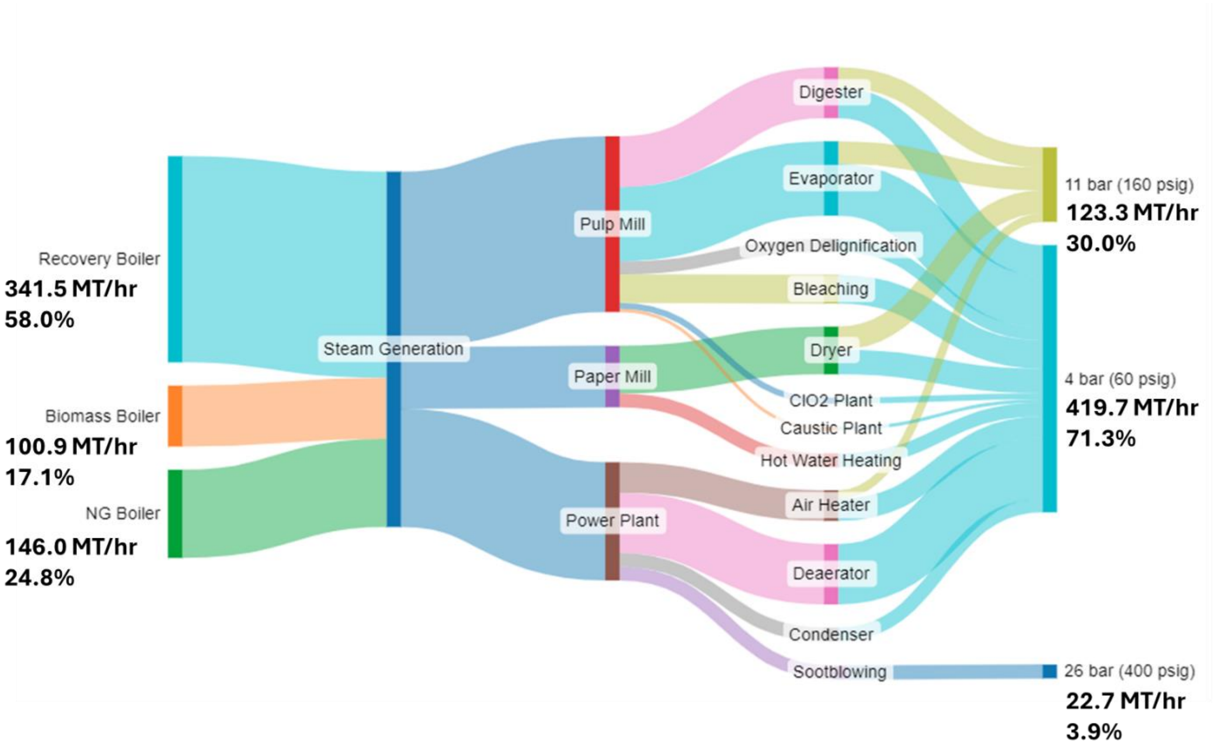


Figure E13. Steam production and use at a 500,877 air-dried metric tons unbleached softwood kraft pulp mill.[d]

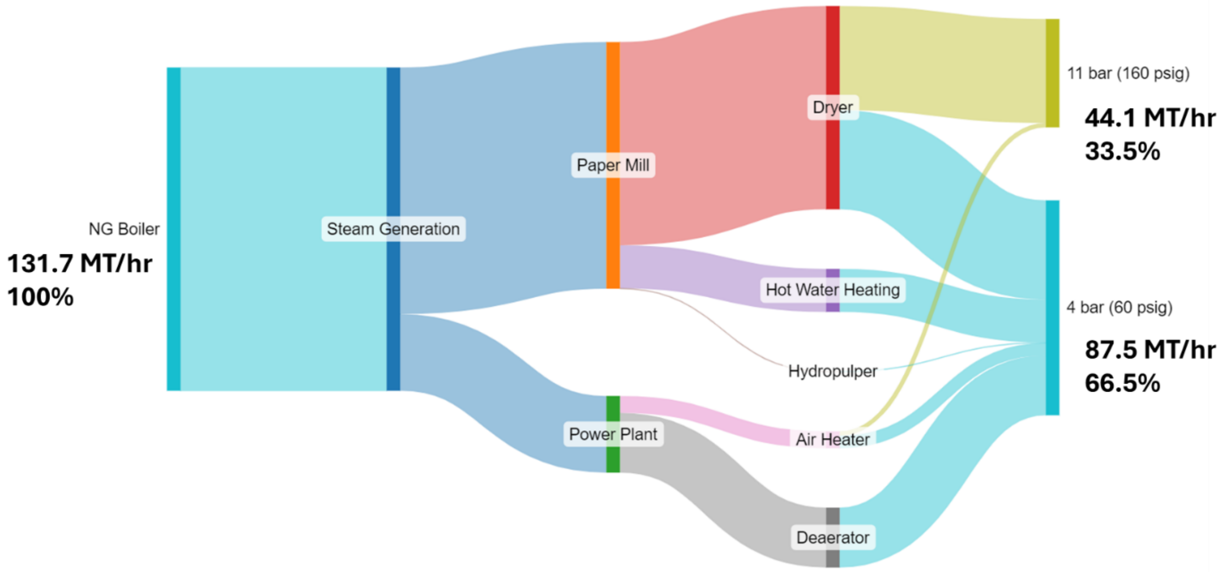


Figure E14. Steam production and use at a 500,877 air-dried metric tons per year paper recycling mill.[d]

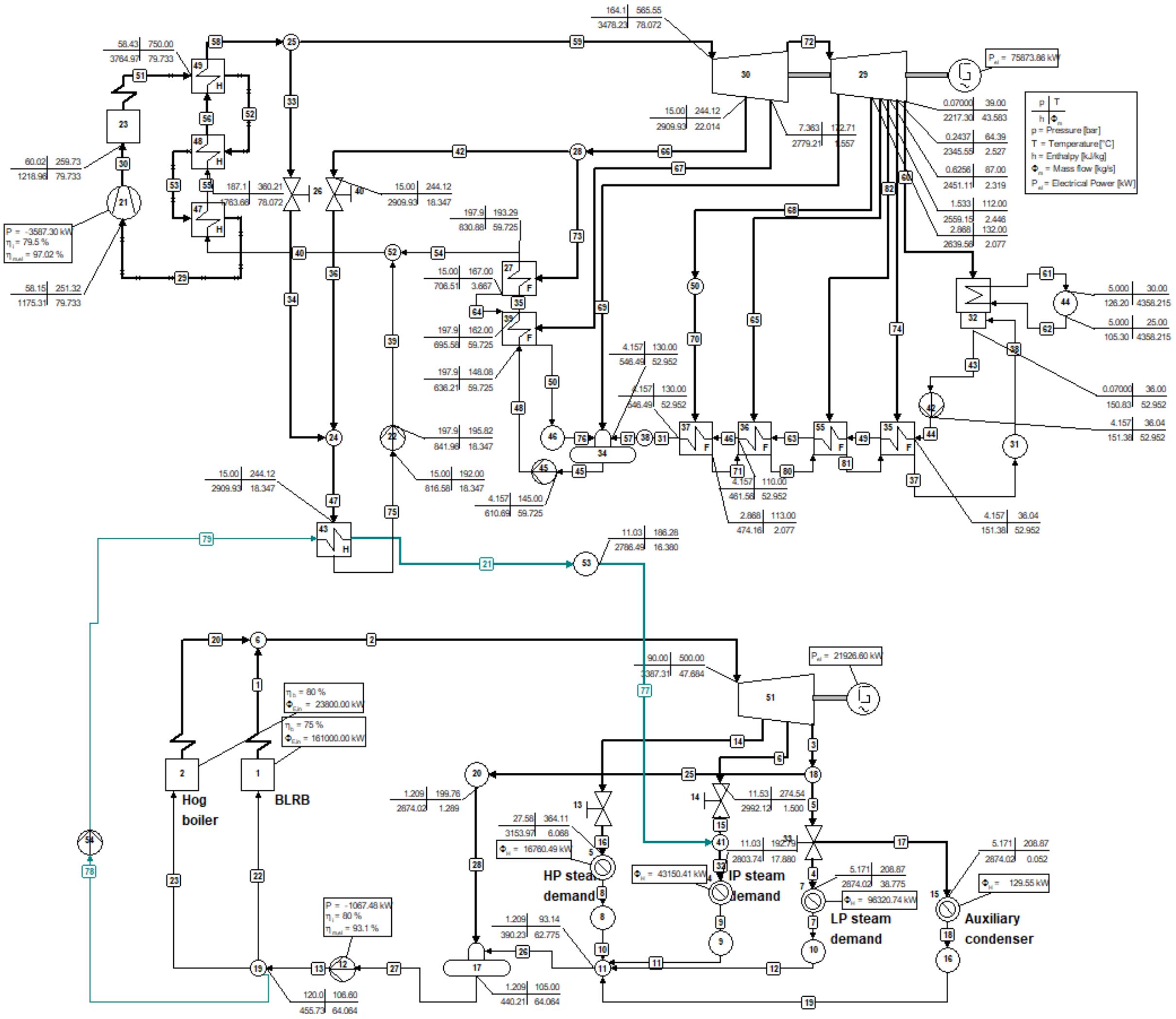


Figure E15. Process diagram of nuclear-sourced process steam integration into a reference pulp mill CHP system (nuclear-sourced process steam system in teal).

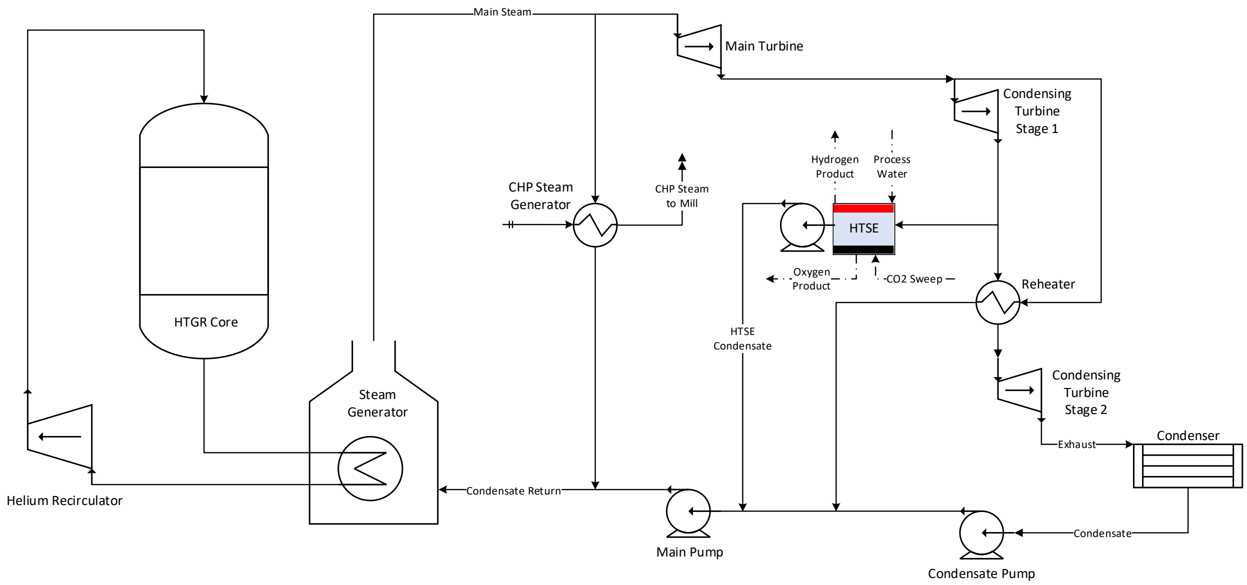


Figure E16. PFD of HTGR cycle for reference Kraft mill complex nuclear integration.