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Revision 1

Technoeconomic Analysis of Product Diversification Options for Sustainability of the Monticello and Prairie Island Nuclear Power Plants

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EXECUTIVE SUMMARY

The objective of this work was to perform technoeconomic analysis (TEA) of hybrid options that could be integrated with light-water reactor (LWR) nuclear power plants (NPPs) in order to improve the viability and sustainability of existing LWRs through product diversification by using nuclear energy, not only to produce grid electricity, but also to produce carbon-free products, such as hydrogen, ammonia, or synthetic fuels. Much of the analysis herein could be generally applied to any LWR (high-temperature steam electrolysis [HTSE] design), but potential hydrogen demand and the optimization of the HTSE was completed with specific collaboration and data from Xcel Energy's Prairie Island (PI) and Monticello Nuclear Generating Stations and the surrounding market and logistics potential in the greater Minneapolis region. Xcel Energy has set aggressive goals with regards to decarbonization, including an 80% reduction in CO_2 emissions from 2005 levels by 2030 and 100% carbon-free energy by 2050. Other TEAs regarding hydrogen production with nuclear energy have been completed in collaboration with other utility companies previously. Besides being the first TEA in this regard specific to Xcel Energy and the surrounding markets and logistics, this TEA adds to previous work by providing the most up-to-date and state-of-the-art HTSE modeling and optimization of the hydrogen-production plant integrated with nuclear power.

This work has developed the following analyses and conclusions:

- **Potential hydrogen market analysis of the greater Minneapolis region and Lifecycle CO_2 emissions analysis** of various hybrid product options that can be integrated with an LWR and produced using carbon-free nuclear energy, including:
    - Hydrogen use in fuel-cell electric vehicles versus conventional transportation fuels.
    - Co-firing of hydrogen in a 30 vol% mixture with natural gas in combustion turbine power plants versus 100% natural gas firing.
    - Hydrogen use in petroleum refineries as an alternative to steam methane reforming (SMR).
    - Hydrogen use in direct reduced iron steel-making process versus conventional coke usage.
    - Hydrogen use in ammonia production versus conventional SMR.
    - Hydrogen and CO_2 feedstock to synthetic fuels processes versus conventional transportation fuels.
- **Detailed state-of-the-art process design and financial analysis of hydrogen production via HTSE integrated with an LWR NPP**.
- **Optimization of the NPP-HTSE plant in a regulated grid environment** where HTSE capital expenses (CAPEX), HTSE capacity (hydrogen demand), and a possible hydrogen-production tax credit (PTC) are used as optimization variables and where the NPP can dispatch electricity to either the grid or to the HTSE plant, depending on the locational marginal pricing (LMP) forecast. An "envelope of profitability" to show conditions under which the NPP-HTSE can be profitable is presented.
- **Modeling of use case analysis** of carbon-free hydrogen produced from nuclear energy such as:
    - Blending of hydrogen in 30 vol% blend
    - Delivery of hydrogen via compression versus liquefaction
    - Ammonia production
    - Synthetic-fuels production (diesel and jet).

Figure E-1 shows an overview of the HTSE and NPP generic design integration with heat off-take via a thermal delivery loop (TDL) from the NPP to the HTSE as designed in the analysis in this report. It is recognized that various iterations of designs for thermal power extraction are being studied, and this configuration may not be the optimal final design. Other design options not included in this report could include removing heat from other areas and adding condensate return to the first NPP feedwater heater

versus the condenser. These alternate design options could prove to be less expensive and more efficient and will be the topic of future studies.

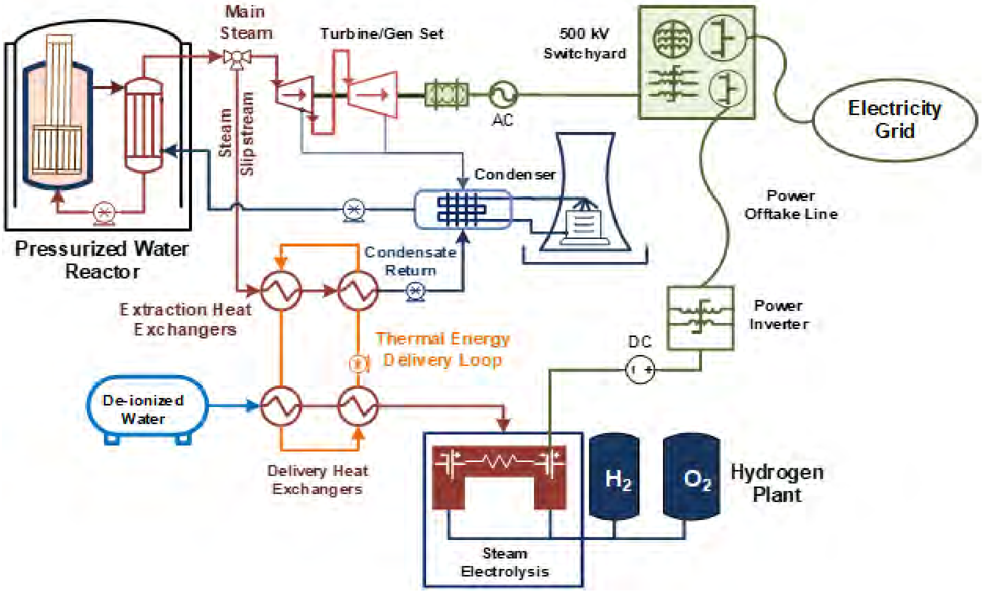

Figure E-1. Overview of HTSE Integrated with an NPP. Equipment added to the NPP include the steam slipstream from the turbine inlet, the TDL, the HTSE hydrogen plant and associated water and electricity supply tie-ins.

A static steady-state levelized cost of hydrogen (LCOH) analysis advanced case, shown in Figure E-2, shows that for a 347 tonne/day hydrogen plant (roughly the size of the output of the Monticello NPP), an NPP-HSTE can produce hydrogen competitively with SMR if the electricity price is just above $20/MWh with no CO_2 credit and assuming a modified 2021 AEO West North Central (WNC) region reference natural gas price case. NPP-HTSE can only be competitive with SMR at $30/MWh electricity price if a hydrogen production credit is considered. The advanced case assumes a lower solid-oxide electrolyzer cell (SOEC) stack cost than the base case, which is based on the Hydrogen Fuel Cell Technology Office (HFTO) current record. This lower stack cost is based on publicly calculated information on state-of-the-art improvements in SOEC capital costs from various vendors.

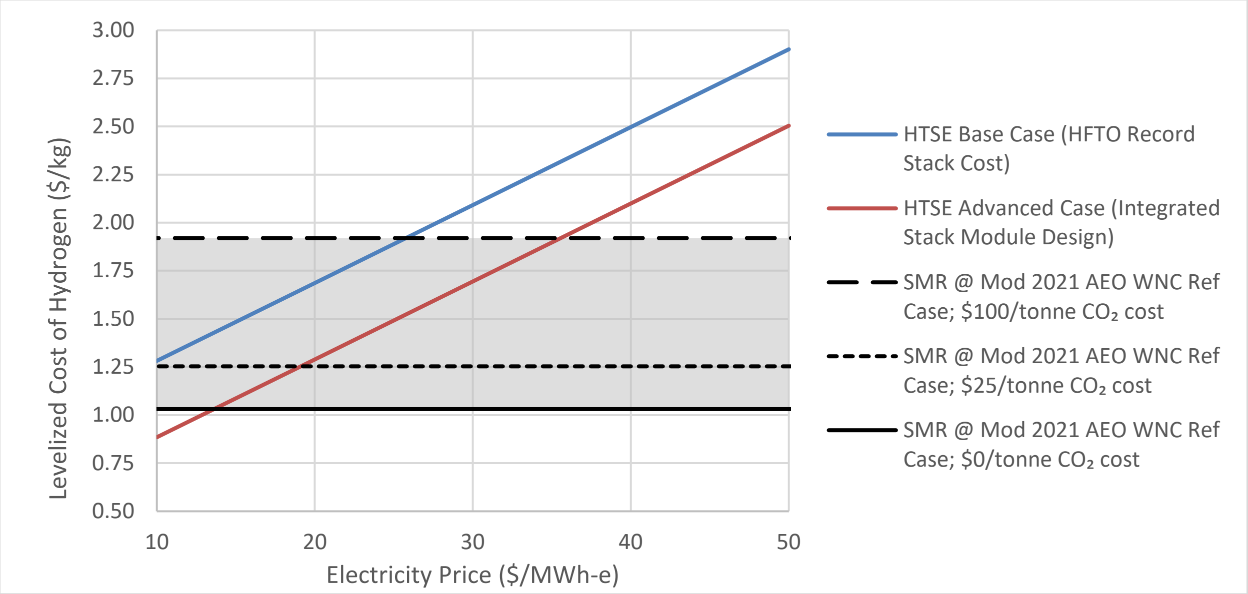

Figure E-2. LCOH of 347 tonne/day HTSE base and advanced cases versus 342 tonne/day SMR with $0, $25/tonne, and $100/tonne CO_2 cost. The HTSE LCOH includes a $0.16/kg adder for the cost of transporting hydrogen product to an off-site customer. SMR natural gas feedstock pricing based on Modified 2021 AEO West North Central (WNC) Region Reference Case.

From the optimization analysis, Figure E-3 shows the variation of the optimization variables HTSE CAPEX, HTSE capacity (H_2 demand), and a clean hydrogen PTC where an NPP-HTSE plant could be profitable based on the analysis assumptions in this report. The dashed lines represent the high HTSE CAPEX case represented by the HFTO Record and the low HTSE CAPEX represented by the advanced case discussed in this report. Figure E-3a shows the envelope of profitability for Prairie Island.

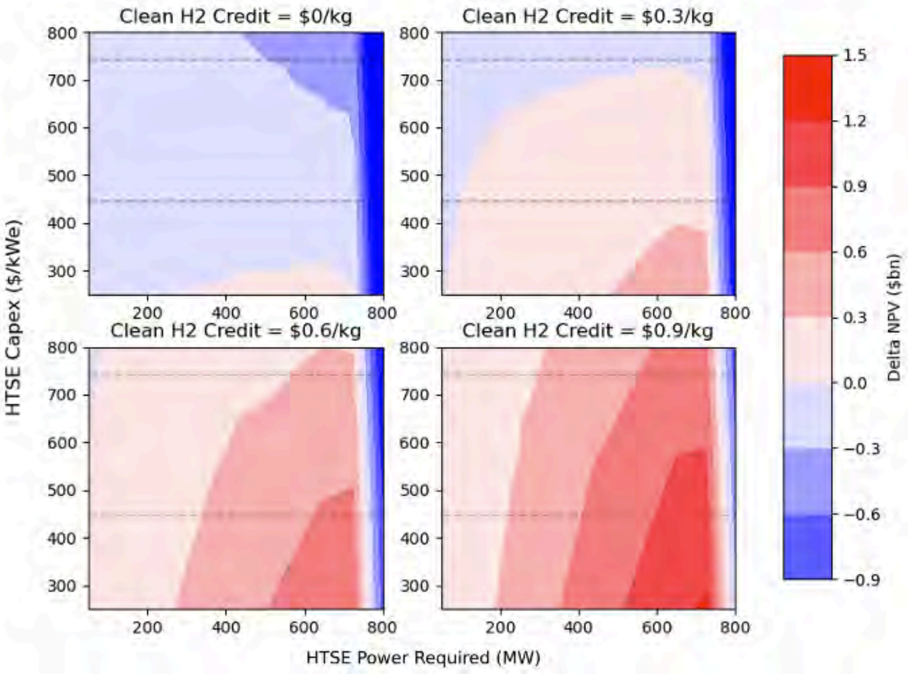

Figure E-3a. HTSE CAPEX (total capital investment), H_2 Demand, carbon-free hydrogen credit and their effect on Δnet present value (NPV) for the NPP-HTSE plant versus business-as-usual (BAU) at Prairie Island. For reference, using the full two reactors of output from PI could produce up to 29,290 kg/hr (703 tonne/day) of H_2 and a single 545 MW reactor could produce up to 14,570 kg/hr (350 tonne/day) of H_2. The horizontal dashed lines show the placement of the base and advanced case HTSE CAPEX corresponding to the high and low scenarios developed in Section 3.

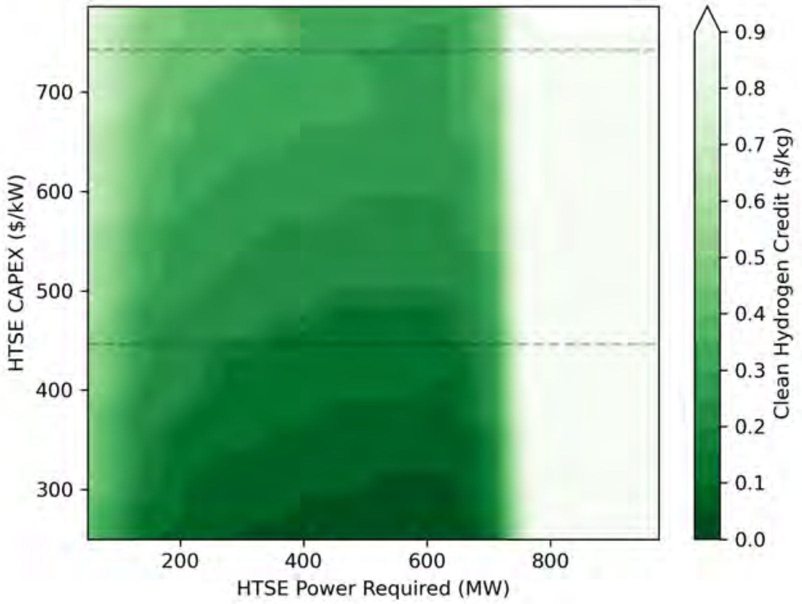

Figure E-3b. Profitable limit surface of HTSE CAPEX (total capital investment), hydrogen demand, and clean-hydrogen credit at Prairie Island. For reference, the maximum energy that PI could provide to an

HTSE could produce up to 29,290 kg/hr (703 tonne/day). A single 545 MW reactor could produce up to 14,570 kg/hr (350 tonne/day). The horizontal dashed lines show the placement of the base and advanced case HTSE CAPEX corresponding to the high and low scenarios developed in Section 3.

The cost of NH_3 production per the NPP-HTSE-NH_3 plant analyzed in this report is shown in Figure E-4 assuming an electricity cost of $30/MWh. The cost of avoided CO_2 is also plotted to show the cost of decarbonization or, alternatively, the hypothetical carbon credit that would make the NPP-HTSE-NH_3 on parity with conventional ammonia production.

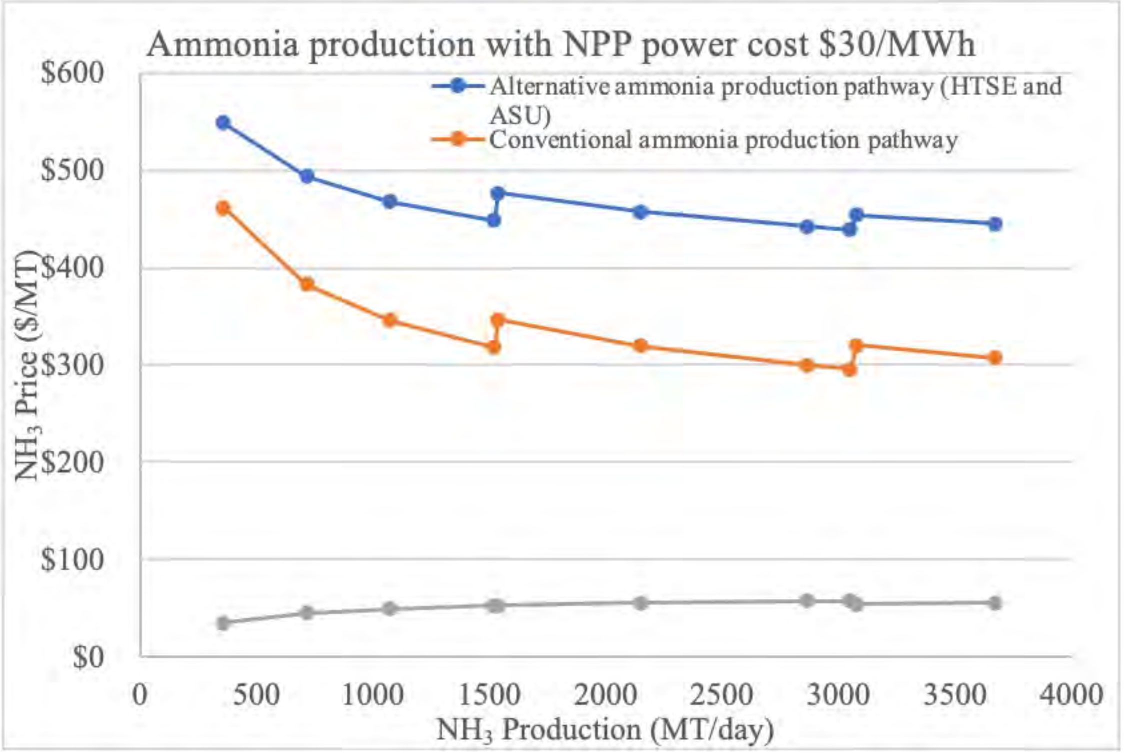

Figure E-4. Conventional and alternative ammonia production price and cost of avoided CO_2 as a function of NH_3 plant capacity (based on an electricity price of $30/MWh).

The CO_2 reduction impact of hydrogen blending with natural gas in natural gas power plants that can be realized is shown in Figure E-5. A 30 vol% mixture of hydrogen with CO_2 results in just over 10% reduction in CO_2. This is because 30 vol% H_2 with natural gas represents only ~9% blending by energy because the volumetric heating value of hydrogen is approximately 30% of the corresponding heating value of natural gas. Although, the potential greenhouse-gas (GHG)-emission reduction for this mixing ratio appears small, the amount of potential CO_2 abatement is significant due to the large contribution of natural-gas generating plants to the U.S. national GHG-emissions inventory.

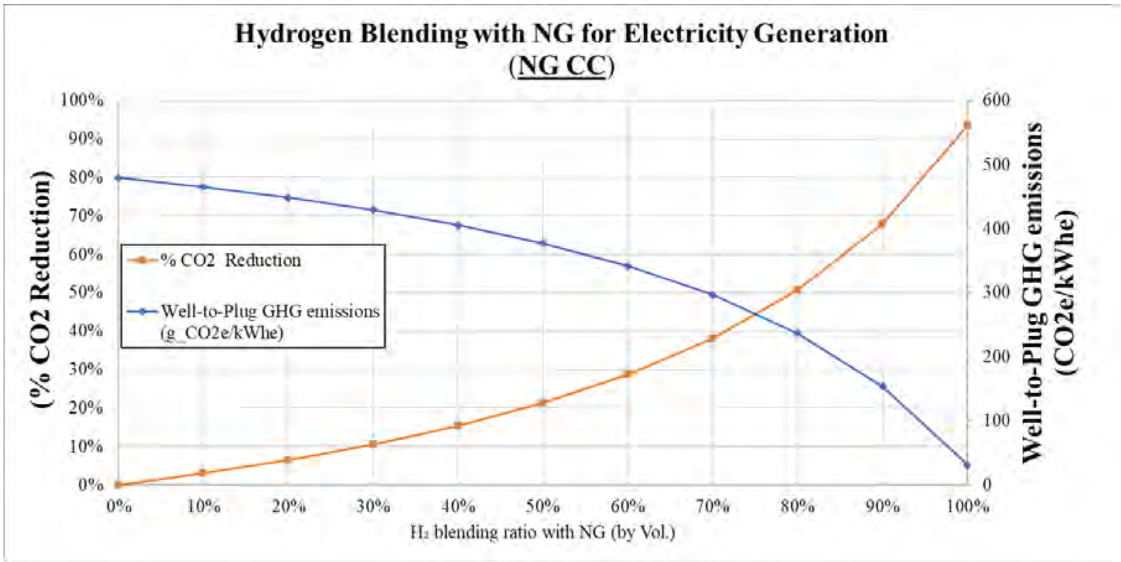

Figure E-5. Hydrogen blending with natural gas at different blending ratios and well-to-pump GHG emissions.

Synthetic fuels analysis using the HTSE + reverse water gas shift (RWGS) + Fischer-Tropsch (FT) pathway is shown below in Figure E-6. This chart shows that with the advanced synfuels case using the advanced HTSE case inputs, a CO_2 feedstock cost of \$17/MT (assuming pure CO_2 from ethanol plants), and the 2050 diesel forecast price, synthetic diesel fuel via this pathway could be competitive with conventional diesel if hydrogen were produced at \$1.14/kg. Today, producing synthetic fuels would have a cost of avoided CO_2 as shown in Table E-1.

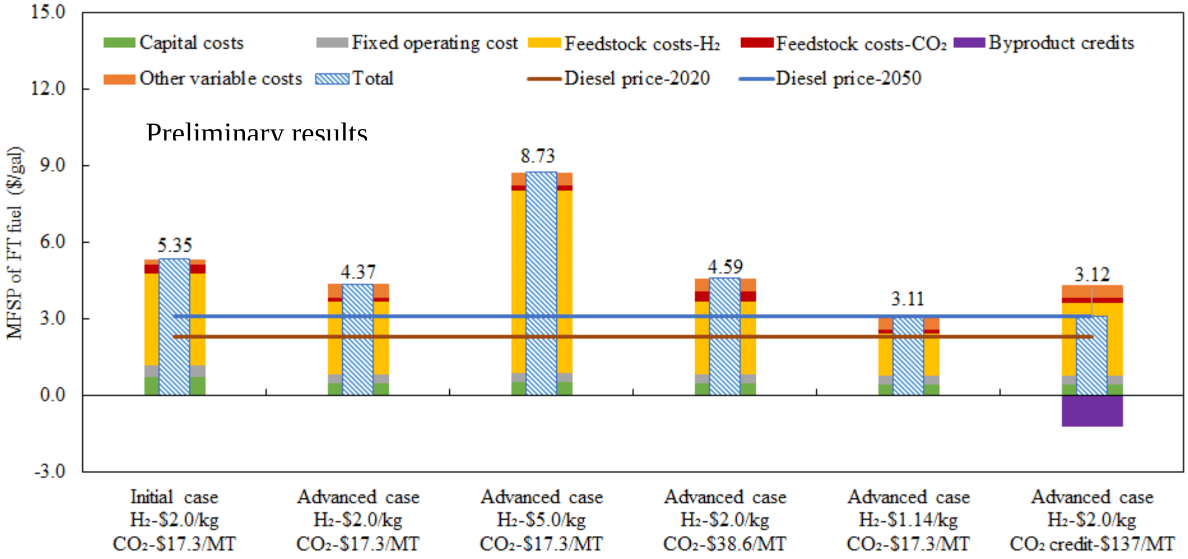

Figure E-6. Synfuels (via HTSE + RWGS + FT) production price at different H_2 and CO_2 price points and break-even scenario

In Table E-1, the hybrid options analyzed in this report are ranked in order of the estimated cost of avoided carbon from lowest to highest.

Cost of avoided of CO_2 is strongly driven by assumptions of key cost drivers such as natural gas prices, nuclear electricity prices, etc. The cost of avoided CO_2 is calculated using the equation below for each application listed in the table. A technology readiness level is estimated for each of the applications provided in the table.

$$Cost\ of\ avoided\ CO_2\left(\frac{\$}{MT}\right)\ = change\ of\ application\ price\ \left(\frac{\$}{MT}\right)\Big/change\ of\ CO_2\ emissions\ \left(\frac{MT\ CO_2}{MT}\right)$$

Table E-1. Hybrid options for integration with LWRs ranked in order of least cost of avoided CO_2 to greatest.

Nuclear-H_2 Applications	Cost of Avoided CO_2 (\$/MT CO_{2e})	Technology Readiness Level (TRL) (basic = 1, fully commercial = 9)	Notes: Nuclear Electricity Price Assumed to be \$30/MWh, Nuclear-H_2 at \$1.93/kg and natural gas pricing based on Modified 2021 AEO West North Central (WNC) Region Reference Case
Ammonia	\$35–58	8–9	Comparing ammonia production facility using nuclear power for air separation unit for N_2 and high-temperature electrolysis (HTE) for H_2 to a conventional ammonia production plant at different production rate. (Section 4.3).
Refineries	\$100	9	Comparing Nuclear-H_2 to H_2 from natural gas SMR at 1.03/kg.
Synfuels	\$137 (Diesel) \$200 (Jet fuel)	2–3	Comparing advanced synfuel production to untaxed diesel prices at \$3.1/gal (2050) and untaxed price of jet fuel \$2.6/gal (2050).
Natural Gas-H_2 blending	\$135–172	6–7	Comparing nuclear-H_2 to energy equivalent price of natural gas on higher heating value (HHV) Btu basis. This cost of avoided CO_2 is for the range of natural gas prices for natural gas electricity generators in the Minnesota's Twin Cities region.
FCEVs	\$55–270	9	Comparing H_2 \$5-7/kg (at dispenser for fuel cell vehicles), per DOE H_2 fueling cost target, to untaxed gasoline price in 2050 (\$2.96/gal), the cost of avoided carbon is very sensitive to assumed H_2 price.

CONTENTS

FIGURES

TABLES

ACRONYMS

AA	acetic acid
AC	alternating current
AEO	Annual Energy Outlook
AG	Aktiengesellschaft
AI	artificial intelligence
ANL	Argonne National Laboratory
APEA	Aspen Process Economic Analyzer
ARES	advanced rail energy storage
ASU	air-separation unit
ATJ	alcohol-to-jet
BASF	Badische Anilin- und SodaFabrik
BAU	business-as-usual
BF	blast furnace
BOF	basic oxygen furnace
BoP	balance-of-plant
BPD	barrels per day
BWR	Boiling Water Reactor
CAES	compressed air energy storage
CAGR	compound annual growth rate
CAPEX	capital expenses
CC	carbon capture
CCC	cryogenic carbon capture
CCS	carbon capture and sequestration
CCU	carbon capture & utilization
CCC	cryogenic carbon capture
CCGT	combined cycle gas-turbine
CDF	cored damage frequency
CEA	Commissionissariat à l'énergie atomique et aux énergies alternatives
CEPCI	Chemical Engineering Plant Cost Index
CFR	codes of federal regulations
CHP	combined heat and power
CIDI	combustion ignition direct injection
CONE	cost of new entrant
CT	combustion turbine
DBA	design basis accident
DCC	direct capital costs
DME	dimethyl ether
DOE	Department of Energy
DRI	direct reduced iron

DRR	demand response resource
DRTS	digital real time simulation
DTU	Denmark Technical University
EAF	Electric arc furnace
EDR	exchanger design & rating
EIA	Energy Information Agency
EIB	EU Investment Bank
EPA	Environmental Protection Agency
EU	European Union
FA	formic acid
FC	fuel cell
FCE	Fuelcell Energy (a fuel-cell company)
FCEV	fuel cell electric vehicle
FCFF	free cash-flow to the firm
FCTO	Fuel Cell Technologies Office
FCV	flow control valves
FEED	front end engineering design
FLC	federal laboratory consortium
FOAK	first-of-a-kind
FSAR	final safety analysis report
FT	Fischer-Tropsch
GHG	greenhouse gas
GPRA	government Performance and Reporting Act
GPWR	generic pressurized water reactor
HD	heavy duty
HDSAM	Hydrogen Delivery Scenario Analysis Model
HDV	heavy-duty vehicle
HES	heat extraction system
HEV	hybrid electric vehicles
HFTO	Hydrogen Fuel-Cell Technology Office
HHV	higher heating value
HTE	high-temperature electrolysis
HTEF	high-temperature electrolysis facility
HTF	heat-transfer fluid
HTGR	high-temperature gas reactor
HTSE	high-temperature electrolysis
ICE	internal combustion engines
ICEV	internal-combustion-engine vehicles
IE	initiating events
IES	integrated energy system
ILS	Integrated Laboratory Scale

INL	Idaho National Laboratory
IRP	Integrated Resource Plan
IRR	internal rate of return
ISO	independent system operator
LAES	liquid Air Energy Storage
LAR	Licensing amendment review
LCA	life-cycle analysis
LCOE	levelized cost of energy
LCOH	levelized cost of hydrogen
LCOS	levelized cost of storage
LDT	light-duty truck
LDV	light-duty vehicles
LERF	large early release frequency
LMP	locational marginal pricing
LMR	load modifying resource
LP	limited partnership
LS	liquid state
LTE	low-temperature electrolysis
LWR	light-water reactor
LWRS	Light Water Reactor Sustainability
MDV	medium-duty vehicles
MFSP	minimum fuel selling price
MHD	medium- and heavy-duty
MISO	Midcontinent Independent System Operator
MSH	main steam header
MSIV	main steam isolation valve
MT	metric tonne
MTO	methanol-to-olefins
MW	megawatt
NaOH	sodium hydroxide
NASA	National Aeronautics and Space Administration
NE	nuclear energy
NGCC	natural gas combined cycle
NHI	nuclear hydrogen initiative
NOAK	nth-of-a-kind
NPP	nuclear power plant
NPV	net present value
NRC	Nuclear Regulatory Commission
NREL	National Renewable Energy Laboratory
O&M	Operations and Maintenance
OEM	original equipment manufacturers

OPEX	operating expenses
ORNL	Oak Ridge National Laboratory
OTSG	once-through steam generator
P&ID	piping and instrumentation diagram
PADD	Petroleum Administration for Defense District
PEM	polymer exchange membrane
PFD	process flow diagram
PHEV	plug-in hybrid electric vehicles
PHS	Pumped Hydro Storage
PI	Prairie Island
PNNL	Pacific Northwest National Laboratory
PRA	probabilistic risk assessment
PRMR	planning reserve margin requirement
PTC	production tax credits
PTW	pump-to-wheels
PUC	public utility commission
PVC	polyvinyl chloride
PWR	Pressurized Water Reactor
R&D	research and development
RAVEN	Risk Analysis Virtual ENvironment
REEDS	Regional Energy Deployment System
RFS	renewable fuel standard
RIN	renewable identification numbers
ROW	right-of-way
RSOFC	reversible solid oxide fuel cell
RWGS	reverse water gas shift
SEL	steam extraction loop
SMGES	solid mass gravitational energy storage
SMR	steam methane reforming
SOEC	solid oxide electrolysis cell
SOFC	solid oxide fuel cell
SPK	synthesized paraffinic kerosine
SSC	system, structure, or component
TCI	total capital investment
TDL	thermal-delivery loop
TEA	technoeconomic analysis
TEAL	Tool for economic analysis
TED	thermal energy delivery
TEDS	Thermal Energy Distribution System
TES	thermal energy storage
TPD	thermal-power delivery

TPE	thermal-power extraction
TRL	technology readiness level
U.S.	United States
UAN	urea ammonium nitrate
UFSAR	Updated Final Safety Analysis Report
VMT	vehicle miles traveled
WACC	weighted average cost of capital
WNC	West North Central
WTG	well to plant gate
WTP	well-to-pump
WTW	well-to-wheels
ZEV	zero-emission vehicles

Technoeconomic Analysis of Product Diversification Options for Sustainability of the Monticello and Prairie Island Nuclear Power Plants

1 INTRODUCTION

With electricity-grid operations undergoing rapid and far-reaching changes as intermittent generation sources such as solar and wind increase, nuclear power plant (NPP) owners and utility companies, as an alternative to turning down plant capacity, need to understand the technical, operational, and human-factors requirements for plant operations that involve varying energy output between electricity production for the grid and providing both thermal and electrical energy directly to an industrial partner to make the best use of NPP capital resources. For example, the NPP could apportion electricity between the grid and an electrolysis plant that produces hydrogen.

Due to off-market incentives supporting the penetration of variable renewables like wind and solar (e.g., production tax credit) and low natural gas prices, in some U.S. regions the average production cost of electricity at times falls below the production cost at NPPs. In these regions, it might be possible for NPPs to recover profitability by using available nuclear power to produce a marketable non-electric product, such as hydrogen. Hydrogen generated in this way may qualify for future credits that would incentivize hydrogen produced with low carbon intensity.

This situation suggests a new paradigm for the use of nuclear energy. With flexible operation and generation, NPPs could distribute energy to an industrial process in a dynamic manner that optimizes the revenue of NPP owners. Studies have shown NPPs may be able to competitively provide the energy required to produce hydrogen and other valuable chemical products.[1,2,3] This may yield a more-advantageous market and revenue position for utilities employed in this market.[3]

The purpose of this work is to provide technoeconomic analysis (TEA) of hybrid configurations to optimize the total system in a regulated market specific to the Monticello and Prairie Island (PI) light-water reactor (LWR) NPPs located in the greater Minneapolis area and operated by Xcel Energy.

Market demand analysis for hydrogen, ammonia, and synthetic fuels, as well as the carbon dioxide supply market specific to the Minnesota region is presented. Demand sources are reported in the distances located from both NPPs considered, as well as relative demand volume and assumed selling prices. The electric market for the Minnesota region has been modeled in a way that takes assumed generation buildout forecasts and grid interactions into account using REEDS and PLEXOS in a cost/benefit system analysis in a regulated market. REEDS is a custom grid-generation modeling software built by the National Renewable Energy Laboratory (NREL). Grid-electricity pricing forecasts have been calculated. This electric and grid market analysis is being performed in a separately funded, but coordinated analysis by NREL that will be reported separately in the near future.[4] The grid analysis performed by NREL and the analysis presented in this report were coordinated, and inputs and assumptions were shared among the respective models.

Hydrogen production via solid oxide electrolysis cells (SOEC)/high-temperature steam electrolysis (HTSE) has been extensively modeled and evaluated, including stack, balance-of-plant (BoP), and a very limited assumption on NPP thermal tie-in capital expenses (CAPEX), operating expenses (OPEX) including utility and energy consumption, design and performance assumptions, current versus future nth-of-a kind (NOAK) technology—using an assumed learning rate—and operating modes. It is recognized that the assumptions of NPP thermal tie-in expenses are very minimally treated here in relation to simple piping expenses. Other costs associated with thermal tie-in to an existing plant such as exact design, safety and hazard analysis, permitting, regulation reviews etc are out of scope of this work. Sensitivity analyses on parameters such as HTSE capacity, electricity price, and possible production tax credits (PTCs) for the life-cycle low-carbon-intensity hydrogen that can be produced is presented. The above calculations are presented along with the normalized cost to produce the hydrogen or levelized cost of hydrogen (LCOH) in dollars per kilogram of hydrogen for various scenarios, and the case is made for the profitability and viability of future looking SOEC/HTSE systems integrated with NPPs.

In-depth analysis has been done on the optimization of the NPP-HTSE system by itself and integrated with the electrical-grid to maximize the possible revenue given the input assumptions. Sensitivity analyses on parameters, such as HTSE capacity, electricity price, and possible PTC, have also been run in the optimization models. The net present value (NPV) of the investment for the NPP-HTSE with various assumptions is reported. Because the electricity price is the single largest cost factor involved in both the LCOH calculation and the NPV of the investment, an in-depth discussion is included on alternative scenarios and paradigm shifts under which the HTSE integrated with an NPP could be operated and treated by the electrical-grid operator and the public utility commission (PUC). These alternative proposals for treatment by the electrical-grid operator and PUC are hypothetical only because it will be up to the utility company operating NPPs to determine and negotiate with their respective regulators and partners any new paradigm related to the interactions with an NPP-HTSE plant.

Specific use cases for the assumed volumes of hydrogen that could be produced by Xcel Energy’s NPPs are also analyzed and presented, such as blending hydrogen with natural gas in natural gas power plants, compression, storage, and liquefaction of hydrogen for transportation to the end user, and other follow-on hybrid integrations with NPPs which could consume the hydrogen and produce low carbon ammonia and synthetic fuels (diesel, jet fuel, motor gasoline).

Additionally, separately funded work has analyzed the methods, detailed modeling of equipment and controls, and safety analysis of thermal-power extraction (TPE) from an NPP to be used in a hybrid, integrated industrial process, such as HTSE. Extracting a small portion of thermal power from an NPP at the inlet of the turbine is a novel concept that is being pioneered, including engineering design as well as lab and simulator testing. Thermal power can be transported short distances and used to provide large amounts of low-temperature energy to industrial processes. A discussion on heat-transfer fluid (HTF) options is included. TPE could give NPPs a competitive advantage in the production of low-carbon, value-added products, such as hydrogen, ammonia, synthetic fuels, and others. A preliminary fault-tree safety analysis probabilistic risk assessment (PRA) for TPE from an NPP is summarized. This includes a discussion commenting on the assessment of safety within the existing licenses of LWRs. This section on TPE technical and safety analysis is summarized in the appendix of this report for completeness of this TEA effort, but the reader is referred to separately produced reports on the subject for a more complete analysis.

The following analyses are treated in this study to show the viability of hybrid integrations with nuclear power:

- Market analysis
    - Hydrogen, ammonia, and synfuels market analysis specific to the greater Minneapolis region
    - Minnesota-region regulated electricity-grid market analysis (detailed in separate NREL report).[4]
- NPP-HTSE—Hybrid integration of HTSE with LWR NPPs
    - HTSE integrated full plant design using industry standard modeling software (Aspen, H_2A), including CAPEX and OPEX, stack and performance assumptions, NOAK technology, operating modes, etc.
    - Optimization of the assumptions and conditions to show the envelope of profitability where HTSE H_2 production, integrated with nuclear power, is viable, including NPV optimization analyses of the HTSE investment integrated with an NPP. Range of valuation of carbon that may increase the viability of NPP-HTSE hydrogen production and other hybrid options

  - Hypothetical alternative proposals for the treatment of the NPP-HTSE plant by the electrical-grid operator and PUC (found in the appendix).
- Use cases for low-carbon hydrogen, produced using nuclear energy
  - Blending of hydrogen with natural gas to be burned in natural gas power plants
  - Compression and storage of hydrogen in trucks or pipelines versus liquefaction of hydrogen for transport to an end-user at various distances
  - NPP-HTSE-NH_3—hybrid integration of the NPP and HTSE with ammonia production versus the standalone NPP-HTSE
  - NPP-HTSE-synfuels—hybrid integration of the NPP and HTSE with synthetic fuels (diesel, jet fuel, motor gasoline) production using low carbon hydrogen.

This work is funded under DOE Office of Nuclear Energy (NE) industry funding opportunity announcement (iFOA), DE-FOA-0001817. Separate funding has also been contributed from the DOE Hydrogen and Fuel Cell Technologies Office (HFTO).

2 DEMAND MARKETS AND LIFECYCLE CO_2 ANALYSIS

This section discusses the U.S. market potential, size, and location for value-added products, such as hydrogen, which could be produced in an integrated facility with NPPs. Life-cycle CO_2 emissions reduction associated with nuclear-produced H_2 for these markets are reported in this section. Following, in its own subsection, the potential hydrogen demand around the PI and Monticello NPPs are categorized and discussed.

2.1 National Potential Hydrogen Demand

2.1.1 Fuel-Cell Electric Vehicles

2.1.1.1 Light-duty vehicle (LDV) applications

For LDV applications, hydrogen demand depends on the size of the vehicle population into which fuel-cell electric vehicles (FCEVs) will be introduced and the success of FCEVs in capturing a share of that population, while assuming simultaneous improvements in battery electric vehicle technologies. Each of these factors contains varying degrees of uncertainty, both in how quickly they may evolve, and how external factors (e.g., vehicle and fuel cost assumptions, availability of hydrogen fueling stations, consumer lifestyles, and preferences) might influence them. Forecasts of future hydrogen demand should be based, to the greatest extent possible, on objective, widely accepted trends, and projections and well-vetted tools and techniques. The analysis utilized a vehicle-choice model to estimate the future market penetration of hydrogen fuel cell vehicles, along with estimates of future fuel economy to calculate potential future hydrogen demand.[5] Hydrogen demand potentials were estimated without considering economic factors, which can be considered an upper bound for the size of the FCEV market and defined as “serviceable consumption potential.”

For this analysis, estimates of FCEV car and light-duty truck (LDT) sales, stock, and hydrogen consumption were developed according to the following process:

- Total FCEV Market Penetration and Sales: FCEV car and LDT sales shares were obtained from prior HFTO analyses consistent with HFTO price targets (Table 2) for delivered hydrogen. Annual numbers of FCEVs sold were derived by applying these shares to EIA forecasts of national LDV sales by year.
- Total FCEV Stock and Hydrogen Use: FCEV car and LDT stock, vehicle miles traveled (VMTs), and hydrogen consumption were estimated by year and summed to produce national totals using Argonne’s VISION model.[6]

Market penetration of FCEVs was estimated using the Market Acceptance of Advanced Automotive Technologies (MA3T) vehicle-choice model developed and maintained by Oak Ridge National Laboratory (ORNL).[7] MA3T estimates market penetration rates or shares of conventional internal-combustion-engine vehicles (ICEVs), FCEVs, hybrid electric vehicles (HEVs), battery-electric vehicles, and plug-in HEVs (PHEVs) of different range capabilities.[8] For this analysis, MA3T was run assuming that HFTO’s FC and hydrogen fuel cost and performance targets will be met in the future (i.e., the “Program Success” case in HFTO’s annual reporting as documented by [9].

Table 2. Projected light-duty vehicle stock penetration for the United States in year 2050 by powertrain from MA3T.[5]

	Fuel Cell Electric	Internal Combustion Engine[a]	Hybrid Electric	Plug-in Hybrid Electric	Battery Electric
Cars	18%	15%	10%	21%	36%
LDTs	26%	11%	7%	20%	36%

a Includes gasoline, diesel, compressed natural gas, and ethanol flex-fuel ICEVs.

2.1.1.1.1 Potential Hydrogen Demand for Light-Duty FCEV

As shown in Table 3, FCEV average fuel economy increases from 54 and 41 miles per gasoline gallon equivalent (mpgge) for cars and light trucks, respectively, in 2015 to 100 and 64 mpgge, respectively, in 2050. FCEVs are assumed to have equivalent fuel economy and to be driven and scrapped (i.e., retired from use) at the same rates as the conventional vehicles they replace. They are also assumed to have the same survival rate and lifetime VMTs (about 167,370 miles) as their conventional counterparts. Table 3 compares our fuel economy assumptions (in mpgge) for FCEVs with those for comparable ICEVs. Fuel economy for ICEVs also improves between 2015 and 2050, rising from an average of 26–43 mpg for ICEV cars, and from 20–30 mpg for ICEV light trucks. These fuel economy assumptions are based on Autonomie's vehicle simulation model estimates (which reflect HFTO program goals) for mid-sized cars and mid-sized SUVs, adjusted to on-road values using factors applied in the GREET model (see Section 2.1.1.3). Based on the vehicle stock estimates at market equilibrium in Table 2 and vehicle fuel economy estimates in Table 3, the future hydrogen consumption by FCEVs was estimated at 4.3 MMT for cars and 7.4 MMT for LDTs, for a total of 11.7 MMT.

Table 3. United States on-road fuel economy of FCEV and ICEV cars and light trucks by model year.[5]

Model Year	Car mpgge (mi/gasoline gal equivalent)		Light Truck mpgge (mi/gasoline gal equivalent)	
	Gasoline (E10) ICEV	FCEV	Gasoline (E10) ICEV	FCEV
2015	26	54	20	41
2020	31	61	23	45
2025	35	72	24	52
2030	37	80	25	55
2035	39	90	28	58
2050	43	100	30	64

The serviceable consumption potential is estimated as the demand if FCEVs constituted 41% of the LDV fleet in 2050 (i.e., 66 million of 163 million cars and 63 million of 153 million light-duty trucks). Fleet penetration was based on the analysis from [10], which estimates a FCEV sales share of 41% for passenger vehicles in 2050 using favorable assumptions. The fleet penetration is applied to the 2050 LDV stock to estimate FCEV stock and a corresponding annual hydrogen demand of 21.4 MMT/yr (10.0 MMT/yr for cars and 11.4 MMT/yr for light-duty trucks).

2.1.1.2 Medium- And Heavy-Duty Truck Applications

Medium- and heavy-duty (MHD) vehicles (MHDVs) are used to move freight and provide various services. They encompass a wide range of sizes and body types and operate on a variety of duty cycles. Contractors, construction workers, and delivery services use medium duty trucks. These vehicles are sometimes called "last-mile" delivery trucks. Heavy duty (HD) trucks are used for moving heavier and larger cargo within urban areas and over short distances, as well as over long distances. MHD trucks account for a significant portion (20–25%) of the energy consumption and air emissions of the U.S. transportation sector. MHDVs, around 11 million trucks and fewer than 1 million buses, represent only 4.5% of the 260 million vehicles on the road nationally.[6] Although they comprise only a small share of the national vehicle population, MHDVs are the second-largest energy consumers and greenhouse gas (GHG) emitters, behind only light- duty vehicles that include passenger cars, sports-utility vehicles, and pickup trucks.[5]

In the United States, nearly half of MHD trucks are used for urban, local, and short-haul operations, with a daily travel distance of less than 200 miles.[11] To calculate hydrogen use by FCEVs at any given time, three key parameters are required for each vehicle class: (1) number of vehicles on the road, (2) annual VMT, and (3) fuel economy or fuel consumption per mile. Because of the lack of reliable vehicle-choice models for MHDV classes, we assumed the market penetration of fuel cell MHDVs to be consistent with that of fuel cell LDVs in 2050 (i.e., ~22% penetration). Then, we calculated annual sales of MHDVs and, using the VISION model, computed annual vehicle stocks, VMT, and energy use.

2.1.1.2.1 Potential Hydrogen Demand by Medium- and Heavy-Duty Fuel Cell Vehicles

As shown in Table 4, composite fuel economies of 33.0 and 14.7 mi/kgH2 were estimated for Class 4 walk-in delivery trucks and Class 8 long-haul sleeper-cab trucks, respectively. Using the VISION model, total annual VMTs for all MDVs (Classes 2b–6) and HDVs (Classes 7–8) in 2050 were estimated at 212 billion and 252 billion for MDVs and HDVs, respectively. The above estimates of fuel economy and total annual VMT, along with the 22% penetration of FCEVs into the MHDV sector, resulted in potential hydrogen consumption of 1.4 and 3.8 MMT by MD and HD fuel cell trucks, respectively. For the entire MHDV sector, total hydrogen demand was estimated to reach 5.2 MMT in 2050.

We estimate the serviceable consumption potential of the MDV and HDV hydrogen market as equivalent to the hydrogen required if 35% of the fleet operated on hydrogen, based on the 2050 sales share estimate from Roadmap to a U.S. Hydrogen Economy,[10] which uses favorable assumptions for FCEVs. Applying this market penetration to the vehicle stock results in 4.2 million FCEVs of a 12 million MDV stock and 2.0 million FCEVs of a 5.7 million HDV stock in 2050. The corresponding annual hydrogen demand is estimated at 8.2 MMT/yr (2.2 MMT/yr for MDVs and 6.0 MMT/yr for HDVs).

Table 4. Parameters relating to fuel efficiency of MHD fuel cell trucks.[5]

Test Vehicle	GVWR Class	Vehicle's Payload (tonnes)	Weighting Strategy for Computing Efficiency	Composite Fuel Economy (mi/kgH$_2$)
Walk-in Truck	4	3.64	Urban drive cycle	33.0
Long-haul Truck	8	16.3	EPA 55 drive cycle	14.7

2.1.1.3 Lifecycle CO$_2$ emissions analysis of H$_2$ from nuclear energy for vehicles

GHG emissions associated with hydrogen production and the delivery and dispensing pathway can be estimated using a well-to-wheels (WTW) analysis with the Argonne National Laboratory's (ANL's) Greenhouse gas, Regulated Emissions, and Energy use in Transportation GREET 2019 model to conduct the life-cycle analysis (LCA).[12] The WTW analysis can be further broken down into well-to-pump (WTP) and pump-to-wheels (PTW) stages. The WTP stage includes fuel production from the primary source of energy (feedstock) to its delivery to the vehicle's energy-storage system (fuel tank). The PTW stage includes fuel consumption during the operation phase of the vehicle to power the vehicle's wheels. The results from WTP and PTW analyses are summed to give the WTW energy use and GHG emissions associated with various vehicle-fuel technologies. WTW analysis was carried out using the GREET 2019[12] model for LDVs, including FCEVs, using various hydrogen-production and delivery pathways and baseline gasoline ICEVs. Fuel economy of 26 mpg was assumed for gasoline ICEVs and 55 mpg gasoline equivalent (ge) for H$_2$ FCEVs. Conventional internal-combustion engines (ICEs) using gasoline and diesel were compared to FCEV's using hydrogen produced from natural gas SMR and nuclear electricity.

The WTW equivalent CO$_2$ emissions per mile for LDVs compared ICEVs using gasoline, FCEVs using hydrogen from SMR, and FCEVs using nuclear-H$_2$. An ICE using gasoline produces 387 g CO$_2$ eq/mile while FCEV using H$_2$ from SMR produces 170 g CO$_2$ eq/mile, and FCEV using H$_2$ from nuclear electricity produces only 33 g CO$_2$ eq/mile, on a WTW basis (Figure 7).

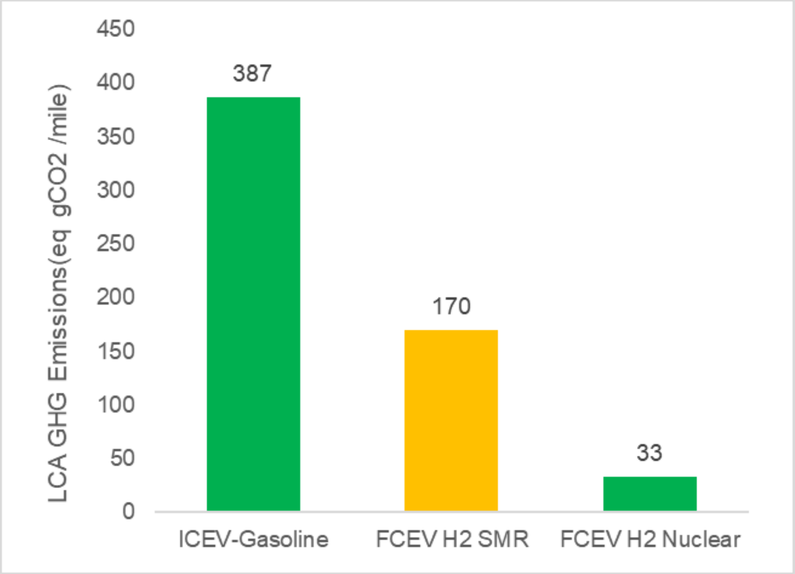

Figure 7. WTW life-cycle GHG emissions results for LDVs.

The WTW eqivalent CO_2 emissions per mile for HDVs were also compared. The conventional heavy-duty ICEV using diesel in compression-ignition direct injection (CIDI) engine produces 1.7 kg CO_2 eq/mile while the heavy-duty FCEV using H_2 from SMR is estimated to produce 0.8 kg CO_2 eq/mile, and the heavy-duty FCEV using nuclear-H_2 produces 0.1 kg CO_2 eq/mile (Figure 8).

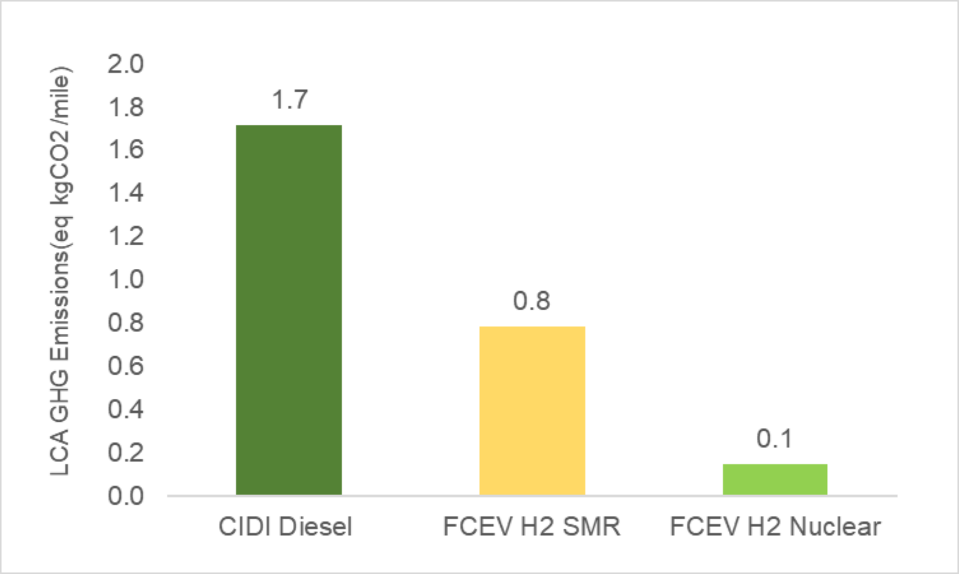

Figure 8. WTW life-cycle GHG-emissions results for HDVs, including CIDI diesel engines.

2.1.2 Co-Firing of Hydrogen with Natural Gas in Combustion Turbines

Another potential use of clean hydrogen produced from NE is its injection into natural gas pipelines for use as a low-carbon green component of a natural gas/hydrogen fuel mix for general heating or for exclusive use in combustion turbines (CTs) for power generation. The potential and technical barriers to mixing H_2 with natural gas are discussed elsewhere.[13] It is important to note that, in this use case, the hydrogen should be considered and evaluated as an energy-storage means, rather than as in the other use cases mentioned in this report, where hydrogen is an intermediate, used for upgrading products to a higher value in the product chain. This is due to the hydrogen being produced using electricity and then stored until a later time when it is again converted back to electricity, with all the associated efficiency losses from conversion, storage, and regeneration. Thus, the firing of hydrogen in turbines to produce electricity should be evaluated against other means of energy storage.

For the purposes of this study, potential demand is estimated for hydrogen by assuming it can be used by natural gas CTs with a volume ratio of 30% hydrogen blended with 70% natural gas. Electricity generators were identified using the data sets from the EIA-860 and EIA-923 forms describing electricity-generator facility locations and fuel use. Figure 9 shows natural gas electricity generators throughout the U.S. and the potential demand if for hydrogen if it were blended at 30 vol% with natural gas. Future planned natural gas green field and coal conversion to natural gas plants are not included.

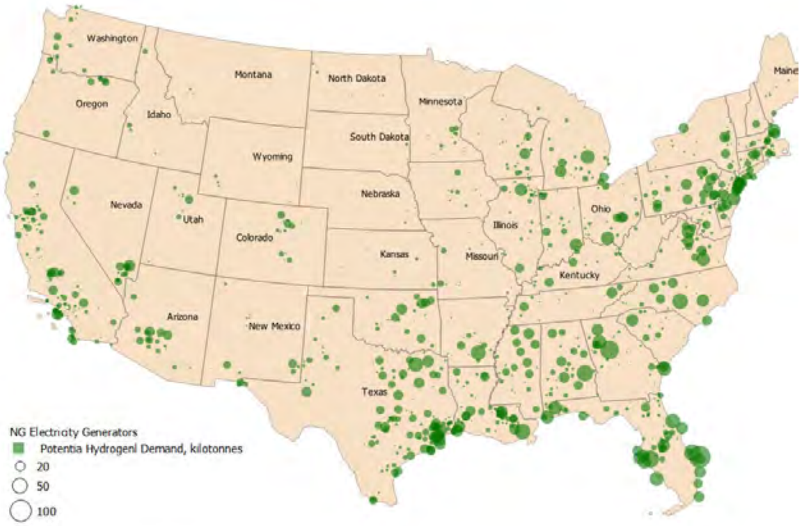

Figure 9. Natural gas power plants in the U.S and potential hydrogen demand if hydrogen were blended at 30 vol% H_2 with natural gas.

2.1.2.1 Lifecycle CO_2 Emissions Analysis of H_2/Natural Gas Blending in Natural Gas Power Plants

The LCA was carried out using the GREET 2020 model to estimate GHG emissions for 100% natural gas, as well as a mixture of 30% hydrogen and 70% natural gas by volume, as fuel supply to the electricity generators. The equivalent CO_2 emissions per kWh of electricity produced and transmitted to end use (i.e., at the wall outlet) are compared in Figure 10. The life-cycle GHG emissions are estimated at 493 g CO_{2e}/kWh when using only natural gas as the feed, and 442 g CO_{2e}/kWh for the mixture of 30% hydrogen and 70% natural gas by volume for different natural gas turbines technology shares. We note that 30 vol% H_2 with natural gas represents only ~9% blending by energy because the volumetric heating value of hydrogen is approximately 30% of the corresponding heating value of natural gas. Although the potential GHG-emission reduction for this mixing ratio appears small, the amount of potential CO_2 abatement is significant due to the large contribution of natural gas generating plants to the U.S. national GHG emissions inventory. Furthermore, future turbine designs that can handle higher mixing ratios, and potentially combust 100% hydrogen, will have the potential to eliminate CO_2 emissions from gas power-generation units. We also note that mixing hydrogen with natural gas in the near term is attractive compared to other new hydrogen end-use applications because it leverages the existing natural gas infrastructure and application end use (i.e., the gas-turbine); thus, less new capital investment may be needed to be compared to building a new hydrogen delivery infrastructure. Some pipeline modifications and restrictions on the percentage of hydrogen may be needed due to various concerns specific to hydrogen such as hydrogen embrittlement, confined spaces, etc.

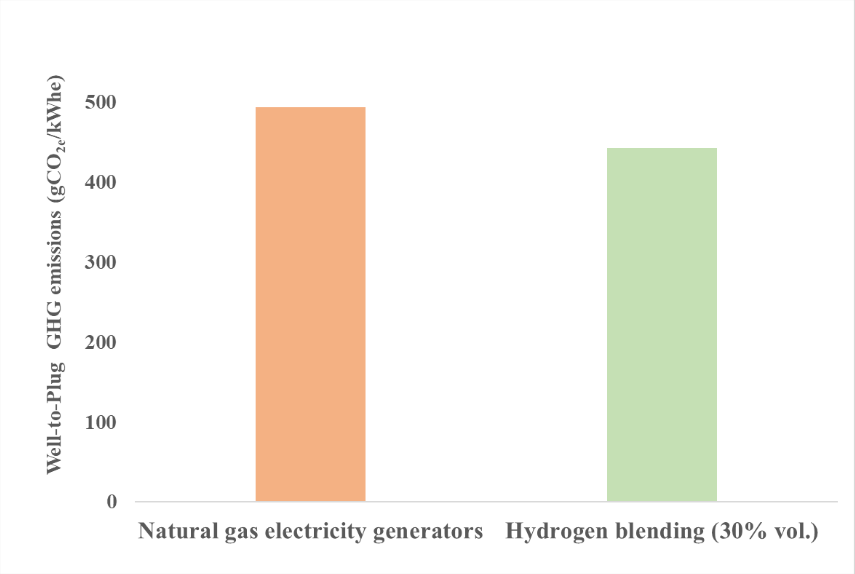

Figure 10. Life-cycle GHG-emission for natural gas electricity generators, well-to-wall outlet analysis.

Figure 11 below shows a trend of the percentage and amount of CO_2 reduction that can be achieved versus the volume percentage mix of H_2 with natural gas in natural gas electricity-generating units.

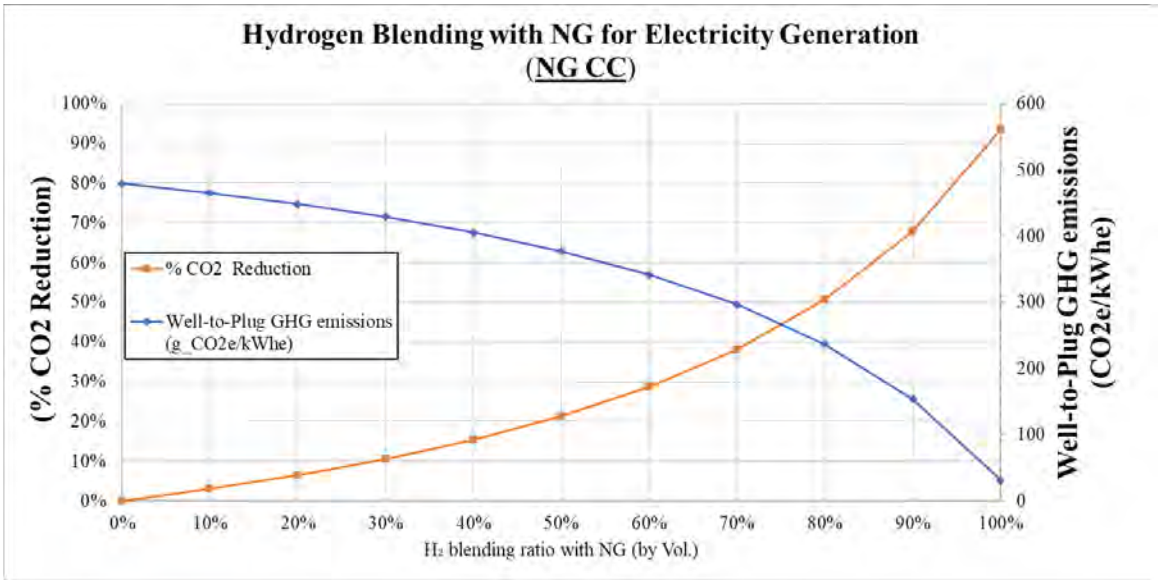

Figure 11. NG-H_2 blending ratio versus achievable percentage of CO_2 reduction and W2W GHG emissions for natural gas power plants.

2.1.3 Petroleum Refineries

Petroleum refineries are currently the most significant user of hydrogen in the U.S., consuming approximately 10 MMT of hydrogen annually, including byproduct hydrogen from naphtha reformers.[14] Approximately one-third of refinery hydrogen used is the byproduct of naphtha-reforming processes while most of the rest of the needed hydrogen is typically produced onsite using the SMR process with natural gas as the feedstock. Some refineries also use hydrogen regional pipelines, which are mostly limited to the Gulf Coast in the U.S. Hydrogen is used primarily for hydrocracking and hydrotreating. (Hydrocracking is used to produce diesel from heavy crude, and hydrotreating is used to remove sulfur from feed, intermediate, and product streams.) Most hydrotreating capacity is used for reducing sulfur in diesel, fluid catalytic cracker feeds, and naphtha streams. Refinery hydrogen demand is, in general, driven by the ratio of gasoline to diesel production, American Petroleum Institute (API) gravity, sulfur content of the petroleum inputs, and the complexity of refinery processing.

Elgowainy et al[5] estimated future hydrogen demand through 2050 for petroleum refining, based on projections of crude inputs and market demand for refinery products from the EIA Annual Energy Outlook,[14] and crude API gravity and sulfur content based on Han and colleague's analysis.[15] The main conclusions are that crude inputs are estimated to increase from 16 to 18 Mbbl/d (with a steeper increase of 9% from 2015 to 2021 and then a more gradual increase to 2050), gasoline output decreases from 8 to 6 Mbbl/d, diesel output increases slightly, and average jet-fuel output increases roughly 0.5 Mbbl/d from about 1.7 to 2.2 Mbbl/d.[4]

Refinery hydrogen demand by Petroleum Administration for Defense District PADD region shown in Figure 12, is projected to increase due to increased ratio of diesel/gasoline demand, stringent sulfur requirements, higher API gravity, and sulfur content for petroleum feedstocks, and increased crude inputs to refineries. In addition to the internal hydrogen production via catalytic reforming of naphtha, the total U.S. hydrogen demand for petroleum refining grows from 5.9 MMT/year in 2017 to an estimated 7.5 MMT/year in 2050. Minnesota fall under PADD 2 region.

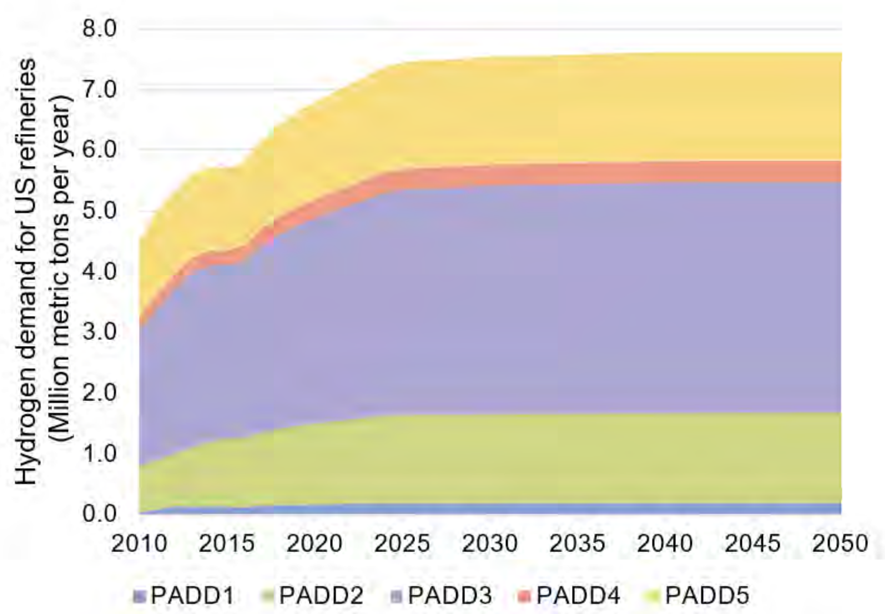

Figure 12. Projected total hydrogen demand for U.S. refineries by PADD, through 2050.

2.1.3.1 Life-Cycle CO_2 Emissions Analysis H_2 from Nuclear versus H_2 from SMR for Refinery Use

Life-cycle emissions for hydrogen production from well to plant gate (WTG) were calculated using GREET 2020. The H_2 production from SMR was compared to production from high-temperature electrolysis (HTE) using SOEC with nuclear electricity; this H_2 can later be used for naphtha reforming processes at the refineries. Figure 13 below shows the well-to-gate CO_{2e} p emissions for H_2 produced from natural gas SMR and HTE (nuclear), which are 9.28 kg CO_{2e} /kg H_2 and 0.15 kg CO_{2e}/kg H_2 respectively.

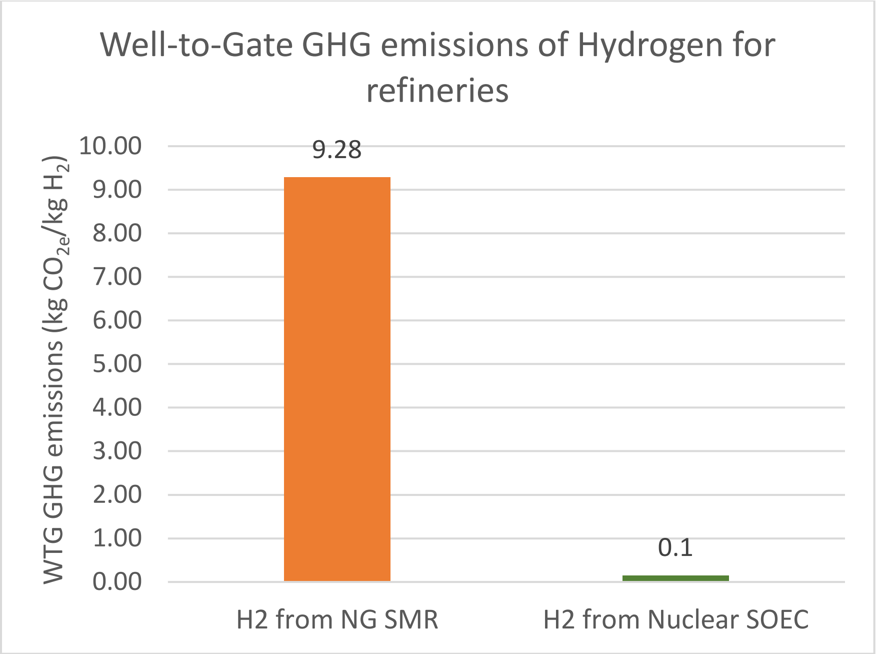

Figure 13. WTG GHG emissions for hydrogen production with SMR and HTE.

2.1.4 Direct Reduced Iron for Metals-Refining and Steel Production

The direct reduction of iron (DRI) is a process developed by Midrex Technologies, Inc., for producing high-purity iron from ore at temperatures below the melting point of iron by reducing the iron oxide ore and driving off oxygen in a reactor using a reducing agent. The reducing agent can be carbon coke, hydrogen, or syngas. In the conventional approach to steel making, iron ore is reduced to pig-iron using coking coal as the reducing agent in a blast furnace (BF), and the pig-iron is then refined and converted to steel in a basic oxygen furnace (BOF). In the DRI process, DRI is converted to steel in an electric arc furnace (EAF), allowing reductions in overall energy use and CO_2 emissions compared to the conventional BF-BOF steel-production process. In the U.S., the amount of steel produced by EAF has been increasing and is expected to continue to grow, mainly due to the increased production of scrap, which can be incorporated in the EAF feed, while the amount produced by BOF is expected to remain relatively flat.[16] Product quality dictates the amount of scrap that can be used in an EAF; the remainder must be made up with pig-iron from a BF-BOF or DRI. Due to its high purity, DRI has the potential to increase the amount of scrap which can be used by EAF relative to using pig-iron from a BF-BOF. The DRI process, using 100% hydrogen as the reducing agent, requires up to 100 kg hydrogen per MT of steel—i.e., a mass ratio of approximately 10%. However, using hydrogen in a blend with natural gas up to 30/70 ratio by energy to produce DRI would not require modifications to the original technology which was developed to work solely with natural gas.[17]

Syngas ($CO_2 + H_2$) could alternatively be produced and supplied to metals plants using CO_2 and water in CO_2 or co-electrolysis. There are a few advantages to using a carbon-containing molecule such as CO in addition to hydrogen in the metals-refining process versus using hydrogen alone. First, different grades of steel require varying amounts of carbon as part of the finished alloy in order to obtain the desired material properties, so carbon will need to be incorporated regardless. Second, some carbon can be consumed in the metals-refining process, releasing energy, and resulting in a more-economic process overall due to the reduced outside-heating requirements versus a metals-refining process using only hydrogen as the reducing agent. This use case will not be further analyzed in this report.

Nuclear and renewable hydrogen could be used to offset natural gas or other fuels in the DRI process. For this analysis, we estimate the potential hydrogen demand for DRI was based on using 30% hydrogen and 70% natural gas on an energy basis.[17] In 2017, U.S. steel consumption was 106.2 MMT, while production was 81.6 MMT. Based on trends in U.S. iron ore production, imports, and exports, we estimate that 68% of the 81.6 MMT of U.S. steel production was in electric arc furnaces (i.e., only 32% was produced in basic oxygen furnaces). The mass of hydrogen required to fully reduce 1 MT of iron ore ranges from 0.08 to 0.12 MT, depending on the technology employed, reaction temperature, and the reaction off-gas available for hydrogen preheating. Hydrogen price affects economic feasibility more strongly than the capital and operating costs of the DRI process. It is estimated that a hydrogen price of $1.2/kg would generate a positive NPV for the (MIDREX_EAF_H_2) DRI technology.[18]

EIA's Annual Energy Outlook projects crude steel production growing to approximately 120 MMT.[14] If all this production were converted to MIDREX_EAF DRI technology using low- cost hydrogen (i.e., no BF plants), the technical potential for hydrogen demand could be on the order of 12 MMT annually.

These estimates are conservative relative to the national estimate of Elgowainy et al. for potential future hydrogen demand of 4 MMT for 30% replacement of natural gas on an energy basis.[5] Their estimates are based on the Annual Energy Outlook projection of 50% growth in U.S. steel production by 2040 and full replacement of iron inputs with those produced by DRI.[16]

2.1.4.1 Lifecycle CO_2 Emissions Analysis of Steel-Making Pathways

The GHG emissions associated with using DRI were assessed by comparing it with conventional BF-BOF and MIDREX EAF. These processes were evaluated using the GREET 2019 model for LCA, to estimate the equivalent CO_2 emissions for each process and highlight the benefits of using nuclear-H_2 in DRI production.

Figure 14 compares the equivalent CO_2 emissions per metric tonne (MT) of steel produced for four possible process steps in the steel-making process: 1) blast furnace/BOF (using coal), 2) MIDREX EAF using 100% natural gas, 3) MIDREX EAF using 70% natural gas and 30% Nuclear H_2, and 4) MIDREX EAF using only nuclear H_2. The GHG emissions from each respectively is: 1.97-MT eq.CO_2 /MT steel from BF, 1.47-ton eq.CO_2 /MT steel from EAF using 100% natural gas, 1.28-MT eq.CO_2 /MT steel from EAF using 70% natural gas and 30% Nuclear H_2, and 0.99-MT eq.CO_2 /MT steel from EAF using only nuclear-H_2.

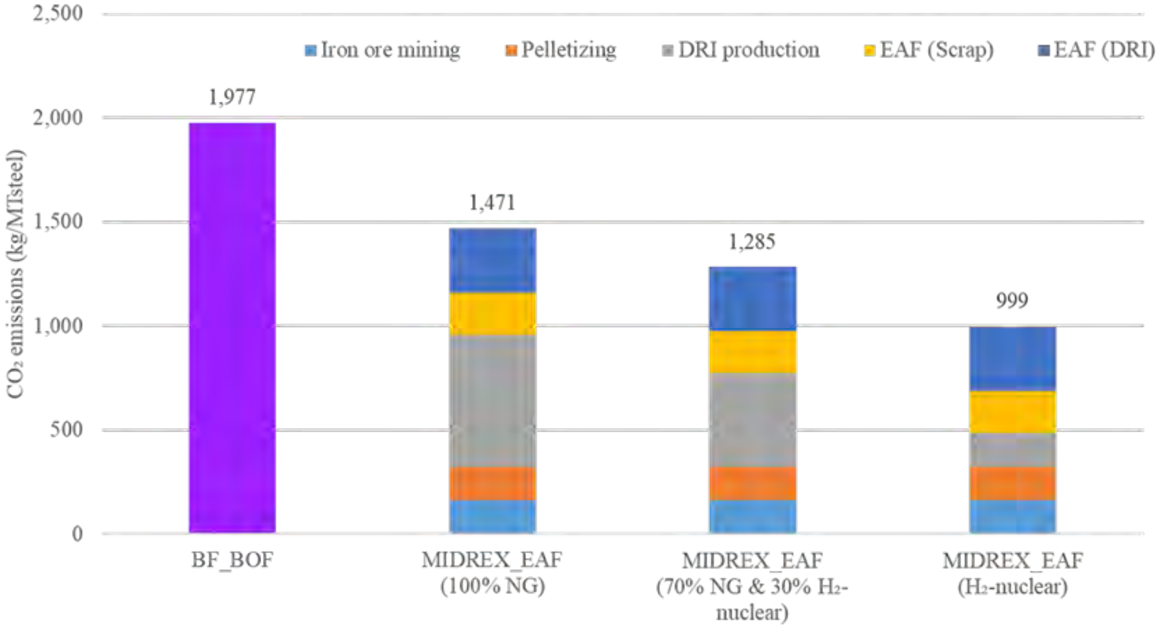

Figure 14. LCA of GHG emissions for various steel-making process options.

2.1.5 Ammonia and Fertilizers

Ammonia is produced by the Haber-Bosch process, which reacts hydrogen, usually produced from natural gas via the SMR process, with nitrogen separated from the air. In 2016, 13.6 MMT/year of ammonia were consumed in the U.S., with 9.8 MMT were produced domestically, while 3.8 MMT were imported,[19] with 12% of consumption being for non-agricultural products, and the remainder used to produce fertilizer products, including anhydrous ammonia, urea, diammonium phosphate, monoammonium phosphate, and nitric acid. The Haber-Bosch process uses hydrogen in a molar ratio of 3 moles H_2 to 2 moles of NH_3; therefore, 0.178 kg of hydrogen are required to produce 1 kg of ammonia. As ammonia is the source of nitrogen in other fertilizer products, we can generalize this as 0.216 kg hydrogen per kilogram of nitrogen in fertilizer.

Currently, 88% of domestic ammonia consumption is associated with fertilizer use. In addition to anhydrous ammonia, fertilizer products that are derived from ammonia include urea, ammonium nitrate, ammonium sulfate, and urea ammonium nitrate (UAN). Ammonia is also used in the production of explosives, plastics, synthetic fibers and resins, and numerous other chemical compounds. Ammonia usage for fertilizer applications is not expected to grow significantly in the coming years; increases in nitrogen fertilizer efficiency contribute to a projected minor decrease in ammonia demand for use in agricultural applications.[20]

A potential ammonia-market growth opportunity exists in the area of ammonia as an energy carrier. Ammonia could be used one day as a transportation fuel in ICEs. Ammonia may be suited for use in marine applications (i.e., industry with requirements for large-scale energy consumption and energy-storage requirements to enable travel of large vessels over long distances; additionally, marine applications would require fewer modifications to fuel-distribution networks to enable distribution of large volumes of an ammonia fuel product to the end user). Ammonia can be stored and transported as a liquid product, which can lead to significant reductions in storage and transportation energy requirements and costs when compared to hydrogen.[20,21]

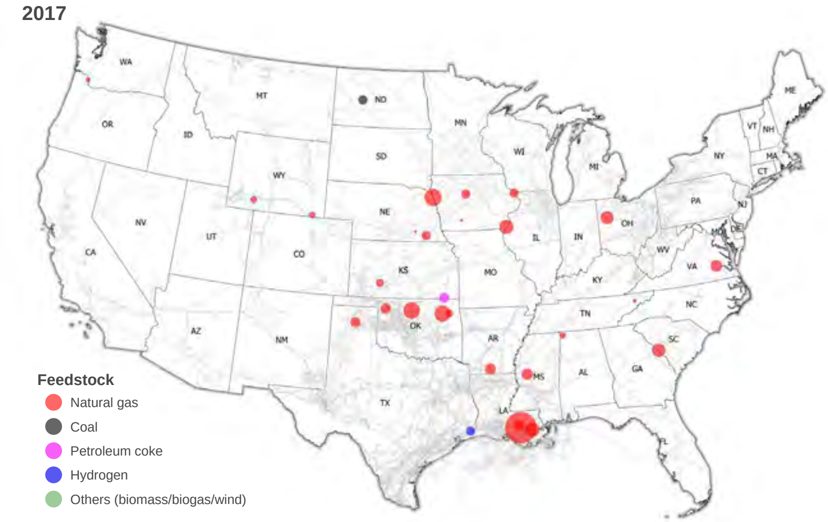

Figure 15. Estimated H_2 demand for U.S. NH_3 production in 2017.

Industry data on existing and planned NH_3 plants in the United States were used to estimate the input hydrogen required for NH_3 production by region (Ammonia Industry 2018). Figure 16 shows an estimated 25% increase in hydrogen demand for NH_3 production between 2017 and 2024. We assumed that domestic hydrogen demand for NH_3 production beyond 2024 would grow by another 15% by 2050, thus increasing annual hydrogen demand to 3.6 MMT.

The import share for ammonia consumption in the U.S. has been declining from 30% in 2019 to 17% in 2018, reflecting the expectation that domestic production may potentially displace all imports in the U.S. market due to the low-cost natural gas and the potential availability of low-cost green hydrogen.

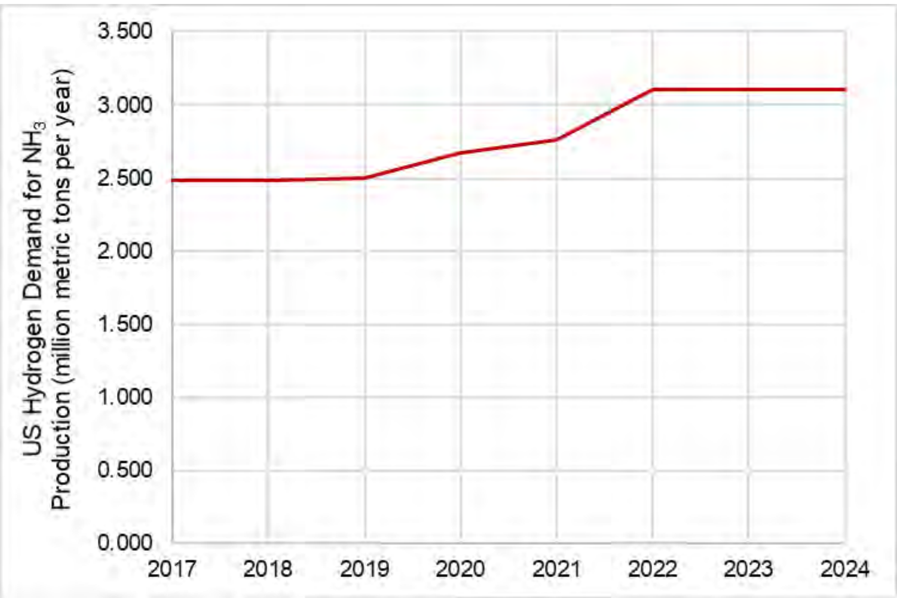

Figure 16. Estimated hydrogen demand for U.S. NH_3 production through 2024 (based on data from Ammonia Industry [2018] through 2024)

2.1.5.1 Life-Cycle CO_2 Emissions Analysis of H_2 from Nuclear Energy for Ammonia Production

To evaluate the environmental benefits and trade-offs for using nuclear-H_2 for ammonia production, the Haber-Bosch process was considered. The GREET 2020 model was used to conduct the LCA for ammonia production. Various production pathways for hydrogen were considered to understand the equivalent CO_2 emissions associated with various ammonia-feedstock sources and production pathways. Figure 17 compares CO_2 emissions from the conventional ammonia-production process to an alternative synthesis process using nuclear heat and electricity for H_2 and ASU in terms of eq CO_2 per MT NH_3. The conventional pathway produces about 2.55 MT CO_2/MT NH_3 while the nuclear for both H_2 and ASU produce 0.06 MT CO_2/MT NH_3, respectively, on a life-cycle basis. Life-cycle analysis results for conventional ammonia production process and alternative ammonia production process (ASU-N_2 and HTE-H_2) were acquired from Liu et al.[22]

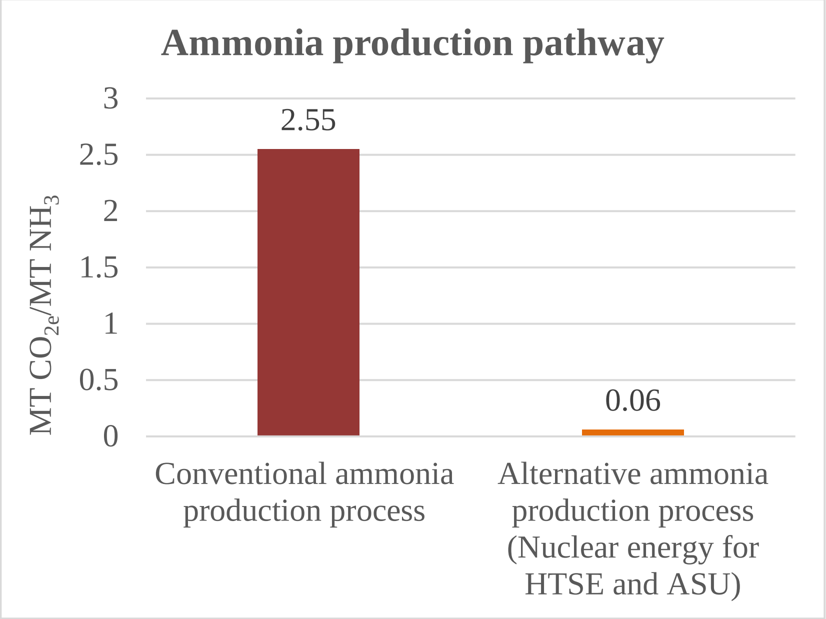

Figure 17. Life-cycle CO_2 emissions for conventional ammonia production versus ammonia production using NE for hydrogen generation and nitrogen separation in an ASU.

2.1.6 Synthetic Fuels

Synthesis gas (syngas) is a mixture of carbon monoxide and hydrogen. It is called syngas because these two molecules can be used to synthesize synthetic fuels (synfuels) and chemicals (synchemicals). The following sections discuss the markets for syngas and synfuels.

Significant quantities of high-purity CO_2 are generated in industry processes such as ethanol-production, SMR used for hydrogen production from natural gas for refining, and ammonia production. These high-concentration CO_2 sources present opportunities for the production of synfuels and synchemicals using a wide variety of pathways while minimizing the cost and energy penalty to capture CO_2 relative to other dilute CO_2 sources (e.g., from flue gases of coal and natural gas power plants). High-demand products that can be produced using CO_2 as a feedstock include methanol, Fischer-Tropsch (FT) diesel, and dimethyl ether (DME). In many of these processes, synthesis gas or syngas (a mixture of CO and H_2) is a key intermediate building block. Depending on the process pathway, hydrogen is used either in the initial reaction with CO_2 or in downstream processes such as refining of the synthetic fuels. In co-electrolysis, CO_2 and water are fed to an electrolyzer to produce CO, H_2 (syngas), and O_2. The syngas is then reacted to form any of a large variety of molecules. Methanol can also be produced by a variety of processes and presents an opportunity because its manufacturing process is relatively simple, and its global market is expected to grow for multiple uses, such as petrochemicals, fuel blending, or as a blend stock for transportation-fuel production. Methanol produced from waste CO_2 streams and hydrogen from clean NE offers a low-carbon alternative to methanol produced via the conventional process using natural gas. The merchant market for CO_2 is currently underused. Of the 100 MMT of CO_2 generated in concentrated forms from ethanol-production and SMR, only 11 MMT of CO_2 is currently used in merchant markets such as food processing and carbonated beverages. Production of synfuels using this quantity of CO_2 would create a maximum hydrogen demand of 14 MMT/year.[23] This leaves a significant CO_2 resource availability which could be used for methanol and synfuel production, depending on the region and company-specific economics, incentives, and decarbonization goals. In this report, we focus only on the potential hydrogen demand for synfuel production from highly concentrated sources of CO_2. The potential hydrogen demand for methanol production from the same CO_2 sources will be of a similar magnitude; thus, producing one chemical or fuel in place of the other will result in similar hydrogen demand considering the same CO_2 resources.

The hydrogen demand for synfuel production can be estimated based on the stoichiometric 1:3 mole ratio of CO_2 to H_2 that is required for the synthesis of FT diesel or DME. The availability of high-purity CO_2 from SMR, associated with merchant hydrogen and ammonia production, and the locations of facilities are based on values reported by facilities to the U.S. Environmental Protection Agency's Greenhouse Gas Reporting Program.[24] The H_2 demand reported here for refineries includes only conventional current hydrogen demand (hydrocracking, etc.) and does not include any future hydrogen demand for synfuel production using CO_2 that could be captured from SMRs that operate in these refineries. High-concentration CO_2 sources from ethanol-production are estimated based on the 1:1 mole ratio of ethanol to CO_2 generated during the conversion of glucose and sucrose in the fermentation process.

In 2017, 15.6 billion gal of ethanol were produced in the U.S.,[25] which generates an estimated 44 MMT of high-purity CO_2. The locations and capacities of ethanol-production facilities are based on an EIA dataset[25] and illustrated in Figure 18, while production by facility is estimated based on the national average capacity-usage rate.

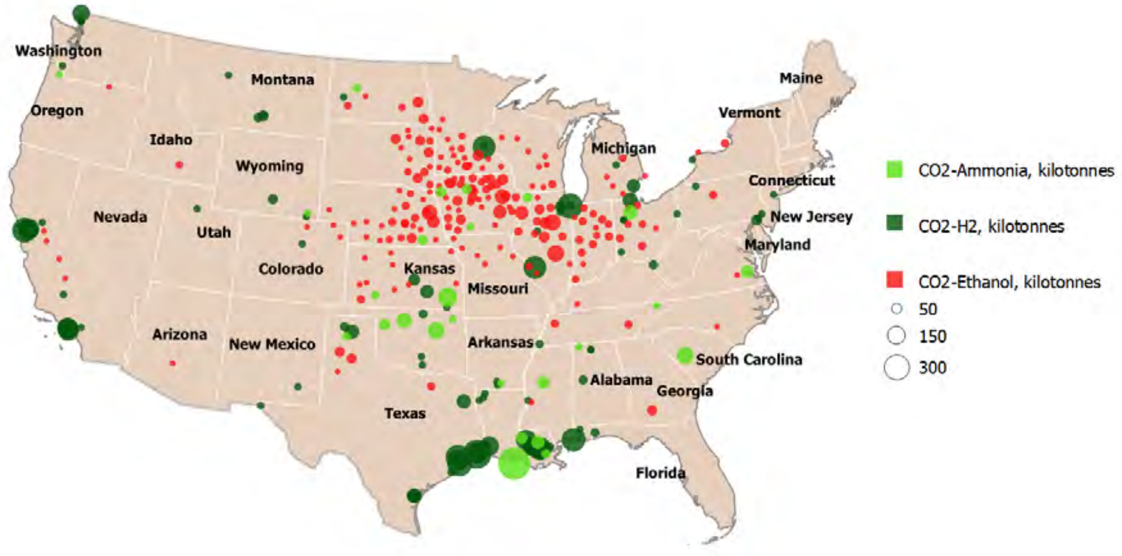

Figure 18. CO_2 sources for use in synfuels production.

Potential hydrogen demand for high-purity CO2 (~44 MMT) from ethanol plants if converted into to near-carbon-neutral synthetic FT fuels would be about 5 MMT which is approximately 2.3 billion gallon of FT fuel. Whereas potential hydrogen demand for producing synthetic methanol would be about 6 MMT, as mentioned in Elgowainy, 2020.[5] On the other hand, if all recoverable CO2 from these sources were utilized for synfuel production, the maximum annual supply of synfuels would be approximately 10 billion gallons. Production of this quantity of synfuels would create a maximum hydrogen demand of 14 MMT/year. Of this 14 MMT/year, 6 MMT hydrogen/year would be needed to convert all of the CO2 from ethanol plants into synfuels, 5.9 MMT H_2/year would be needed to convert all CO2 from refinery SMR plants into synfuels, and 2.1 MMT/year would be needed to convert all CO_2 from NH_3 plants into synfuels.[5]

While most ethanol-production is clustered in Midwest states, ammonia plants are located in a broader area, mainly in the Midwest, Gulf Coast, and Southeast, while other SMR plants are located near petroleum refineries, mostly along the Gulf Coast and near San Francisco, Los Angeles, Chicago, Detroit, Minneapolis, St. Louis, and Toledo. High-purity CO_2 sources for synfuel production are shown in the demand tables at the end of Section 2 for Monticello and PI NPPs, associated with ethanol plants. Ammonia plants are not listed in the demand tables as there are none within 100 miles of the PI and Monticello NPPs.

2.1.6.1 Life-Cycle CO_2 emissions analysis of H_2 from nuclear energy for synthetic transportation fuels

The GREET 2020 model was used to estimate GHG emissions assuming captured CO_2 and nuclear-H_2 for producing these synfuels. A recently published study on the production of electro-fuels from renewable H_2 and waste CO_2 streams are of increasing interest because of their CO_2 emission-reduction potentials compared to fossil counterparts.[26] This study evaluated the WTW GHG emissions of FT fuels from various electrolytic H_2 pathways and CO_2 sources, using various process designs (i.e., with and without H_2 recycle) and system boundaries. The FT fuel-synthesis process was modeled using Aspen Plus, which showed 45% of the carbon in CO_2 can be chemically bound up or fixed in the FT fuel, with a fuel production energy efficiency of 58%. Using nuclear electricity, stand-alone FT fuel production from various plant designs can reduce WTW GHG emissions by 90–108%, relative to petroleum fuels. The FT fuel nuclear LTE-H_2 recycle case was the base case for producing FT fuel using CO_2 and H_2 from nuclear sources considered.[26]

Figure 19 compares the GHG emission to produce conventional fuels, such as gasoline and jet and diesel fuels, to highlight the benefits of the FT pathway using nuclear H_2. The GHG emissions per megajoule for various fuels like gasoline, jet fuel, diesel fuel, and FT fuel (using nuclear H_2) are 93, 86, 91 and 9 g CO_2 eq./MJ, respectively (referred to as base synfuels case). The land-use change emissions for ethanol are associated with the process by which human activities transform the natural landscape, referring to how the land has been used, usually emphasizing the functional role of land for economic activities.

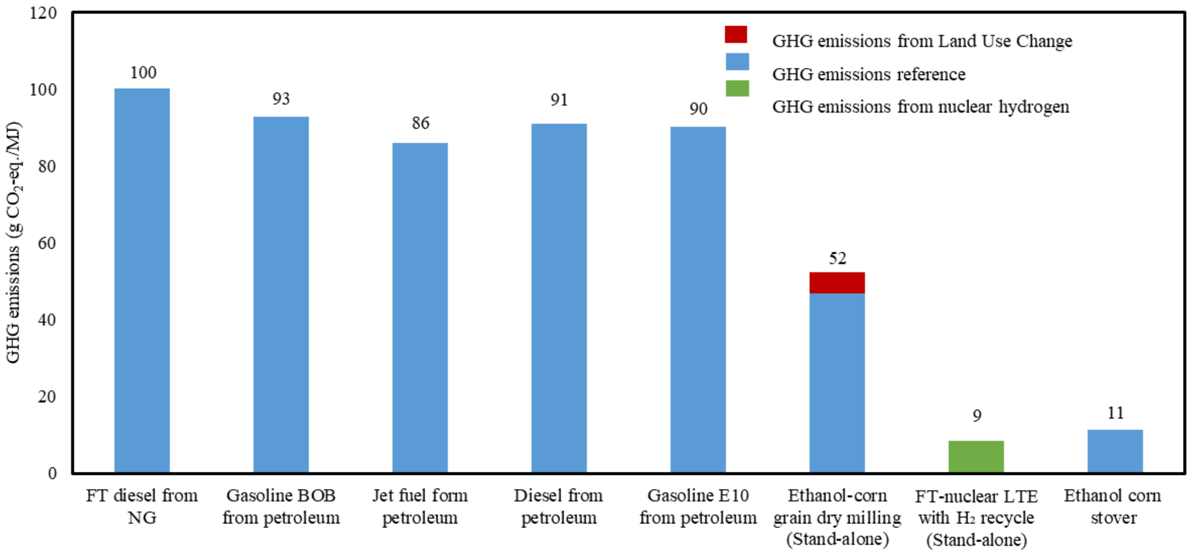

Figure 19. WTW total-CO_{2e} emissions for gasoline and jet, diesel, and FT fuel (produced using nuclear H_2). Before oxygenate blending (BOB), is motor gasoline before the required amounts of oxygenate (ethanol) have been blended.

2.1.7 Summary of U.S. Potential Hydrogen Demand Forecast

Table 5 summarizes the potential hydrogen demand for the U.S. forecasted in the year 2050.

Table 5. Summary of U.S. hydrogen demand potential in 2050.

Application	Potential Hydrogen Consumption[5] [MMT/yr.]
Petroleum Refineries	7.5
Ammonia	2.5
Synfuels from Ethanol - CO_2	5
Injection to natural gas Infrastructure	16
Iron Reduction and Steelmaking	8
Light-Duty FCEVs (Cars and Trucks)	21
Medium-Duty + Heavy-Duty FCEVs	8

2.2 Regional Potential Hydrogen Demand in the Minnesota Region

The Xcel Energy Monticello and PI NPPs provide opportunities for producing near-zero-carbon hydrogen and other non-electric products for various potential markets. The potential cumulative current and future hydrogen demands out to 2030 in the regions surrounding these NPPs are examined and evaluated in the following sections leveraging methods used to estimate national potential hydrogen demand in Sections 2.1–2.1.7.

2.2.1 Prairie Island Nuclear Power Plant

The PI NPP, shown in Figure 20, is a 1,100 MW facility with two pressurized water reactors (PWRs). It is located about 40 miles southeast of Minneapolis-St. Paul, in Red Wing, Minnesota. Tables in Appendix F show the cumulative potential hydrogen demand within 100 miles of Prairie Island Nuclear Generating Plant.

Figure 20. Prairie Island Nuclear Generating Plant in Welch, Minnesota.

Current hydrogen demand near the Prairie Island Generating Plant is predominantly from the Marathon Petroleum Corp. St. Paul Park and Koch Industries, Inc., Flint Hills Resources, Pine Bend refineries located in Saint Paul, Minnesota, and Rosemont, Minnesota, respectively, within 30 miles of the NPP. The combined hydrogen demand from these two refineries is up to 310 MT/day. The rest of the potential near-term demand, 82 MT/day, is associated with the co-combustion of hydrogen with natural gas in 38 gas-powered electricity generators located within 100 miles driving distance from Prairie Island Generating Plant. The "natural gas electricity generators" in the following tables comprise the hydrogen demand calculated for each of these electricity generators if they were to use a mixture of 30 vol% H_2 mixed with natural gas. The cumulative potential hydrogen demand by type and distance near the Prairie Island Generating Plant are plotted and mapped in Figure 21 and Figure 22.

The future potential hydrogen demand for the Prairie Island Generating Station is about 905 MT/day, from potential markets within 100 miles of the NPP. The majority of the future potential hydrogen demand is from the two refineries previously mentioned and their associated SMRs, which have a combined demand for 400 MT/day of hydrogen for the refining process. Although not considered in this report, note that if CO_2 were to be separated from the SMR effluents at the refineries considered, this could be used as an additional future opportunity to produce synthetic fuels requiring more hydrogen demand. The five ethanol plants, Al-Corn Clean Fuel at Claremont; Guardian Energy, LLC, at Janesville; Pro Corn, LLC, at Preston; Big River Resources Boyceville, LLC, in Boyceville; and Heartland Corn Products in Winthrop are located within 100 miles of the Prairie Island Generating Station. Each of these is a source of high–purity CO_2 from which CO_2 could be separated and used to produce synthetic fuels requiring hydrogen at potentially 410 MT/day for synthetic fuels manufacturing process. The majority of the potential hydrogen demand is in the area northwest of the Prairie Island Plant, suggesting a potentially large-scale hydrogen infrastructure that could serve multiple demands in that region, thus reducing the cost of hydrogen delivery per unit of hydrogen demand in each location (assuming all demand can be realized within a narrow timeframe).

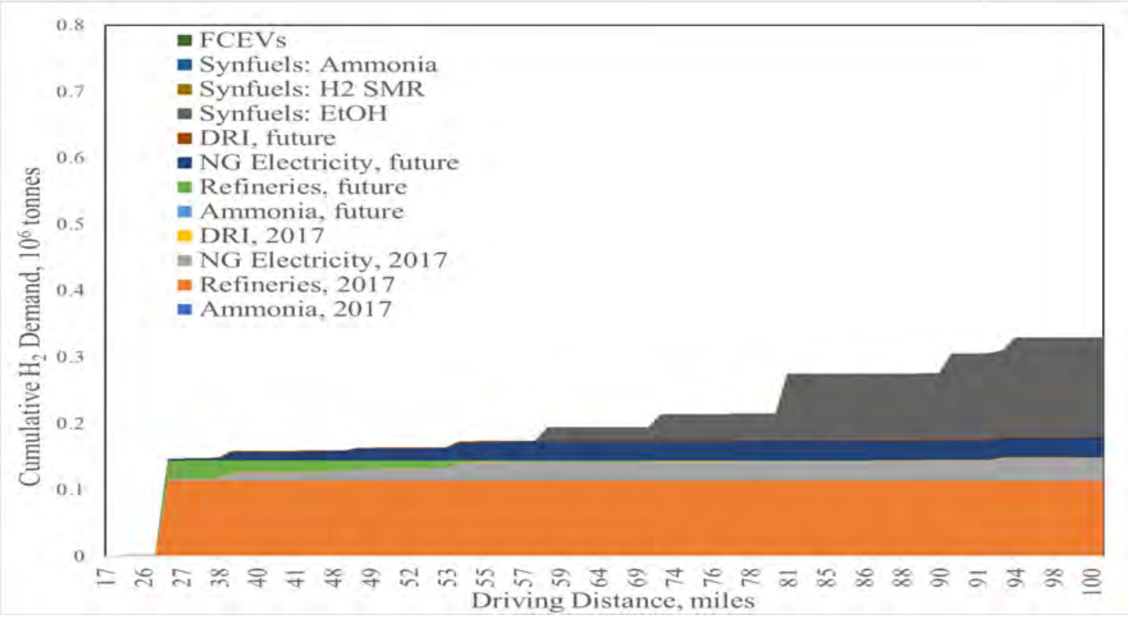

Figure 21. Cumulative potential hydrogen demand by type and distance near the Prairie Island Generating Plant.

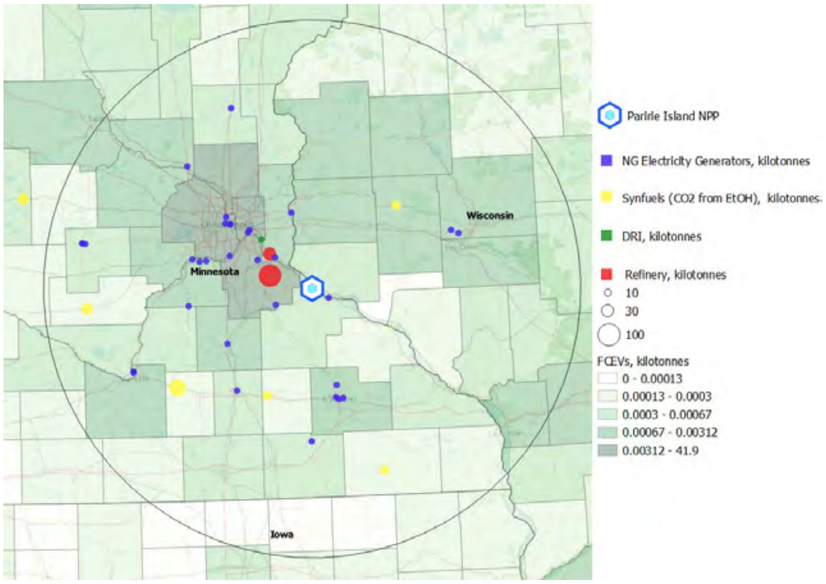

Figure 22. Future potential hydrogen demand near the Prairie Island Generating Plant.

2.2.2 Monticello Nuclear Power Plant

The Monticello NPP, shown in Figure 23, is a 647 MW facility with one boiling water reactor (BWR) located along the Mississippi river, northwest of Minneapolis-St. Paul, in Monticello, Minnesota. Figure 24 and Figure 25 show the cumulative potential hydrogen demand within 100 miles of the Monticello Nuclear Generating Plant.

Figure 23. Monticello Nuclear Generating Plant in Monticello, Minnesota.

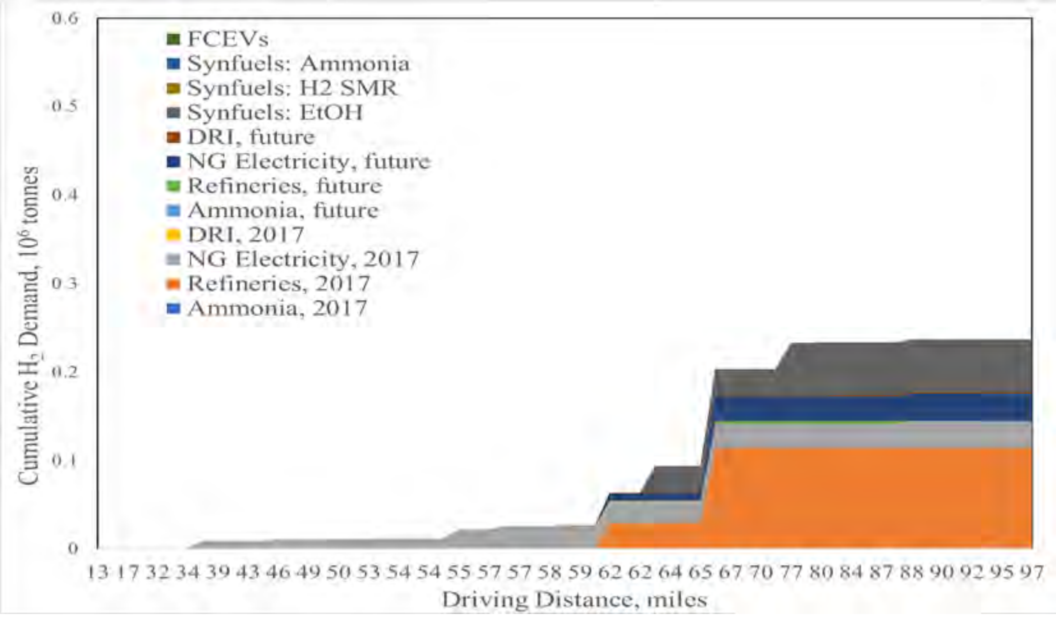

Figure 24. Cumulative potential hydrogen demand by type and distance near the Monticello Nuclear Generating Plant.

Current and near-term hydrogen demand near the Monticello facility depends mainly on the co-combustion of hydrogen with natural gas for electricity generation and on two refineries which add to this demand (assuming such demand will not be satisfied with Prairie Island plant). Tables showing the hydrogen demand surrounding Monticello are found in Appendix F. The cumulative near-term potential hydrogen demand for this location is 400 MT/day. About 27 natural gas electricity generators are located within 100 miles of this facility, and these have a combined potential hydrogen demand of 85 MT/day. Two refineries, Marathon Petroleum Corp. St. Paul Park, and Koch Industries, Inc., Flint Hills Resources, Pine Bend, have an estimated hydrogen demand of 310 MT/day. The cumulative potential hydrogen demand by type and distance near the Monticello generating station are plotted and mapped in Figure 24 and Figure 25 respectively.

Future hydrogen demand near Monticello's location would be mostly for synthetic-fuels production, petroleum refineries, and for co-combustion of hydrogen with natural gas. The natural gas electricity generators column reflects the hydrogen demand calculated for each of these electricity generators if they were to use a mixture of 30 vol% H2 with natural gas. Synthetic-fuel-producing facilities using CO_2 only from ethanol facilities within 100 miles would have a combined future potential demand of 164 MT/day. The two refineries increase the potential demand for hydrogen around the Monticello region, with a total of about 400 MT/day. The cumulative future hydrogen potential demand for Monticello NPP will be about 650 MT/day within 100 miles.

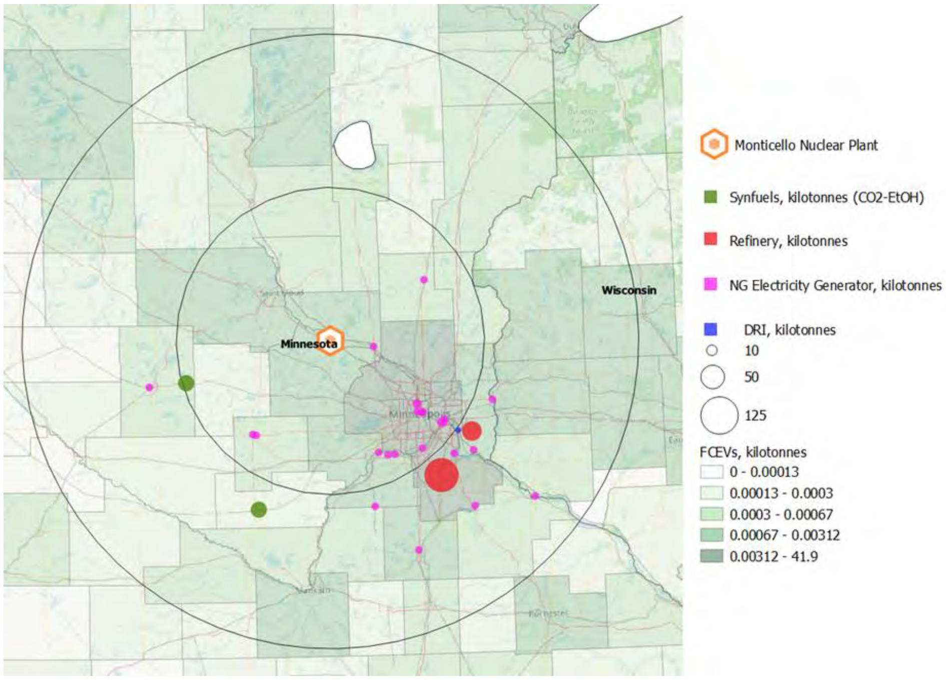

Figure 25. Future potential hydrogen demand near the Monticello Generating Plant.

2.2.3 Overlapping Demand Between Prairie Island and Monticello NPPs

Xcel Energy's nuclear facilities in Minnesota (at Monticello and Prairie Island) were evaluated for potential hydrogen demand for different markets. Because of the proximity of these two power plants, there is significant overlap of the estimated potential H_2 demand for these two NPPs. To recommend which NPP should provide hydrogen to these markets, demand size and distance were assessed for the overlapping markets, so as to facilitate lowest transportation cost, which has major impact on the breakeven cost of hydrogen.

The majority of the potential current and future hydrogen demand near Xcel Energy's plants is in the Minneapolis/St. Paul area. The overlapping demand is presented in Appendix F, with the type and distances within 50 miles of the respective nuclear facilities. As shown in Figure 26, the majority of the overlapping demand is closer to Xcel's PI NPP. This overlapping cumulative potential future demand is about 337 kMT/ year (Figure 26).

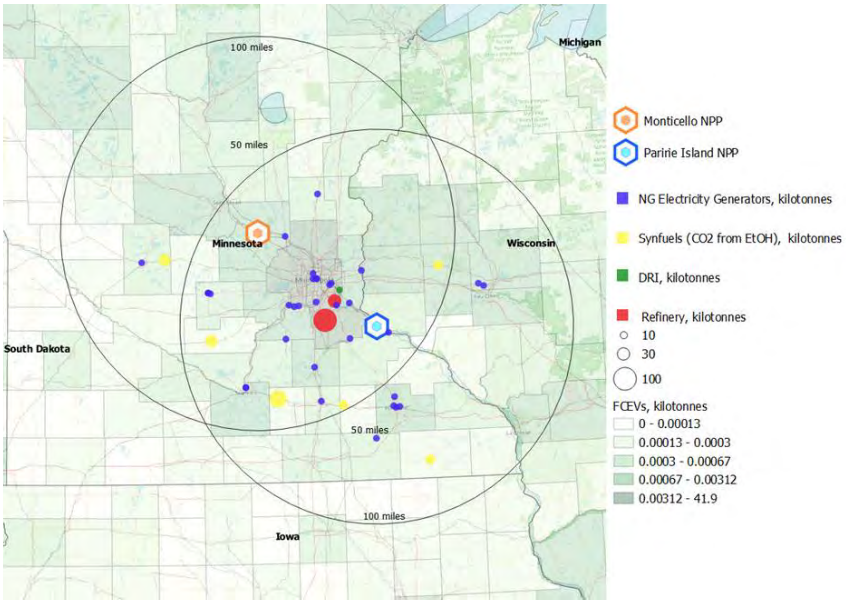

Figure 26. Overlapping future potential hydrogen demand withing 50 miles of Prairie Island and Monticello power plants.

2.2.4 Transportation of Hydrogen Using Pipelines for Xcel Energy’s NPPs

In order to reach these hydrogen markets, a network of pipelines is proposed, starting from the PI and Monticello NPPs respectively, which can cover these locations and 90% of the total overlapping demand. These pipeline proposals are based on current generating portfolios and may shift as coal power shuts down and other generating sources are added, replaced, or modified. Hypothetical pipelines are planned adjacent to the roads in the region and are sized according to the amount of hydrogen required for these markets. The demand location covered by both pipeline networks is in the table below; labels are used to link these demand points.

Prairie Island NPP’s pipeline network covers 17 concurrent demand locations, which include two refineries and several natural gas electricity generators. The total hydrogen demand served by this pipeline network is 172 kMT/year. The pipeline starts from PI NPP goes to Koch Industries, Inc., Flint Hills Resources Pine Bend Refinery (27 mi), continues to Marathon Petroleum Corp., St. Paul Park Refinery (16 mi), and then splits into two branches to serve the rest of the demand locations (see Figure 27).

Monticello NPP’s pipeline network covers five concurrent demand locations: synfuel production and natural gas electricity generators with about 60 kMT/year total potential hydrogen demand. The pipeline starts from Monticello NPP and splits at a central location, labeled as 1, to serve all the demand location on its route (see Figure 28). The cost of H2 transportation using these pipelines is calculated using data provided in Appendix F. The network for both pipelines can be traced with link locations in Table 6. The pipeline size and its levelized cost of hydrogen delivery are calculated based on ANL’s Hydrogen Delivery Scenario Analysis Model (HDSAM).[27] Depending on the size and location of demand, the hydrogen delivery cost ranges between \$0.05–0.25/kg, assuming that a pipeline network will be built to concurrently serve all potential hydrogen demand locations. The potential hydrogen demand curves for the Prairie Island and Monticello power plants after adjusting to account for the hydrogen pipeline delivery cost are provided below in Figure 27 and Figure 28, respectively.

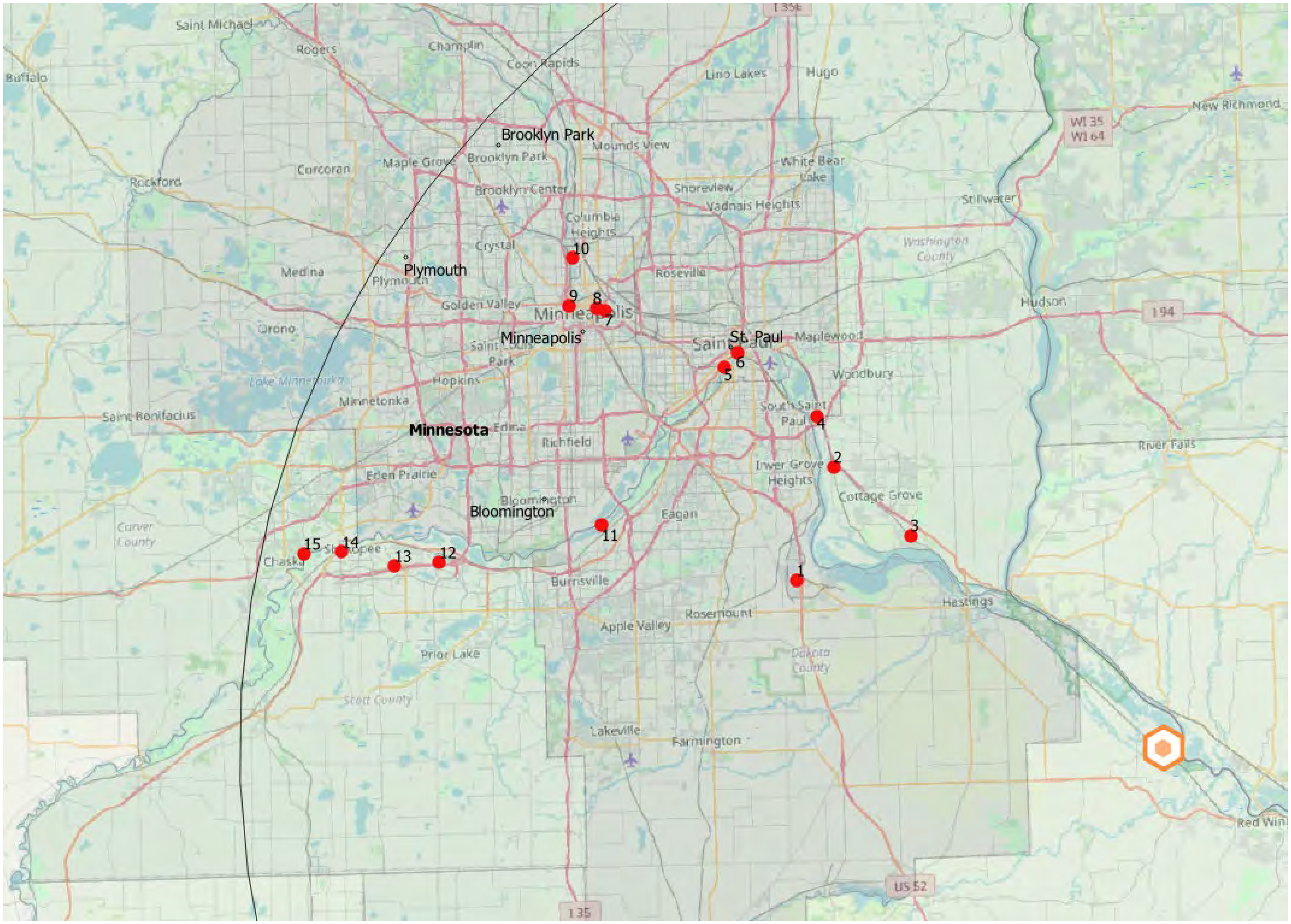

Figure 27. Demand locations with labels covered by the pipeline network for Prairie Island Nuclear Power Plant. Red dots: location for H_2 demand).

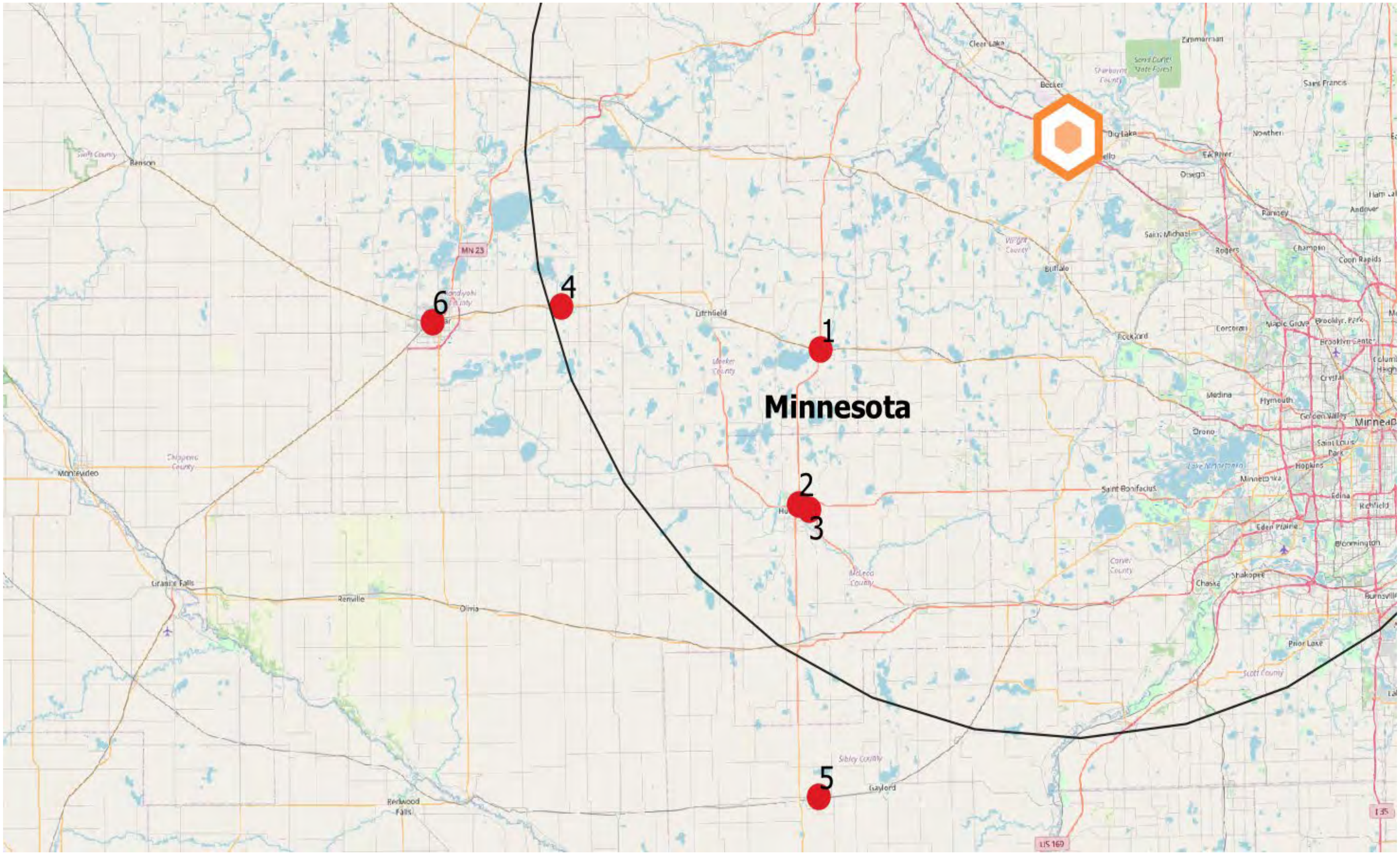

Figure 28. Demand locations with labels covered by the pipeline network for Monticello NPP. Red dots: location for H_2 demand.

Table 6. Pipeline network covered for transportation near Xcel Energy’s Monticello and Prairie Island NPPs.

Pipeline Link	Point	From	To	Pipeline Distance, Miles	Future Potential Demand, Kilotonnes
Prairie Island NPP Pipeline network					
0	1	PI NPP	Koch Industries Inc Flint Hills Resources Pine Bend Refinery	27	107.8
1	2	Koch Industries, Inc., Flint Hills Resources Pine Bend Refinery	Marathon Petroleum Corp. St. Paul Park Refinery	15	36.6
2	3	Marathon Petroleum Corp. St. Paul Park Refinery	LSP-Cottage Grove LP: Cottage Grove Operating Services LLC	6.3	1.7
3	4	Marathon Petroleum Corp. St. Paul Park Refinery	Gerdau Long Steel North America–St. Paul	3.2	1.4
4	5	Gerdau Long Steel North America, St. Paul	High Bridge: Northern States Power Co–Minnesota	10.6	10.2
5	6	High Bridge: Northern States Power Co, Minnesota	St. Paul Cogeneration: St. Paul Cogeneration LLC	2.3	0.4
6	7	St. Paul Cogeneration: St. Paul Cogeneration, LLC	Univ Minnesota CHP Plant: Veolia Energy	10.3	0.83
7	8	Univ Minnesota CHP Plant: Veolia Energy	Southeast Steam Plant: Veolia Energy	1.3	0.38
8	9	Southeast Steam Plant: Veolia Energy	Covanta Hennepin Energy: Covanta Energy Co	2.4	0.03
9	10	Covanta Hennepin Energy: Covanta Energy Co	Riverside (Minnesota): Northern States Power Co–Minnesota	3.9	8.48
1	11	Koch Industries Inc, Saint Paul	Black Dog: Northern States Power Co–Minnesota	17.7	3.68
11	12	Black Dog: Northern States Power Co., Minnesota	Blue Lake: Northern States Power Co–Minnesota	11.7	0.65
12	13	Blue Lake: Northern States Power Co., Minnesota	Shakopee Energy Park: Minnesota Municipal Power Agency	4.5	0.06
13	14	Shakopee Energy Park: Minnesota Municipal Power Agency	Koda Biomass Plant: Koda Energy LLC	3.3	0.16
14	15	Koda Biomass Plant: Koda Energy LLC	Minnesota River: Minnesota Municipal Power Agency	5.3	0.004
Monticello NPP pipeline network					
0	1	Monticello NPP	Fictitious Pipeline location/ Link	42	NA
1	2	Fictitious Pipeline location	Hutchinson Plant #1: Hutchinson Utilities Commission	12.6	0.06

Pipeline Link	Point	From	To	Pipeline Distance, Miles	Future Potential Demand, Kilotonnes
2	3	Hutchinson Plant #1: Hutchinson Utilities Commission	Hutchinson Plant #2: Hutchinson Utilities Commission	1.5	0.14
3	5	Hutchinson Plant #2: Hutchinson Utilities Commission	Heartland Corn Products, Winthrop	36.5	30
1	4	Fictitious Pipeline location	Bushmills Ethanol Inc, Atwater	28.8	30
4	6	Bushmills Ethanol Inc, Atwater	Willmar: Willmar Municipal Utilities	12.2	0.06

2.2.5 Delivery Cost Adjusted Demand Curves for Prairie Island and Monticello NPPs

Hydrogen-demand curves for Prairie Island and Monticello NPPs are shown in Figure 29 and Figure 30, respectively. The H_2 price for each demand type and location is adjusted to account for carbon credits and delivery costs. The carbon credits were calculated based on an assumed carbon tax of $22.20 per short ton of CO_2, based on the Xcel Energy 2019 IRP, and the estimated carbon reduction when using nuclear-H_2 to displace the carbon emission for different products through their conventional pathways. The hydrogen prices were calculated by assuming H_2 being delivered to demand points using the pipeline network in Figure 29 and Figure 30 for Prairie Island and Monticello NPPs respectively. The amount of H_2 delivered to each demand location is different; therefore, the cost of H_2 delivery (transmission) is unique to each demand location.

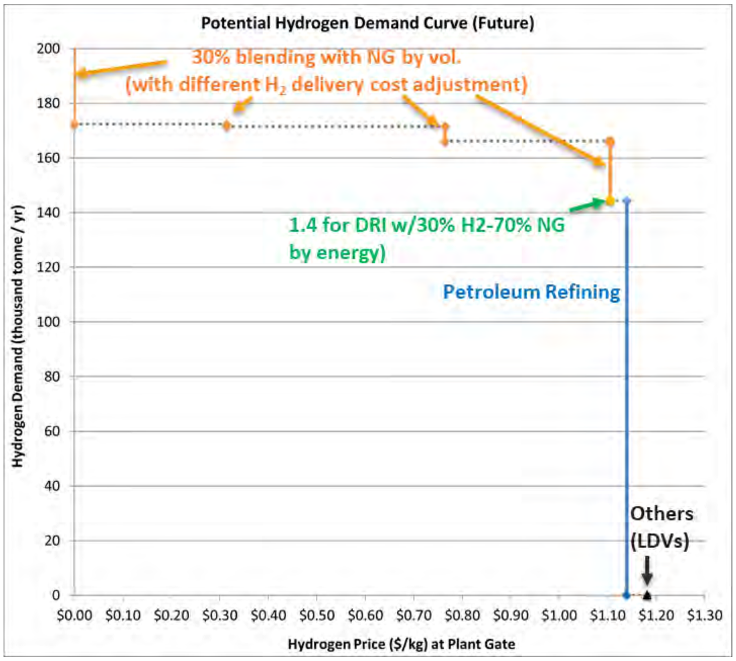

Figure 29. Potential hydrogen demand curve near Prairie Island Nuclear Generating Plant.

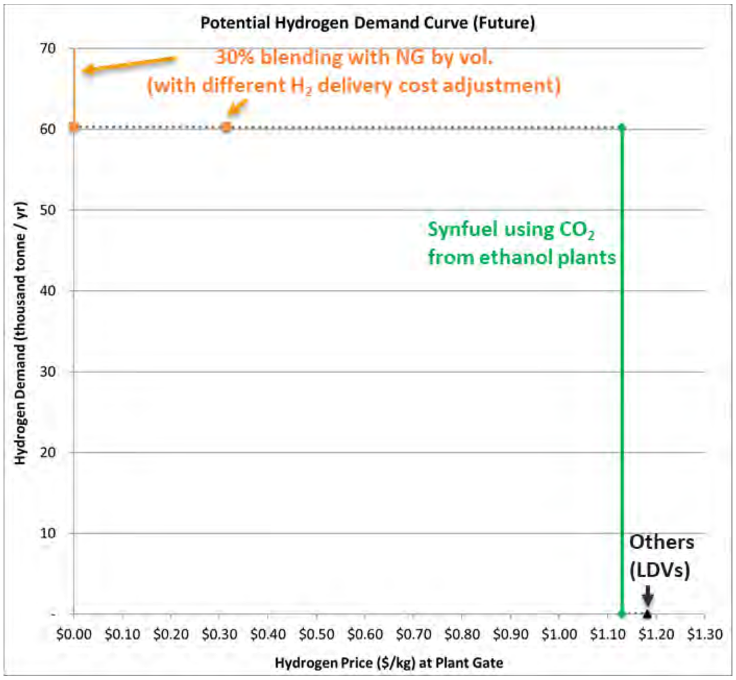

Figure 30. Potential hydrogen demand curve near Monticello Nuclear Generating Plant.

For natural gas generators and DRI, hydrogen is assumed to replace natural gas, so the breakeven price for H_2 has to compete with natural gas pricing on an energy basis. For petroleum refineries, SMR-H_2 is assumed to be replaced by LWR-Nuclear-H_2 and thus has to compete with natural gas-SMR-H_2 price ($1.03 per kg). For Synfuels produced using Nuclear H_2 it is assumed that these synfuels would replace conventional diesel and will compete with diesel prices (2050, $3.1 per gal). Similar for FCEVs the Gasoline- ICEVs are assumed to replace FCEVs using Nuclear- H_2.

These breakeven prices were then adjusted to account for carbon credits and hydrogen delivery prices to construct the demand curve for Prairie Island and Monticello NPPs using the equation below. Table 7 shows the breakeven H_2 prices and CO_2 credits for different demand types. As mentioned earlier, the H_2 delivery prices are unique to each demand location depending on the delivered amount and distance from nuclear power plant. The H_2 transmission cost was calculated using ANL's HDSAM by assuming a pipeline network that connects the NPPs to the various surrounding demand location. These demand curves have been used for further analysis in the economic dispatch and optimization section of the report.

The equation for calculating the cost adjusted hydrogen price used in the demand curves.

$$Hydrogen\ price\ \left(\frac{\$}{kg}\right) = Breakeven\ price\ for\ the\ specific\ market\ \left(\frac{\$}{kg}\right) + \ Carbon\ credits\ \left(\frac{\$}{kg}\right) - Hydrogen\ Transmission\ cost\ \left(\frac{\$}{kg}\right)$$

Table 7. Breakeven H_2 prices and CO_2 credits for hydrogen demand markets.

Demand Markets	Fuel being replaced using Nuclear H_2	Breakeven H_2 Price	*CO_2 Credits ($/kg)	Notes regarding breakeven H_2 price
NG Generators	Natural Gas	$0.53	$0.28	H_2 vs natural gas on BTU basis
DRI	Natural Gas	$0.53	$0. 58	DRI w/30% H_2 in syngas vs 100% natural gas
Synfuels	Conventional Diesel	$1.14	$0.22	H_2 for FT to breakeven with untaxed diesel price ($3.1/gal) (2050)
FCEVs	Conventional Gasoline	$1.03	$0.27	FCEV vs. Gasoline ICEV

*CO_2 credits of $22.20 per short ton was assumed per the Xcel Energy 2019 IRP.

The transmission cost for H_2 to each demand point was calculated using HDSAM.[27] The CO_2 credits in Table 7 above are calculated based on the assumed carbon tax of $22.20 per short ton of CO_2 and the difference in life-cycle carbon intensity between conventional products and the equivalent low-carbon products facilitated by nuclear-hydrogen production. Table 8 below shows the magnitude of the CO_2 credits per kilogram of hydrogen for a range of carbon tax between $0 and $200 per short ton of CO_2. This table shows the credit that nuclear-hydrogen can accrue per kilogram of hydrogen used in various applications at various CO_2 prices due to the associated life-cycle carbon-intensity reduction in these applications compared to the carbon intensity of their conventional pathways. As an example, the nuclear-hydrogen can be produced competitively with a premium of $2.43/kg over the cost of SMR-H_2 when the CO_2 price is at $200/short ton.

Table 8. CO_2 credits per kg of hydrogen for a range of carbon taxes per short ton of CO_2.

CO_2 price/short ton	Nuclear H_2 CO_2 vs. natural gas SMR_H_2	Nuclear H_2 vs. natural gas on energy basis	DRI w/30% H_2 in syngas vs. BF-BOF	Synfuel from Nuclear H_2 vs. conventional diesel	H_2 FCEV vs. conventional gasoline ICEV
$0.00	$0.00	$0.00	$0.00	$0.00	$0.00
$22.20	$0.27	$0.28	$0.58	$0.22	$0.27
$50.00	$0.61	$0.63	$1.30	$0.50	$0.61
$100.00	$1.21	$1.27	$2.61	$0.99	$1.21
$150.00	$1.82	$1.90	$3.91	$1.49	$1.82
$200.00	$2.43	$2.54	$5.21	$1.98	$2.43

3 LWR-HTSE HYDROGEN PLANT DESIGN MODEL DEVELOPMENT

This section presents a detailed engineering plant-design model and analysis for the integration of hydrogen production via SOEC/HTSE with an LWR NPP. This analysis represents original and groundbreaking modeling and analysis of current state-of-the-art HTSE technology integrated with nuclear power as well as forecasted performance improvements of HTSE technology. This section also describes various sensitivity studies on the cost to produce hydrogen (i.e., the LCOH) as well as a detailed optimization and discounted cash flow (DCF) NPV analysis. Thus, not only current and forecasted technologies are modeled, but the sensitivity and optimization studies give a sense for what is possible with the improvement of the process-input parameters. Plant design, analysis, and optimization of the input parameters, results, costs, and benefits of HTSE integrated with nuclear power is the main objective of this report.

A small portion of heat from an LWR is diverted to provide heat to the HTSE process, which can significantly increase its process efficiency. A detailed process and control model of both the thermal-delivery loop (TDL) and the nuclear reactor dynamics for TPE from nuclear power have been separately performed, and the status of this ongoing analysis is summarized in the appendix of this report. For the current analysis, a simplified model of TPE with all the necessary details was used. Figure 31 shows a diagram of the HTSE integrated with an NPP in a generic layout as designed and analyzed in this report. It is recognized that various iterations of designs for thermal power extraction are being studied and this configuration may not be the optimal final design. Other design options not included in this report could include removing after the high-pressure turbine, eliminating the TDL to use NPP steam to directly heat treated HTSE feedwater, decreasing the distance between the steam extraction and the HTSE, and returning condensate to the first NPP feedwater heater versus to the condenser. The scope of the analysis for this report is only for the TPE via the TDL and the hydrogen plant, as shown in Figure 31. The nuclear reactor dynamics and control of the TPE are rigorously modeled elsewhere, as mentioned in the TPE section in the appendix of this report. Other sections of this report deal with hydrogen transportation logistics and downstream use-case analysis of the produced hydrogen.

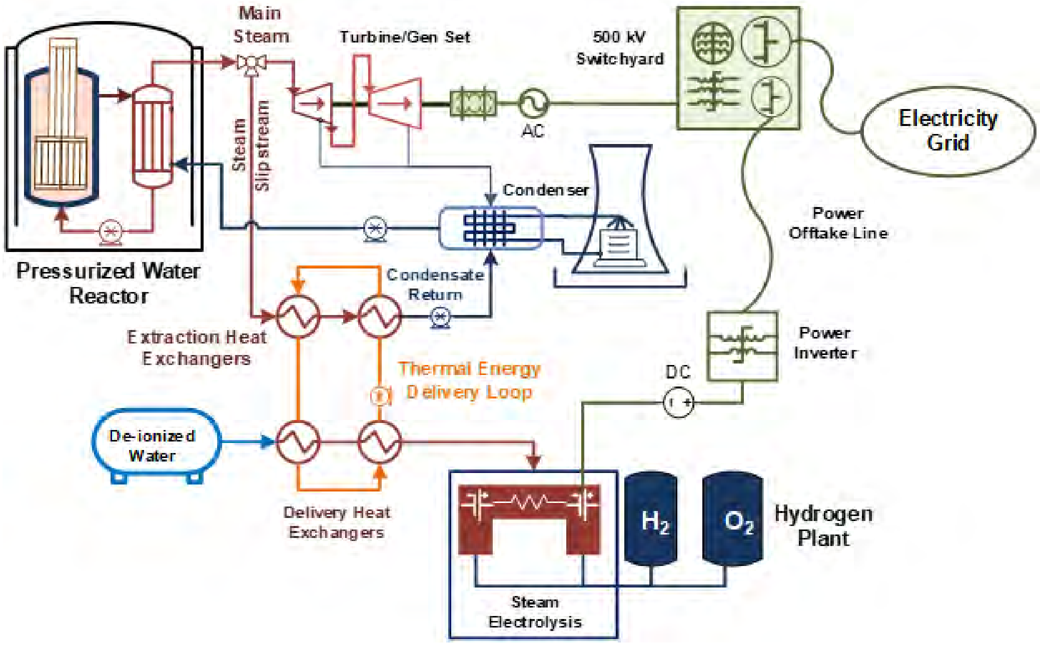

Figure 31. Overview of HTSE integrated with an NPP. Equipment added to the NPP includes the steam slip stream from the turbine inlet, the TDL, the HTSE hydrogen plant, and associated water and electricity supply tie-ins.

A process model of the NPP-HTSE was rigorously developed using AspenTech HYSYS to model the CAPEX and OPEX) as well as energy and utility requirements. This model could be used to provide inputs to front end engineering design (FEED) in a follow-on analysis. Outputs from the process model were used in the hydrogen production analysis model (H_2A) to determine the overall costs of hydrogen production.

SOEC/HTSE is a developing technology that is currently at a lower commercial scale than polymer electrolyte membrane (PEM) low-temperature electrolysis (LTE). But HTSE has advantages over LTE that will only amplify as the technology continues to develop. The greatest advantage of HTSE includes higher efficiency of hydrogen production and, therefore, reduction in the cost to produce hydrogen, especially when integrated with nuclear power.

As the name suggests, HTSE is operated at a higher temperature than LTE, which thermodynamically drives a higher reaction rate to the desired hydrogen product. LTE uses expensive catalysts to drive the hydrogen-production reaction rate. When integrated with nuclear power, HTSE can achieve cost reduction by using low-cost heat from the nuclear reactor to overcome the heat of vaporization of the water. Although the NPP heat is considered low grade at a temperature of up to 300°C, the NPP heat is used to overcome the large amount of latent heat energy needed to vaporize large volumes of water. Following vaporization, heat recuperation and topping heaters can be used to supply the sensible heat needed to raise the steam to HTSE operating temperature.

The TDL modeled as part of this study includes only major assumptions of equipment capital costs. It does not include cost allowance for NPP tie-ins, downtime, detailed control equipment for the TDL, or any nuclear reactor controls or regulatory reviews. Thus, the cost of the thermal integration is expected to be higher than estimated here; actual costs of thermal integration with a nuclear reactor will be more accurate coming from a utility company performing a separate study to include those costs.

The following analysis discusses the inputs, assumptions, methodology, and results, as well as various sensitivity studies, and concludes with a detailed optimization of the HTSE integrated with nuclear power in a regulated grid market.

3.1 Process-Modeling Design Basis

An LWR-integrated HTSE process model was developed using AspenTech HYSYS simulation software[28] for the purposes of (1) determining HTSE process energy requirements, (2) computing hydrogen production rates and the corresponding feed-water flow rate requirements, (3) establishing equipment-sizing parameters in support of capital-cost analysis, and (4) determining the maximum capacity HTSE plant that could be coupled with a specified LWR NPP.

A process-flow diagram (PFD) of the HYSYS model main HTSE process area is shown in Figure 32. This figure highlights the location of the SOEC stacks, the steam generator used to vaporize process-feedwater stream using nuclear process heat, the high-temperature electrical topping heaters, and the high- and low-temperature recuperators used to provide process-heat integration. Descriptions of the process subsystems included in the process model are included in Section 3.1.1. Process operating conditions and equipment performance specifications are detailed in Section 3.1.2.

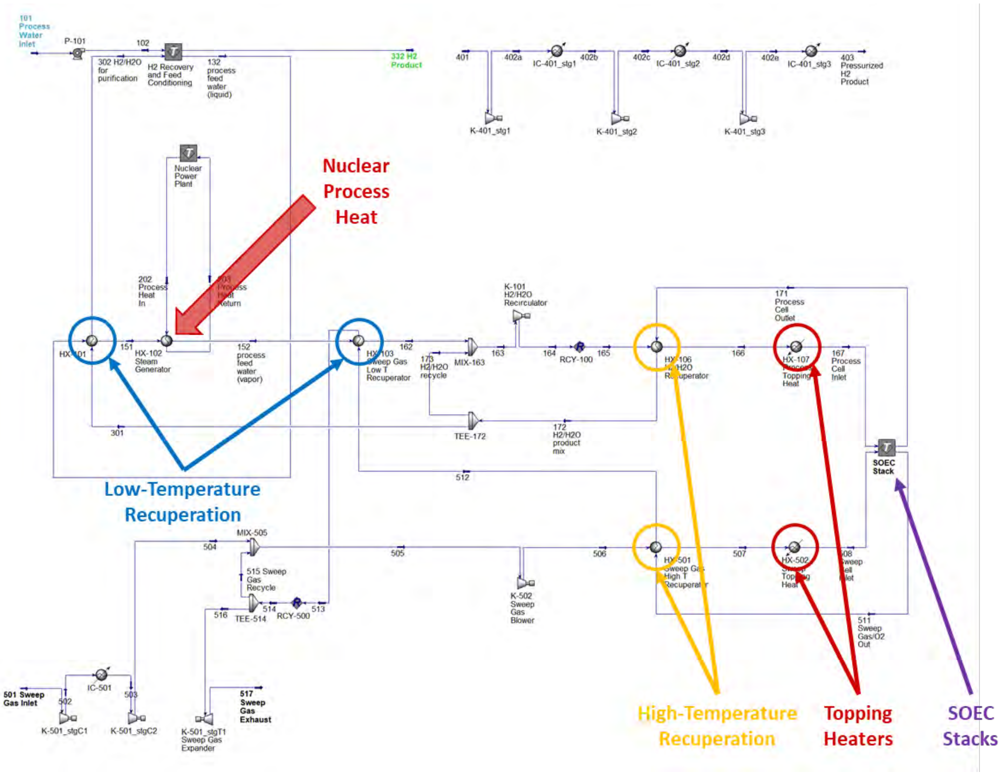

Figure 32. HTSE-process flow diagram.

3.1.1 Process Overview

The HTSE system evaluated includes several major process systems. These systems include (1) HTSE system, (2) feed and utility system, (3) air sweep-gas system, (4) hydrogen/steam system, (5) hydrogen purification system, (6) nuclear process-heat delivery system, (7) multistage product compression, and (8) the control system. A description of each of these process systems is included in the sections below. Process-flow diagrams with each of the separate process systems identified are included in the Appendix.

3.1.1.1 HTSE system

The HTSE system includes the SOEC stacks, the high-temperature recuperators, the trim heaters, and the insulated containment vessel that houses the stack array and provides pressure containment. The HTSE system also includes electrical-power distribution as well as instrumentation required to maintain the specified stack-operating conditions. The HTSE system recuperating heat exchangers are used to transfer heat from the high-temperature stack's outlet streams to the lower-temperature stack inlet streams; use of recuperators allows the T <300°C heat supplied by the LWR to be used primarily for feedwater vaporization (at temperatures in the 100–200°C range) because this heat is not available at a sufficient temperature to heat the hydrogen/steam stack inlet-gas mixture to the stack operating temperature range (700–800°C). The HTSE system electrical trim heaters adjust the temperature of the steam/hydrogen mixture entering the stack from the recuperator outlet temperature to the specified operating temperature.

3.1.1.2 Feedwater and utility systems

The feed and utility system includes the process components necessary to prepare and stage a clean, demineralized feedwater stream (separate from the NPP's process and steam-cycle water) for use in the HTSE process, including water filtration, purification, and storage, as well as the cooling and electrical-power-distribution systems needed to support HTSE process operation. Cooling towers are included to provide process cooling duty (used in the hydrogen purification system for cooling the process-gas streams and for providing compressor cooling). The feed and utility system also includes electrical-power transmission and distribution equipment to provide electrical-power connections between the nuclear plant and the HTSE site, transforming the power from the NPP substation voltage to the rectifier input voltage, inverting the AC power from the transmission system to DC power for use in the SOEC stacks and the bus bars for distributing the high-amperage current from the rectifier to the stacks.

3.1.1.3 Air sweep-gas system

During HTSE process operation, pure oxygen is generated on the anode side of the SOEC stacks. Because the stacks operate at elevated temperatures (700–800°C), oxidation of the SOEC system materials of construction is an operational issue if the oxygen concentration is not reduced. An air sweep-gas stream is used to dilute and evacuate high-concentration oxygen from the HTSE system. The sweep-gas system delivers the air sweep-gas stream to the stack at the specified operating temperature and pressure to minimize any thermal or pressure gradients between the anode and cathode sides of each cell, which reduces mechanical stresses on the cells. The enriched-oxygen air sweep-gas stream is released to the atmosphere following expansion through a pressure-recovery turbine to capture the energy in the stream. Because the flow rate of the sweep-gas outlet stream is greater than the flow rate of the sweep-gas inlet stream (due to the addition of oxygen in the stack), the net-power requirements of the sweep-gas compressor/expander are negligible in comparison with other HTSE-system power demands.

3.1.1.4 Hydrogen/steam system

The hydrogen/steam system vaporizes the feedwater stream and mixes the resulting steam with the specified quantity of recycled hydrogen exiting the stack. Low-temperature recuperators, the feedwater steam generator, high-temperature gas blowers, and piping/manifolds necessary to recycle a portion of the stack product gas comprise the hydrogen/steam system. (The presence of hydrogen in the stack inlet gas stream maintains reducing conditions important in minimizing SOEC degradation.) The low-temperature recuperators are used to preheat the liquid-phase feedwater while simultaneously cooling the H_2/H_2O mixture en route to the hydrogen-purification system.

3.1.1.5 Hydrogen purification system

The hydrogen/steam process-gas mixture in the stack-outlet stream flows through high- and low-temperature recuperators in the HTSE and hydrogen/steam systems to cool the stream to a temperature near the dew point. The hydrogen purification system uses multiple stages of cooling and compression to progressively condense a greater fraction of the water from the stream. In addition to using cooling water as a heat sink for the hydrogen purification system’s cooling operations, preheating the purified process feedwater provides a useful cooling duty for cooling/condensing steam from the hydrogen/steam process gas mixture. The hydrogen purification system is configured to cool and compress the hydrogen product stream to a temperature of 20°C and a pressure of 20 bar, which results in a 99.9% pure hydrogen product gas stream.

3.1.1.6 Nuclear process heat delivery system

In addition to electrical power, the NPP can provide a source of low-cost process heat for use in the HTSE process. Although, as previously mentioned, steam from the existing LWR fleet does not have a sufficiently high temperature to provide direct heat input to the SOEC stack, LWR nuclear process heat is well-suited to provide the thermal energy required to vaporize the clean, demineralized HTSE process feedwater (separate from the NPP’s process and steam-cycle water). In the case where a gigawatt-scale HTSE plant is coupled with an NPP such that the HTSE plant consumes all the plants energy output (both thermal and electrical), 5–10% of the nuclear plant steam flow is required to provide the heat duty required for vaporization of the HTSE process feedwater.

NPP steam is diverted from a location upstream of the steam Rankine-cycle high-pressure turbine into the steam-extraction loop (SEL). A series of heat exchangers are used to condense and subcool the nuclear plant steam to transfer heat to a HTF in the TDL. The TDL is a closed-loop heat-transfer system that uses steam or synthetic heat-transfer oil to transfer nuclear process heat between the NPP and the HTSE process. The present analysis specifies the use of a synthetic heat-transfer oil (such as Therminol-66 or DowTherm) in the TDL.

Safety considerations require that the nuclear and HTSE plant sites be physically separated to minimize the risks to the nuclear plant associated with the possible detonation of the hydrogen produced by the HTSE plant. The TDL HTF transports the nuclear process heat from the nuclear plant to the HTSE plant (a distance of 1.0 kilometer is specified in the current analysis), where it is distributed between an array of heat exchangers (one per HTSE modular block) that serve as the HTSE process feedwater steam generators. The cooled TDL HTF is then returned to the nuclear plant via the TDL return piping, where fluid subsequently flows through a pump that provides the pressure differential required to recirculate the HTF through the TDL.

3.1.1.7 Multistage product compression

The purified hydrogen product exits the hydrogen-purification system at a pressure of approximately 20 bar, which is not sufficient for cost-effective hydrogen storage or transportation. Additionally, 20 bar is too low a pressure for hydrogen end-use applications, including FCEV refueling stations or input into a Haber-Bosch ammonia-synthesis process. The purified hydrogen product exiting the hydrogen-purification system is therefore sent to the multistage product compression system, where it is compressed to a pressure of 70 bar, which is a suitable pressure for injection into a distribution pipeline, or input to a hydrogen-storage system (additional compression would be required for high-pressure hydrogen-storage applications). Injection into a hydrogen-gas distribution pipeline is the primary application intended for the 70-bar high-pressure hydrogen product specified in this analysis.

3.1.1.8 Control system

The control system includes a control building and multiple operator centers for use in monitoring and controlling the HTSE process. Because the instrumentation costs for individual process unit operations are included in the Aspen Process Economic Analyzer’s (APEA’s) installed-equipment costs (and cost allowances are made for other sensors and instrumentation), the control-system capital costs are limited to those for the control building and operator centers. The HTSE control system will also be required to interface with the NPP control system. In order to avoid conflicts and increased regulations associated with the NPP control system, the HTSE control system will most likely be kept isolated from the NPP control system other than the ability of the NPP operator to shut down the HTSE at any time for any reason.

3.1.2 Equipment and Operating Condition Specifications

The HTSE process model is based on a stack operating temperature of 800°C and thermoneutral operating voltage of 1.29 V/cell. The steam inlet concentration is specified as 90 mol%, with 10 mole% hydrogen included to maintain reducing conditions. A detailed listing of HTSE-process operating-condition specifications is provided in Table 9. The stack operating pressure and steam usage are parameters that impact the energy consumption for the BoP operations; Section 3.1.2.1 describes a parametric analysis of the operating pressure and steam usage to obtain system design specifications that result in energy-efficient process operation.

BoP equipment specifications are listed in Table 10. As detailed in Table 10 the system design basis includes purification of the hydrogen product to 99.9 mol% hydrogen and compression to a pressure of 69 bar, which is a pressure suitable for injection into a transportation pipeline. The system design basis specifically does not include hydrogen-storage capacity or storage compressors because the specifications of this equipment are a result of the dispatch optimization analysis discussed in Section 4.

Table 9. HTSE and related subsystem process operating condition specifications.

Parameter	Value	Reference or Note
Stack operating temperature	800°C	O'Brien et al 2020[29]
Stack operating pressure	5 bars	See Section 3.1.2.1
Operating mode	Constant V	
Cell voltage	1.29 V/cell	Thermo-neutral stack operating point
Current density	1.0 A/cm²	
Stack inlet H_2O composition	90 mol%	O'Brien et al 2020[29]
Steam utilization	80%	See Section 3.1.2.1
HTSE modular-block capacity	25 MW-dc	1000x capacity increase[29]
Cell area per modular block	1945.4 m²	
Sweep gas	Air	O'Brien et al 2020[29]
Sweep-gas inlet flow rate	Flow set to achieve 40 mol% O_2 in anode outlet stream	
Stack service life	4 years	HFTO Hydrogen Production Record 20006[30]
Stack degradation rate	1.2%/1000 hr	Degradation rate calculated to match stack end-of-life performance of 67% projected from H_2A v3.101 future central SOEC default case. 1.2%/1000 hr degradation rate is very conservative as current indications are that suppliers may be able to achieve at least half this rate of degradation, this would lead to an increased service life and decreased maintenance and replacement cost
Stack replacement schedule	Annual stack replacements completed to restore design production capacity	Based on H_2A model stack replacement cost calculations

Table 10. BoP equipment specifications.

Parameter	Value	Reference or Note
Heat Exchangers		
Heat exchanger ΔP: TDL, feedwater heating, low temp recuperators	ΔP set using inlet pressure dependent correlation	ΔP correlation adapted from AspenTech Exchanger Design & Rating (EDR) software allowable pressure drop specification
Heat exchanger ΔP: High-temperature recuperators, intercoolers, cooling water utility exchangers	ΔP set to 2% of exchanger inlet pressure	
Heat exchanger minimum temperature approach	20°C in TDL; 15°C in HTSE process	Larger ΔT specified in TDL exchangers to provide additional flexibility for varying LWR and/or HTSE operating conditions
Cooling water utility	20°C supply T; 34°C return T	
Compression		
Compressor adiabatic efficiency	80%	
Compressor pressure ratio per stage	~1.5 max	
Product Recovery		
H_2 product recovery stage pressures (approximate)	5, 10, 20 bars	Approximately equal compression ratios between stages
H_2 product purity	99.9 mol%	
H_2 product pressure	69 bars (1000 psi)	Purified hydrogen product compressed to final pressure using high-pressure multistage compressor
Thermal-Delivery Loop		
TDL HTF	Therminol-66	O'Brien et al 2017[31]; Frick et al 2019[32]
TDL transport distance	1.0 km	Vedros et al 2020[33]
Maximum HTF velocity	3.0 m/s	Basis for pipe diameter calculations

3.1.2.1 Stack operating conditions selection

Because the majority of hydrogen production costs are generally associated with energy input, it is important for the HTSE system normal-operating mode to correspond to operating conditions that minimize process-energy costs and, also, equipment capital costs. Steam usage and stack operating pressure are two parameters that have a significant impact on the system energy consumption.

Increases in the stack operating pressure decrease the compression-energy requirements in the hydrogen purification system. Because the steam-generator pressure can be elevated through the use of liquid-phase pumps, the energy requirements for increasing the stack operating pressure are low. However, increases in the stack operating pressure will require process vessels to be rated for higher operating pressures, which increases capital costs. Additionally, increasing the stack operating pressure increases the Nernst (open cell) potential, which has the effect of increasing the stack input power requirements.

The steam usage has a direct impact on the HTSE system cooling and thermal-energy input requirements. References [31], [34], [35], [36], and [37] indicate that HTSE steam usage typically ranges anywhere from 40 to 90%. The lower the steam usage, the greater the quantity of unreacted steam exiting the stack. Because the unreacted steam must be condensed in the hydrogen-purification system and is then recycled to the steam generator, a low steam-usage value results in increased system cooling and thermal-energy input requirements. Although the use of a very high steam-usage operating specification would minimize the process-cooling and thermal-energy input requirements, there are practical upper limits on this parameter due to mass-transfer limitations associated with delivering the steam reactant to the active sites on the electrolysis cathode. Additionally, the presence of excess steam in the cells has the effect of lowering the Nernst potential, which has the effect of reducing the stack's input-power requirements.

A parametric analysis of the impact of the stack operating pressure and steam usage on process-energy requirements was completed using the HYSYS HTSE process model. In this analysis, the stack operating pressure was varied from 1 to 10 bar absolute pressure, and the steam usage was varied from 60 to 80%. Current technology steam utilization could already be as high as 80%. Not considering improvements to the technology itself but only improvements to process controls and process optimization, the steam utilization could possibly increase to nearly 90% in the near future.

The effect of the stack operating pressure and steam usage on the system's electrical-energy consumption are shown in Figure 33. Over the range of conditions evaluated, the stack operating pressure has the greatest effect on electrical-energy consumption. Increases in stack operating pressure result in decreases in the electrical-energy consumption. Higher steam usage results in lower energy consumption for all pressures evaluated. At a stack operating pressure of approximately 5 bar, the energy savings associated with increasing the stack operating pressure become less pronounced.

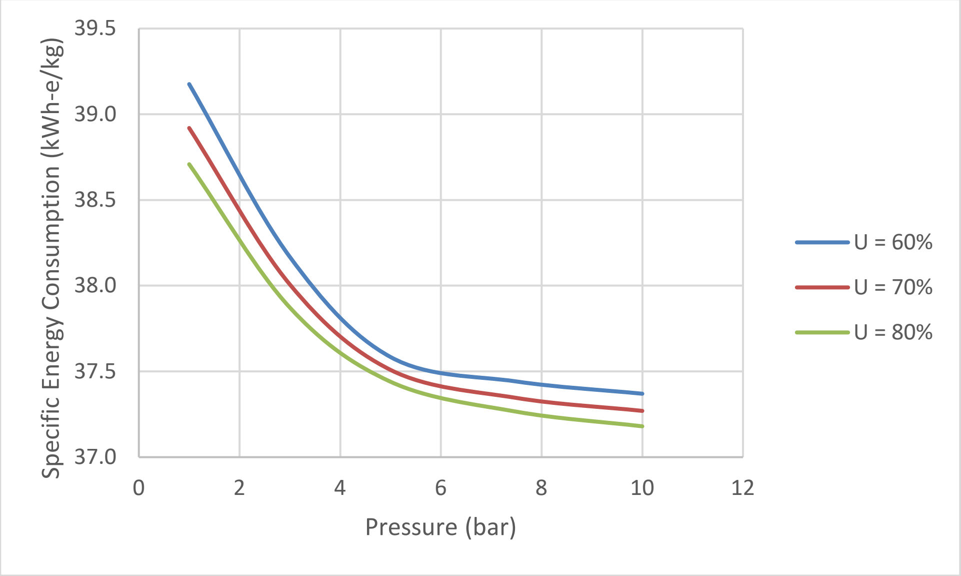

Figure 33. Electrical-energy consumption as function of stack operating pressure with steam usage as a parameter.

The system's thermal-energy consumption is most strongly affected by steam usage, as shown in Figure 34. As previously described, increased steam usage decreases the quantity of unreacted steam exiting the stack that must be condensed and revaporized. Increasing the steam usage (and thereby decreasing the steam-recycle rate) therefore has a direct impact on reducing the thermal-energy requirements of the process, as well as reducing its cooling requirements. Increases in the steam usage result in a nearly linear decrease in the thermal-energy consumption over the range of values evaluated.

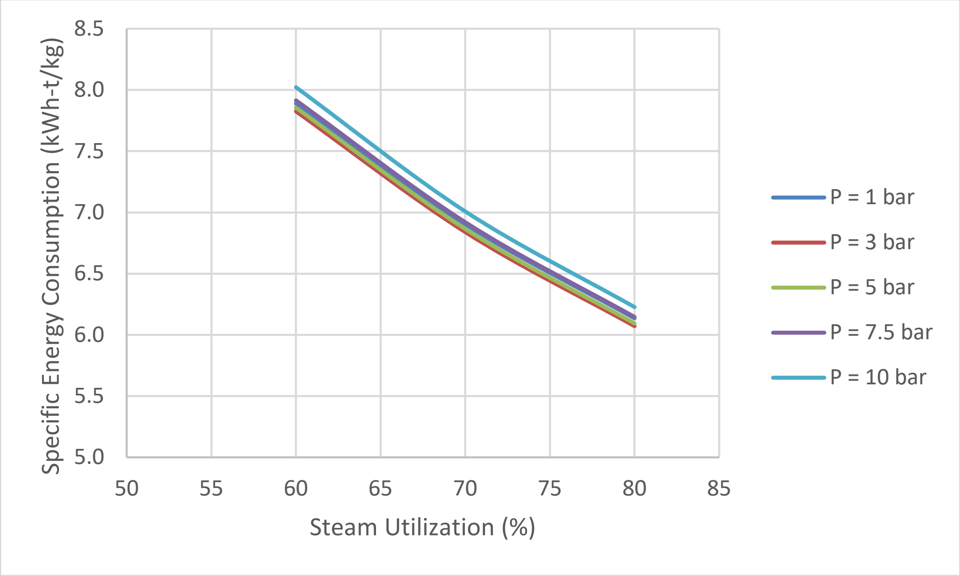

Figure 34. Thermal-energy consumption as function of steam usage with stack operating pressure as a parameter.

The HTSE system efficiency is a metric that includes both thermal- and electrical-energy consumption. The impact of the stack operating pressure and steam usage on the HTSE system efficiency is plotted in Figure 35. Increases in steam usage increase system efficiency at all conditions evaluated. Increases in system pressure result in a significant increase in system efficiency up to a pressure of approximately 5 bar, where there is a "knee" in the curve, and further increases in system pressure return a lower increase in system efficiency.

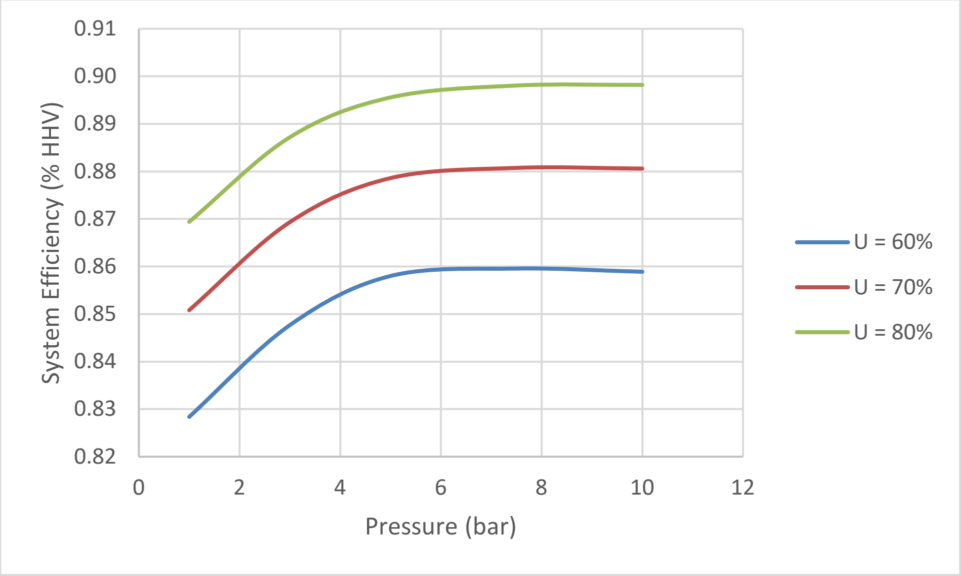

Figure 35. HTSE system efficiency as function of stack operating pressure with steam usage as a parameter.

The HTSE system, hydrogen/steam system, and air sweep-gas system will all be required to have components rated for the specified stack operating pressure. Therefore, a 5-bar operating pressure was selected as the system design point; this operating pressure will achieve near-optimal system efficiency without incurring the additional capital costs that would be incident to further increases in pressure ratings of the relevant process equipment. A steam usage of 80% was selected as the system design point, based on the significant decreases in system thermal-energy consumption associated with elevated steam usage predicted by the parametric analysis. The value of 80% steam usage is within the range of conditions that have been demonstrated and/or are suggested as practical by numerous literature sources.[31,34,35,36,37]

3.1.2.2 Normal operation

The HTSE plant's normal operating mode is used for hydrogen production. HTSE-system normal operation is characterized by the conditions specified in Table 9. During normal operations, the LWR plant dispatches a rated quantity of electrical power and process heat to the HTSE plant to support hydrogen-production operations. The HTSE most likely will be operated to produce hydrogen during periods of off-peak electrical-market demand.

3.1.2.3 Hot-standby mode

A grid-integrated HTSE plant is expected to operate in the normal operating mode for the majority of the time, with interruptions in hydrogen production generally occurring for up to several hours per day during peak electricity-demand periods. During the interruptions in hydrogen production, the HTSE plant would be operated in a hot-standby mode. The hot-standby mode would cease hydrogen-production operations in order that maximal energy output from the nuclear plant could be dispatched to the electrical-grid. Because the HTSE plant would need to be quickly brought back online at the end of the period of peak-electrical demand, the hot-standby mode is designed to maintain HTSE process conditions necessary to support a rapid resumption of hydrogen-production operations. This involves the continued circulation of process fluids to keep the process equipment operational and at temperatures, pressures, etc. Because the hot-standby operating mode continues to circulate process and HTFs, the HTSE process-energy requirements are not eliminated during hot-standby mode. Instead, both electrical power and thermal power input requirements remain, albeit at a much lower rate than during normal operations.

A dedicated HTSE plant would not be expected to use the hot-standby operating mode. Instead, a dedicated HTSE plant would maximize the time spent operating in the normal operating mode (resulting in a high capacity factor) in order to fully use the capital investment and provide a steady supply of hydrogen to customers. The hot-standby mode is, therefore, only applicable in the evaluation of grid-integrated HTSE plant-operating scenarios (e.g., RAVEN model-optimization analyses).

To estimate hot-standby mode electrical- and thermal-energy demands, the HYSYS model was modified with a set of operating conditions representative of the hot-standby energy mode. The HYSYS model hot-standby energy-requirement estimates represent steady-state operation of hot-standby mode. The process model hot-standby operating mode was configured using the following specifications:

- Stack operating temperature, pressure, inlet composition maintained at nominal values using TDL steam generator, process, and sweep-gas blowers, electric topping heaters, and recycle of stack outlet H_2/H_2O process gas and H_2 from product recovery.
- All of the design-point stack-inlet flow rate (process gas and sweep gas); stack process gas inlet composition maintained at design point specification (10 mol% hydrogen in process gas-inlet stream to maintain reducing process conditions; process-gas recycle provides a source of hydrogen in stack inlet stream).
- None of the stack power during hot-standby mode; stack outlet composition is equal to stack inlet composition during hot-standby operating mode (no steam usage); since greater than 90% of design point process electrical power requirements are used to power stack operations. This specification represents the largest single reduction in process-energy consumption associated with the hot-standby operating mode.
- The flow rate of steam entering the product-purification area is maintained at the design point value (the total H_2 + H_2O flow rate entering the product-purification area during hot-standby operations decreases because no hydrogen production occurs). Maintaining the design-point steam-inlet flow rate minimizes changes in heat load to the product-recovery area condensing equipment. Because the steam condensate collected in the product recovery area serves as the steam-generator feedwater source during hot-standby mode, the steam-generator thermal load is decreased accordingly (the net process feedwater input requirements are zero during hot-standby mode because no hydrogen is produced; the hot-standby process thermal-energy demands correspond to the heat needed to vaporize the process water that is recirculated between the stack and the product-recovery area).
- The high-pressure product compressor is shut down because no hydrogen is produced during hot-standby operation, and all hydrogen exiting the product purification area is recycled to the stack inlet.

3.1.2.4 SOEC performance degradation

Actual annual hydrogen production may vary from the design production capacity for several reasons. In a grid-integrated HTSE system, annual production will be reduced in proportion to the time that the HTSE plant is taken offline so that power generation from the nuclear plant can be dispatched to the electrical-grid. In a dedicated hydrogen-production HTSE system, any plant outages (due to maintenance, NPP refueling, etc.) will reduce the HTSE plant capacity factor such that the actual annual production rate is less than the design production rate. In addition to plant outages and/or interruptions in production activities, the HTSE plant's hydrogen-production capacity is also affected by cell-performance degradation that occurs over the service life of each SOEC stack.

The design basis specifies constant voltage mode; therefore, cell degradation results in a decrease in the electrical current that passes through the cell during normal operations. Decreased current results in decreased stack power consumption and a proportional decrease in stack hydrogen production. Therefore, cell degradation results in a decrease in the overall HTSE operating-capacity factor beyond the reductions in capacity factor associated with HTSE plant standby and outage periods.

The HFTO Hydrogen Production Record used as the data source for the base case HTSE analysis specifies a 4-year stack service life.[30] A degradation rate of 1.2%/1000 hr with an end-of-life stack performance of 67% was specified in this analysis; this value corresponds to the stack end-of-life performance projected from the degradation rate specified in the H_2A future central SOEC model version 3.101.[38]

Based on the specified degradation rate, the production capacity would be reduced to 90.6% at the end of 1 year of operation. The system design basis specifies that annual stack replacements will be performed to restore design production capacity. When the stack performance is averaged over the annual replacement schedule, the actual system production rate is calculated as 95.3% of the design production rate. Multiplication of this factor with the percentage of time within each operating year that the HTSE plant is online provides the net operating capacity factor.

3.1.3 HTSE Process Model Performance Estimates

The LWR-HTSE process-material balances, process-energy requirements, and process efficiency are summarized in Table 11. LWR-HTSE process summaries are provided for a design based on use of both Prairie Island nuclear units as well as for a design based on the single Monticello nuclear unit.

Table 11. LWR-HTSE process summary.

Parameter	Prairie Island 2 Nuclear Unit Design	Monticello 1 Nuclear Unit Design	Notes
Plant Design Capacity	1032 MW-e	597.3 MW-e	
Design Hydrogen Design Capacity	662 tonnes/day	383 tonnes/day	
Availability Factor	95%	95%	HTSE plant operating time; corresponds to availability of nuclear plant [time]/[time]
Cell Degradation Factor	95.3%	95.3%	Adjustment to production rate due to cell degradation
Operating Capacity Factor	90.5%	90.5%	Ratio of actual production rate to design production rate. Calculated as product of availability and cell degradation factors.
Actual Hydrogen-Production Rate	599 tonnes/day	347 tonnes/day	
Process Power Requirement, Normal Electrical (design condition) Thermal (design condition)	 1032 MW-e 177.5 MW-t	 597.3 MW-e 102.7 MW-t	
Process Power Requirement, Hot-Standby Electrical Thermal	 8.8 MW-e 33.7 MW-t	 5.1 MW-e 19.5 MW-t	
Specific Energy Consumption Electrical Thermal	 37.4 kWh-e/kg H_2 6.4 kWh-t/kg H_2	 37.4 kWh-e/kg H_2 6.4 kWh-t/kg H_2	
System H_2 Production Efficiency	88.9% higher heating value (HHV) basis	88.9% HHV basis	Energy content of product H_2 divided by electrical energy equivalent input
Utilities Process Water Feed Rate Cooling Water Circulation Rate	 69 kg/s [1.1 k gpm] 1334 kg/s [21 k gpm]	 40 kg/s [0.6 kgpm] 772 kg/s [12 kgpm]	

Electrical and thermal power requirements by equipment type are shown for the Prairie Island LWR-HTSE process design in Figure 36 and for the Monticello LWR-HTSE design in Figure 37.

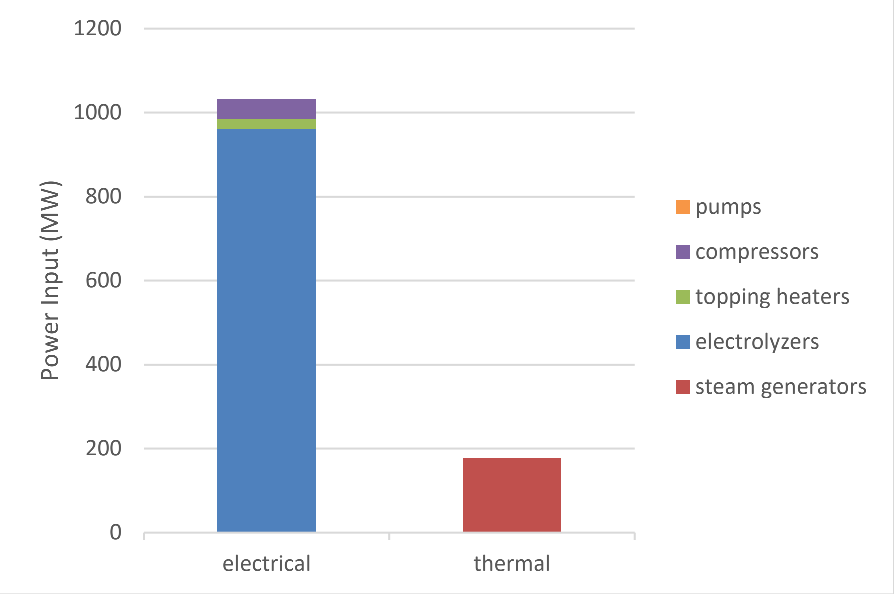

Figure 36. Prairie Island LWR-HTSE electrical- and thermal-power requirements (design point).

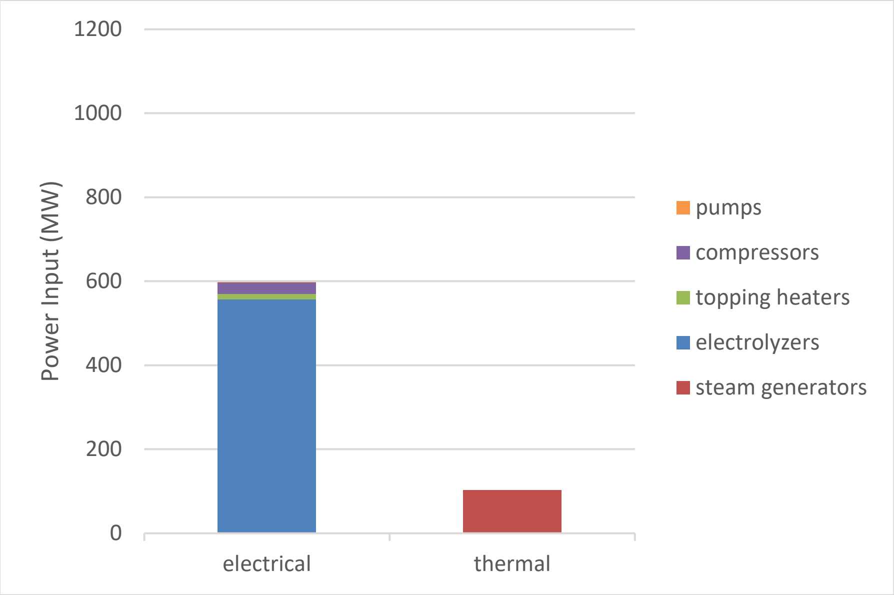

Figure 37. Monticello LWR-HTSE electrical- and thermal-power requirements (design point).

3.1.4 HTSE Process Design Considerations

3.1.4.1 LWR/HTSE integration

3.1.4.1.1 Design Basis (for Preliminary Design and Cost Estimation Purposes)

In the appendix is found a full listing of the equipment used to establish estimates of the system’s capital costs. A subset of the HTSE system design equipment that exists at the interface of the LWR/HTSE systems is listed in Table 12. Table 12 also includes equipment with functionality that may exist separately in both the LWR and the HTSE plant. Although this equipment is included in the INL HTSE system design basis, it is possible that the LWR systems identified (water purification, process cooling, and process control) may be modified for use with the HTSE installation such that purchase and installation of separate HTSE-specific equipment items is not required.

The HTSE design basis described herein is subject to change based on NPP facility selection, TDL HTF, and SOEC technology selected for prospective final system design. Considerations which may impact the final system design, including the reinjection point for the SEL condensate, the number of LWR units from which nuclear process heat is extracted, and the TDL HTF selection are discussed in additional detail in Section 3.1.4.2.

Table 12. LWR/HTSE system interface equipment (the list includes NPP water-purification and cooling-system equipment that could potentially be leveraged for HTSE system operations).

Equipment	System
Backup Electric Boiler	Nuclear Process Heat Delivery System
PIPE-201 Nuclear Process Heat Piping (supply)	Nuclear Process Heat Delivery System
PIPE-202 Nuclear Process Heat Piping (return)	Nuclear Process Heat Delivery System
P-201 Nuclear Process Heat Circulation Pump	Nuclear Process Heat Delivery System
HX-201 Nuclear Process Heat TDL HX	Nuclear Process Heat Delivery System
HX-202 Nuclear Process Heat TDL HX	Nuclear Process Heat Delivery System
Therminol-66 HTF	Nuclear Process Heat Delivery System
Rectifier/Power Supply	Electrical Power Transport & Distribution System
Disconnect Switch	Electrical Power Transport & Distribution System
Transformer	Electrical Power Transport & Distribution System
Switch Board	Electrical Power Transport & Distribution System
DC Bus Power Distribution	Electrical Power Transport & Distribution System
Power Pole Lines	Electrical Power Transport & Distribution System
Purified Water Storage Tank	Feedwater Purification & Storage System
PIPE-801 Feed Water Supply Piping	Feedwater Purification & Storage System
P-801 Feed Water Supply Pump	Feedwater Purification & Storage System
Water Pretreatment Filter/Softener System	Feedwater Purification & Storage System
Water Treatment RO/EDI System	Feedwater Purification & Storage System
PIPE-901 Cooling Water Supply Piping	Process Cooling System
PIPE-902 Cooling Water Return Piping	Process Cooling System
P-901 Cooling Water Recirculation Pump	Process Cooling System
CT-901 Cooling Tower	Process Cooling System
CB-101 Control Building	Control System
OC-101 Operator Center	Control System

3.1.4.1.1.1 Nuclear Process Heat

The TDL and associated heat exchangers are included in the INL HTSE system design basis. The TDL heat exchangers transfer heat from the NPP steam to the TDL HTF. The SEL piping on the NPP side of the TDL heat exchangers (and the costs of installing this system or modifying existing systems to establish this functionality) is NOT included in the INL HTSE system design basis. Figure 38 provides a simplified diagram of a nuclear plant power block with an SEL (Streams 711, 712, and 713), the TDL heat exchangers (HX-201 and HX-202), the TDL (200 number category streams) and the interface with the HTSE-process feedwater-heating system (Streams 151 and 152, HX-102).

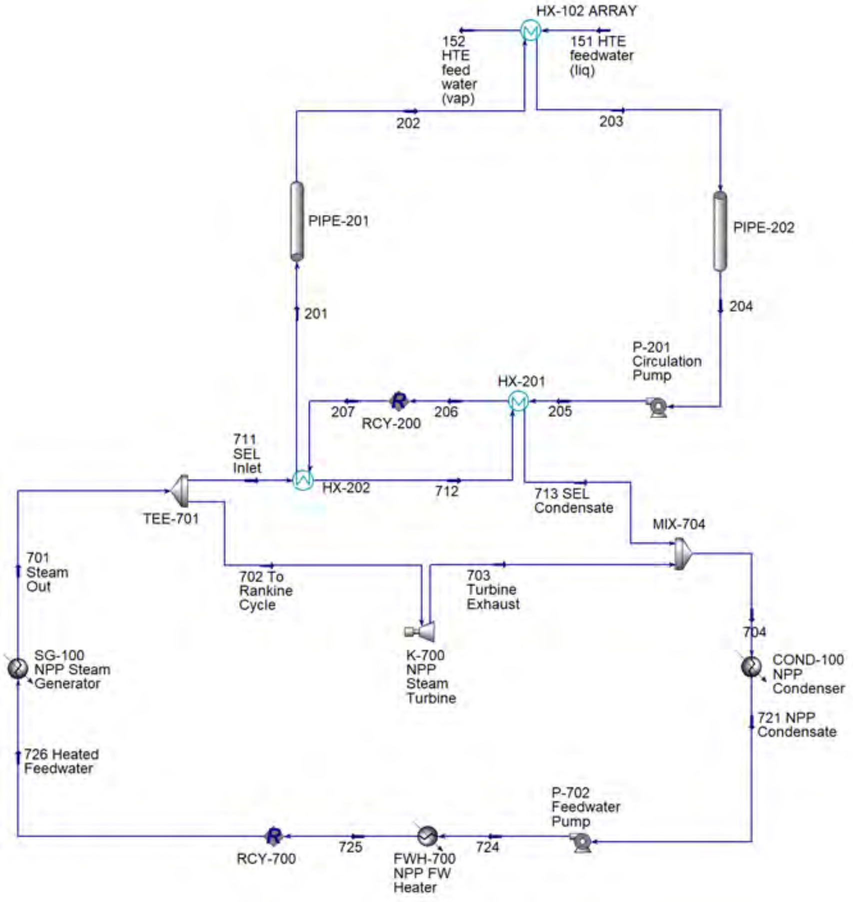

Figure 38. TDL integration with NPP (simplified NPP model).

INL analysis indicates that approximately 5% of the total steam flow rate produced by the NPP steam generators would be required to meet the HTSE thermal demands (187 MW-t of the 2 × 1684 MW-t thermal output from the two reactor units comprising the Xcel Energy Prairie Island Nuclear Power Plant[39] would be required for a full-scale 1086 MW-ac HTSE plant installation). A full-scale 1086 MW-ac HTSE plant installation would use the maximal energy output available from both units of the PI NPP. The INL HTSE design basis specifies that a single set of heat exchangers would interface with one unit of the NPP and that an ancillary boiler would be used to provide the HTSE process heat during NPP plant outages (e.g., refueling).

3.1.4.1.1.2 Electrical Power

The INL HTSE system design basis includes electrical transmission lines necessary to deliver power from the NPP to the HTSE site and transformers for stepping down the AC power from 20 kV to the rectifier supply voltage (assumed ~4 kV for equipment costing purposes). Note that 20 kV is the voltage of the PI generator output, and in the current PI plant configuration this power is stepped up to 345/161 kV for transmission.[40] If equipment for distribution of 20 kV power is not present, then this equipment will need to be retrofitted to the PI substation, or the INL system design basis will require modification to include step-down transformers with the proper operating specification. Future work would consider pulling power from the transmission grid at 345/161 kV to keep the hydrogen plant and the NPP generator decoupled and avoid having NPP perturbations affect the hydrogen plant. The INL

HTSE system design basis also includes power inverters (rectifiers) for converting the AC power to DC power and DC bus bars for distributing the power from the inverters to each of the HTSE modules.

3.1.4.1.1.3 Control Center

INL's HTSE-system design basis includes the costs of a control building with operator stations for monitoring and control of the HTSE process systems. This system may be redundant in the event that the NPP control system is ultimately used to provide seamless control between NPP and HTSE-system operations. In either case, the additional capability for control of the HTSE system must be considered, and the costs listed in INL's HTSE-system design basis provide an initial estimate for the purposes of the current analysis.

3.1.4.1.1.4 Water Purification

INL's HTSE-system design basis includes feedwater pretreatment and purified feedwater storage capacity. If the Xcel Prairie Island or Monticello plants include water-pretreatment equipment and storage capacity sufficient to supply the HTSE plant, these equipment items can be removed from INL's HTSE system cost estimate. However, if the NPP water-treatment system were used to supply purified feedwater to the HTSE plant, an additional pipeline would be required to transport the purified feedwater from the NPP site to the HTSE process site. The cost of such a pipeline is not currently included in INL's HTSE-system design basis.

3.1.4.1.1.5 Process Cooling

INL HTSE process-modeling analysis indicates that process-cooling capacity is required to provide a heat sink for the hydrogen-purification subprocess (which removes steam from the H_2/H_2O mixture exiting the stacks by cooling and compressing the product-gas mixture). INL's HTSE-system design basis includes a cooling-tower installation to provide this capacity. Alternatively, cooling water from the Xcel Prairie Island or Monticello plant cooling systems could be used to provide the required HTSE process cooling duty. If the existing cooling systems were to be used, additional cooling-water supply and return lines would be required to transport the cooling water between the cooling-water source (whether based on use of river water or cooling towers) and the HTSE site.

3.1.4.1.2 Cost Items Excluded from HTSE System Design Basis

The following is a list of cost items that are specifically NOT included in the INL HTSE-system design basis. These items are excluded from the present analysis due to insufficient information and deferral to the expertise of the nuclear plant operators and/or future studies that perform detailed evaluations of the NPP system-modification requirements and costs.

- Nuclear plant modification (pipes/valves to divert steam to TDL heat exchanger)
- NPP instrumentation and control system modifications to enable nuclear plant to vary distribution of steam between the power cycle and nuclear process-heat applications (e.g., HTSE)
- Leak monitoring and detection equipment (i.e., equipment and systems for detection of radioactive components that could have escaped from the NPP primary or secondary steam loops)
- Substation modifications to divert electrical power to the HTSE process instead of, or in addition to, the electrical-grid
- Regulatory costs—i.e., cost of obtaining any additional permits necessary to operate the NPP in variable electricity/hydrogen-dispatch mode
- Expenses and lost revenues due to any NPP shutdown, de-rating, or interruption of service or operations required to implement process modifications.

3.1.4.2 Thermal delivery loop design parameters requiring further investigation

3.1.4.2.1 Options for Steam-Extraction Loop Condensate Return

A detailed diagram of the PI NPP cycle/TDL integration is shown in Figure 40, which illustrates an SEL configuration in which several possible SEL-condensate return points are visible. SEL condensate could be returned upstream of the condenser (MIX-172), to a location in the low-pressure boiler-feedwater heating train (e.g. MIX-184), or to a location in the high-pressure boiler-feedwater heating train (e.g., MIX-188). The INL system design basis specifies return of the SEL condensate to the point upstream of the condenser because NPP condensers are built with excess design capacity, suitable for handling excess steam input associated with plant startup and shutdown, plant trips, etc., and are designed to be able to robustly absorb heat release associated with transient plant operations. Although the nuclear plant's operating efficiency could be incrementally improved by returning the SEL condensate to a point in the feedwater heating train with similar temperature and pressure (which would avoid cooling the SEL condensate in the condenser only to reheat it in the feedwater heating train), this configuration would increase the system's operating complexity as well as retrofit costs; therefore, it was not considered in the current analysis.

3.1.4.2.2 Nuclear Process Heat Extraction from Multiple Reactor Units

This analysis specifies TDL heat-exchanger equipment with heat-transfer capacity to remove the quantity of nuclear process heat required by an HTSE plant during normal operations. Because several heat exchangers operating in parallel comprise the nuclear plant side of the TDL, these heat exchangers could be distributed between both units of the PI NPP to use approximately 5% of the steam flow from each unit, thereby providing the HTSE plant with the required nuclear process heat. The approximate value of 5% of the steam flow from each nuclear unit assumes that the HTSE plant is sized to use near maximal thermal- and electrical-power output from the nuclear plant during normal operations.

Alternatively, all heat exchangers on the nuclear plant side of the TDL could be assigned to a single unit of the PI NPP, which would result in approximately 10% of the steam from the nuclear unit being used to supply HTSE thermal demands (with the second nuclear unit being used to supply only electrical power to the HTSE plant). Although this configuration would reduce the nuclear plant's retrofit costs for HTSE installation (by only modifying one nuclear unit, instead of two), there are several operational disadvantages associated with a configuration that extracts heat from only one nuclear unit. First, because nuclear process heat would only be provided by a single nuclear unit, a greater fraction of the unit's total steam generation would have to be diverted to the TDL to meet the HTSE-process's thermal energy demands. Diverting a larger fraction of the steam output from a single nuclear unit away from this unit's steam turbines is expected to have a larger impact on steam-cycle efficiency than would diverting a smaller fraction of steam output from two separate nuclear units. Second, if all nuclear process heat were sourced from a single unit, the HTSE plant would be left without a nuclear process heat source during that unit's refueling shutdown period, which occurs approximately every 24 months and may last for several weeks. If the nuclear plant side of the TDL were configured such that the nuclear process heat could be extracted from either unit of a two-unit nuclear plant, the staggered nature of the refueling operations would allow the nuclear plant to continue to provide nuclear process heat to the HTSE plant during refueling (although electricity required to operate the HTSE plant at full capacity would have to be sourced from the grid).

This analysis specifies capital costs for TDL heat-exchange equipment necessary to meet the HTSE thermal demands at the design point (i.e., the capital costs include heat exchangers with only the surface area required to meet HTSE thermal demands during NPP normal operations). The analysis assumes that the heat exchangers on the nuclear plant side of the TDL would be distributed between all units of the nuclear plant (two units in the case of Prairie Island or one unit in the case of Monticello). Because no excess heat-exchanger area is specified in this analysis (i.e., the nuclear process-heat load could not be fully shifted to Unit 2 of Prairie Island during refueling of Unit 1), the HTSE process design specifications presented in this report include ancillary electric steam-generator equipment for use during nuclear plant refueling. For Prairie Island, the ancillary electric steam generator could replace nuclear process-heat output of one unit during refueling of the second. For Monticello, the ancillary electric steam generator could provide all thermal-energy input needed for HTSE-process feedwater vaporization during refueling. The capital costs for the ancillary steam-generator capacity are included in the cost estimates. As mentioned above, it is anticipated that the electrical-grid would provide the power to operate the ancillary steam generators during nuclear plant refueling.

3.1.4.2.3 HTF Selection and Implications on System Design, Cost, and Operations

Therminol-66 was selected as the HTF for non-proprietary system design. Use of Therminol-66 or another synthetic heat-transfer oil (such as DowTherm) decreases the operational complexity of the system because the TDL heat exchangers will not experience phase change on both the hot and cold sides of the TDL heat-exchanger network.

Although use of steam as the TDL HTF would present process-control challenges due to phase change on both the hot and cold sides of the TDL heat exchanger network, a water and steam-based design could decrease TDL capital costs (Figure 34):

- Heat-transfer coefficients associated with steam vaporization and condensation are generally higher than those for sensible heat transfer associated with a heat-transfer oil, resulting in reduced heat exchanger area (and cost) for the water and steam system
- HTF costs are significantly lower for a water and steam system than for a synthetic heat-transfer oil-based system
- The enthalpy flow associated with water and steam vaporization and condensation is significantly higher than that for the synthetic-oil sensible heat transfer; therefore, the mass-flow rate required to transport a specified quantity of nuclear process heat can be significantly lower for water and steam than for a synthetic oil. The lower HTF mass-flow rate for a water and steam design results in a TDL system with smaller-diameter, less-expensive piping.

A water and steam TDL design is compatible with methods used for detection of radioactive contaminants that may have escaped from the NPP primary or secondary loops. Equivalent protocols for detection of radioactive components in synthetic heat-transfer oils would have to be determined in engineering design of an actual system.

INL is currently investigating heat-exchanger network configurations and control strategies that could be implemented to allow use of a water and steam based TDL design. It is anticipated that successful development and testing of a robust water and steam TDL-system design would result in the HTSE system’s design being adapted to use water and steam as the TDL working fluid.

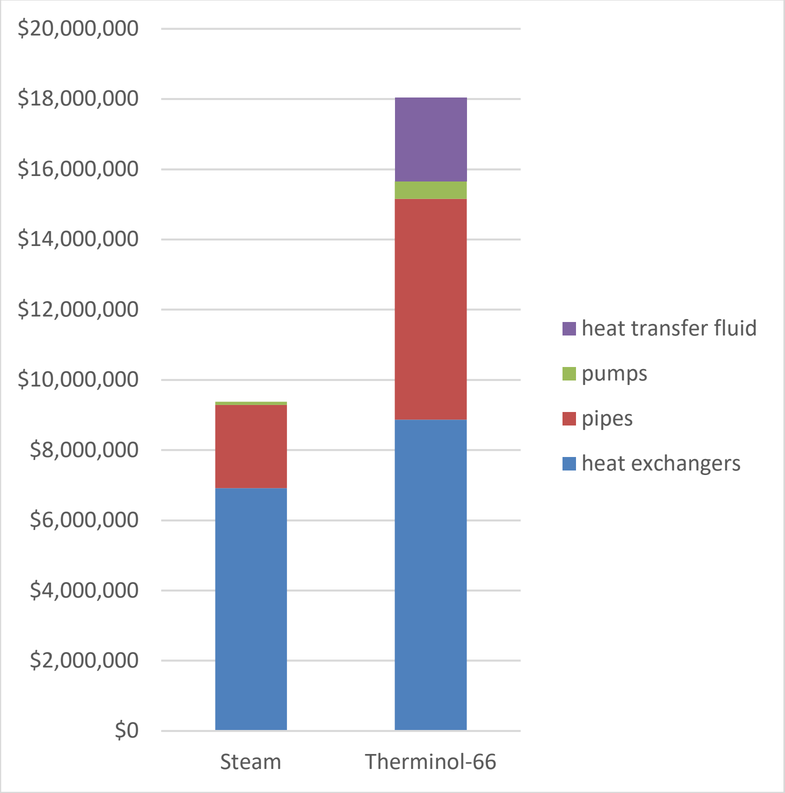

Figure 39. Steam versus synthetic heat transfer oil TDL capital costs for a gigawatt-scale LWR/HTSE plant installation.

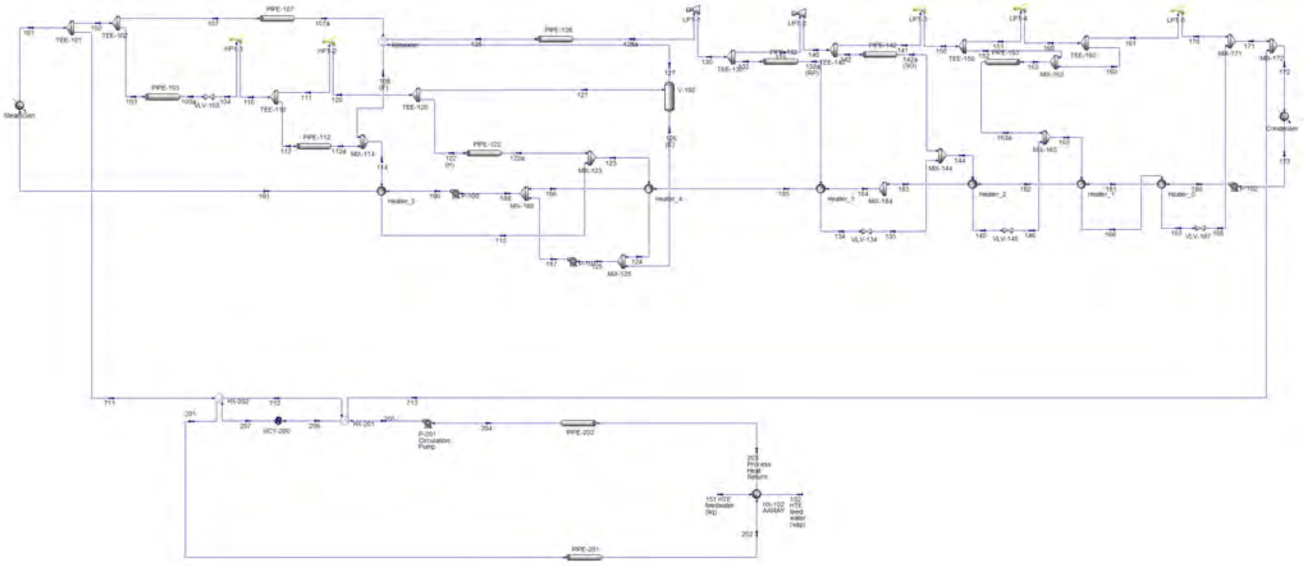

Figure 40. TD integration with NPP (detailed NPP model).

3.2 HTSE Process Capital and Operating Costs

3.2.1 HTSE Process Capital Costs

3.2.1.1 Capital cost estimation methodology

3.2.1.1.1 Modular Equipment

The analysis assumes that the HTSE plant is constructed using a modular concept. This concept involves use of multiple HTSE modular units operating in parallel to achieve the specified hydrogen-production capacity. The basis for this analysis specifies each modular unit has an electrolysis capacity of 25 MW-dc. The modular units include the equipment that comprises the HTSE: the air sweep-gas, hydrogen/steam, and hydrogen-purification systems. The HTSE modules, therefore, include the stacks and many BoP system components such as feedwater pumps, feedwater preheating equipment, steam generators, recuperators, topping heaters, product-purification equipment (compressors, gas coolers, knock-out drums), sweep-gas system.

Modular construction is a logical approach for the HTSE process, which inherently involves many individual SOEC stacks operating in parallel. The modular-construction technique provides a way to group each of the SOEC stacks into functional units that can be deployed and serviced practically. It is envisioned that the HTSE modules produced by a given SOEC manufacturer would adhere to a standardized design, and the modules would be mass-produced at an industrial manufacturing facility. The standard design would reduce indirect costs (i.e., engineering and process design) associated with deploying HTSE plants, and mass production of the modules would lead to cost savings through economies of mass production (i.e., use of standardized components purchased in bulk quantities and assembly-line fabrication and construction of modular systems to reduce equipment-installation costs).

The system components included in each of the HTSE modular blocks are identified as modular equipment, and a specific methodology is applied for estimating the modular system-component costs as a function of plant capacity. The BoP equipment components included in each modular HTSE block introduce additional thermal and/or electrical power demands such that the total power requirements for each HTSE block exceed 25 MW-e total power input (e.g., the total power requirements for an HTSE plant configuration with 40 modular units will exceed 1 GW-e of power input). The modular equipment components represent the majority of the plant infrastructure for the design cases considered in the INL HTSE system design analysis.

Cost reductions associated with a large-scale modular HTSE plant are estimated through use a of a learning-curve relationship to account for economies of mass production. A learning rate of 95% (which corresponds to a 5% cost reduction with every doubling of the number of units produced) was specified in the development of the cost-versus-capacity correlation developed in this analysis. The learning curve was applied to the installed costs of the modular process-equipment components. The learning-curve cost reductions are applied on a module-by-module basis, meaning that it is assumed that the economy of mass production cost savings is compounded as a greater number of complete modules have been constructed by the manufacturer.

3.2.1.1.1.1 First- and nth-of-a-kind plant construction

A gigawatt-scale HTSE plant has not yet been constructed; therefore, capital-cost reductions are expected from the first-of-a-kind (FOAK) plant installation to a NOAK plant installation. The cost versus capacity curve for the HTSE modular blocks (the modular components) was determined based on a learning-curve relationship.

For a FOAK HTSE plant, the modules deployed would be among the first manufactured, and it is assumed that cost reductions would be realized immediately (impacting the cost of the second, third, etc. modules deployed in a single large-scale HTSE process installation). For an NOAK plant, many HTSE modules will have been previously manufactured and deployed, and the most significant learning-curve-related cost savings will have been realized. Therefore, for the NOAK plant, the learning curve has “flattened out” such that there are minimal cost savings between the successively installed modules that comprise the overall HTSE plant.

Different modular system component costs apply for FOAK vs NOAK plants. To estimate the modular equipment costs for each of these cases, the following methodology was used: First, equipment-sizing parameters were determined based on the results of the AspenTech HYSYS HTSE process simulation. Next, installed equipment costs were estimated using APEA software[41] and/or scaled based on data reported in previous HTSE process evaluations.[30][42][43][44][45] The costs of the HTSE modular block components were evaluated at a capacity of 25 MW. Finally, a learning curve was applied to determine how the installed capital costs could decrease as a function of the number of modular HTSE units manufactured for the FOAK and NOAK scenarios. For both scenarios, a learning rate of 95% was specified.

3.2.1.1.1.2 First-of-a-Kind Plant Construction

For the FOAK scenario, the modular system component costs are the cumulative sum of all 25 MWe HTSE modular blocks installed to achieve the specified plant capacity. The cost of each HTSE modular block is lower than the previous due to the learning effects, so the total cost is equal to the sum of all blocks installed. As an example, a FOAK plant with 8 × 25 MWe HTSE modular blocks would pay the cumulative cost for all eight HTSE blocks, where the eighth HTSE block is characterized by three doublings in the number of units produced ($2^3 = 8$), such that the unit cost of the eighth HTSE block is $8^{\mathrm{log2}}(0.95) = 0.95^3$ = 85.7% of the first unit. This cost relationship is applied to each of the HTSE blocks that comprise the FOAK plant such that, in the eight modular-block example case, the cumulative cost of all eight units is 7.26 times the cost of the first unit. The FOAK equipment cost vs plant capacity curve is provided in Section 3.2.2.

3.2.1.1.1.3 Nth-of-a-Kind Plant Construction

For a NOAK plant, the most significant learning effects have been realized in the production of the previous modules, such that each additional module manufactured has essentially the same cost for a given large-scale HTSE process installation; i.e., each modular HTSE block has an equal cost due to the low slope of the learning curve at large N. For this analysis, the NOAK plant is assumed to correspond to N = 100 previous HTSE block installations (i.e., 2.5 GWe of HTSE plant capacity previously installed). All HTSE blocks installed for the NOAK plant therefore have the same cost; i.e., the modular equipment unit cost is independent of plant scale. In this analysis the SOEC stack costs are assumed to remain constant at the specified value; the learning-curve cost reductions are applied to all other balance-of-module and/or BoP equipment components identified as “modular.” The NOAK equipment cost vs plant capacity curve is provided in Section 3.2.2.

3.2.1.1.2 Scalable Equipment

The feed and utility, NPH delivery, multistage product compression, and control systems are constructed of equipment classified as “scalable” equipment components.

Scalable equipment design and costs will be dependent on the overall scale of the HTSE process installation. As noted above, the TDL used to transport thermal energy from the NPP to the array of HTSE modules is a scalable plant component. The size and capacity of the TDL heat exchangers, pipes, and pump used to circulate the fluid will depend on the capacity of the HTSE plant. In contrast to the HTSE modules, it is envisioned that one TDL, instead of multiple parallel units, will be used to transport the thermal energy from the NPP to the HTSE plant. The capital costs of the TDL equipment will therefore scale in the conventional sense: equipment with increased capacity is more cost-effective on a unit-cost basis.

To determine the dependence of the scalable equipment component costs on the HTSE plant capacity, several steps were performed. First, HYSYS process-modeling software was used to establish multiple sets of HTSE plant-design specifications over a range of plant capacities (25 to 1150 MW). This activity provided equipment-sizing parameters (heat-exchanger area, pipe diameter, pump-driver power, etc.) for each of the scalable equipment components as a function of plant capacity. Next, APEA software was used to evaluate scalable equipment installed costs for each of the plant capacities evaluated (which ranged, as mentioned, from 25 to 1150 MW). Item-specific scaling exponents for each of the scalable-equipment components were then determined from the capacity vs installed capital-cost analysis (based on the APEA estimates of equipment cost as a function of capacity), or specified per the corresponding data source (for components with costs obtained from sources other than APEA). Finally, the individual scalable-equipment component costs were summed to establish a total scalable equipment cost versus capacity data set, and this data set was then used to derive a power-law correlation to predict total scalable equipment costs as a function of plant capacity.

3.2.1.1.2.1 Purified Hydrogen Product Compressor Capital Costs

The HTSE plant design includes compressor costs associated with pressurizing the purified hydrogen product from ~20 to ~70 bar. Storage and transportation compressor costs are derived from ANL's HDSAM model. The compressor-cost equation included below for low-pressure storage or transportation compression is used to determine the cost of the purified product compressors.

$$CAPEX_{comp} = \$40{,}500 \times \left(P_{comp}\right)^{0.46}$$

In this equation, P_{comp} is the compressor-power input in units of kW and the predicted CAPEX is the uninstalled equipment cost. An installation factor of 1.3 is applied to the uninstalled compressor cost to obtain the total installed equipment cost.

Report INL/EXT-20-57885,[46] which also references the HDSAM compressor-cost correlations, indicates that low-pressure storage is characterized by pressures ranging from 150 to 500 bar. The process-model-specified outlet pressure of ~70 bar is well below this pressure range, and the low-pressure storage and transportation compressor-cost correlation is therefore applicable.

3.2.1.1.2.2 Sweep-Gas System and Hydrogen/Steam System Compressor Capital Costs

The HTSE process model also includes compressors for compression of the H_2/H_2O mixture and sweep gas to the stack. Because these compressors are not specific to storage or transportation applications the APEA compressor-cost estimates were used for the compressors internal to the HTSE-process design. The APEA cost estimates used for the H_2/H_2O compression are somewhat more conservative (i.e., represent higher estimated cost) than the HDSAM cost estimates.

3.2.1.1.3 Indirect Costs

An indirect cost multiplier of 1.294 is applied to the installed capital costs predicted by the equation (see Table 13). The indirect costs include site preparation, engineering and design, project contingency, contractor's and legal fees, and land. The engineering and design and process-continency values assumed were reduced from the default H_2A values on the basis that reductions to these costs would be realized as a result of the use of modular process construction technology (use of a standardized design would decrease engineering and design costs as well as the risks associated with the deployment of a standardized design).

Table 13. Indirect-cost multipliers.

Indirect Cost Category	HFTO Hydrogen-Production Record	INL HTSE Process Analysis
Site Preparation	2%	2%
Engineering and Design	10%	2.3%*
Process Contingency	15% total	1.6%*
Project Contingency		7.2%
Contractor’s Fee	15% total	10%
Legal Fee		5%
Land	<1%	1%
Cumulative Multiplier	1.421	1.294

* NOAK plant specifications were obtained by applying an 80% learning curve to value in HFTO Record.

3.2.1.1.4 Total Capital Investment

A total capital-investment cost versus plant-capacity correlation was derived by evaluating seven data points within the specified range of HTSE plant capacities. Each data point includes the sum of all modular installed-equipment costs, scalable installed-equipment costs, and indirect costs. A correlation for the total capital investment was derived by fitting the resulting cost versus capacity data set using a power law relation.

3.2.1.2 Prairie Island and Monticello LWR-HTSE Estimated Process Capital Costs

As described above in Section 3.1 the HTSE system evaluated includes several major process systems. Individual equipment components included in each of these systems are identified in the equipment table included in the Appendix.

Capital costs reported correspond to the maximum HTSE plant capacity that could be supported by the two Prairie Island and one Monticello NPP units. For PI, this corresponds to 38 × 25 MW HTSE units or 950 MW-dc of electrolysis capacity, with total plant power consumption of 1032 MW-ac (accounting for the power consumption associated with the BoP equipment) and a design point hydrogen-production rate of 662 tonnes/day. For Monticello this corresponds to 22 × 25 MW-dc HTSE units (550 MW-dc of electrolysis capacity) with 597.3 MW-ac total power consumption and a design point hydrogen-production rate of 383 tonnes/day.

Capital cost summary tables for the Prairie Island and Monticello LWR-HTSE process designs are included in Table 14 and Table 15, respectively. The LWR-HTSE capital cost estimates for FOAK and NOAK plant types are presented graphically in Figure 41 for Prairie Island and Figure 42 for Monticello. Capital costs for each of the equipment components within the LWR/HTSE plant boundary limits are obtained from sources.[30,41,42,43,44,45] All capital costs were indexed to 2020 dollars using the Chemical Engineering Plant Cost Index (CEPCI).

Table 14. Capital cost summary for Prairie Island HTSE plant (NOAK plant type; max HTSE process energy consumption from both nuclear plant units).

		2020 Dollar Basis	% of total	% of total
Direct capital costs	HTSE system[α]	\$275,993,520	47.4%	36.7%
	Balancing gas system	\$0	0.0%	0.0%
	Feed and utility system	\$143,204,425	24.6%	19.0%
	Sweep gas system	\$49,801,557	8.6%	6.6%
	Hydrogen/steam system	\$16,099,161	2.8%	2.1%
	Hydrogen purification	\$69,107,918	11.9%	9.2%
	Nuclear steam delivery	\$21,733,877	3.7%	2.9%
	H_2 compression and storage	\$5,336,379	0.9%	0.7%
	Control center	\$828,104	0.1%	0.1%
	Total	\$582,104,942	100.0%	77.3%
Indirect depreciable capital costs	Site preparation	\$11,642,099		1.5%
	Engineering and design	\$13,217,372		1.8%
	Contingencies and contractor's fee	\$109,453,559		14.5%
	Legal fee	\$29,105,247		3.9%
	Total	\$163,418,277		21.7%
Total depreciable capital costs		\$745,523,219		99.0%
Non-depreciable capital costs	Land	\$7,455,232		1.0%
Total capital investment		\$752,978,451		100%
Total capital investment (\$/kW)		730		

[α] Based on HTSE stack capital cost specification of \$155/kW-dc.[30]

Table 15. Capital cost summary for Monticello HTSE plant (NOAK plant type; max HTSE process-energy consumption from single nuclear plant unit).

		2020 Dollar Basis	% Of Total	% Of Total
Direct capital costs	HTSE systemα	$159,785,722	46.6%	36.0%
	Balancing gas system	$0	0.0%	0.0%
	Feed and utility system	$85,246,759	24.9%	19.2%
	Sweep gas system	$28,832,481	8.4%	6.5%
	Hydrogen/steam system	$9,320,567	2.7%	2.1%
	Hydrogen purification	$40,009,847	11.7%	9.0%
	Nuclear steam delivery	$14,850,478	4.3%	3.3%
	H2 compression and storage	$3,839,465	1.1%	0.9%
	Control center	$828,104	0.2%	0.2%
	Total	$342,713,423	100.0%	77.3%
Indirect depreciable capital costs	Site preparation	$6,854,268		1.5%
	Engineering and design	$7,781,708		1.8%
	Contingencies and contractor's fee	$64,440,621		14.5%
	Legal fee	$17,135,671		3.9%
	Total	$96,212,268		21.7%
Total depreciable capital costs		$438,925,691		99.0%
Non-depreciable capital costs	Land	$4,389,257		1.0%
Total capital investment		$443,314,948		100%
Total capital investment ($/kW)		742		

[α] Based on HTSE stack capital cost specification of $155/kW-dc.[30]

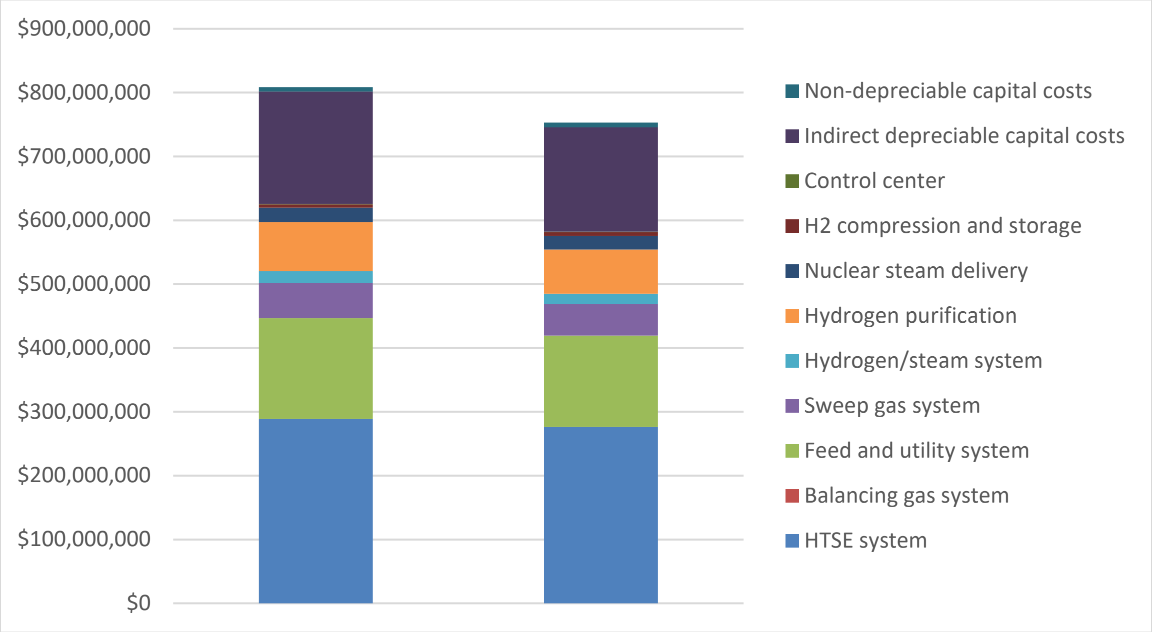

Figure 41. FOAK- and NOAK-plant capital cost estimates for Prairie Island LWR-HTSE plant (HTSE stack capital cost specification of $155/kW-dc).[30]

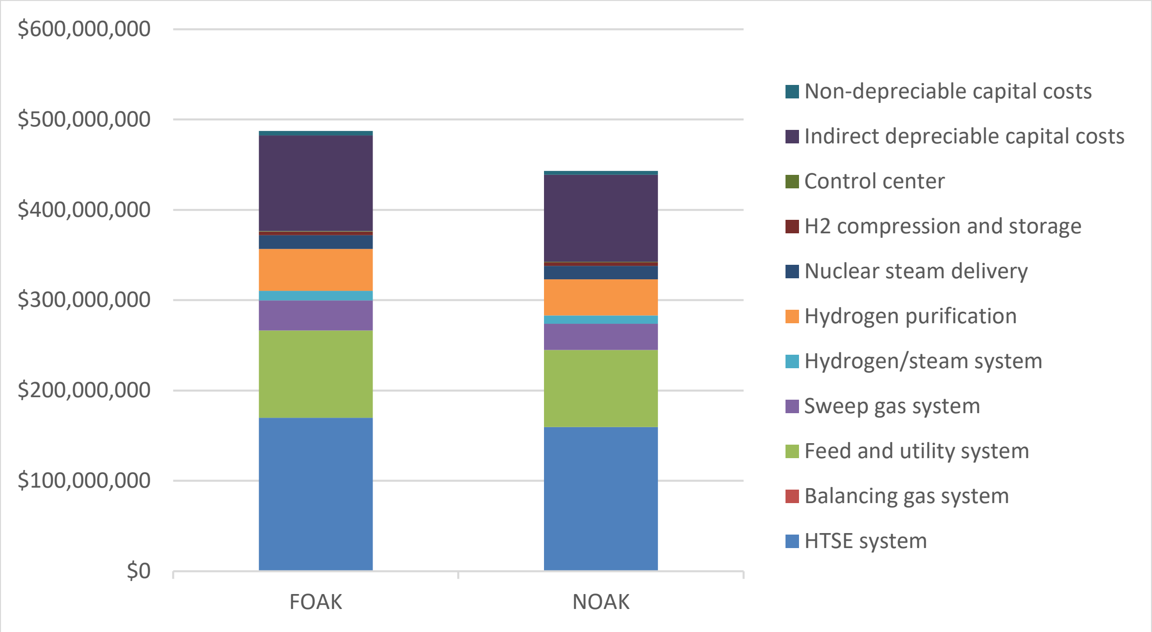

Figure 42. FOAK- and NOAK-plant cost estimates for Monticello LWR-HTSE plant (HTSE stack capital cost specification of $155/kW-dc).[30]

3.2.1.3 Generalized HTSE Process Capital Cost Correlation

A generalized HTSE cost correlation was developed to estimate plant capital costs as a function of plant capacity. This capital cost correlation is a key input to the grid-integrated LWR-HTSE plant-optimization analyses described elsewhere in this report. Figure 43 and Figure 44, are graphical representations of the unit capital costs for FOAK and NOAK LWR-HTSE plants, respectively. The capital cost curves include contributions from modular equipment, scalable equipment, and indirect costs (see Section 3.2.1.1 above for more information on these equipment categorizations). The capital cost correlation estimates the capital costs of the HTSE process areas described in Section 3.1.1. Note that retrofit costs required for the LWR to interface with the TDL system are not included in the cost estimates (see Section 3.1.4.1.2 for additional information).

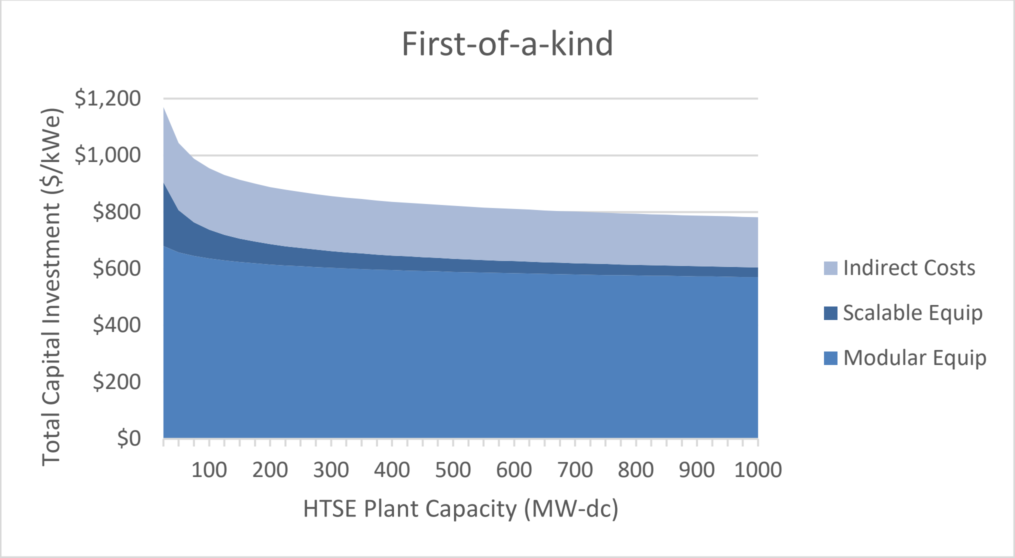

Figure 43. Total capital investment as a function of plant capacity for a FOAK HTSE plant (HTSE stack capital-cost specification of $155/kW-dc).[30]

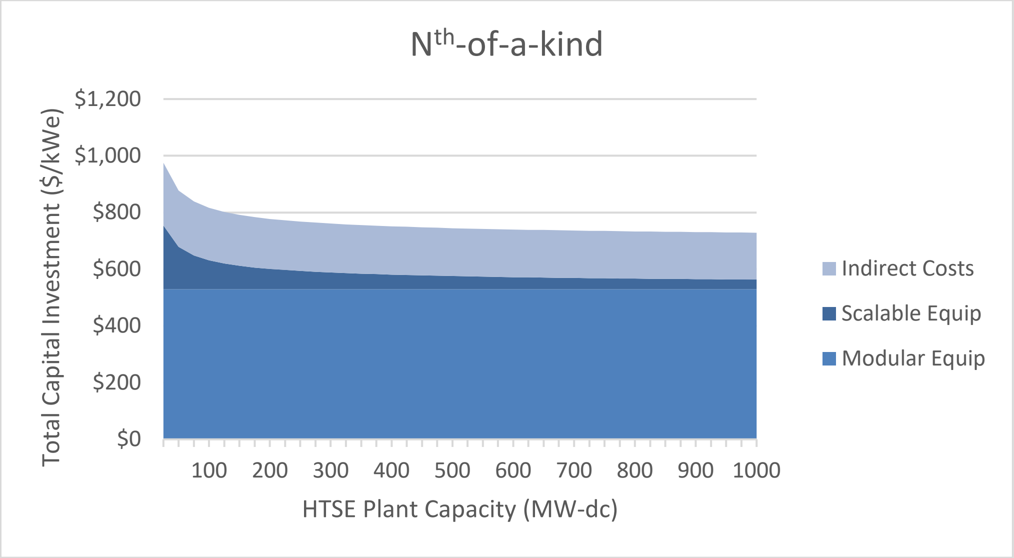

Figure 44. Total capital investment as a function of plant capacity for an NOAK HTSE plant (HTSE stack capital-cost specification of $155/kW-dc).[30]

Data from the capital-cost evaluation of FOAK and NOAK plant types over a range of plant capacities was regressed to develop an equation for use in estimating HTSE plant total capital investment as a function of plant capacity (in MW-ac). The correlation includes terms to account for capital-cost contributions from modular- and scalable-equipment components. The indirect-cost multiplier includes contributions from the cost categories already described. HTSE system direct capital costs can be estimated by setting the indirect-cost multiplier equal to a value of one. The HTSE capital-cost correlation based on the HTSE process analysis is presented below. The values of each of the cost-correlation parameters are included in Table 16 and Table 17.

$$TCI = m(a_{scalable}P^{n_{scalable}} + a_{modular}P^{n_{modular}})$$

where

TCI= Total Capital Investment ($/kWe)

P = HTSE system power (MWe)

m = indirect cost multiplier = 1.2383

$a_{scalable}$ = scalable equipment cost coefficient

$n_{scalable}$ = scalable equipment scaling exponent

$a_{modular}$ = modular equipment cost coefficient

$n_{modular}$ = modular equipment scaling exponent,

Table 16. LWR-HTSE capital-cost correlation parameters (HTSE stack capital cost specification of $155/kW-dc,[30] results in 2016 dollars).

	FOAK	NOAK
$a_{scalable}$	719.2	719.2
$n_{scalable}$	-0.504	-0.504
$a_{modular}$	542.7	377.3
$n_{modular}$	-0.043	0

Table 17. LWR-HTSE capital cost correlation parameters (HTSE stack capital cost specification of $155/kW-dc,[30] results in 2020 dollars).

	FOAK	NOAK
$a_{scalable}$	799.1	799.1
$n_{scalable}$	-0.505	-0.505
$a_{modular}$	596.4	414.5
$n_{modular}$	-0.043	0

3.2.1.3.1 SOEC Technology Readiness Level Represented by Capital Cost Analysis

Capital-cost analysis was performed using publicly available and vendor-specific proprietary stack-cost estimates. The public data case uses the projected current hydrogen-production case from the DOE HFTO Hydrogen-Production Record #20006[30] as the basis for stack capital cost ($155/kW) and stack service life (4 years). The HFTO Hydrogen-Production Record projected current hydrogen-production case capital costs apply to SOEC stack technology available at the present time (technology year 2019 in the HFTO Record) for a 50 tonne H_2 per day HTSE plant. While these costs were used as the basis for the public stack-cost analysis, it is worth noting that SOEC stack-manufacturing capacity would need to support installation of a gigawatt-scale (>600 tonne/day hydrogen production) HTSE plant by year 2026 to support startup of an HTSE plant in year 2027 as specified in the dedicated hydrogen-production LCOH analysis presented below (Section 3.3).

The BoP components are, in general, commercial technology, and the pricing information specified for these components corresponds to the current time. Therefore, the technology readiness level schedule questions apply primarily to the stacks (e.g., commercial realization of the specified prices and degradation rates specified).

3.2.2 HTSE Process Operations and Maintenance Costs

HTSE process operations and maintenance (O&M) costs were calculated according to the input specifications listed in Table 18. The O&M cost calculations include a stack service life of 4 years, with annual stack replacements to restore the plant's production capacity to the design value at the start of each operating year. Plant-maintenance costs also include an annual cost of 0.5% of the total direct depreciable costs for unplanned equipment replacements (stack and BoP equipment). The O&M costs do not include an allowance for the 100% replacement of the BoP after 20 years since the cash-flow analyses in this report specify a 20-year project duration.

Table 18. HTSE process O&M cost-estimate basis.

Category	Prairie Island LWR-HTSE plant	Monticello LWR-HTSE plant	Reference or Note
Fixed Operating Costs			
Total Plant Staff	15 (corresponds to 662 tonne/day design hydrogen-production capacity)	13 (corresponds to 383 tonne/day design hydrogen-production capacity)	8 person plant staff for a 50 tonne/day plant assumed[30]; 0.25 scaling exponent for varying plant capacity[49]
Burdened labor cost	$60/hr	$60/hr	
G&A rate/costs	20% of labor	20% of labor	
Licensing, permits, and Fees	N/A	N/A	
Property Tax and Insurance	2% of TCI per year	2% of TCI per year	
Rent	N/A	N/A	
Maintenance and Repairs	0.5% of direct capital costs (DCC) per year	0.5% of DCC per year	
Replacement Costs	0.5% total capital for annual unplanned replacements. 25% annual stack replacement	0.5% total capital for annual unplanned replacements; 25% annual stack replacement	0.5% unplanned replacement costs per year. Full stack replacement every 4 years at specified stack capital cost. Full system replacement at inflated DCC value every 20 years
Process Water **Cooling Water**	$2.00/k-gal $0.02/k-gal	$2.00/k-gal $0.02/k-gal	Cooling water cost is for make-up and chemical treatment

O&M cost estimates for the Prairie Island LWR-HTSE plant and the Monticello LWR-HTSE plant are provided in Table 19 and Table 20, respectively. O&M estimates for both plants correspond to the HTSE plant capacities detailed in Table 11 (both cases correspond to maximal hydrogen-production based on the use of energy available from both Prairie Island nuclear units and the single Monticello nuclear unit).

Table 19. Prairie Island LWR-HTSE annual O&M costs in 2020 dollars.

Fixed O&M Costs					
Burdened labor cost, including overhead	$60.00	$/hour	15	FTEs	$1,904,331
G&A rate	20%	% of labor cost			$380,866
Property Tax and Insurance	2%	% of total capital investment			$15,032,015
Production Maintenance and Repairs	0.50%	% of installed direct capital costs			$2,905,199
Total Fixed O&M					$20,222,411 ($19.6/kWe-yr)
Variable O&M Costs					
Replacement Costs					
Annual Stack Replacement Percentage[α]	25.00%	% of design capacity			$36,812,500[α]
Total Unplanned Replacement	0.50%	% of total direct depreciable costs/year			$3,720,796
Electricity	30	$/MWh-e	1,032	MW-e	$245,435,321
Nuclear process heat	10.4[β]	$/MWh-t	177	MW-t	$14,606,972
Process Water	2	$/k-gal	1,566	k-gal/day	$1,034,528
Cooling Water (make-up and chemical treatment)	0.02	$/k-gal	30,504	k-gal/day	$201,577
Total Variable O&M (including energy costs)					$301,811,695 ($36.89/MWe-hr)
Total Variable O&M (excluding energy costs)					$41,769,401 ($5.11/MWe-hr)

[α] Based on HTSE stack capital cost specification of $155/kW-dc[30]

[β] Based on a thermal-to-electrical conversion efficiency of 34.6%

Table 20. Monticello LWR-HTSE annual O&M costs in 2020 dollars.

Fixed O&M Costs					
Burdened labor cost, including overhead	$60.00	$/hour	13	FTEs	$1,661,125
G&A rate	20%	% of labor cost			$332,225
Property Tax and Insurance	2%	% of total capital investment			$8,873,918
Production Maintenance and Repairs	0.50%	% of installed direct capital costs			$1,715,040
Total Fixed O&M					$12,582,307 ($21.1/kWe-yr)
Variable O&M Costs					
Replacement Costs					
Annual Stack Replacement Percentage[α]	25.00%	% of design capacity			$21,312,500[α]
Total Unplanned Replacement	0.50%	% of total direct depreciable costs/year			$2,196,514
Electricity	30	$/MWh-e	597	MW-e	$142,094,133
Nuclear process heat	9.69[β]	$/MWh-t	103	MW-t	$7,894,520
Process Water	2	$/k-gal	906	k-gal/day	$598,937
Cooling Water (make-up and chemical treatment)	0.02	$/k-gal	17,660	k-gal/day	$116,703
Total Variable O&M (including energy costs)					$174,213,307 ($36.78/MWe-hr)
Total Variable O&M (excluding energy costs)					$24,224,654 ($5.11/MWe-hr)

[α] Based on HTSE stack capital cost specification of $155/kW-dc[30]
[β] Based on a thermal-to-electrical conversion efficiency of 32.3%

3.3 Constant Hydrogen Production (Non-Grid-Integrated) LCOH Production Analysis

One business case for LWR-HTSE plants is to operate in a constant hydrogen-production (non-grid-integrated) mode. The analysis presented here does not account for grid impacts or interactions. It considers the LWR-HTSE plant isolated and standalone as a limiting case. Even if a utility company does not intend to operate in this manner to produce hydrogen, this analysis, and these results are still useful in that they show the bounding / limiting scenario of full hydrogen production without grid interactions. It should be noted that operating in this manner would affect local grid node pricing and therefore the reader should understand that electricity pricing of the regional area would be affected in ways that are not presented or accounted for in this section. The dispatch optimization in Section 4 and a separate forthcoming report that will later be released by NREL provide insight into the NPP-HTSE profitability with grid interactions taken into account when the NPP is allowed to switch between sending electricity to the grid and hydrogen production.

An LWR-HTSE plant configured for constant hydrogen production requires that the LWR nuclear plant provide a constant supply of heat and power to the HTSE plant; therefore, the LWR plant would no longer dispatch electrical power to the grid as part of routine operations. The constant hydrogen-production configuration would simplify the HTSE process-operating scheme and reduce capital expenditures required prior to HTSE plant startup (use of hot-standby operating mode, hydrogen storage, and replacement of removed electrical generation capacity are not required or considered in this analysis). Because the nuclear plant would no longer dispatch electrical power to the grid, transient operating conditions associated with entering and exiting HTSE process hot-standby mode (and the associated transient system operations) would also be significantly reduced.

While there are advantages associated with a simplified, non-grid-integrated LWR-HTSE process operating scheme, it provides fewer potential revenue streams—i.e., electrical power dispatch is no longer an option, which also eliminates potential income from capacity-market payments—and decreased ability to operate the plant in a manner that allows the production of the product (hydrogen or electrical power) with the highest profit margin in any given time period.

Hydrogen-production costs for an LWR-HTSE system configured for constant hydrogen (non-grid-integrated) production were evaluated. The DOE H_2A model[48] was configured with the LWR-HTSE process-performance parameters described in Section 3.1.3, the capital costs described in Section 3.2.1.2, the O&M costs described in Section 3.2.2, and the project financial-input parameters listed in Table 21 to calculate the non-grid-integrated LWR-HTSE plant LCOH.

Table 21. LWR-HTSE constant hydrogen production LCOH analysis input parameters.

Parameter	Value
Start-up year	2027
Length of construction period	1 year
Start-up time	1 year
Plant life	20 years
Depreciation schedule	15-year MACRS
% Equity financing	40%
Interest rate on debt	5%
Debt period	20 years
% of fixed operating costs during start-up	75%
% of revenues during start-up	75%
% of variable operating costs during start-up	75%
Decommissioning costs (% of TDC)	10%
Salvage value (% of TCI)	10%
Inflation rate	1.9%
After-tax real internal rate of return (IRR)	9.9%
State taxes	6%
Federal taxes	21%
Indirect costs	
Site preparation (% of DCC)	2%
Engineering and design (% of DCC)	2.3%
Process contingency (% of DCC)	1.6%
Project contingency (% of DCC)	7.2%
Contractor’s fee (% of DCC)	10%
Legal fee (% of DCC)	5%
Land (% of TDC)	1%
Plant type (NOAK)	NOAK
Learning rate for modular equipment cost reduction	95%
NOAK plant stack cost	
Base HTSE Case: HFTO Record Current Technology	$155/kW-dc stack cost “Current Case” in reference[41]
Advanced HTSE case: Integrated stack module design	$35/kW-dc stack module (stack and balance-of-module components included)[56]

The LCOH analysis results for a baseline case with an HTSE plant providing actual hydrogen-production capacity of 347 tonnes hydrogen per day (with a design capacity of 383 tonnes/day) and an energy price of $30/MWh-e are presented in Figure 45. An HTSE plant of this capacity would use 597 MW-ac of total power input (550 MW-dc stack power input) and would use the majority of the Monticello NPP energy output. The LCOH for this baseline case is $1.93/kg (in 2020 dollars). It is apparent from this figure that the largest contributor to the LCOH is the energy cost.

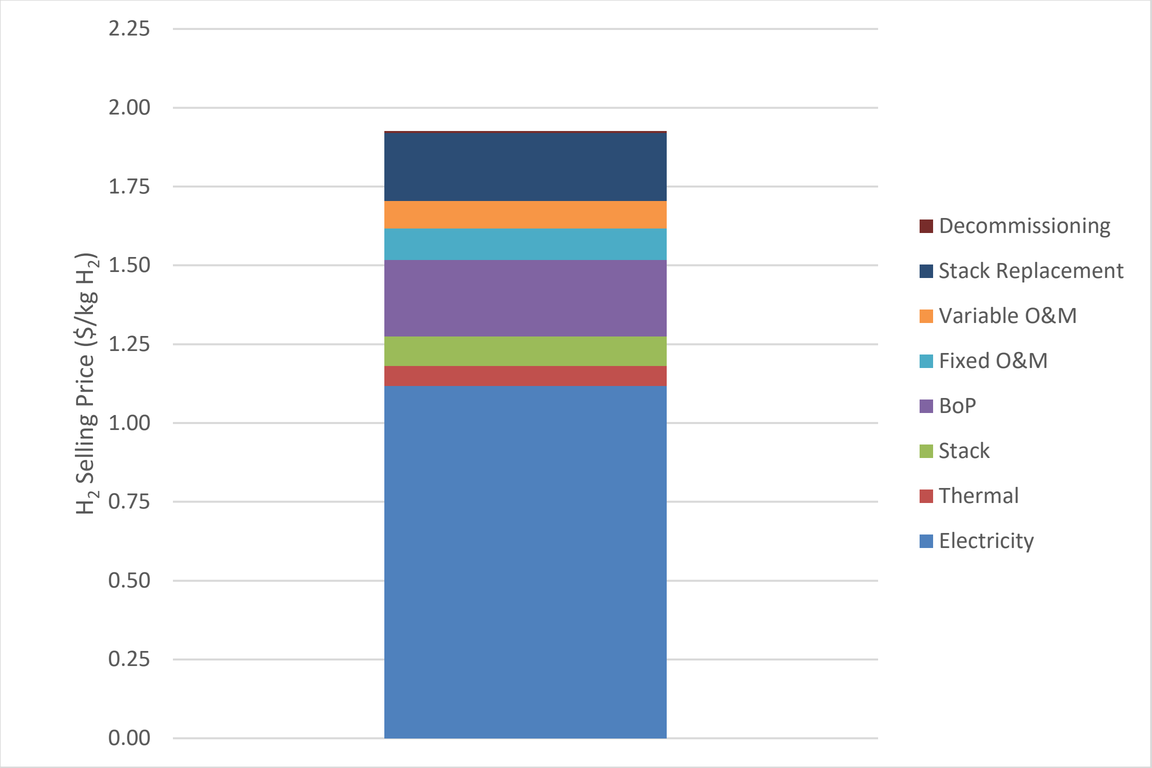

Figure 45. LCOH cost components for an NOAK constant hydrogen production LWR-HTSE system configuration with 347 tonnes per day actual hydrogen-production capacity (383 tonnes/day design capacity), stack cost of $155/kW-dc, and an energy price of $30/MWh-e.

3.3.1 Sensitivity Analyses

Sensitivity analysis was performed to evaluate the impact of energy price and other key variables on the LCOH production. Again, as described at the first of Section 3.3, this analysis is a limiting/bounding analysis that shows the resulting possibilities if an NPP were completely dedicated to hydrogen production. The impacts to the regional grid market pricing of dedicating an LWR to hydrogen production is not considered in this analysis. The interactions of an NPP-HTSE complex that is able to dispatch between sending electricity to the grid and to the hydrogen plant is not considered here but is considered in the optimization of Section 4. A set of base conditions for the sensitivity analysis is listed in Table 22. These correspond to an HTSE plant with high-value stack costs and a stack service life consistent with the HFTO Hydrogen Production Record[30] and a design capacity that would use a total energy input similar to that provided by using the Xcel Energy Monticello NPP as a dedicated energy source. Also discussed in this section, and the next is an advanced case that has a low-value stack cost based on publicly available information and calculations that represent the state-of-the-art technology performance based on predictions by various SOEC suppliers for the near term.

Table 22. Base high-value stack cost and advanced low-value stack cost cases: conditions for HTSE sensitivity analysis.

Parameter	Base Case, High Stack Cost	Advanced Case, Low Stack Cost
Electricity Price	$30/MWh-e	-
Stack Cost	$155/kW-dc	$27/kW-dc
IRR	9.9%	-
Learning Rate for Modular Equipment Cost Reductions	95%	-
Stack Service Life	4 years	7 years
Previous HTSE Plant Installations	100 (NOAK plant type)	-
Plant Design Capacity	383 tonnes/day (550 MW-dc stack power input; 597 MW-ac system power input)	-

Figure 46 is a tornado chart that illustrates the LCOH sensitivity to the variables listed in Table 22 for the base high stack value case. Each of the sensitivity variables shown in the tornado chart is manipulated individually while all other variables are kept constant at the base values listed in Table 22. The sensitivity-variable lower bound, base value, and upper bound are listed in brackets next to the chart axis labels. The upper and lower bounds selected for each of the variables are expected to bracket the conditions that could characterize an LWR-based HTSE plant installation within an approximately 5-year timeline (or once the manufacturing capacity to support HTSE plant installations of the specified size are available). The results presented in Figure 46 are sorted such that the variables that result in the largest net change in LCOH are positioned at the top of the chart.

It can be observed from Figure 46 that, as expected, the specified changes in electricity price and stack cost have the largest impact on LCOH. Note that the range of electricity prices evaluated represents expected trends in future electricity market pricing (as well as typical LWR O&M costs). A higher value for the stack price is not included in the sensitivity analysis since the base value corresponds to the HFTO Hydrogen Production Record[30] current technology case and is already considered a high case. The low value of $27/kW corresponds to the advanced case value computed from costs reported by Tang et al.[44] for an SOEC stack module designed for manufacture in a mass production facility. The basis for the $27/kW stack cost calculation is included in Table 23.

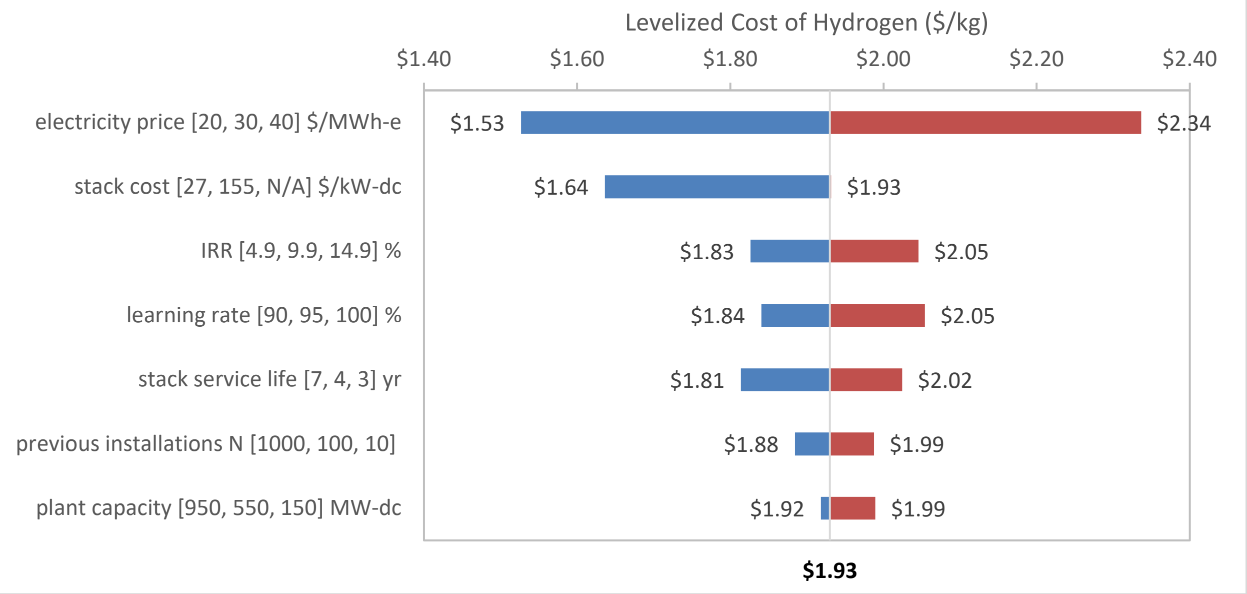

Figure 46. Sensitivity of LCOH to selected constant hydrogen production case input parameters. Base high stack cost of $155/kW-dc results in LCOH value of $1.93/kg corresponds to an HTSE plant with design capacity of 383 tonne/day (550 MW-dc stack power, 597 MW-ac system power), which corresponds to the maximum HTSE plant capacity that could be coupled to Monticello NPP.

Table 23. Basis for advanced case low-value stack-cost sensitivity-analysis input specification.

Parameter	Value	Note
Stacks/module	40	
Cells/stack	350	
Cells/module	14,000	Calculated based on values reported in [48]
Cell area	81 cm²	
Current Density	1.25 A/cm²	
Operating voltage	1.35 V/cell	
Current	101.25 A	Calculated based on values reported in [48]
Stack power	47.8 kW	Calculated based on values reported in [48]
Module Power	1913.6 kW	Calculated based on values reported in [48]
Stack cost	$1300/stack ($27/kW)	Value read from Tang et al Figure 26 [48]
Module cost	$15,000/module	Value read from Tang et al Figure 27 [48]
System cost	$67,000/system ($35/kW)	System includes 40 stacks and one module

Figure 47 provides additional information on the sensitivity of LCOH to the energy price. It can be observed from this figure that a $10/MWh-e decrease in the price of the energy obtained from the LWR results in approximately a $0.40/kg decrease in hydrogen-production cost. The LWR energy cost available to the HTSE plant is therefore a key variable in determining the economic viability of an LWR-HTSE hydrogen-production plant. The thermal-energy unit price is calculated by applying the nuclear plant thermal-conversion efficiency to the electrical-power price.

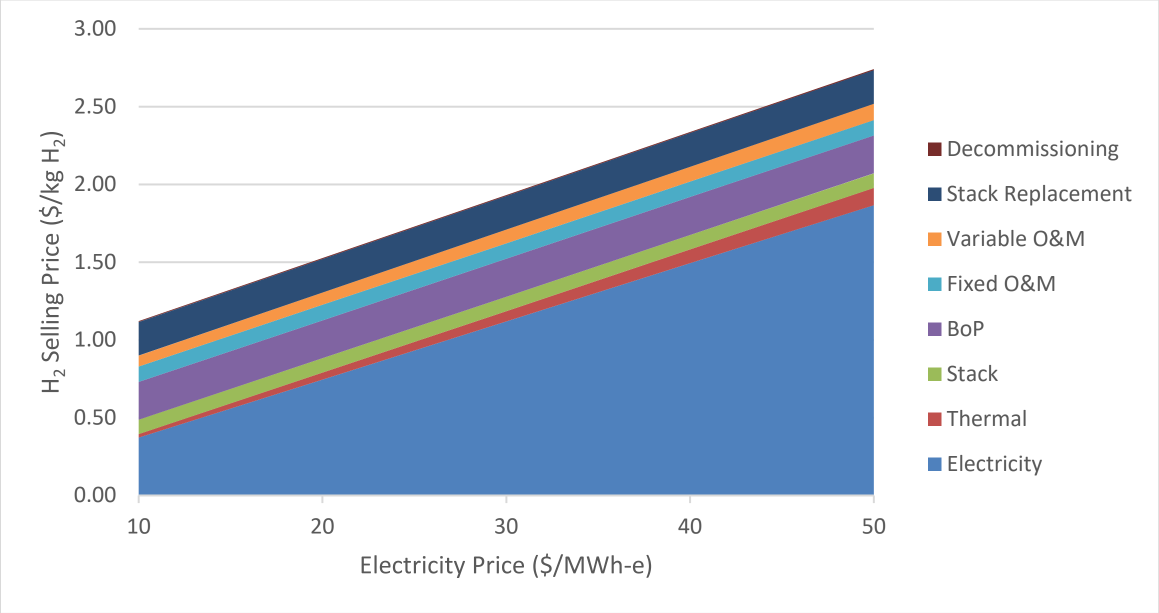

Figure 47. LCOH as function of energy price for an NOAK constant hydrogen production HTSE plant, base high stack cost (\$155/kW-dc stack cost) with 347 tonnes/day actual hydrogen production capacity (383 tonnes/day design capacity).

In addition to the energy costs, Figure 46 indicates that the HTSE system capital costs also have a significant impact on the LCOH. The HTSE system capital costs provide a direct contribution to the LCOH via the initial capital investment associated with the stack and BoP, but also result in an indirect contribution to the LCOH by affecting the magnitude of the O&M costs as described in Section 3.2.2 (stack-replacement costs, maintenance costs, property tax and insurance costs, etc., are a function of the capital costs).

Figure 48 provides a plot of the LCOH sensitivity to HTSE plant capital costs. In this figure, the upper horizontal axis displays the HTSE plant total capital investment while the lower horizontal axis displays the corresponding modular-equipment capital costs. The total capital investment (upper horizontal axis) includes the capital costs of the HTSE plant, TDL, electrical transmission from the nuclear plant to the HTSE plant, feed/utility system costs, as well as indirect costs. The modular equipment (lower horizontal axis) includes the modular HTSE and BoP units that are expected to be manufactured offsite before being transported to the LWR site for installation and integration with the site-specific LWR-HTSE infrastructure (i.e., the equipment and services that account for the additional costs represented by the total capital investment).

The capital-cost sensitivity analysis shown in Figure 48 is based on a stack-cost specification equal to 30% of the modular-equipment costs (the stack cost provides contributions to the LCOH from both the initial cost of plant construction as well as through recurring stack-replacement costs). Stack service life determines the frequency with which stacks must be replaced. Figure 48 includes curves that indicate the estimated LCOH for HTSE plants constructed using SOEC stacks with 4 and 7-year service lives. The data identify a point that corresponds to the LCOH and total capital investment of an HTSE plant with specifications that correspond to sensitivity-analysis base values identified in Table 22 (e.g., 550 MW-dc HTSE plant design capacity, \$155/kW stack price, 4-year stack service life, \$30/MWh-e electricity price).

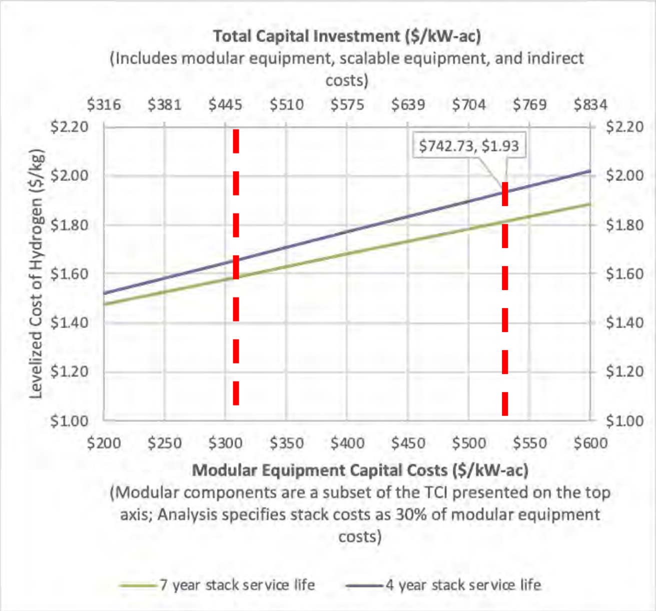

Figure 48. LCOH sensitivity to HTSE plant capital cost and stack service life. Stack costs specified as 30% of modular equipment costs. HTSE plant design capacity of 383 tonne/day (550 MW-dc) with $30/MWh-e electricity price (see Table 22 for listing of additional input parameter specifications). Dashed lines correspond to the base and advanced case.

Another HTSE-process design specification that impacts LCOH is the HTSE plant capacity. As shown in Section 3.2.1.2, the HTSE process unit capital costs (reported in $/kW-e) decrease with increasing plant capacity. Increasing HTSE plant capacity by addition of a single HTSE modular block unit (25 MW-dc per block) will result in a larger decrease in the unit ($/kW) HTSE capital costs for a small-capacity plant than for a large-capacity plant due to the effect of economies of scale becoming less significant at larger plant capacity. This effect results in a decrease in the slope of the unit capital cost versus plant capacity curves with increasing plant capacity as shown for FOAK plant types in Figure 43 and NOAK plant types in Figure 44. A plot of LCOH versus plant capacity is provided in Figure 49. While increased unit capital costs are a contributing factor behind elevated LCOH for small capacity HTSE plants, the primary driver for the higher LCOH observed for small capacity plants is the increased fixed O&M cost associated with the proportionately higher number of plant staff members per unit of hydrogen production (the total number of plant staff members is adjusted for plant capacity according to a scaling exponent of 0.25 per the H_2A centralized SOEC hydrogen production demo models). The LCOH sensitivity to plant capacity shown in Figure 46 indicates that there is a greater impact from decreasing the plant capacity than from increasing the plant capacity (relative to the base-capacity value). This is consistent with the decrease in the slope of the LCOH versus plant capacity shown in Figure 49.

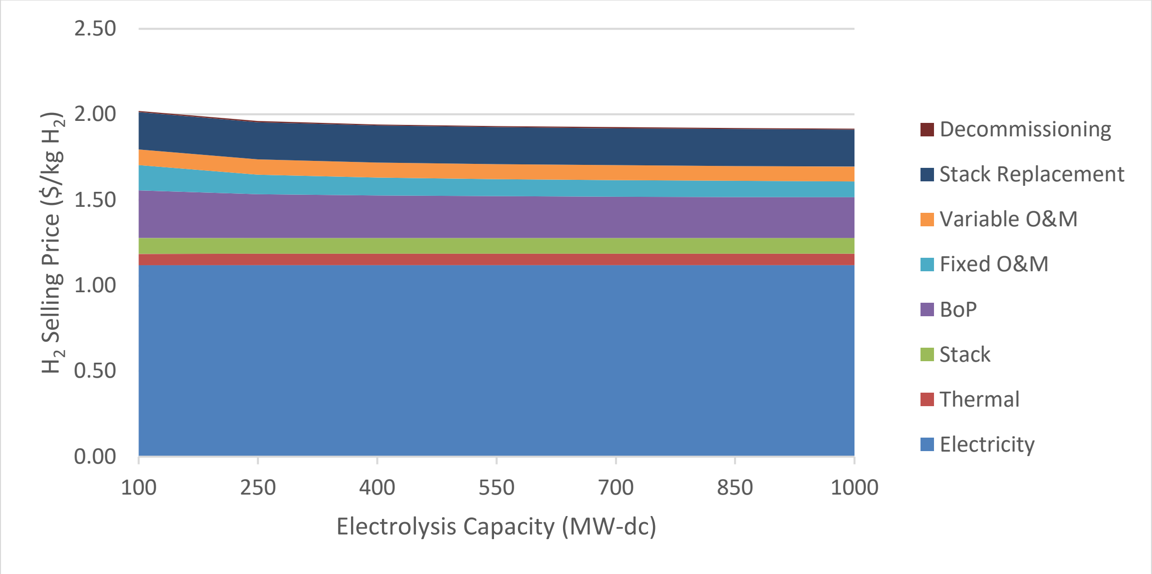

Figure 49. LCOH as function of LWR plant capacity (NOAK plant type, $30/MWh-e energy price, $155/kW-dc stack cost)

3.3.2 Comparison of HTSE and Steam Methane Reforming

3.3.2.1 Natural Gas Steam Methane Reforming

The incumbent competitor to an NPP-HTSE hydrogen plant is natural gas SMR. As a result, the highest hydrogen price as a function of demand size will be determined by the economies of scale that an SMR plant can achieve. It should be noted, however, that the economics of natural gas plants are very different from those of an NPP-HTSE. SMR LCOH costs were calculated using the H_2A model[38] with input parameters defined in Table 24. Baseline SMR plant installed capital costs of $132,500,000 (in 2020 dollars) for a 380 tonne hydrogen/day production plant were scaled using a scaling exponent of 0.6. The SMR plant capital costs include an additional hydrogen-product compressor to provide a product gas with an output pressure of 69 bar, which is equivalent to that specified for the HTSE plant.

Table 24. H_2A model input parameters for SMR LCOH analysis.

Input Parameter	Value
Natural gas price	Varies
Plant capacity	Varies
Startup year	2030
Construction period	3 years
Startup period	1 year
Plant life	20 years
Depreciation schedule	15-year MACRS
Equity financing	60%
Interest rate on debt	3.70%
Fixed operating costs during startup	75%
Variable operating costs during startup	75%
Revenues during startup period	50%
Decommissioning costs	10%

Input Parameter	Value
Salvage value	10%
Inflation rate	1.9%
After-tax real IRR	7.0%
State tax rate	6%
Federal tax rate	21%
Total tax rate	25.74%

SMR-plant capital costs were extrapolated (using the 6/10 scaling factor) outside the H_2A recommended range for plant capacity values below 235 tonne/day. For plant capacities above, the H_2A recommended upper limit for scaling capacity of 425 MT H_2/day; the capital-cost calculations were modified to account for use of multiple process trains; i.e., the economic benefits associated with economies of scale are limited to the equipment sizes associated with a 425 tonne/day plant capacity. This modification prevents economy-of-scale capital-cost reductions from being applied to predict costs for equipment that would be impractical to construct or transport.

3.3.3 Impact of Natural Gas Price

While fuel costs are low for an NPP, they are the main contributor for a natural gas plant. While natural gas prices are currently very low, they have historically seen much variability. As a result, four conditions are considered in this subsection: (1) a medium gas price (which corresponds to the U.S. EIA 2021 Annual Energy Outlook (AEO)[48] West North Central (WNC) Region Reference Case), (2) a low price corresponding to the EIA 2021 AEO WNC Region High Oil and Gas Supply Case, (3) a high natural gas price corresponding to the 2021 AEO WNC Region Low Oil and Gas Supply Case, and (4) a modified version of the 2021 AEO West North C Region Reference Case. A plot of each of these natural gas price projections versus time is shown in Figure 50.

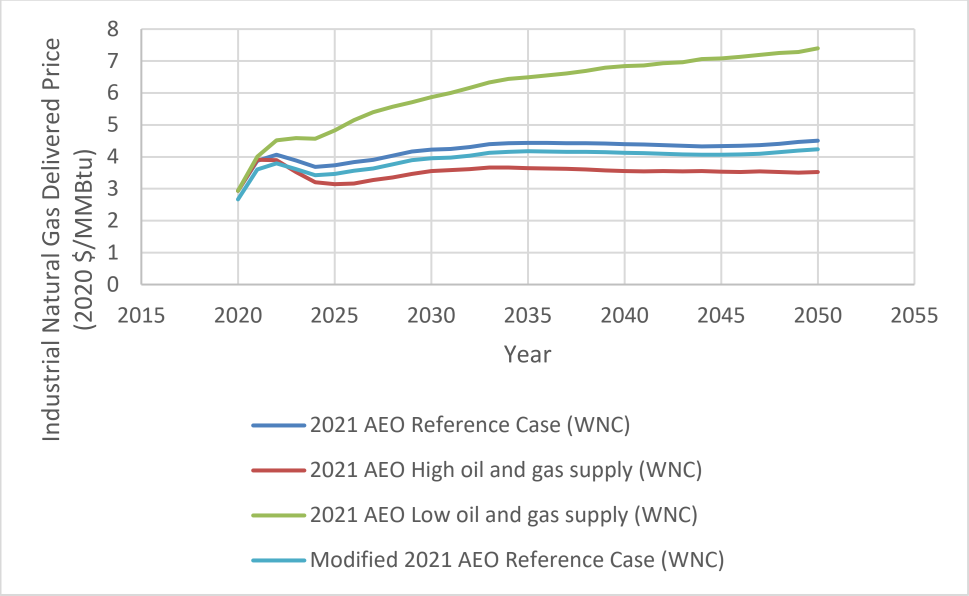

Figure 50. Projected natural gas pricing in the West North Central Region as reported in selected EIA 2021 Annual Energy Outlook Analysis Cases.[48] The basis for the SMR hydrogen production cost analysis presented in this report is a modified natural gas price projection in which the 2021 AEO West North Central Region Reference Case is offset (indicated by the light blue line).

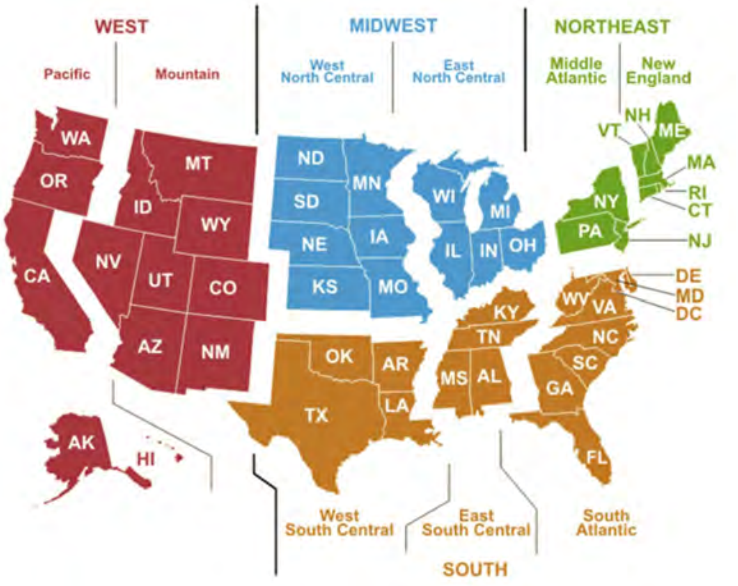

Figure 51. US Census regions and divisions.[49]

Figure 52 includes plots of SMR LCOH for each of the EIA 2020 AEO natural gas price cases as a function of SMR plant capacity. The plant-capacity scaling range recommended by the H_2A model falls between the vertical dotted lines. As previously described, the H_2A model was modified to account for use of multiple process trains for SMR plant capacities above the suggested plant capacity. As a result of the H_2A model modification, minimal additional LCOH reductions due to economies of scale are realized for plant design capacities exceeding 425 MT/day (382.5 MT/day actual production). The LCOH values corresponding to SMR plant actual production capacities of 382.5 MT/day, therefore, represent the SMR price floor, at which point the economies of scale have been maximized and minimal LCOH reductions can be achieved from increases in plant capacity. Beyond this point, the natural gas price is the primary driver of the SMR LCOH.

The SMR H_2A model-default plant-design capacity of 380 MT/day (342 MT/day actual production) was chosen as the basis for comparison with HTSE plant LCOH. This SMR plant capacity is very near the point on the LCOH vs capacity curve where the SMR LCOH price floor is achieved. It provides a comparable level of hydrogen production to the 550 MW-dc HTSE plant (383 MT/day design capacity, 347 MT/day actual production) specified in the sensitivity-analysis base case.

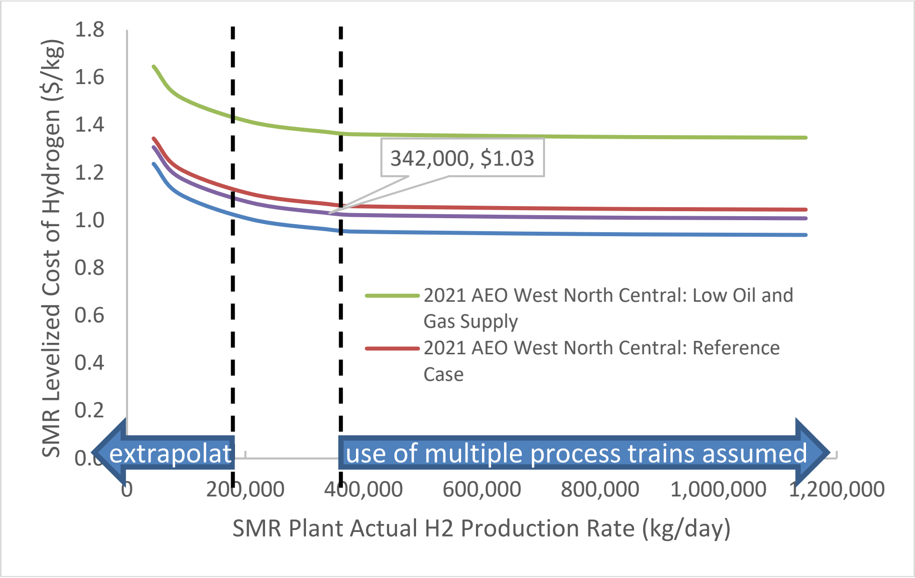

Figure 52. LCOH of SMR-based hydrogen production as a function of plant capacity and natural gas pricing. Industrial natural gas pricing based on EIA 2021 AEO WNC Region Reference Case, Low and High Oil and Gas Supply Cases, as well as Modified Reference Case.

3.3.4 Impact of Carbon Tax

The current and future price of hydrogen is tied to the natural gas industry, specifically the availability of natural gas in the U.S. market and the possible future implementation of a carbon tax or credit system. While the impact of the price of natural gas was considered previously, this section will investigate the implementation of carbon taxes. In theory, this could be both in the form of a traditional tax, or as a function of the cost of carbon sequestration. A low value of \$25/tonne-CO_2 corresponds to the 2025 anticipated rate in the Minnesota market.[50] Some studies even envision prices as high as \$100/tonne-CO_2 to reach deep decarbonization.[51] This was selected as the high value.

The values specified for the cost of CO_2 were applied to the SMR LCOH to translate this cost to an increase in the market price for hydrogen (based on SMR production). Using estimates from NREL/TP-570-27637, the life-cycle emissions from an SMR plant can be calculated at around 8.9 kg-CO_2/kg-H_2.[52] For the low and high carbon-tax rates, this corresponds to an added \$0.22/kg-H_2 and \$0.89/kg-H_2, respectively.

If SMR plants were to implement carbon capture (CC) the resulting LCOH of SMR + CC could be in the range of \$0.48 to \$0.99/kg-H_2 or a carbon capture cost of \$0.25/kg-CO_2 to \$0.90/kg-CO_2.[53,54,55]

3.3.5 Comparison of HTSE and SMR LCOH

Two HTSE cases were defined for the purposes of comparing HTSE and SMR LCOH. The HTSE cases include (1) a Base Case identical to the case defined by the base condition for the sensitivity analyses in Section 3.3.1 and (2) an Advanced Case that incorporates improvements to the stack cost and stack service life. Specifications for the HTSE cases are provided in Table 25. The SMR LCOH based on a modified version of the 2021 AEO West North Central Region Reference Case, in which all years of the AEO WNC Region Reference Case projection are offset, is plotted with zero, \$25/tonne, and \$100/tonne cost of CO_2 in Figure 53. The LCOH for both HTSE cases is plotted as a function of the electricity cost (the sensitivity variable with the greatest impact on HTSE LCOH). Hydrogen transportation costs of \$0.16/kg are included in the HTSE LCOH shown in Figure 53. No transportation costs are included for the SMR cases based on the assumption that the hydrogen produced by an SMR plant would be consumed by a customer co-located with the SMR plant (e.g., a petroleum refinery).

Table 25. Definition of HTSE cases used for LCOH comparison with natural gas SMR.

	HTSE Base Case: HFTO Record Stack Cost[α]	HTSE Advanced Case: Integrated Stack Module Design[β]	Reference or Note
HTSE plant design capacity (MW-e)	550 (stacks) 597 (system)	550 (stacks) 597 (system)	Corresponds to maximum Monticello HTSE plant capacity
HTSE Plant Hydrogen Production Rate (tonnes/day)	383 (design) 347 (actual)	383 (design) 347 (actual)	Net capacity factor of 90.5% (includes contributions from plant down time and cell degradation)
Electricity price ($/MWh-e)	varies	varies	Electricity price is a sensitivity analysis input parameter
Stack cost ($/kW-dc)	155	35	Base Case cost is for stack only; advanced case is stack module (stack plus balance-of-module)
Direct capital cost ($/kW-ac)	574	345	Based on a 95% learning rate per Table 21
Total capital investment ($/kW-ac)	742	446	Based on a 95% learning rate per Table 21
Real IRR (%)	9.9%	9.9%	Table 21
Stack Service Life (years)	4	7	Correspond to Current and Future Cases
Number of previous HTSE plant installations, N	100	100	NOAK Plant Type
LCOH with energy price of $30/MWh-e ($/kg-H_2)	1.93	1.53	Nuclear plant thermal efficiency used to derive corresponding thermal energy price

[α] The HTSE Base Case is equivalent to the sensitivity analysis base conditions listed in Table 22

[β] The HTSE Advanced Case incorporates improvements to both the stack cost and stack service life relative to the sensitivity analysis base conditions; this case is therefore not represented in the sensitivity analyses presented in Section 3.3.1.

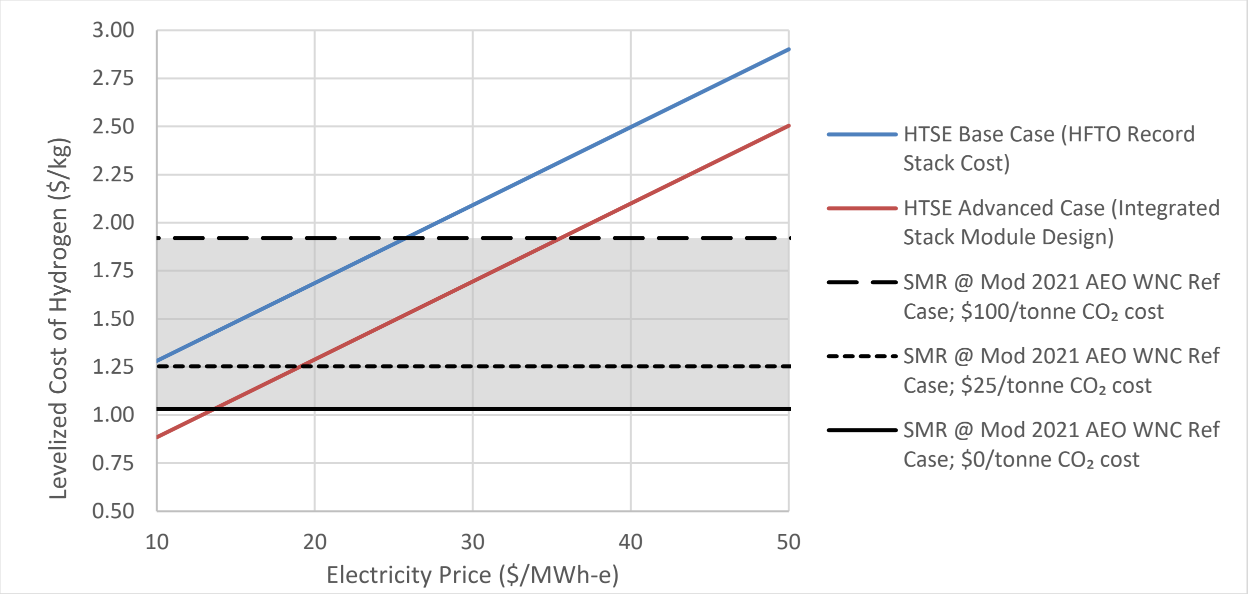

Figure 53. LCOH of 347 tonne/day HTSE base and advanced cases versus 342 tonne/day SMR with $0, $25/tonne, and $100/tonne CO_2 cost. SMR and HTSE case definitions provided in Table 24 and Table 25, respectively. The HTSE LCOH includes a $0.16/kg adder for the cost of transporting hydrogen product to an off-site customer. SMR natural gas feedstock pricing based on Modified 2021 AEO WNC Region Reference Case.

It is apparent from Figure 53 that, with electricity pricing of $30/MWh-e, an SOEC stack cost of $155/kW (HTSE base case), and no cost for SMR CO_2 emissions, it would not be possible for a constant-hydrogen-production mode LWR-HTSE plant to complete with an SMR plant purely on a hydrogen-production cost basis. Availability of SOEC technology consistent with the HTSE Advanced Case would allow the HTSE plant to produce hydrogen at a cost competitive with SMR if the nuclear plant could sell electricity to the HTSE plant at a price of $13.6/MWh-e. This electricity price is lower than the current or projected future O&M costs for existing LWRs, suggesting that it is unlikely that an LWR powered HTSE plant could produce hydrogen at a lower price than SMR for scenarios in which natural gas pricing is aligned with the modified 2021 AEO WNC Region price projection and no carbon tax and/or clean hydrogen production credit exists.

An HTSE production credit for avoided carbon emissions, or a cost of carbon added to the SMR hydrogen-production cost could significantly change the economic favorability of hydrogen production via LWR-HTSE. As indicated in Figure 53, the presence of a $25/tonne or $100/tonne carbon tax would increase the range of electricity pricing for which an LWR-HTSE plant could be cost competitive with natural gas SMR.

At natural gas pricing consistent with the modified 2021 AEO WNC Reference Case natural gas price projection with a $25/tonne CO_2 tax, an HTSE plant could produce hydrogen at a price competitive with SMR at an electricity price of $9.2/MWh-e for the HTSE Base Case or $19/MWh-e for the HTSE Advanced Case (the HTSE LCOH includes a $0.16/kg transportation adder). Although these electricity prices are lower than current O&M costs for many LWR nuclear plants and are also lower than the average electricity pricing in many markets—if the LWR provides power to the HTSE plant it does so at the opportunity cost of not selling this power to the electricity market—both LWR NPP O&M costs and future electricity prices are expected to decrease in the coming decade. Based on recent analyses, however, it is unlikely that LWR O&M costs are likely to decrease significantly below $20/MWh-e such that a $25/tonne CO_2 penalty or production credit would be insufficient to enable cost competitive LWR/HTSE-based hydrogen production.

An increase in the clean hydrogen production credit or CO_2 emission penalty to \$100/tonne CO_2 would increase the electricity price at which an LWR could sell power to an HTSE plant for cost competitive hydrogen production to \$25.9/MWh-e for the base HTSE Base Case and \$35.7/MWh-e for the HTSE Advanced Case (the HTSE LCOH includes a \$0.16/kg transportation adder). Many LWR plants could operate profitably today by selling power to an HTSE plant at a price greater than \$30/MWh-e, and as additional cost saving measures are implemented at LWR plants it is expected that many could operate profitably with future electricity sales prices of around \$25/MWh-e or greater.

If future natural gas pricing is higher than the modified 2021 AEO WNC Region Reference Case projection used in this analysis SMR hydrogen production costs would increase, and the electricity prices at which HTSE could be cost competitive with SMR would increase accordingly. However, if future natural gas pricing is lower than the modified 20201 AEO WNC Region Reference Case projection used in this analysis, decreases to the electricity prices identified above would be required to enable the HTSE plant to produce hydrogen at a cost competitive with natural gas SMR. Plots of LWR-HTSE LCOH versus natural gas SMR LCOH that include the 2021 AEO WNC Region Low Oil & Gas Availability Case (high natural gas prices), Modified 2021 AEO WNC Region Reference Case, and 2021 AEO WNC Region High Oil & Gas Availability Case are included in Section F-6 of Appendix D.

3.4 Summary of HTSE Process & Status Financial Analysis

A gigawatt-scale LWR-HTSE process design model was built and used to evaluate some basic steady state constant hydrogen production scenarios. The evaluation determined that an HTSE, scaled to match the energy output of an LWR plant of 597 MW-e, would require approximately 5% of the LWR total steam flow to provide the process-heat input needed to vaporize the HTSE process feedwater. The analysis specified use of Therminol-66 as the HTF to transfer nuclear process heat an assumed distance of 1 km to the HTSE plant. The HTSE plant was determined to have specific electricity and thermal energy requirements of 37.4 kWh-e/kg-H_2 and 6.4 kWh-t/kg-H_2 respectively. The HTSE plant efficiency was calculated as 88.9% on an HHV basis. Two SOEC technology cases were considered in evaluating HTSE LCOH, under basic steady state constant hydrogen production:

1. The HTSE Base Case is projected from an assumed stack-cost specification of \$155/kW-dc consistent with that reported for current SOEC technology in the DOE HFTO Hydrogen Production Record [30]. The base case also uses the HFTO Record stack service life specification of 4 years. This analysis includes annual stack replacements to restore the HTSE plant design-capacity rating at the start of each operating year. The base case NOAK HTSE plant with a hydrogen-production design capacity of 383 tonne H2/day (597 MW-e) has DCC of \$574/kW-ac (includes assumptions on HTSE plant equipment and nuclear plant heat- and power-delivery equipment) and a total capital investment of \$742/kW-ac (includes project indirect costs in addition to DCCs listed above). When energy from the LWR is purchased at a price of \$30/MWh-e (the nuclear plant's thermal efficiency is used to derive corresponding thermal-energy price), the base case HTSE plant is able to produce hydrogen at an LCOH of \$1.93/kg, which does not include product storage or transportation costs.

2. The HTSE Advanced Case uses a stack module (stack plus balance-of-module) cost specification of \$35/kW by comparison of publicly available information from various SOEC vendors. The advanced-case stack service life is specified as 7 years, consistent with current best-in-class SOEC technology. The advanced case NOAK HTSE plant with a hydrogen-production design capacity of 383 tonne/day (597 MW-e) has DCC of \$345/kW-ac and a total capital investment of \$446/kW-ac. When energy from the LWR is purchased at a price of \$30/MWh-e, the advanced case HTSE plant is able to produce hydrogen at an LCOH of \$1.53/kg (excluding storage and transport costs).

A summary of the assumptions and results for the Base and Advanced HTSE cases is shown below in Table 26.

Table 26. Summary of Base and Advanced HTSE Model Design Cases.

	Base Case	Advanced Case	Notes
Power Utilization	587 MW-e	587 MW-e	
H_2 Production	347 tonne/day production 383 tonne/day design	347 tonne/day production 383 tonne/day design	
Efficiency (HHV)	88.9%	88.9%	Includes both thermal- and electrical-energy consumption
Operating Pressure	5 bar	5 bar	Based on maximizing system efficiency by trending operating pressure and steam utilization versus system efficiency
Steam Utilization (conversion of reactant steam)	80%	80%	
Electricity Required	37.4 kWh-e/kg-H_2	37.4 kWh-e/kg-H_2	
Thermal Energy Required	6.4 kWh-t/kg-H_2	6.4 kWh-t/kg-H_2	
Technology Horizon	NOAK, 95% learning rate	NOAK, 95% learning rate	95% corresponds to a 5% cost reduction with every doubling of the number of units produced
Stack Cost	$155/kW-dc	$35/kW-dc	
Service Life	4 years	7 years	Assumes annual stack replacements to restore the HTSE plant design-capacity rating at the start of each operating year
Direct Capital Cost	$574/kW-ac	$345/kW-ac	
Total Capital Investment	$742/kW-ac	$446/kW-ac	
Levelized Cost of H_2 (HTSE)	$1.93/kg	$1.53/kg	At $30/MWh electricity cost. Excluding storage and transport costs. If a carbon tax of $25/ton or $100/ton were in place, this would raise the competitor SMR LCOH by $0.22/kg-$H_2$ and $0.89/kg-$H_2$ respectively

The advanced HTSE case represents SOEC-vendor stack-pricing estimates and best-in-class cell-degradation-rate performance. It is expected that SOEC technology will be aligned with the advanced HTSE case within the HTSE plant construction schedule evaluated in this analysis (i.e., start of plant construction in 2026, with plant startup in 2027). The LCOH corresponding to the advanced LCOH case is, therefore, most applicable for the purposes of comparison with natural gas SMR. The combination of decreased stack-module cost and increased stack service life enables the HTSE Advanced Case to achieve a $0.40/kg reduction in LCOH relative to the HTSE base case across the range of electricity prices evaluated (Figure 53).

A sensitivity analysis was completed to evaluate the impact of several key process and economic parameters on the HTSE LCOH. The upper and lower bounds for each of the input parameters were selected to correspond to expected technology advancement and/or variation in market conditions. Based on the selected range over which the sensitivity variables were perturbed, the parameters that have the greatest impact on LCOH are energy price and SOEC stack cost. A second set of variables including the learning rate (for decreases in modular equipment costs as a function of the number of units produced by the equipment manufacturer), stack service life, and IRR have a medium impact on the LCOH. Once NOAK plant status has been achieved (defined as previous deployment of N = 100 count of 25 MW-e modular blocks, or 2.5 GW-e of production capacity) and a base plant capacity of several hundred MW is considered, perturbations to these variables have a less-pronounced impact on LCOH than the sensitivity variables identified above. Additional results and observations from the sensitivity analysis are listed below:

- Electricity price is a major cost driver of HTSE LCOH. A decrease of \$10/MWh-e in the price of the energy obtained from the LWR would result in approximately a \$0.40/kg decrease in the HTSE hydrogen production cost.
- Stack costs are also a major driver of the HTSE LCOH. The stack costs contribute to the initial plant construction costs as well as the HTSE plant O&M costs (for stack replacement). There is a significant difference between the values of the stack cost specified by DOE HFTO for a current-technology hydrogen-production cost evaluation [30] versus the stack cost that specific SOEC vendors have reported would be possible using current technology with manufacturing capacity of several hundred megawatts per year. Therefore, a prospective HTSE plant developer could significantly reduce uncertainties in hydrogen-production cost by obtaining competitive project-specific stack and system pricing information from SOEC vendors.
- The learning rate affects the HTSE plant modular-equipment capital costs. Variation in the learning rate of ±5% have a moderate impact on LCOH relative to the other sensitivity variables evaluated. Planned expansions in vendor-specific manufacturing capacity could affect the learning rate that is realized as establishment of large-scale SOEC manufacturing capacity continues in the coming years.
- Provided a NOAK HTSE plant is installed at a large scale (several hundred megawatts), scalable plant components (nuclear process-heat delivery, electrical-power distribution, utilities, etc.) will have achieved sufficient economies of scale and modular HTSE process components will have obtained cost reductions through economies of mass production. Therefore, a relatively minor impact to the LCOH is obtained from the HTSE plant capacity specification over a range from several hundred megawatts to gigawatt-scale HTSE.

A comparison of LWR-HTSE and natural gas SMR LCOH was performed to identify cases where HTSE could produce hydrogen at a cost competitive with SMR. The SMR LCOH is highly dependent on natural gas pricing. Use of a modified 2021 AEO WNC Region Reference Case natural gas price projection results in an LCOH estimate of \$1.03/kg for a natural gas SMR plant with a design capacity of 380 tonne/day (342 tonne/day actual production rate).

Because hydrogen produced via SMR is associated with significant carbon emissions, some customers may be willing to pay a price premium for carbon-free, “green” hydrogen or that a price on carbon could increase the effective cost of SMR-derived hydrogen. The natural gas SMR LCOH is increased by approximately \$0.01/kg for every \$1/MT-CO_2 tax that is applied. Specifically, the calculations described in Section 3.3.2 indicate that a carbon tax of \$25/tonne-CO_2 would result in an increase in the natural gas SMR LCOH of \$0.22/kg. In addition to the electricity price and SOEC stack cost/service life, the presence of a CO_2 tax is one of the most significant drivers that could determine the profitability of hydrogen production via HTSE relative to SMR.

The analysis concludes that advanced HTSE technology (e.g., the advanced HTSE case), a low electricity price (e.g., the advanced HTSE case requires an electricity price of $13.6/MWh-e to compete with natural gas SMR in the absence of a carbon tax), and/or a green hydrogen production credit or carbon tax on CO_2 emissions from natural gas SMR would be required for HTSE to be cost competitive with SMR. The following section goes beyond the basic steady state constant hydrogen production scenarios presented here and uses the HTSE model outputs developed here to run an optimization in order to further characterize the conditions under which profitability with LWR-HTSE hydrogen production can be obtained.

As mentioned in the introduction to this section, the design and analysis approach in this section represents a generic snapshot of the possible design configuration of integrating an HTSE hydrogen plant with an NPP. Other configurations are possible and could be analyzed in future work. These future iterations of designs for thermal power extraction from an NPP are being. Other design options not included in this report could include removing after the high-pressure turbine, eliminating the TDL to utilize NPP steam to directly heat treated HTSE feedwater, decreasing the distance between the steam extraction and the HTSE and condensate return to the first NPP feedwater heater versus to the condenser. Also, hydrogen storage may be located offsite. These alternate design options could prove to be less expensive and more efficient and will be the topic of future studies. One of these possible designs is shown in Figure 54.

Additionally, the design and control has not been optimized in this report but that would happen as the design matures towards commercialization.

Future work would also consider pulling power from the transmission grid at 345/161 kV in order to keep the hydrogen plant and the NPP generator decoupled and avoid having NPP perturbations affecting the hydrogen plant.

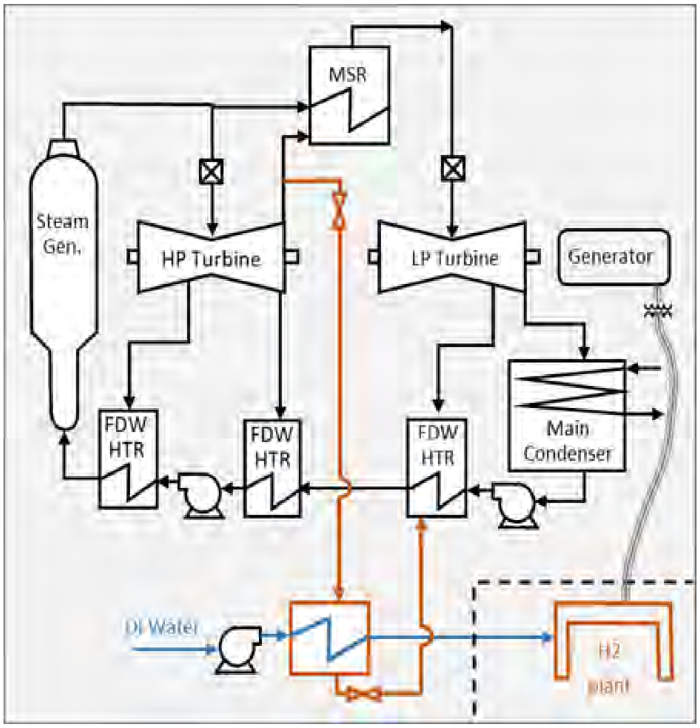

Figure 54. Alternate design configurations for thermal power extraction from NPP to HTSE. This design as marked in orange is not addressed in this report but could be the topic of future studies.

4 LWR-HTSE ECONOMIC DISPATCH AND OPTIMIZATION

The previous sections developed and characterized the HTSE design model, including process design conditions, power and utility requirements and capital and OPEX. The previous section also showed the results of a static steady-state constant hydrogen production sensitivity analysis. This section is a further and more detailed analysis which uses the HTSE model parameters developed in the previous section as well as the hydrogen demand curves for the Minnesota region developed in Section 2 as well as grid electricity locational marginal pricing (LMP) forecasts separately developed by NREL. This section develops the optimization of the NPP-HTSE plant results which show the envelope of profitability—i.e., the conditions under which the system can be profitable.

In the optimization model, parameters are varied within certain constraints, and the effect on the NPV is recorded. The system is economically dispatched by allowing the model to send NPP electricity to either the grid or the HTSE plant, depending on LMP forecasts and other parameters in order to maximize the NPV. NPP-HTSE NPV is calculated and compared to a business-as-usual (BAU) NPP NPV. The advantage of the ∆NPV approach is that the economics of the HTSE system can easily and quickly be compared to the BAU. Conveniently, only cash flows that change between the BAU NPP and NPP-HTSE cases need to be represented. It should be noted that this type of analysis only quantifies profitability relative to BAU, not the overall profitability of the NPP-HTSE.

The purpose of the economic dispatch and optimization portion of this project is to quantify the effects of economic parameters on the overall profitability of the nuclear with HTSE system. A model was developed to perform the yearly economic dispatch, track all the economic parameters, and explore several variables and assess their effect on NPV. The dispatcher looks at the LMP and decides to dispatch in one of three dispatch modes, 1) fill H_2 storage (charge mode), 2) deplete H_2 storage (discharge mode), 3) meet H_2 demand only (meet demand mode). The HTSE system is effectively acting as a demand-response system that can shift its load to hours of low electricity price and maximize electricity production in high-price hours. The current analysis assumes the NPP-HTSE plant is a price taker, meaning that the model does not change the price of electricity as the amount of HTSE load changes. Future work may involve modeling assuming price feedback in price maker scenarios.

Optimization variables that were allowed to be varied by the model include 1) HTSE capital cost (total capital investment), 2) HTSE capacity (hydrogen demand), and 3) possible PTCs for producing carbon-free hydrogen with NE.

In this analysis, it is assumed that the hydrogen demand set forth must be met in each hour of the year. This is a capacity contract scenario where hypothetically the amount of hydrogen to be produced and delivered has been agreed to ahead of time between the HTSE plant and the H_2 users. Another scenario not considered here would be an intermediate gas company fulfilling H_2 demand contracts and supplementing H_2 from the NPP-HTSE into the supply.

4.1 Optimization-Model Formulation

The dispatch and optimization model was developed using the Risk Analysis Virtual Environment (RAVEN) framework which has been developed at INL[a]. RAVEN is a multipurpose code for regression analysis, optimization, uncertainty quantification, and data analysis. This dispatch model uses RAVEN's conjugate gradient optimization, sensitivity analysis, and multi-level run features. The model also leverages an economic plugin called the TEAL (Tool for Economic AnaLysis) to track financial parameters throughout the lifetime of an HTSE facility. TEAL is a plugin to extend RAVEN's economic capabilities with the ability to track and discount cash flows, apply taxes or depreciation, and calculate economic parameters such as NPV or IRR.

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[a] For more information on RAVEN, see https://github.com/idaholab/raven/wiki.

The model operates in a two-loop configuration. The outer loop varies HTSE capital cost, HTSE demand, and a clean hydrogen credit. This allows the user to explore the intersection of these three variables and their effect on profitability. The model inner loop performs the economic dispatch of electricity vs hydrogen, optimizes hydrogen-storage parameters, and calculates the NPV. A schematic of the model architecture is shown in Figure 55. The optimization model is explained in more depth in subsequent sections.

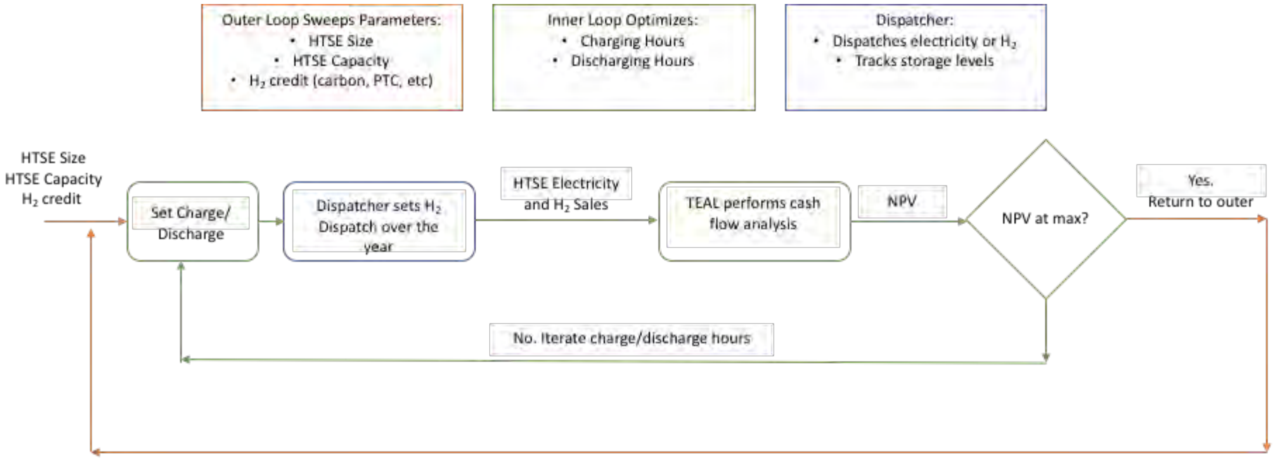

Figure 55. Economic dispatch and optimization model schematic.

4.1.1 Optimization Model Inputs

The optimization model inputs can be put into two parameters: physical and economic. The physical inputs come from the HTSE model developed in Section 0.details the physical inputs to the optimization model.

Table 27. Physical inputs to dispatch and optimization model.

Input Parameter	Value
Electricity Requirement	37.4 kWh-e/kg H_2
Thermal Requirement	6.4 kWh-t/kg H_2
Electrical Hot-Standby	0.9% of HTSE MW_e
Thermal Hot-Standby	3.2% of HTSE MW_{th}
Cell Degradation Factor	0.953
NPP Thermal Efficiency	0.346
NPP Capacity (PI)	1096 MW
NPP Capacity (Monticello)	671 MW

The HTSE economic inputs already discussed are given in Table 28.

Table 28. Economic inputs to dispatch and optimization model from Section 0.

Input Parameter	Value or Equation
HTSE CAPEX ($/kW) (Total Capital Investment)	Varied between 250 and 850 $/kW
HTSE Variable Operating and Maintenance Cost ($/MWh)	$5.20 * \text{HTSE\_Capacity}^{-0.004}$

HTSE Fixed Operating and Maintenance Cost ($/kW-yr)	75.51 * HTSE_Capacity$^{-0.208}$

Hydrogen-storage costs were taken from the DOE Hydrogen HFTO Record.[30] Figure 56 shows the cost of various hydrogen storage technologies. Underground pipe storage is assumed. The correlation for the underground pipe storage is shown in Equation 3.5.1 with the coefficients in Table 29.

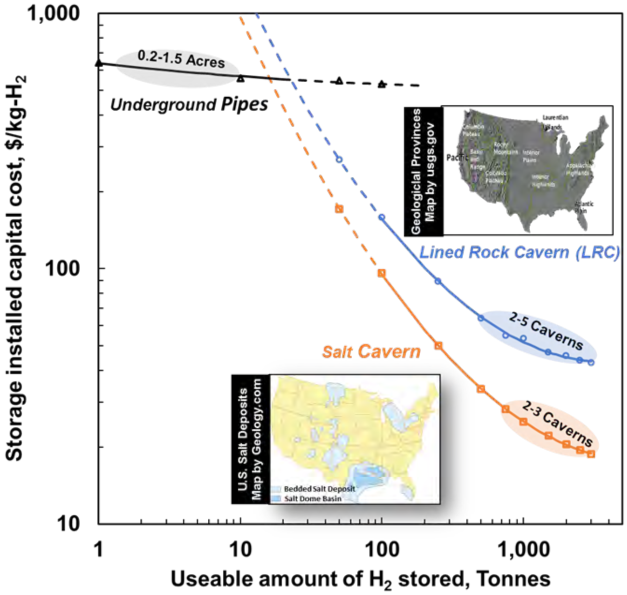

Figure 56. Hydrogen storage cost curves.

$$Storage\ Capex = \exp(a(\ln(m))^2 - b\ln(m) + c)$$

Table 29. Coefficients for storage capital cost.

Storage	a	b	c
Underground pipe storage	0.0041617	0.060369	6.4581
Underground lined rock caverns	0.095803	1.5868	10.332
Underground salt caverns	0.092548	1.6432	10.161

In addition to economic inputs, electricity- and hydrogen-market prices need to be provided to perform the dispatch between electricity and hydrogen. The electricity LMPs were output by several PLEXOS dispatch model runs by NREL. PLEXOS is a commercial dispatch software that can use an input set of electricity-generating units in a given region to find the LMP at various nodes in the optimization model, accounting for electricity-generator clearing price and transmission-congestion effects. The specific LMP profiles that are input into the optimization model represent three different model years, 2026, 2030, and 2034, at an hourly resolution. The LMPs represent wholesale electricity prices. There are different sets of LMP time histories for the two different plant locations.

The hydrogen market is quantified for each plant location by the demand curves from Figure 29 Figure 30 given in Section 2.2.4. The hydrogen sale price is correlated with demand based on the hydrogen users in the area and their distance from the specific NPP. Prairie Island has more hydrogen demand in the area, meaning that PI could sell hydrogen at a higher price relative to Monticello. The hydrogen markets are very shallow, meaning that if the demand reaches a certain point, the sale price drops dramatically due to hydrogen market saturation.

4.1.2 Dispatch Logic

Within the inner loop, the dispatch routine receives the hydrogen demand, HTSE capital costs (in $/MW), and clean-hydrogen credit from the outer loop, as well as the input physical and economic parameters discussed in Section 4.1.1. The inner loop also introduces the hydrogen storage charge and discharge hours to be optimized by RAVEN. Charge hours are the number of hours the HTSE expects to overproduce and send excess hydrogen to storage. Conversely, the discharge hours represent the amount of time expected that the HTSE will reduce its load and use storage to meet hydrogen demand. It is assumed that the hydrogen demand must be met in each hour of the year.

The dispatcher in the model first sets the physical sizes of H_2 storage and HTSE according to the inputs. The HTSE total capacity is the hydrogen demand plus any oversize that will be used to fill hydrogen storage during grid electricity off-peak hours. The total HTSE capacity is constrained by the capacity of the nuclear plant with which it is associated. The oversize excess capacity corresponds to the ratio of storage charge to discharge hours. The HTSE must meet the amount of hydrogen required to meet demand while also filling the hydrogen storage. Additionally, the storage size is set by the discharge hours, where the system will need to have the storage capacity to meet demand in the all the discharge hours. An adder of 4 hours is used in the dispatch model to increase the storage margin in case of multiple discharge events in a row. Equations for each of these calculated physical parameters are given below, where D_{H2} is the H_2 demand passed into the dispatch by the outer loop.

$$Cap_{HTSE} = D_{H2}(MW) + D_{H2}(MW) * \frac{hrs_{charge}}{hrs_{discharge}}$$

$$Oversize = D_{H2}(MW) * \frac{hrs_{charge}}{hrs_{discharge}}$$

$$Cap_{storage} = D_{h2}(kg) * (hrs_{discharge} + hrs_{margin_{adder}})$$

The dispatcher looks at the LMP and decides to dispatch in one of three dispatch modes, 1) fill H_2 storage (charge mode), 2) deplete H_2 storage (discharge mode), 3) meet H_2 demand only (meet demand mode). When the LMP is low, the dispatcher will choose charge mode and produce more hydrogen than is required to meet demand, prioritizing producing and storing hydrogen over sending electricity to the grid. When the LMPs are high, the dispatcher will operate in discharge mode, using hydrogen from storage to meet demand and maximizing the amount of electricity sold to the grid. In meet demand mode, the HTSE only produces the amount of hydrogen required to meet hydrogen demand, selling the remaining electricity from the NPP to the grid. This means that the HTSE system is effectively acting as a demand-response system that can shift its load to hours of low electricity price and maximize electricity production in high-price hours.

To decide when the charge and discharge events occur, the dispatcher searches LMPs over a 24-hour period looking for the minimum and maximum. The dispatcher sets a block of hours surrounding the maximum LMP equal to the number of discharge hours. The dispatcher then looks to set a block around the minimum LMP equal to the number of charge hours. The charge is allowed to happen in multiple blocks of time, but the discharge block must be contiguous. This means that in a 4-hour discharge scenario, the dispatcher finds the highest LMP hours and backfills in another 3 hours around it, for a total of four. The charge hours then use the same process around the lowest LMP. It is possible for the discharge event to happen before or after the charge event. In the case in which sufficient storage is not available to meet hydrogen demand, the dispatcher prioritizes meeting hydrogen demand and reduces the amount of electricity sold to the grid. This scheme effectively works as peak-shaving arbitrage.

Figure 57 shows this dispatch strategy over a 4-day period. Note that when the LMP is high, the storage flows are negative, meaning hydrogen is leaving the storage tank to be sold. At that same time, hydrogen generation goes to zero and the electricity sold to the grid is maximized. Surrounding the times of low electricity price, the HTSE overproduces hydrogen and sends the excess to storage. The duration of these charge and discharge periods is optimized by the model and made consistent throughout the year.

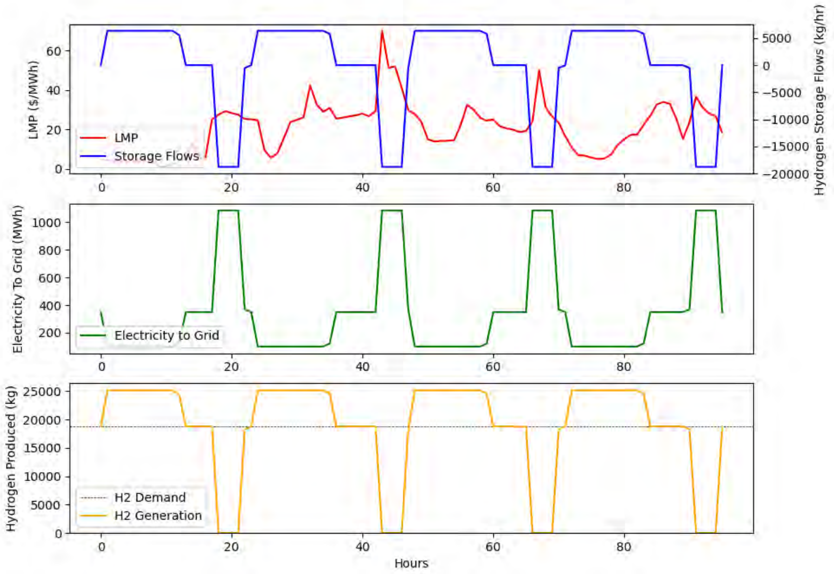

Figure 57. Demonstration of model dispatch logic over a 4-day period.

This approach assumes the NPP-HTSE plant is a price taker, meaning that the model does not change the price of electricity as the amount of HTSE load changes. Future work may involve modeling assuming price feedback in price maker scenarios. The LMPs were generated with the nuclear plant in a standard operating mode without the HTSE. In reality, the load of the HTSE, and thus the reduction in capacity that the nuclear plant has available to bid into the day-ahead electricity market, could change the clearing price of electricity, i.e. become a price maker. This price taker assumption is generally less error prone in changing generators that are small in comparison to all generators that bid.

4.1.3 Economic Inputs

With the yearly dispatch for each of the three input years (2026, 2030, 2034), the hourly electricity production profile, hourly hydrogen-production profile, and the equipment costs are used to compare NPP-HTSE NPV to a BAU nuclear NPV. This comparison provides a ΔNPV value that can be used to assess profitability. A positive ΔNPV means that the NPP-HTSE combined system is more profitable relative to running the NPP as usual. A negative ΔNPV means that the NPP + HTSE makes less money than the NPP running in standard operation. An ΔNPV equal to 0 is the breakeven point at which the operation of the NPP-HTSE is roughly equivalent in value to the BAU case.

$$\Delta NPV = NPV_{NPP+HTSE} - NPV_{BAU}$$

The advantage of the ΔNPV approach is that the economics of the HTSE system can easily and quickly be compared to the BAU. Conveniently, only cash flows that change between the BAU NPP and NPP-HTSE cases need to be represented. Nuclear plant cash flows that do not change between cases, such as nuclear fixed-O&M or capital expenditures, do not need to be quantified. This simplifies the simulation and reduces the uncertainty that would be present in a standard, absolute NPV analysis. It should be noted that this type of analysis only quantifies profitability relative to BAU, not the overall profitability of the NPP-HTSE.

The cash-flow parameters in Table 30 are fed to the RAVEN plugin TEAL to calculate the NPV. Hydrogen sales are calculated according to the amount delivered and the demand curve for the corresponding plant. The electricity sales are calculated by multiplying the amount of electricity to the grid in each hour by the respective LMP.

Table 30. Cash flows for ΔNPV calculation.

NPV w/ Nuclear + HTSE	NPV w/ Nuclear BAU
HTSE Capital Cost	Electricity Sales (2026, 2030, 2034)
HTSE Fixed Operating and Maintenance	Extra Capacity Payment
HTSE Variable Operating and Maintenance	
Hydrogen Storage CAPEX	
Electricity Sales (2026, 2030, 2034)	
Hydrogen Sales (2026, 2030, 2034)	

An extra capacity payment is accounted for in BAU NPV. This payment is meant to quantify the amount of capacity lost to HTSE hot standby. Regardless of hydrogen-production level, some hot-standby amount will decrease the capacity of the nuclear plant. This lost capacity will require a different generator to replace the capacity.

To quantify the value of lost capacity, several capacity-expansion cases were run to find system cost for various scenarios.[4] These scenarios varied the level of nuclear capacity lost and let the resource-planning model replace capacity. Scenarios where nuclear capacity was replaced, and carbon goals were met, were fit to a curve in order to back out the approximate cost per MW-yr of replacing capacity. This number is higher than the cost of new entrant (CONE) because it requires low- or zero-carbon replacements rather than a simple gas-turbine replacement. As an example, the range of HTSE capacities for Prairie Island generally requires less than 20 MW hot standby.

The financial parameters are shown in Table 31. The project life is 25 years, starting in 2026. The 2026 LMPs are applied for the first 4 years, 2030 for the next 4 years, and the remaining years use the 2034 LMPs.

Table 31. Financial parameters.

Parameter	Value
WACC	7.09%
Federal Tax Rate	21.0%
State Tax Rate	9.8%
Tax Deduction	-2.06%
Capital Depreciation	MACRS, 15yrs
Project Lifetime	25 years

4.1.4 Optimization Model Assumptions

Several key assumptions that went into this optimization model are itemized here.

- **Storage arbitrage operation**: The hydrogen storage operates in a daily arbitrage fashion. The model dispatcher finds the highest LMP in the day and depletes hydrogen storage (discharges) according to the number of determined discharge hours. Conversely, the model dispatcher finds the lowest LMP in the day and fills hydrogen storage (charges) during those hours according to the charge duration. This effectively shifts the load from high-electricity-price times to low-price times so that electricity sale during high-price hours can be maximized.

- **Price-taker dispatch**: The dispatch model operates in a price-taker fashion, assuming that the change in electricity delivery due to HTSE load does not change electricity prices. Future work will be ongoing to quantify the effect of this assumption and extend the model to make it more consonant with the price-maker approach.
- **NPP refueling**: NPP refueling assumes that each plant operates at 50% capacity during the refueling period (3 weeks, exact time dependent on the NPP). During refueling, the HTSE uses as much energy from the NPP as it can and buys the remaining electricity from the grid. Additional capital cost is associated with the extra resistance heaters needed to make this possible. The arbitrage is not performed during refueling, meaning the HTSE meets hydrogen demand but does not produce excess for storage or shift load to storage. The storage amount in the refueling period is held static. Refueling is also accounted for in the BAU case.
- **Wholesale electricity prices**: The dispatch is performed with wholesale electricity price LMPs. Any buying or selling of electricity is based on these wholesale prices.
- **Model dispatch prioritizes hydrogen**: The model dispatcher prioritizes meeting hydrogen demand over selling electricity. This is to simulate something like a hydrogen purchase agreement or a consumer who uses the hydrogen for chemical or industrial processes that needs hydrogen at all times.
- **Hydrogen demand is a constant hourly amount through lifetime**.
- **Hydrogen price is a single price, based on demand:** The hydrogen price is founded on the demand curves developed in Section 2, based on the input hydrogen demand parameter. Those demand curves include a \$22/ton CO_2 credit already baked into the hydrogen price per the Xcel Energy IRP planning for a CO_2 credit in this amount. The demand sets the price, and that price remains constant throughout the optimization lifetime. Sensitivities on hydrogen market selling price can be inferred by studying the clean hydrogen credit sensitivities.
- **Clean-hydrogen credit:** A clean-hydrogen credit is applied as a sensitivity. In the model, this is applied as an adder to the sale price. In reality, this would have the same effect as a PTC or a carbon tax on competitor hydrogen that raises hydrogen prices in the market. Sensitivities on hydrogen market selling price can be inferred by studying the clean hydrogen credit sensitivities.
- **Degradation as a capital cost adder:** The year-over-year electrolysis-cell degradation was found to effectively reduce the capacity of the hydrogen production facility to 95.3% of the starting value at the beginning of the year. The O&M costs account for the replacement year over year. This degradation was accounted for in the optimization model by scaling the capacity up 4.9% (HTSE Capacity/0.953) in the cost and O&M calculations. The capacity available for producing hydrogen is fixed at the degraded value, rather than quantifying the complexity of degradation effects from hour to hour. This has the effect of producing a more-conservative estimate because the capacity in the beginning of the year would be able to produce slightly more than this model accounts for.
- **Capacity payment and/or replacement covers changes from HTSE hot standby only:** It is assumed that the rated capacity of the nuclear plant is only decremented by the constant hot-standby amount. The plant operates in a different mode from a baseline BAU case, but the capacity (minus HTSE hot standby) is still available, especially in high-price hours when it draws on the storage. This is semi-idealized because there may be instances that storage is insufficient, and hydrogen is

prioritized, causing the NPP to miss electricity sale in high-price, high-demand events. Further work should be done to provide insights on this issue.

- **No down time for HTSE cells**: Downtime for HTSE cell replacement or other maintenance is not included in the hourly dispatch.

4.2 Dispatch Optimization Results

The optimization model facilitated the exploration of several parameters to understand their effects on profitability when comparing the NPP-HTSE with the BAU case. The storage-dispatch hours were optimized in the inner loop to maximize ΔNPV. The hydrogen demand, hydrogen CAPEX, and clean-hydrogen credit were varied in the outer loop to understand the area where the system is profitable or the 'envelope of profitability'.

The outer loop used an advanced sampling technique called the limit-surface search, which is available in RAVEN. This technique zeros in on areas where the ΔNPV changes from negative to positive. As it does, the model bounds the actual points where the transition from unprofitable to profitable occurs. This means that a projection of the profitable region of this surface can quickly show what combination of the three variables—HTSE CAPEX, hydrogen demand (HTSE capacity), and green-hydrogen credit—will yield a profitable system. This envelope of profitability will show the tradeoff between these three variables in a profitable system. This is valuable for the decision-maker to study these charts and to analyze hypothetical scenarios where these variables will change and be able to infer the rough ΔNPV under those scenarios.

4.2.1 Prairie Island

4.2.1.1 Storage Dynamics and Size

First, it is important to understand the underlying optimization that occurs under each combination of capacity, CAPEX, and clean hydrogen credit. The hydrogen storage charge and discharge hours are optimized by RAVEN in each run. The hydrogen discharge hours effectively set the storage size and the ratio of discharge to charge sets the HTSE oversize that, with hydrogen demand, sets overall HTSE capacity.

The plots show the effect of HTSE capital cost on both charge and discharge hours. The ΔNPV was maximized at 4 discharge hours in the lower HTSE CAPEX case and 2 hours in the high. The system trades higher HTSE capacity and storage capital costs for selling more electricity at high-price hours. The high CAPEX plot shows that the system prefers to miss high electricity price times and minimize HTSE size when the HTSE capital cost is high. Lower capital cost creates more value in storage arbitrage; thus, building a slightly larger HTSE to facilitate a larger storage capacity is advantageous. This is true to a point, seen after the ΔNPV peaks and starts to drop, for example, in the 5th and 6th discharge hours in the lower-CAPEX case. The value of arbitrage is diminished with each additional discharge hour because each subsequent hour will have a lower LMP than the previously captured one. This means that selling electricity in that time is not worth the increase in HTSE size and storage size.

While the charge hours have a peak, the optimization shows that it is advantageous to maximize discharge hours. In Figure 58, each line corresponds to a different number of charge hours. In both capital-cost cases, the ΔNPV is maximized with the largest number of charge hours. This dynamic is observed because charge hours are inversely proportional to HTSE capacity oversize. If there are more hours to overproduce and store H_2, then the requirement for discharge can be met with less capacity. Minimizing HTSE capacity was advantageous to system economics because it is such a large driver of cost.

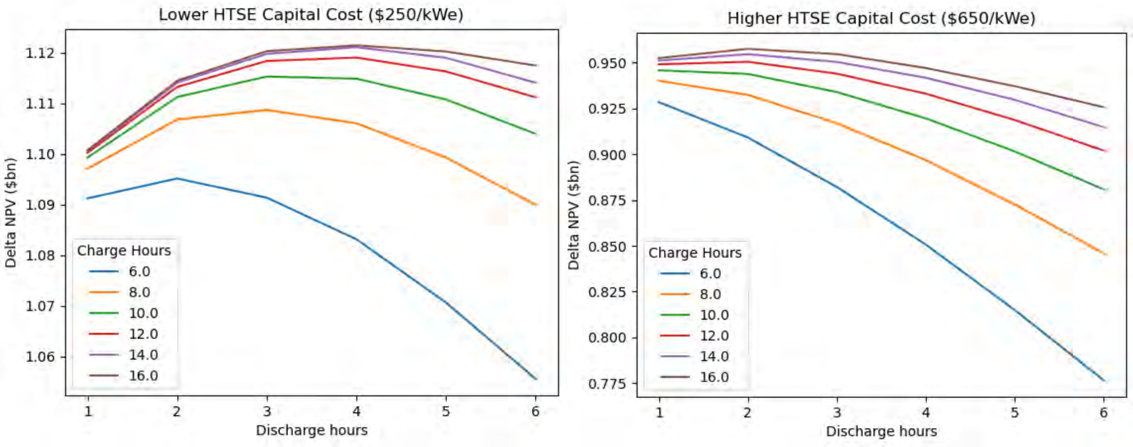

Figure 58. Effect of HTSE capital cost on storage charge and discharge hours on ΔNPV for the project 25-year lifetime at Prairie Island.

The plots in Figure 58 were run with two different HTSE capital costs, but with the same green-hydrogen credit and the same hydrogen demand. The green credit does not affect storage size because the hydrogen sale is dictated by hydrogen demand, not arbitrage. The dispatch prioritizes hydrogen, so the same amount is sold in cases with different hydrogen credits, regardless of arbitrage. Similarly, different hydrogen demand does not have an effect on the storage because hydrogen is prioritized and sold at the same rate regardless of arbitrage.

This effect of capital cost on storage charge and discharge lengths could create a tradeoff if there are so few discharge hours that the nuclear capacity requires more replacement capacity. That effect is not quantified in this analysis but should be considered when deciding on HTSE operational modes. In some cases, there may be enough margin to build larger storage systems and flex the HTSE at a non-optimal configuration to avoid extra replacement capacity. This ultimately will depend on the operational mode, the HTSE CAPEX provided by manufacturers, and how much extra capacity might need to be built elsewhere in the system.

4.2.1.2 Demand, CAPEX, and Clean Credit

Understanding the relationship between HTSE CAPEX, hydrogen demand, and clean-hydrogen sale credits are important to understanding when the PI NPP-HTSE system is profitable compared to BAU. Figure 59 shows four different static cases for a hypothetical HTSE plant installed at PI. These four cases help illustrate several important effects. Note that for each combination of the three variables, storage charge, and discharge hours are optimized in the inner loop.

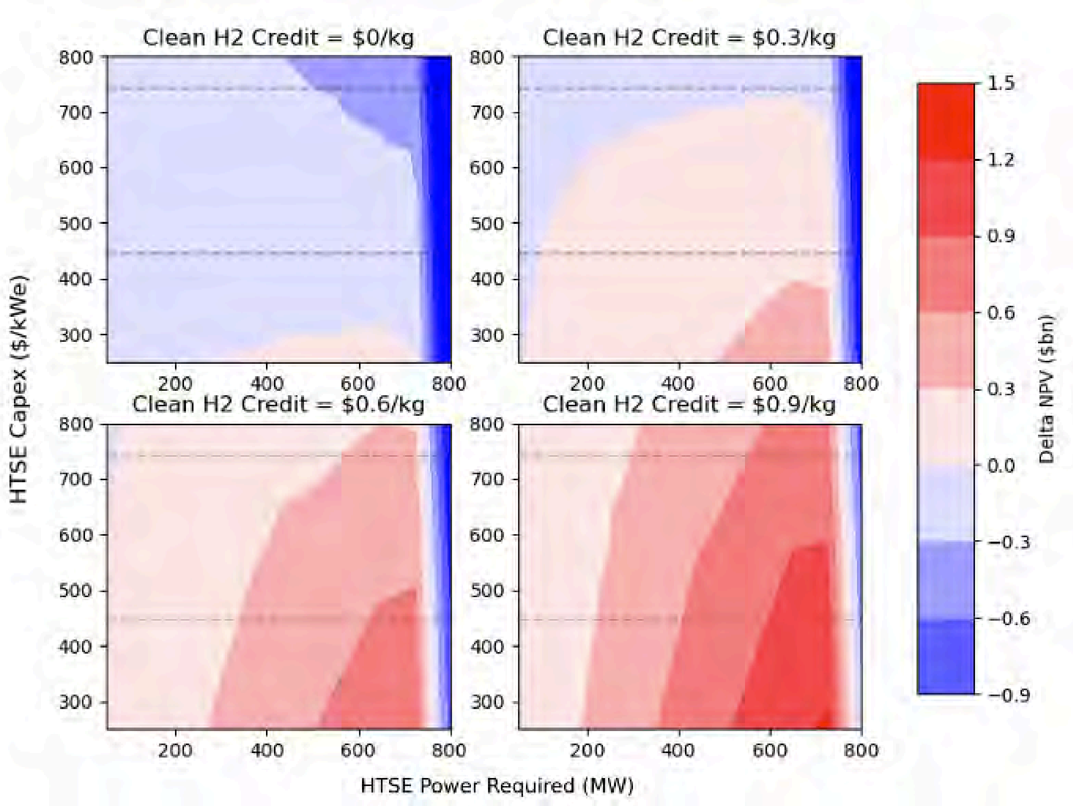

Figure 59. HTSE CAPEX (total capital investment), H_2 Demand, carbon-free hydrogen credit and their effect on ΔNPV for the NPP-HTSE plant versus BAU at Prairie Island. For reference, using the full two reactors of output from PI could produce up to 29,290 kg/hr (703 tonne/day) of H_2 and a single 545 MW reactor could produce up to 14,570 kg/hr (350 tonne/day) of H_2. The horizontal dashed lines show the placement of the base and advanced case HTSE CAPEX corresponding to the high and low scenarios previously developed.

Adding an HTSE to PI yields a profitable system in a small region at a very low capital cost. This region is where the H_2 demand is maximized, but the market is not saturated. In the hydrogen-demand curve for the region surrounding PI, a drop off occurs when hydrogen supply saturates the demand market as the demand market goes to the next lowest price tier, pushing the price of hydrogen from $1.14/kg to $1.10/kg. At $1.10/kg, there is insufficient profit from hydrogen sales to justify larger HTSE facilities. At even higher hydrogen demands, the H_2 is sold for even less, driving ΔNPV even lower at higher demand values. The lower sale price makes the system unable to recover the capital and other costs, meaning the system is less profitable than BAU. This effect can still be seen in cases with clean hydrogen credits.

By adding a clean hydrogen credit, the system becomes more profitable with higher CAPEX values as shown in Figure 59. With credits of $0.70/kg, the system is more profitable than BAU for every tested CAPEX before the market is saturated. The optimal ΔNPVs are still found in the region where the hydrogen demand maximizes the amount sold at the $1.32/kg base-hydrogen price. It is conceivable that small hydrogen credits below $0.50/kg would flip a large portion of possible HTSE facilities at PI from not profitable to profitable relative to BAU.

Figure 59 also demonstrates that the HTSE at PI will be limited by market size rather than NPP capacity. The shallow H_2 market causes a significant drop off in profitability before the HTSE is large enough to take all the energy that PI can provide (recall this is up to 29,290 kg/hr of hydrogen), but as mentioned before, the price of hydrogen drops significantly after supply exceeds 17,600 kg/hr. This means that there is still capacity at PI to provide electricity to the grid 100% of the time. At a hydrogen-production rate of 17,600 kg/hr, PI still has approximately 340 MW to send to the grid on a consistent basis, depending on storage capacity and extra HTSE capacity used for flexing. Keep in mind that these results apply to existing H_2 demand as presented in the assumptions in Section 2 of this report. If larger amounts of carbon free H_2 market demand materialize then the assumptions of the H_2 market and thus this analysis would change.

4.2.1.3 Envelope of Profitability

The previous section gives an example of the economic implications of three varied parameters, but the coarseness of sampling does not quantify what exact combinations would make the system profitable. To remedy this coarseness in sampling, RAVEN's limit-surface search capability was used. The limited-surface search creates boundaries around the actual points where the ΔNPV changes from positive to negative. The search is shown in Figure 60. The plot is a three-dimensional (3D) figure in which the varied parameters exist, each on its own axis. The red dots represent a negative ΔNPV, and the green represents a positive ΔNPV. The limited-surface sampling strategy explored the space and honed in on the transitionary spots at which the sign of ΔNPV changes. This sign change means the system went from not profitable (negative) to profitable (positive) compared to BAU.

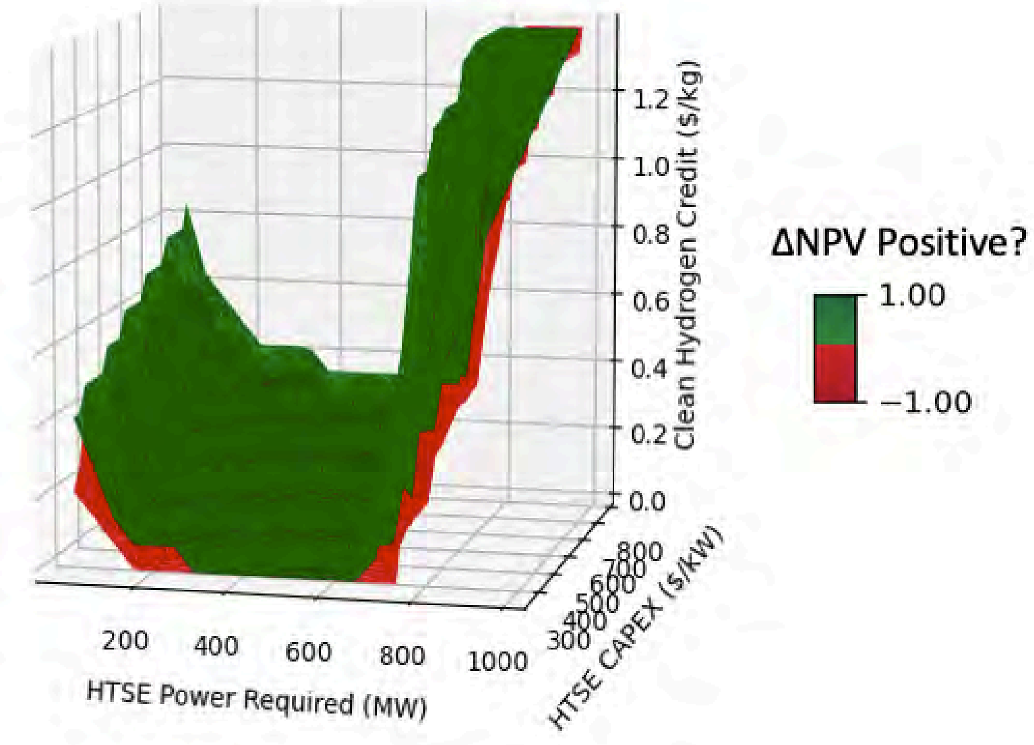

Figure 60. Limited-surface search exploring H_2 delivered, HTSE CAPEX (total capital investment), and clean-hydrogen credits at Prairie Island. The green surface represents a positive ΔNPV relative to BAU and the red surface represents a negative ΔNPV relative to BAU.

While the 3D visualization is useful for seeing the shape of the profitability transition region, it is difficult to understand the detailed interaction between the three degrees of freedom. To visualize this, the green surface can be projected onto a two-dimensional (2D) plane, as shown in Figure 61. This surface represents the boundary of profitable configurations. Representing the limit surface in this manner allows for a visualization of the relationship between the sampled variables. A reader can quickly see what is required for a profitable NPP- HTSE system relative to BAU.

Observing one or two parameters in Figure 61 gives an indication as to what the remaining degrees of freedom should be. For example, if a manufacturer is able to build the HTSE system at $500/kW total capital investment, and the hydrogen demand that Prairie Island expects to meet is 10,000 kg, then a modest production tax or other clean credit of ~$0.05/kg or greater is required to be profitable when compared to nuclear plant BAU. Because Figure 61 is the projection of a profitable limit surface, every combination of points on this plot is a break-even point. Any improvement on these degrees of freedom, such as lower than expected capital cost at the fixed demand and credit price, will improve the overall ΔNPV. For comparison purposes, Figure 61 also displays the two HTSE CAPEX scenarios (base case and advanced case) considered in the HTSE model development in Section 3. The high value, $742/kW, represents the public HFTO Record value (base case). The low value, $446/kW, is a near term forecasted capability based on an aggregate of publicly available data from HTSE stack manufacturers (advanced case).

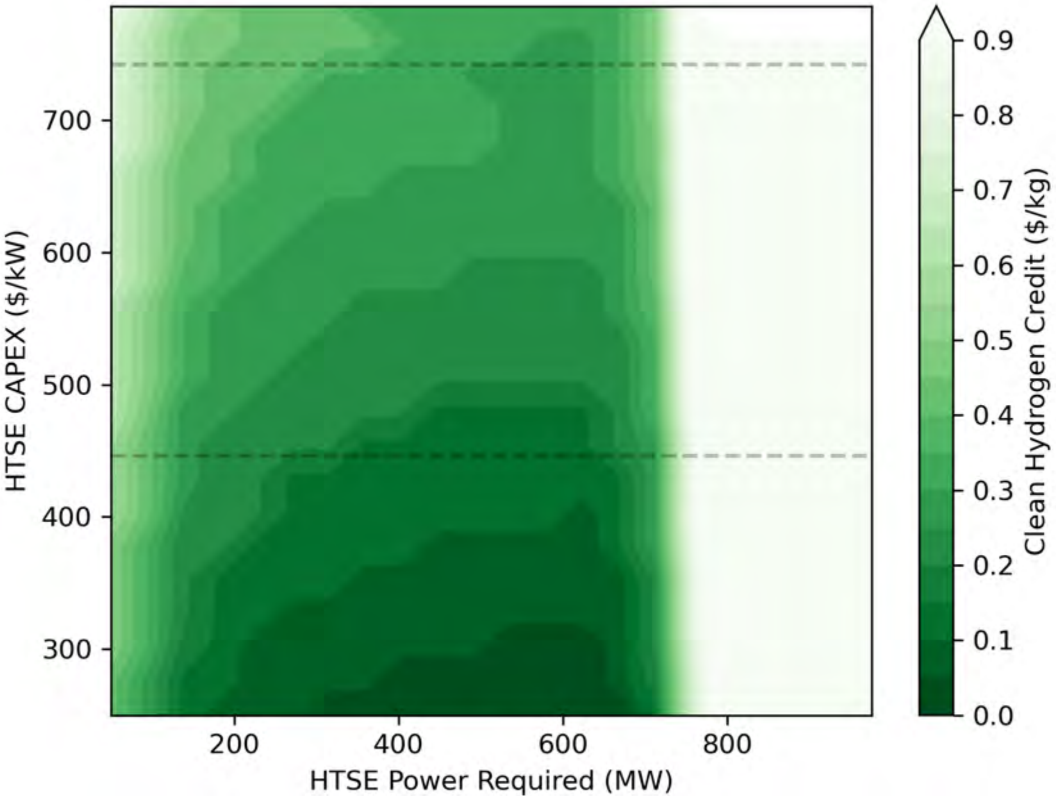

Figure 61. Profitable limit surface of HTSE CAPEX (total capital investment), hydrogen demand, and clean-hydrogen credit at Prairie Island. For reference, the maximum energy that PI could provide to an HTSE could produce up to 29,290 kg/hr (703 tonne/day). A single 545 MW reactor could produce up to 14,570 kg/hr (350 tonne/day). The horizontal dashed lines show the placement of the base and advanced case HTSE CAPEX corresponding to the high and low scenarios developed in Section 3.

These results show that an HTSE at PI could be profitable versus BAU without any hydrogen credits at lower HTSE total capital-cost values. The dynamic where maximizing demand without saturating the market is shown in Figure 61, similar to what was discussed in the previous section. Once the HTSE provides more than 17,600 kg/hr (422 tonne/day), the price of hydrogen drops, and hydrogen credits are required to bring the system to profitability. This result suggests that new hydrogen demand development in excess of all existing possible hydrogen demand would have to be created in order to use the full capacity of PI to produce hydrogen. Note also that this analysis assumes that PI could fill all existing demand when in reality this demand will be shared by other sources. Hydrogen credits are also required to make the system profitable at total HTSE system capital costs over $500/kW at the largest hydrogen delivery before saturation. It is possible to be competitive to BAU at lower hydrogen-delivery rates without any hydrogen credit, but lower capital costs are required. Clean credits of $1/kg or less are required in each configuration before the market saturation occurs to make the system profitable compared to BAU. Larger clean hydrogen credits are required if the hydrogen demand requires over 740 MW and the hydrogen price decreases.

4.2.2 Monticello

4.2.2.1 Storage Dynamics and Size

Monticello produces storage sizing dynamics similar to PI's. Figure 62 shows that lower HTSE capital costs facilitate more discharge hours. Lower capital costs would lead to more hydrogen storage capacity and a larger HTSE to overproduce and store hydrogen. The hydrogen production amount and clean-hydrogen credit were fixed for both of these plots. Much like Prairie Island, lower capital costs lead to hydrogen storage-tank sizes capable of meeting the demand for 4 hours while higher capital-cost cases point to hydrogen storage capacity in the range of 2–3 hours of demand as more desirable.

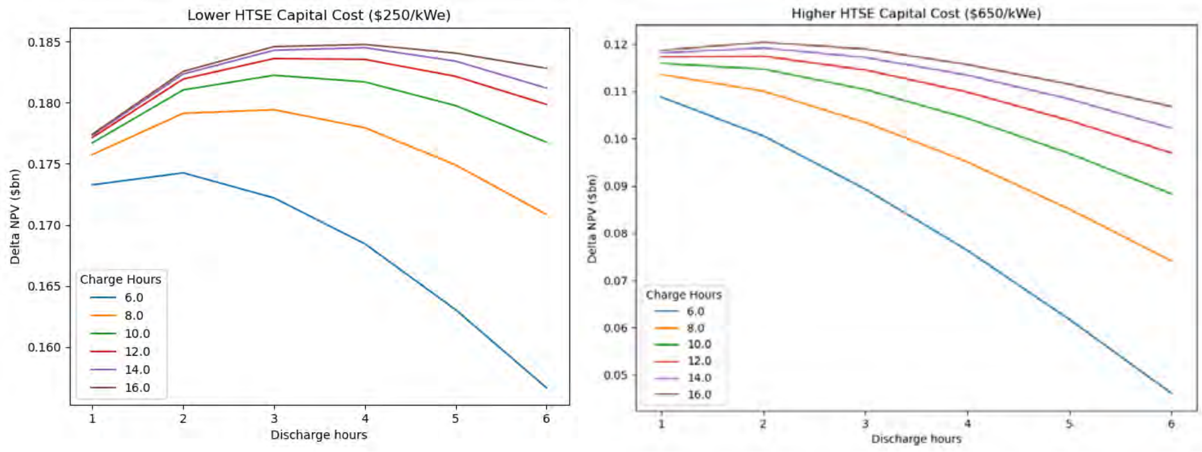

Figure 62. Effect of HTSE capital cost on storage charge and discharge hours on ΔNPV for the project's 25-year lifetime at Monticello.

4.2.2.2 Demand, CAPEX, and Clean Credit

Similar to PI, the hydrogen delivery, CAPEX, and clean hydrogen credit were varied for the Monticello NPP. These three parameters and their effect on ΔNPV is shown in Figure 63. Monticello has a much-smaller potential hydrogen market, which means that profitability is only possible in much-smaller HTSE sizes.

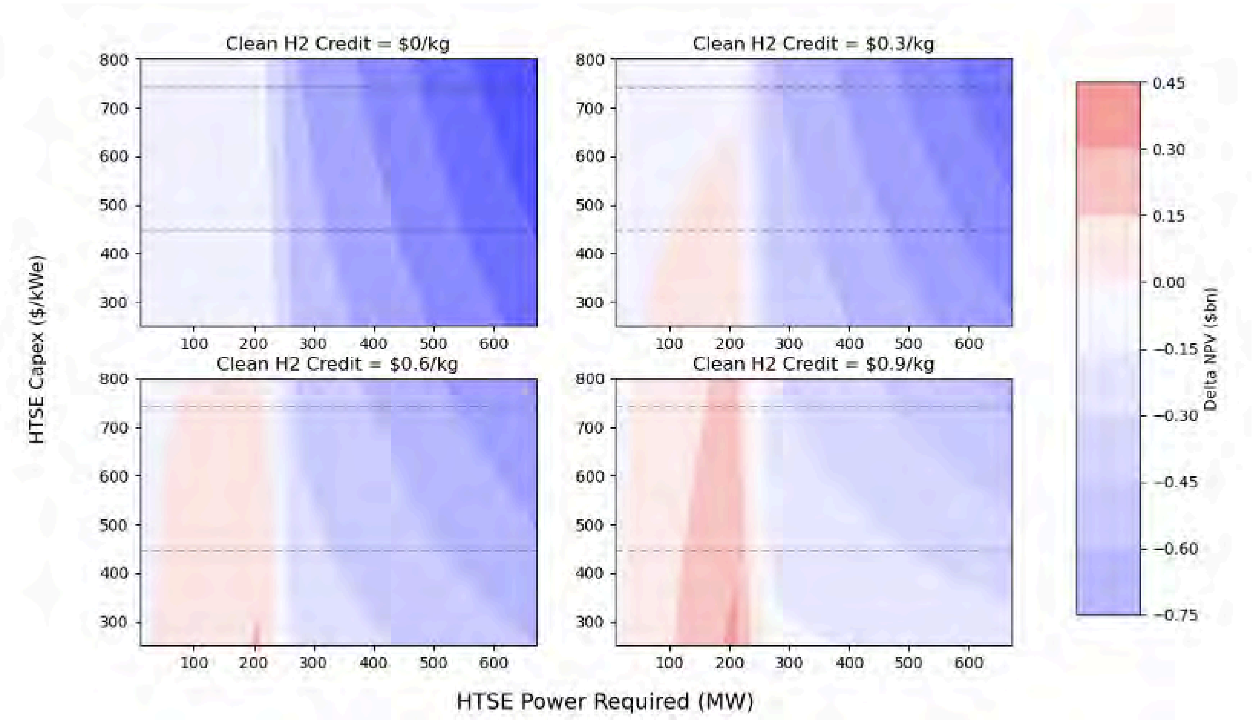

Figure 63. HTSE CAPEX (total capital investment), H_2 demand, clean-hydrogen credit and their effect on ΔNPV versus BAU at Monticello. For reference, the maximum energy that Monticello could provide to an HTSE could produce up to 17,930 kg/hr (430 tonne/day). The horizontal dashed lines show the placement of the base and advanced case HTSE CAPEX corresponding to the high and low scenarios developed in Section 3.

No combination of CAPEX and hydrogen production amount made the Monticello plant profitable versus BAU without a PTC or clean hydrogen credit. The system started to see profitability at low CAPEX and optimal demand with a credit of $0.30/kg. This differed from PI, where a small range of CAPEX and demands were profitable without any PTC due to the higher demand with more room before saturation for the area surrounding PI.

4.2.2.3 Envelope of Profitability

The envelope of profitability for the Monticello NPP was developed with the same limit surface search approach that was used in the Prairie Island analysis. Figure 64 shows the limit surface for the Monticello NPP produced by varying the amount of hydrogen delivered, HTSE CAPEX (total capital investment), and clean-hydrogen credit. Adding an HTSE has a smaller band of profitability at Monticello versus BAU than at Prairie Island. The analysis shows that some type of PTC or clean credit is required to make Monticello profitable at all demand and CAPEX points investigated. This is due to the smaller hydrogen market and a slightly lower hydrogen sale price.

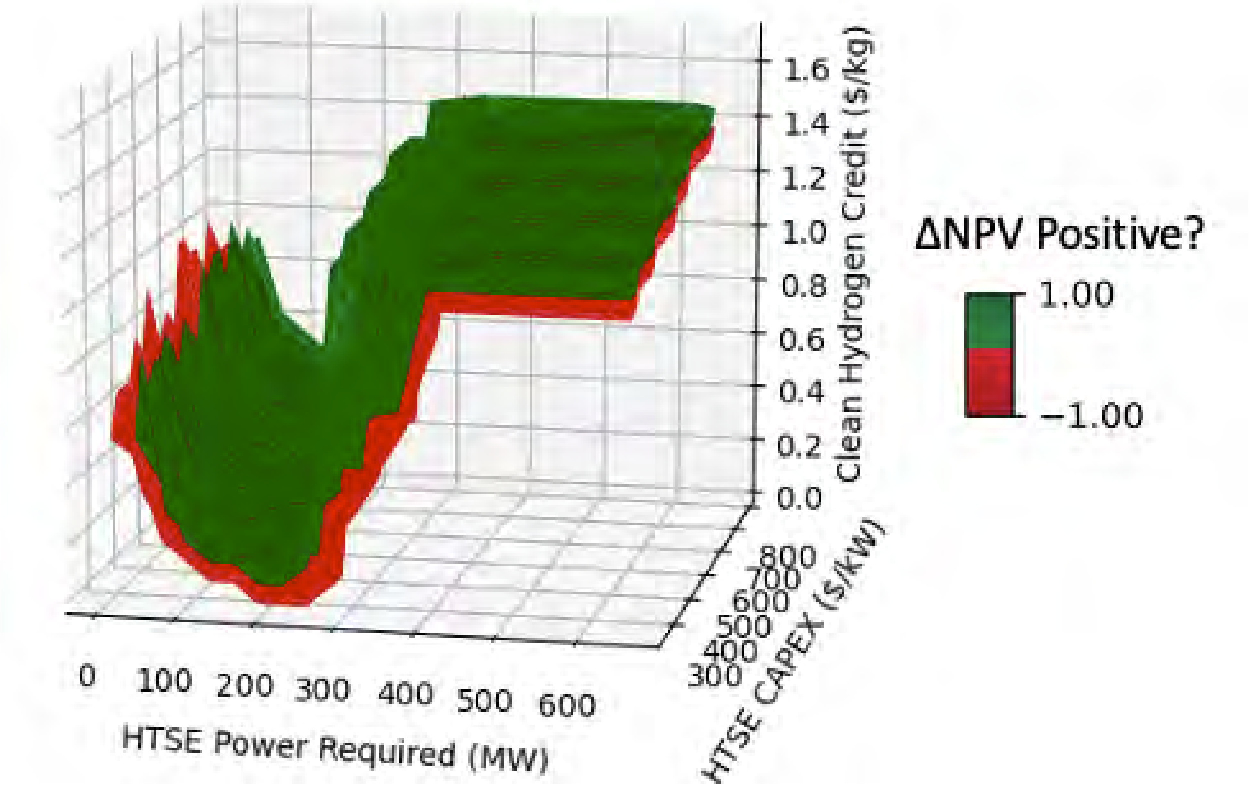

Figure 64. Limit surface search exploring H_2 delivered, HTSE CAPEX (total capital investment), and clean-hydrogen credit at Monticello. The green surface represents a positive ΔNPV relative to BAU, and the red surface represents a negative ΔNPV relative to BAU.

As was done for PI, the profitable surface was extracted from Figure 64 and plotted on Figure 65, which shows the combinations of credit, total HTSE system CAPEX, and hydrogen delivery amount that make the system break even relative to BAU. For comparison purposes, Figure 65 also displays the two HTSE total CAPEX values used in the constant hydrogen production sensitivity analysis in Section 3. The high base case value of \$742/kW represents the public HFTO Record value. The low advanced case value of \$446/kW is an aggregate of publicly available data from HTSE stack manufacturers. The white space represents no break-even configurations for the combination of parameters investigated. This plot shows that even at hydrogen-delivery amounts prior to market, a PTC of between \$0.07/kg and \$0.50/kg is required for positive ΔNPV versus BAU.

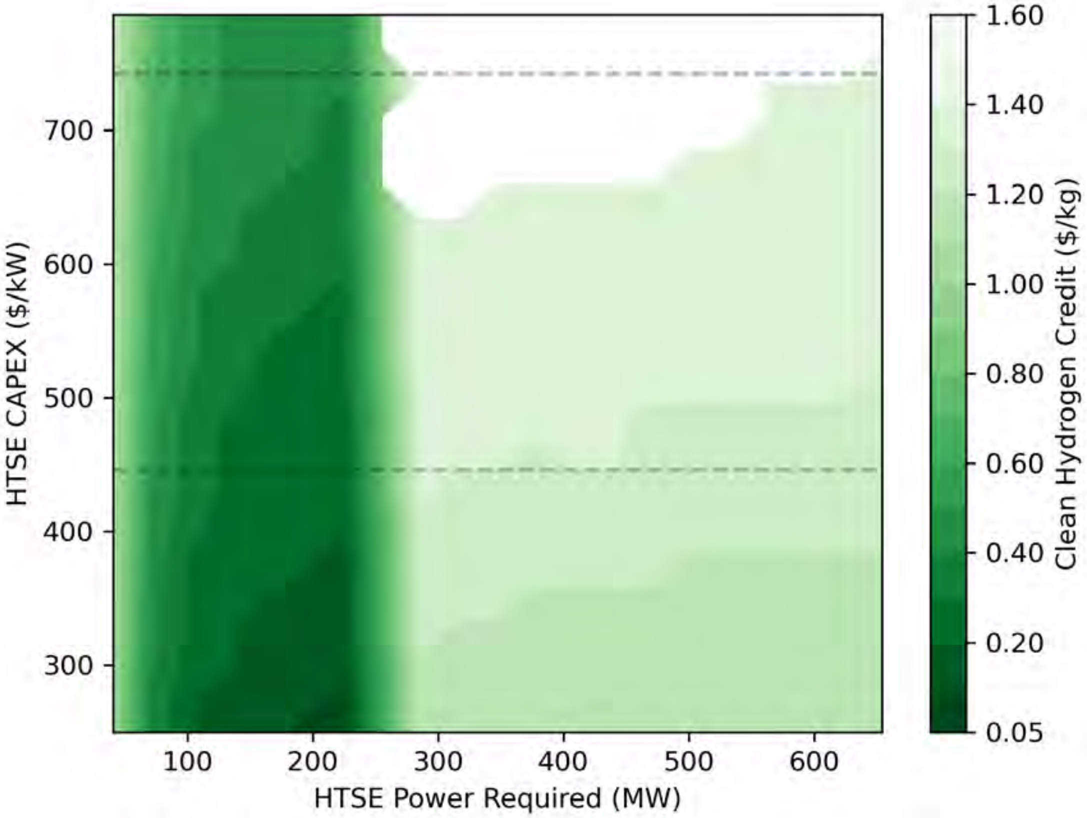

Figure 65. Profitable limit surface versus BAU of HTSE CAPEX (total capital investment), hydrogen demand, and clean hydrogen credit at Monticello. For reference, the maximum energy that Monticello could provide to an HTSE could produce up to 17,930 kg/hr (430 tonne/day). The horizontal dashed lines show the placement of the base and advanced case HTSE CAPEX corresponding to the high and low scenarios developed in Section 3.

4.2.3 Discussion of HTSE Optimization Results

The results of the economic and dispatch-optimization modeling show that a hybrid nuclear and HTSE approach can be profitable versus BAU at Prairie Island and Monticello NPPs. The tradeoff between the amount of hydrogen delivered, HTSE CAPEX, and any clean-hydrogen credits is important in each instance to understand what sizes and economic parameters are needed to be profitable over BAU.

HTSE profitability is highly dependent on the local hydrogen market. The optimal hydrogen delivery was to commit as much hydrogen as possible per hour without saturating the market and reducing the sale price of hydrogen. The hydrogen market surrounding Prairie Island was found to be larger than that around Monticello, so hydrogen delivery amounts and, thus, HTSE sizes could be larger at PI than Monticello. Monticello optimal hydrogen-delivery amounts were nearly three times less than PI on a kg/hr basis. This hydrogen market result could be further investigated by looking at hydrogen market sensitivities or understanding the hydrogen sale contracts into which the hybrid nuclear and hydrogen system might enter. This simulation could be extended into the future to include more hydrogen-market and demand projections throughout the 25-year life of the plant, rather than using a static hydrogen curve. That would allow the model to capture increases in hydrogen demand from new users that may come online in the 2030s.

In addition to dependance on the hydrogen market, system profitability is driven by the total sale price of hydrogen. The plants general required some type of adder above what the hydrogen could be sold for in the market. This adder could come in the form of a PTC, a carbon tax on competitors, or simply an end user willing to pay more for carbon-free hydrogen. Adder amounts ranged from \$0-\$0.9/kg at Prairie Island and \$0.07-\$1.6/kg at Monticello. For context, the Wyden energy tax credit bill that cleared the senate finance committee seeks to implement up to \$3/kg in a PTCs or 30% tax break on capital as an investment tax credit.[56] Either option would make both PI and Monticello profitable versus BAU in this analysis. While this bill is not law, it serves as an example that current legislative efforts for hydrogen PTCs would be enough to make the systems profitable in this analysis.

HTSE capital costs had a large effect on profitability, and the limited-surface search captured a large range of possibilities. The DOE-HTFO record suggests a total capital investment cost around ~\$700/kW, which would require some type of clean-hydrogen credit in all configurations at both plants. Several electrolyzer manufacturers maintain that their capital costs could be less than \$500/kW in the near term, meaning that Prairie Island could run without a PTC or clean-hydrogen credit and be profitable versus BAU. This was not the case in Monticello, where a credit of \$0.60/kg would still be needed for profitable operations versus BAU. The amount of credits that the NPP-HTSE system needs would be highly dependent on HTSE capital costs, but credits of more than \$1.00/kg and \$1.50/kg at PI and Monticello, respectively, would make the hybrid system competitive in all total CAPEX values under \$800/kW prior to the saturation of their respective hydrogen markets.

Monticello and PI have differences in their profitability mainly due to their hydrogen markets and their sizes. The capital cost, on a \$/kW basis, is reduced as the plant gets larger (see Figure 44). If the hydrogen market can handle the delivery from the bigger plant, then it is advantageous to maximize the size of the plant. This could change if hydrogen demand grows in the region around Monticello, but would still be limited to its smaller thermal output compared to PI.

4.2.4 Future Work on Economic Dispatch Optimization

Several extensions of this work are either ongoing or proposed. One focuses on the interaction and feedback between the cycling HTSE load and the LMPs used for making the dispatch decision. As the HTSE uses energy load from the NPP, the price that the NPP bids into the wholesale electrical market changes, potentially changing the electricity clearing price. That electricity price is what is used to make the dispatch decision between electricity and hydrogen, so a variable LMP creates a feedback loop in the dispatch. Capturing this feedback loop would move this optimization analysis away from a price-taker approach to more of a price-maker simulation. Modeling each node via standard PLEXOS production modeling on the system is complex and would hamper the ability to optimize storage components or sample many different options for the economic parameters. Large-scale electric system dispatch models such as PLEXOS also lack the capability to have a secondary hydrogen market with storage. Modeling price-maker scenarios in the modeling platform described in this report would be more straightforward and more agile to test various conditions.

Two efforts are currently underway to investigate the price-taker model assumption and move toward capturing the HTSE feedback on LMP. The first approach is to use a large-scale dispatch model, such as PLEXOS, that can capture the clearing price and transmission congestion at many nodes in the system and then use these outputs to bias the LMPs in the more-nimble optimization model detailed in this report. This approach will use new hourly LMP profiles produced from PLEXOS by varying the amount of nuclear plant bids adjusted for HTSE runtime and will be contained in the forthcoming NREL report. These LMP profiles can be compared to the previous ones to understand the degree to which LMP changes due to HTSE energy usage. If the effect is large, then the dispatch model's input LMPs can be biased according to the LMP changes between the original LMP and the new adjusted LMP.

A more-robust method at INL is currently in development to capture the change in LMP using artificial intelligence (AI). The AI uses a recurrent neural network and long short-term memory structure to capture the interaction between changes in NPP electricity sales and the LMP. The training data set comes from several large PLEXOS runs over the entire Northern States power region. This AI can then be iteratively queried in each hour of dispatch by the optimization model. This will be able to capture any change in LMP while still being able to track and apply hydrogen storage and run the optimization quickly to optimize inputs.

In addition to moving from a price-taker to a price-maker approach, this work could be extended to quantify the effect on the overall system cost and the effect of more flexible NPPs on other generators, like renewables. This would require a more-complete modeling of the rest of the system, rather than just modeling the NPP and HTSE as in the current optimization. This could be done by running the optimization as is, then feeding the dynamics into a large-scale capacity expansion or dispatch model, depending on the desired time horizons. Running with existing capacity expansion or dispatch models would likely require some structuring within their respective code bases to allow for nuclear to be run in this operational mode and to track the value of the hydrogen sold. The effect of the HTSE on the overall system would be helpful for stakeholders to understand the impact this system would have on variable-renewable-energy deployment, battery deployment, and clean-energy goals.

Analysis of this system in the context of the entire system would also be interesting as a means of investigating the inverse correlation between HTSE capital costs and storage usage. The most profitable configurations have low HTSE capital costs, which lower the investment capital, but also create a more-flexible system because oversizing the HTSE to facilitate storage is more advantageous. This flexibility could be an asset for reducing overall system costs and deploying more variable renewable energies. The systemwide analysis could quantify the value of flexibility and potentially justify building larger storage systems.

4.2.5 Summary

A summary of the findings of this optimization section is reported here.

- A PTC at Prairie Island is not needed to be profitable versus BAU when CAPEX is below $300/kw for varied amounts of hydrogen delivery.
- HTSE at Monticello is a more difficult economic proposition mostly due to the smaller hydrogen market near the plant. The smaller hydrogen market means that PTC of at least $0.07/kg is required to make the system more profitable than BAU.
- The profitability of both systems is highly dependent on the hydrogen market that surrounds the plant. Providing too much hydrogen saturates the market and drops the price of hydrogen, making it difficult for the NPP-HTSE to be profitable versus BAU.
- A matrix of profitable configurations was provided for both PI and Monticello. This matrix provides minimum requirements of HTSE CAPEX, hydrogen demand, and clean-hydrogen credit for the system to be profitable compared to BAU.
- Storage-tank sizes for both systems fluctuate between 2–3 and 5 hours of storage, depending on CAPEX. A higher CAPEX depresses storage because more storage requires a larger HTSE. One hour of storage is assumed to cover hydrogen-demand requirements for 1 hour.
- This analysis is from the perspective of an NPP-HTSE plant. Extending this analysis to a systemwide approach to investigate the value of flexible operation on other generators and their deployment and overall system cost would be advantageous. This could be done by extending this optimization structure to perform dispatch with other generators modeled. Another approach would be to use a capacity expansion model with PI and Monticello integrated energy system (IES) included.

5 FINANCIAL EVALUATIONS OF OTHER HYBRID INTEGRATIONS WITH LWRs

The analyses presented in this section show evaluations further down the hydrogen value chain by presenting use-case scenarios for the hydrogen generated from the NPP-HTSE plant already analyzed and presented. These use-case scenarios for hydrogen include:

- Blending of hydrogen with natural gas to be burned in natural gas power plants
- Compression and storage of hydrogen in trucks or pipelines versus liquefaction of hydrogen for transport to an end user at various distances
- NPP-HTSE-NH_3: hybrid integration of the NPP and HTSE with ammonia production versus the standalone NPP-HTSE
- NPP-HTSE-Synfuels: hybrid integration of the NPP and HTSE with synthetic fuels—diesel, jet fuel, and motor gasoline—production using low-carbon hydrogen versus the standalone NPP-HTSE.

Other integrations—such as using nuclear power for running a cryogenic refrigerant cycle, chlor-alkali process, and formic-acid production—are introduced here briefly, but will be analyzed more fully in efforts separate from this project. These integrations approach the idea of an "energy park" by using nuclear power as the central element. High density baseload electricity and thermal energy from an NPP could be highly integrated with various industrial facilities in close proximity to create maximum synergy and efficiency of operations. This concept is already used extensively in large industrial settings—e.g., petrochemical facilities. In modern petrochemical facilities, heat and material streams are passed from one unit to another to the extent possible, such that one unit's waste material or heat is another unit's treasure, so to speak. In this way, waste in the form of heat or material, is minimized to the greatest extent possible to create a much more-efficient and profitable facility overall. In order for this concept to work, there must be an established general interest among the facilities. For example, if each process unit in the complex were to maximize their own profitability without regard for profitability of the complex as a whole, then the complex would lose to the benefit of individual-unit profitability. Thus, the energy park analyzes the profitability of the complex of integrated process units to assure maximum profitability of the whole, not necessarily the parts at the expense of the whole. Since LWR's provide very large amounts of energy and already have their CAPEX fully depreciated, they make excellent candidates for exploring advantageous energy-park configurations.

5.1 Hydrogen Blending with Natural Gas for Use in Natural Gas Power Plants

To better understand this scenario, natural gas and H_2 blending can be evaluated in terms of cost of avoided CO_{2e} per MT. The cost of avoided CO_{2e} was calculated using the equation below, the change in life cycle GHG emissions are from the Section 2.2.1. The energy equivalent prices of H_2 from the natural gas prices (Xcel owned natural gas power generators) were compared with the nuclear H_2 price of \$1.93/kg (from the HTSE modeling). Energy equivalent price were considered as we are replacing natural gas energy with H_2, so H_2 price is about \$0.2-\$0.5/kg for the natural gas price of \$2-\$4/1000 ft^3.

Natural gas prices were compared to energy equivalent prices of hydrogen and avoided cost of CO_{2e} (replacing natural gas by nuclear H_2) for Xcel Energy's High Bridge Generating Plant. Natural gas prices were provided by Xcel energy for their natural gas electricity generation plants in the Twin Cities area. The avoided cost of CO_{2e} was calculated using a H_2 price of \$1.89/kg, estimated using the HTSE model from the previous section.

$$Cost\ of\ avoided\ CO_2\left(\frac{\$}{MT}\right)\ = change\ of\ price\ of\ energy\ \left(\frac{\$}{MT}\right)\Big/ change\ of\ CO_2\ emissions\ \left(\frac{MT\ CO_2}{MT}\right)$$

The cost of avoided CO_2 per ton is about $150 over a period of 2 years; similar results were also obtained for Cannon Falls and LS power stations.

5.2 Hydrogen Storage and Delivery

When hydrogen is produced in large quantities at central locations, a reliable hydrogen-delivery infrastructure is needed to transport it from the production plant to the demand location or other points of use, like refueling stations. Hydrogen is currently transported in two physical forms, liquid and gaseous. It is delivered in gaseous form using tube-trailers and pipelines, and in liquid form using cryogenic tankers. Three major pathways that are considered commercially viable options for hydrogen delivery were discussed in a recently published book chapter.[57,58]

5.2.1 Gaseous Hydrogen Delivery Using a Pipeline-Delivery Pathway

It is economical to have a pipeline delivery pathway when there is a large enough demand to warrant the construction of dedicated transmission and distribution pipelines. Figure 66 shows the gaseous delivery pathway of hydrogen with pipeline supply. A gaseous pipeline-delivery pathway receives hydrogen from a central production plant, along with a geologic/physical storage facility which provides backup supply in the event of production plant outages and acts as a buffer against seasonal demand variations. A compressor conveys hydrogen from the production plant and geologic storage to the transmission pipeline, which in turn transports hydrogen to the city gate or distribution terminal. The distribution pipeline brings hydrogen to the demand locations. If hydrogen is produced at the city gate or distribution terminal, a compressor is needed to pressurize hydrogen from what is called a semi-central production plant to the distribution-pipeline network.

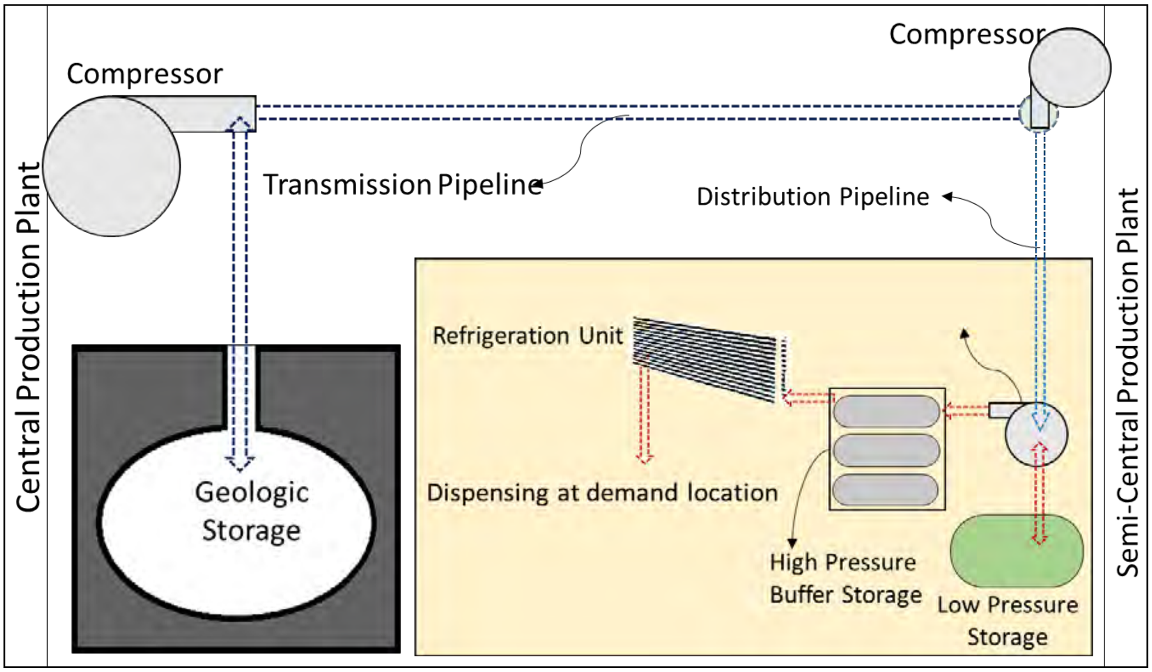

Figure 66. Schematic of gaseous hydrogen-delivery pathway with pipeline supply.

5.2.1.1 Cost estimates

Pipeline construction is a major investment that depends on large and consistent demand to recover the cost. Pipeline construction costs include material costs, labor costs, right-of-way (ROW) cost and miscellaneous costs. A team from Pacific Northwest National Laboratory (PNNL) has developed equations for material, labor, and ROW costs, based on published data for natural gas pipeline construction. While the equations have been developed for each region, average costs for the entire U.S. can be obtained from the relationships shown in Table 32.[59]

Table 32. Relationships used to estimate U.S. pipeline construction cost by component.

Cost Component	Cost Estimating Equation (in 2009 dollars)
Material	63027*e^(0.0697*D)
Labor	$-51.393 * D^2 + 43523 * D + 16171$
ROW	$-9E\text{-}13 * D^2 + 4417.1 * D + 164241$
Miscellaneous Costs	37% of the sum of material, labor, and ROW costs
D represents Pipeline diameter and should be defined in inches The resulting cost estimates are in 2009 dollars Assumes a 10% cost premium for hydrogen pipelines versus natural gas pipelines	

For hydrogen pipelines, cost is assumed to be about 10% higher than the estimates shown in Table 33. Construction costs for transmission pipelines of different diameters are shown in Figure 67.[60]

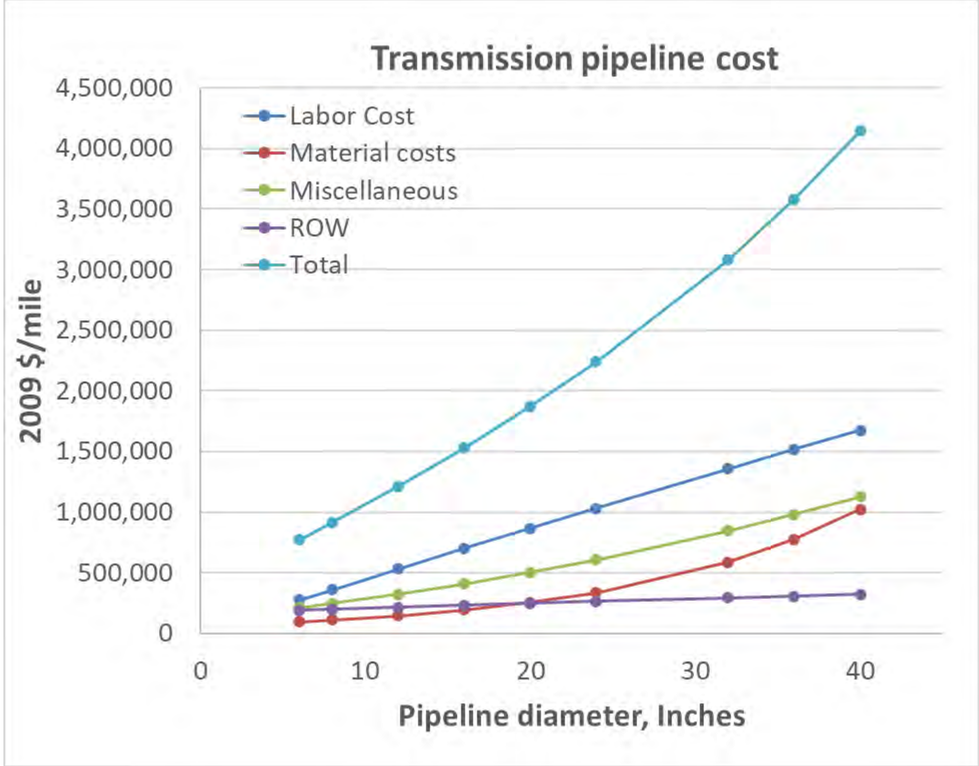

Figure 67. Construction costs for transmission pipelines of different diameters.

Most pipelines used to transport industrial and domestic gas are constructed of steel. The estimated costs of transmission pipelines are shown in Table 33.[61]

Table 33. Estimated costs for transmission pipelines.

Pipeline	2011 Status
Transmission ($ per mile for 6-40 in pipelines, excluding ROW)	$765–4,500K

5.2.1.2 Attributes

5.2.1.2.1 High Cost

As mentioned above, the labor cost of pipeline construction is about 50% of the total cost and is a major barrier to installing hydrogen pipelines.[61] Labor costs are a function of the time required to lay and weld pipe sections. Steel is limited by its weight; hence, a given expanse of steel pipeline will require more pipe sections to be transported to the job site and more welds. Innovative packaging and joining techniques are required to bring down construction time and, consequently, labor costs.

5.2.1.2.2 Hydrogen and Material Interactions

The interaction between hydrogen and pipe materials is not well understood at high operating pressures, especially when pressure cycling is involved. High pressures and pressure cycling also affect the durability of materials. Efforts should be made to develop new coatings to prevent the embrittlement of steel pipelines.

5.2.1.2.3 Compression

Reciprocating compressors are currently used to provide high throughput for the transmission of hydrogen to industrial sites. The reciprocating compressors presently used are costly and can contaminate hydrogen with lubricants, thereby degrading fuel-cell performance. The embrittlement associated with hydrogen service equipment requires special materials. Possible solutions include new lubricant-free compression technologies that can provide high throughput and avoid contamination or low-cost hydrogen-purification processes that reduce the impact of purification on delivery cost.

5.2.1.2.4 Geologic Storage

Though hydrogen has been stored at low cost and in large quantities in geologic storage facilities, leakage and contamination are significant risks, high pressures may create operational challenges, and development costs can be high. Geologic storage needs a cushion gas (minimum amount of gas that must be left in storage) which for natural gas is about 15% of the storage capacity.[61] Leakage due to permeation into the surrounding rock may be unavoidable and may result in significant increases in storage cost. Contamination may require a post-withdrawal purification step. The low energy density of hydrogen requires higher storage pressures than typically maintained for natural gas. The effects of those pressures per se, of cycling (i.e., filling and withdrawing hydrogen) at higher pressure, and of the reactivity of hydrogen with rock formations are not known. Another major challenge is the lack of suitable geologic formations in certain regions, potentially limiting the feasibility of geologic storage in certain geographic locations.

5.2.2 Liquid-Hydrogen Delivery Pathway

Figure 68 shows that hydrogen produced at a central production plant may be transported to the distribution terminal via transmission pipelines, where it is liquefied and loaded into liquid tankers. In the case of semi-central production, hydrogen is produced and liquefied at the distribution terminal. The daily capacity of the liquefier equals or exceeds the average daily demand. Cryogenic storage tanks usually hold 5–7 days of liquefier production. A pump draws liquid hydrogen from the cryogenic storage tank to the liquid tankers. The liquid tanker is transported to the demand location, where it is emptied into a cryogenic storage tank and used further.

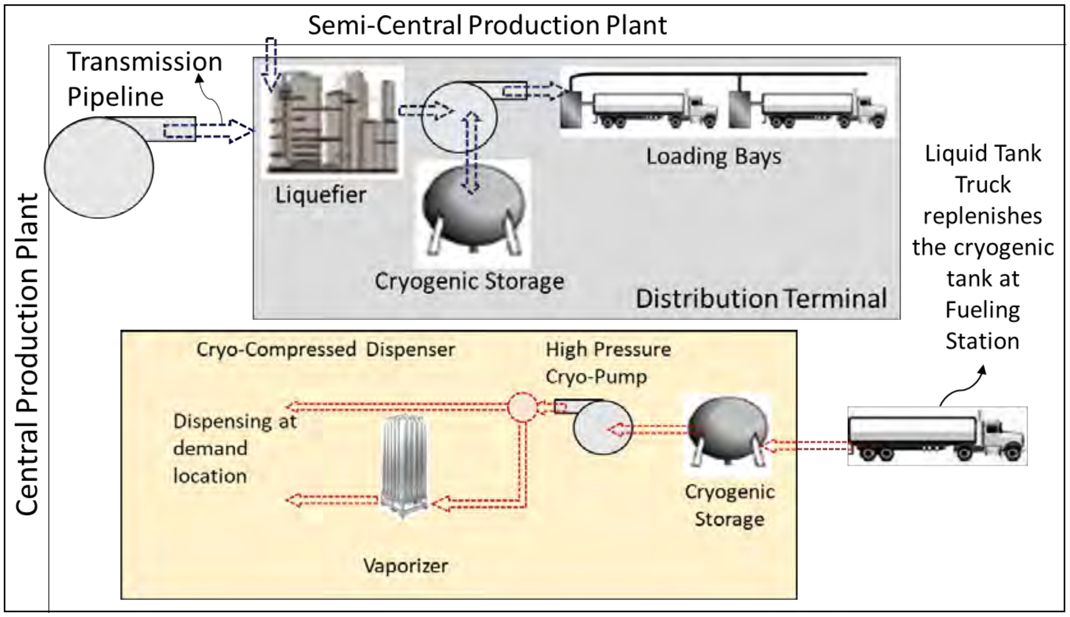

Figure 68. Schematic of the liquid-delivery pathway.

5.2.2.1 Cost estimates

Hydrogen exists in liquid state below 20 K at atmospheric pressure. Achieving such low temperatures is energy intensive and expensive. The liquefaction energy represents about 1/3 of the total energy content of the liquefied hydrogen. The liquefaction process accounts for more than \$1 per kg of hydrogen cost. The liquid hydrogen must be stored at 20 K in jacketed stainless steel vacuum tanks. Heat leaks through the tank walls, resulting in the vaporization of liquid hydrogen (also called boil-off). The boil-off can be recovered or vented to the atmosphere to avoid pressure buildup in the storage tank. The boil-off rate from the tank can be minimized with a high volume-to-surface ratio. Most stationary liquid-hydrogen tanks are spherical in shape to minimize the boil-off rate. The boil-off rate varies from 0.4% to 0.06% per day for 50 m^3 and 20,000 m^3 volume tanks, respectively.[61] Additionally, there is significant boil-off loss during unloading at the demand location (pumping hydrogen from the liquid tanker to an on-site cryogenic tank). To minimize these boil-off losses, the number of deliveries should be limited, and delivery routes should be planned accordingly.

Liquefying hydrogen increases its volumetric mass and energy densities many-fold. The capacity of the liquid tanker is about 4 metric tons, 5–6 times the capacity of a composite tube-trailer and 15–20 times the capacity of a steel tube-trailer. DOE’s 2011 estimate of the cost of liquid-hydrogen tankers is shown in Table 34.[60] It should be noted that liquid-tanker technology is considered mature; hence, DOE sets no 2020 targets.

Table 34. Estimated cost and key features of liquid hydrogen tankers.[60]

Liquid Tanker[a]	**2011 Status**
Operating Pressure (bar)	5
Capacity (kg)	4300
Capital Cost (2007 dollars)	\$720K

[a] Cryogenic tank and trailer (excludes truck cab)

5.2.2.2 Attributes

Boil-off is unavoidable, and methods must be developed to either minimize it or reuse the captured energy elsewhere. Boil-off from pumps unloading fuel from the tanker to site storage tanks can account for an up to 5% loss (by volume) in the amount transferred.[61] These losses should be minimized by more-efficient component designs and improvements such as submerged pistons.

Underground storage minimizes setback distances and is a preferred option. However, its high cost may offset the savings associated with lower land areas.

During the liquefaction of hydrogen, H_2 is cooled to 20 K in a multistage process, which is energy-intensive. The exothermic conversion of ortho- to para-hydrogen consumes a significant amount of energy as well. The total energy consumption to liquefy gaseous hydrogen from atmospheric conditions is approximately 10–15 kWh/kgH_2.[61] Technologies like magnetic or acoustic liquefaction may reduce the energy required, and they need to be investigated. Other storage options, like high-pressure cryo-compressed storage tanks, can reduce the energy demand by avoiding the exothermic conversion of ortho- to para-hydrogen by allowing hydrogen storage at temperatures between 80 and 200 K.[61]

5.3 Ammonia-Plant Financial Analysis

5.3.1 Overview of Ammonia Production

The markets and demand for ammonia have been previously discussed in the demand-analysis section of this report. In the Haber-Bosch NH_3 synthesis process, 3 moles H_2 are combined with 1 mole N_2 to produce 2 moles NH_3. The synthesis of ammonia using low-carbon-intensity hydrogen represents an opportunity to carry hydrogen further down the value chain at the NPP. Ammonia is more easily stored and transported as a liquid than hydrogen.

The Haber-Bosch process uses a heterogeneous metal catalyst to overcome the high activation energy associated with the dissociation of triple-bonded N_2 molecules. The process is operated at high temperature (i.e., 400–500°C) to increase the rate of reaction and high-pressure (>100 bar) to shift the reaction equilibrium in favor of the products (the net ammonia synthesis reaction produces two moles of gaseous product from four moles of gaseous reactants). Multiple passes through the synthesis reactors are required to achieve the targeted overall conversion level. The reaction of nitrogen and hydrogen to produce ammonia is exothermic; for this reason, process cooling is required to maintain the target reaction equilibrium. As a side note, an approximate analysis was carried out to determine whether exothermic heat from an ammonia plant receiving a hydrogen feed from an HTSE could supply the full heat requirements to the HTSE. It was found that the ammonia plant could supply roughly 40% of the energy. Thus, an NPP is still valuable to provide heat to the HTSE versus the ammonia plant. This was a very rough preliminary analysis and will not be discussed further.

5.3.2 Ammonia Synthesis Process Evaluation

For ammonia process modeling, a detailed and previously completed model by Rick Wood in TEV-666[62] was used. The model was recreated from the documentation and validated in Aspen HYSYS. Wood investigated ammonia synthesis and ammonia-derivative production using several process configurations, including conventional gas-to-ammonia as well as a configuration using a high-temperature gas reactor (HTGR) as the primary energy source. Results from the conventional gas-to-ammonia process design are used as a baseline case in this analysis. The nuclear-integrated process design is modified to evaluate use of an LWR as the primary energy source for ammonia production.

The reference ammonia synthesis plant has a daily ammonia-production capacity of 3,048 metric tonnes per day.[62] This quantity of ammonia-production requires an H_2 feed rate of 550 metric tonnes per day and an N_2 feed rate of 2,532 metric tonnes per day. This plant capacity is used as the baseline value for the conventional and nuclear-integrated ammonia-synthesis process designs discussed below.

5.3.2.1 Conventional Natural Gas to Ammonia

5.3.2.1.1 Overview

Conventional ammonia production was evaluated based on processes using natural gas to provide the energy and material feedstocks. SMR is used to generate hydrogen for use in the process, and excess heat from SMR, water-gas shift, ammonia synthesis, and ammonia-derivative synthesis processes is used to generate steam for electrical-power generation. The electrical-power generation does not fully offset the process electrical-power requirements. The electrical power needed to balance the net process electrical-power demand must be purchased from the electrical-grid.

The SMR process is associated with significant carbon emissions. The conventional gas-to-ammonia process design based on the analysis is presented in TEV-666.[62] A portion of the carbon released by the SMR process is there described as being used as a carbon source for the production of urea; however, the anhydrous ammonia process design evaluated herein does not include production of ammonia derivatives (e.g., urea and ammonium nitrate) produced by the process analysis reported in TEV-666. Therefore, carbon emitted by the SMR process is released to the atmosphere, resulting in a significant carbon footprint for the ammonia produced by the conventional gas-to-ammonia case.

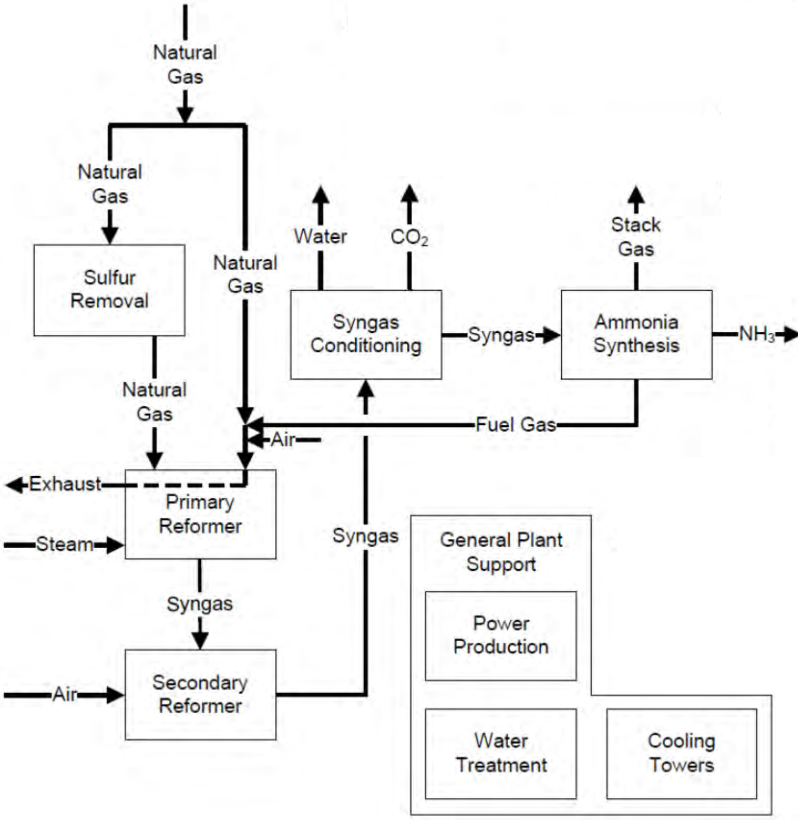

Figure 69. A block flow diagram for the conventional natural gas-to-ammonia process. Adapted from TEV-666.[62]

5.3.2.1.2 Process Area Descriptions[62]

5.3.2.1.2.1 Natural gas purification and reforming

A description of the natural gas purification and steam reforming process areas is provided in TEV-666.[62] The process description from TEV-666 is quoted below for the reader's convenience:

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

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

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

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

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

5.3.2.1.2.2 Syngas conditioning

The following description of the syngas conditioning process area is provided in TEV-666.[62]

Syngas cleaning and conditioning for the natural gas-fed case is similar to the coal-fed case. However, the following changes are made for the natural gas-fed case:

1. Water saturation is not necessary prior to shift conversion. By introducing sufficient steam to the reforming section of the plant, the syngas entering the high-temperature shift converter contains sufficient moisture for shift conversion. The steam-to-dry gas ratio of the syngas is 0.54, which is above the minimum requirement for the catalyst (0.40).
2. The mercury sorbent bed is not required.
3. Because sulfur has been removed prior to reforming, Selexol is used for CO_2 removal only. The Selexol unit is operated to remove the majority of the CO_2; hence, a [pressure swing adsorption] unit is not required.
4. The resulting syngas contains H_2 and N_2 in a molar ratio of 3:1. Small amounts of argon (0.3 mol%) and methane (1.1%) are also present in the cleaned syngas.

5.3.2.1.2.3 *Ammonia synthesis*

Ammonia-synthesis operations are consistent with the process design provided in TEV-666. **Error! Bookmark not defined.** The ammonia-synthesis process description from TEV-666 is reproduced for the reader's convenience:

Syngas feeding the ammonia synthesis unit has been previously adjusted to achieve the H_2/N_2 molar ratio 3.0. Incoming feed gas is compressed to 3,000 psi. Unreacted recycle gas is mixed with the fresh feed gas and preheated by cross exchanging with hot reactor effluent gases. Equilibrium conversion is assumed in the ammonia converters for the following reaction:[63]

$N_2 + 3H_2 \rightarrow 2NH_3$

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

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

5.3.2.1.2.4 *Power generation*

High- (1515 psi), medium- (165 psi) and low-pressure (30 psi) steams generated throughout the plant are sent to steam turbines to generate electrical power. The turbine exhaust is condensed and mixed with condensate return from the plant. Makeup water is added to provide the necessary flow to boiler feedwater pumps.[62]

5.3.2.1.2.5 *Process cooling*

Process cooling is provided using conventional cooling towers. The evaporation rate, drift, and blowdown are based on values suggested by [73].

5.3.2.1.2.6 *Water treatment*

Water treatment is simplistically modeled using a variety of separation blocks. The process model accounts for water inlet and outlet flows in each of the process areas to maintain a water balance for the overall plant.[62]

5.3.2.1.3 Material and Energy Balances

Conventional gas-to-ammonia process material and energy balances were derived from TEV-666 for a modified plant configuration that produces anhydrous ammonia as the final product. Process areas and unit operations required for the production of ammonia derivatives (urea and ammonium nitrate) and intermediate products (nitric acid) are not included in the conventional gas-to-ammonia process design. The conventional natural gas-to-ammonia process design operates in a continuous operating mode with a specified capacity factor of 0.92 (consistent with TEV-666).

5.3.2.1.4 Capital Costs

Capital-cost estimates for all conventional gas-to-ammonia production process areas are obtained from TEV-666. The capital costs for process areas associated with the production of ammonia derivatives (urea and ammonium nitrate) and intermediate products (nitric acid) were excluded from the analysis because anhydrous ammonia is the final product of the conventional gas-to-ammonia process design evaluated. The capital costs reported in TEV-666 were adjusted to 2020 dollars using the CEPCI composite plant index.

The conventional gas-to-ammonia process design is based on an ammonia production rate equal to that specified in TEV-666 (Table 35). For sensitivity analyses of ammonia production costs as a function of plant design capacity, capital costs were estimated by scaling the capital costs reported in TEV-666 by the capacity ratio with a 0.6 scaling exponent. The capacity of each subprocess area train was limited to the maximum value listed in TEV-666.

Table 35. Conventional gas-to-ammonia process capital expenses.

Process Area	Capacity		Direct Capital Cost	Total Capital Cost	Note or Reference
Steam Methane Reforming	88	MM SCFD	$103,300,000	$134,090,000	
Water Gas Shift Reactor	57,039	lbmol/hr	$34,134,000	$44,306,000	
Selexol	41,720	lbmol/hr	$40,344,000	$52,366,000	
Methanation	3,360	ton/day	$8,271,200	$10,736,000	
Subcritical CO_2 Compression			$0	$0	
Supercritical CO_2 Compression			$0	$0	
Ammonia Synthesis	3,360	ton/day	$258,220,000	$335,170,000	
Urea Synthesis			$0	$0	
Nitric Acid Synthesis			$0	$0	
Ammonium Nitrate Synthesis			$0	$0	
Steam Turbines	29	MW	$15,332,000	$19,900,000	Scaled based on values reported in [62]*
Cooling Tower	93,964	gpm	$2,835,000	$3,680,000	Scaled based on values reported in [62]*
Water Systems			$32,833,000	$42,618,000	7.1% of TDCC
Piping			$32,833,000	$42,618,000	7.1% of TDCC
Instrumentation and Control			$12,024,000	$15,607,000	2.6% of TDCC
Electrical Systems			$36,995,000	$48,020,000	8.0% of TDCC
Civil/Structural/ Buildings			$42,545,000	$55,223,000	9.2% of TDCC
Ammonia Pipeline			$0	$0	Cost of ammonia transport pipeline not included
Total			$619,666,200	$804,334,000	

* Conservatively scaled only by plant scaling factor (reduction in process area capacity due to alternative plant configuration not considered)

5.3.2.1.5 Operating Costs

Conventional gas-to-ammonia process direct and indirect operating costs were calculated using the methodology of TEV-666 (Table 36). Direct operating costs include materials (natural gas feedstock, water treatment chemicals, solvent, and catalyst replacement), utilities (electrical power, process, and cooling water makeup), royalties, and labor and maintenance. Indirect operating costs include overhead, insurance, and taxes. Carbon sequestration costs are not included.

The cooling water makeup and treatment costs were adjusted from those reported in TEV-666 to account for the decrease in cooling load associated with the process configuration that produces only anhydrous ammonia (the cooling loads associated with the urea synthesis, nitric-acid synthesis, and ammonium nitrate synthesis process areas were excluded from this analysis).

Table 36. Conventional gas-to-ammonia process operating expenses.

Direct Costs Materials	Price	Unit	Consumed	Unit	Annual Cost	Notes
Average Natural Gas	4.11	$/MSCF	88	MMSCFD	$121,307,000	
Makeup H2O Treatment	0.02	$/k-gal	2695	k-gal/day	$18,000	
Wastewater Treatment	1.32	$/k-gal	1179	k-gal/day	$522,000	
H2S Catalyst	700	$/ft^3	0.04	ft^3/day	$9,000	
Zinc Oxide	300	$/ft^3	0.344	ft^3/day	$35,000	
Preforming Catalyst	2350	$/ft^3	0	ft^3/day	$0	
Primary SMR Catalyst	750	$/ft^3	0.16	ft^3/day	$40,000	
Secondary SMR Catalyst	650	$/ft^3	0.04	ft^3/day	$9,000	
HTS Catalyst	380	$/ft^3	0.19	ft^3/day	$24,000	
LTS Catalyst	600	$/ft^3	0.15	ft^3/day	$30,000	
Selexol Solvent	2.57	$/gal	8.14	gal/day	$7,000	
Methanation Catalyst	700	$/ft^3	0.042	ft^3/day	$10,000	
Ammonia Synthesis Catalyst	775	$/ft^3	0.029	ft^3/day	$8,000	
Utilities						
Electricity	30	$/MW-h	60.66	MW-e	$14,660,000	
Consumption			80.32	MW-e		
Generation			-19.66	MW-e		
Water	0.05	$/k-gal	2696	k-gal/day	$45,000	
Royalties					$1,213,000	
Labor and Maintenance, non-nuclear					$33,380,000	1.15% of TCI for labor, 3% of TCI for maintenance
Indirect Costs						
Overhead					$21,697,000	65% of labor and maintenance costs
Insurance and Taxes					$12,065,000	1.5% of TCI
Total Manufacturing Costs					$205,079,000	

* TCI does not include costs associated with LWR or HTSE systems.

5.3.2.1.6 Revenues

No ammonia derivatives, such as urea or ammonium nitrate, are produced by the conventional gas-to-ammonia process design. Therefore, no revenue streams are included for ammonia-derivative products. The anhydrous ammonia selling price was determined by manipulating the ammonia price to the value that resulted in an NPV equal to zero. This represents the minimum ammonia selling price that would have to be realized for the plant to meet the financial obligations associated with recovering the capital investment costs, covering the operating costs, and providing the specified return on investment to the project financial investors (as compared conventional revenues, see Table 37 and Table 38).

Table 37. Financial parameters for ammonia-production analysis.

Parameter	Value	Note
WACC	7.09%	Value provided by Xcel during project kickoff meeting in Minnesota
Minnesota corporate tax	9.80%	
Federal tax	21.00%	
Tax deductibility	-2.06%	
Inflation	2.50%	
Startup year	2027	
Plant operating life	20 years	

Table 38. Conventional gas-to-ammonia process annual revenues.

Product	Price	Produced	Annual Revenue
Oxygen	\$45.86/tonne	0 tonne/day	\$0
Ammonia	\$295.67/tonne	3048 tonnes/day	\$302,581,000
Total			\$302,581,000

5.3.2.2 Nuclear-integrated ammonia production

5.3.2.2.1 Overview

Green-ammonia production was evaluated using hydrogen sourced from HTSE, as designed and presented in this report, and nitrogen sourced from cryogenic ASUs. An LWR is specified as the energy source for the ammonia production process (including the heat and power for the HTSE process and the power for the ASU). Because water, air, and energy from the LWR are the primary process inputs, the nuclear ammonia-production process produces carbon-free green-ammonia. Figure 70 provides a block flow diagram of the LWR-based ammonia-production process. The proposed process includes unit operations for hydrogen generation via HTSE, nitrogen generation via cryogenic air separation, syngas purification, ammonia synthesis, power generation, cooling towers, and water treatment. Because excess heat from the ammonia synthesis process is used in the HTSE process for feedwater vaporization, the process design requires that the HTSE and ammonia plants be located at the same site.

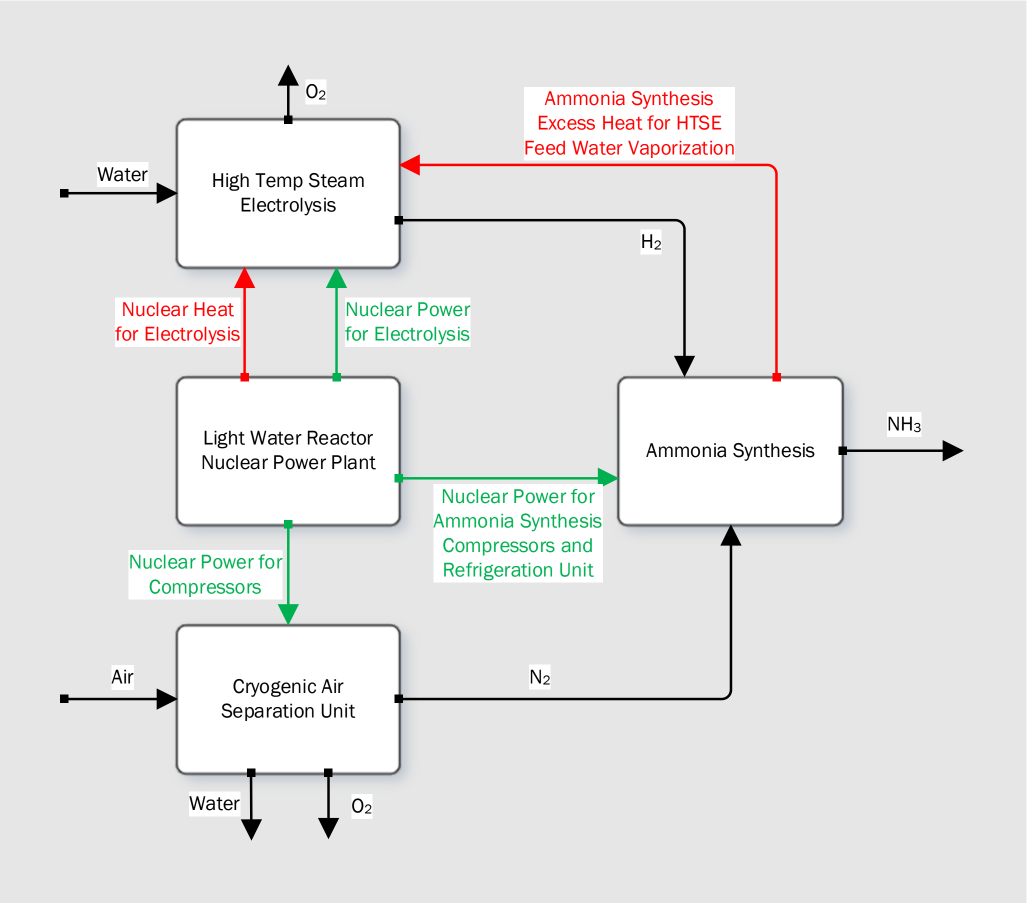

Figure 70. LWR-based anhydrous ammonia production process block flow diagram.

5.3.2.2.2 Process Descriptions

The following section provides descriptions of each of the process areas included in the LWR- HTSE ASU ammonia-production process. HYSYS model process flow diagrams and the accompanying stream tables are included in Appendix E.

5.3.2.2.2.1 Hydrogen production via high-temperature steam electrolysis

HTSE is implemented to generate hydrogen and oxygen from water. Thermal energy and electrical power required to drive the HTSE process are provided by an LWR NPP. The SOEC stacks operate at an elevated temperature of ~750°C, which exceeds the temperature at which heat can be supplied by the LWR. Therefore, heat from the LWR is used to vaporize the HTSE process feedwater at a temperature below the LWR process-heat supply temperature. Electrical power from the LWR is used to power HTSE process pumps, compressors, topping heaters, and SOEC stacks.

Following vaporization of the feedwater, recuperators heat the feed stream via heat exchange with the high-temperature SOEC-stack outlet stream. Electrical topping heaters are used to provide the final heating required to bring the feed stream to the specified SOEC-stack operating temperature. The SOEC stack is assumed to operate at the thermoneutral voltage such that electrolysis operations do not lead to heating or cooling of the products, and no heat addition or removal is required to maintain the specified stack operating temperature.

Use of an air sweep-gas stream is specified for removal of the oxygen product from the SOEC stack. The oxygen enriched sweep-gas stream exiting the process is vented to the atmosphere. Hydrogen produced via electrolysis exits the SOEC stack along with unreacted steam. The hydrogen product is separated from the steam by cooling (using recuperation and cooling water) and pressurizing the stack outlet stream to condense the unreacted steam, which is subsequently recycled back to the process feedwater inlet. The purified hydrogen product is compressed to a pressure of 49 bar specified for blending with nitrogen to produce the ammonia synthesis-gas feed stream.

5.3.2.2.2.2 Nitrogen production via cryogenic air separation unit

Nitrogen and oxygen are produced via a standard cryogenic Linde-type ASU that uses two distillation columns and extensive heat exchange in a cold box.[65] The ASU capacity is defined by N_2 demand for ammonia synthesis. The excess O_2 produced could potentially be sold as an end product from this process.

5.3.2.2.2.3 Ammonia synthesis

Ammonia synthesis for LWR-HTSE ASU case is identical to the conventional natural-gas-fed case.

5.3.2.2.2.4 Process cooling

The cooling-water system for the LWR-HTSE ASU ammonia-process design is identical to that of the conventional natural-gas-to-ammonia process design.

5.3.2.2.2.5 Water treatment

Water treatment for the LWR-HTSE ASU ammonia-process design is identical to that of the conventional natural-gas-fed ammonia production process design.

5.3.2.2.3 Material and Energy Balances

A HYSYS LWR-HTSE ASU ammonia-production process model was developed based on the ammonia-production process design reported in TEV-666.[62] The LWR-HTSE ASU ammonia-production process configuration is modified from the process design presented in [62] in order to produce anhydrous ammonia as the final product. The process areas and unit operations required for production of urea and ammonium nitrate (i.e., ammonia derivatives) are not included in the LWR-HTSE ASU process configuration.

In [62], steam generation associated with ammonia synthesis is used for power generation; in the current analysis, all excess steam generation from ammonia-production operations is sent to the HTSE process area to supplement the LWR heat input for HTSE process-feedwater vaporization. The quantity of steam available from the ammonia-synthesis process operation is not sufficient to fully replace the heat input from the LWR power plant, but it does significantly reduce the quantity of nuclear process heat that must be exported from the LWR to provide HTSE process heat input.

The LWR-HTSE ASU ammonia-process design operates in a continuous operating mode, with a specified capacity factor of 90.5% (consistent with the HTSE plant operations).

5.3.2.2.4 Capital Costs

HTSE-process capital-cost estimates are obtained from the analysis presented in Section 3.2.1 of this report. Capital-cost estimates for all other process areas are obtained from [62]. The capital costs reported in TEV-666 are adjusted to 2020 dollars using the CEPCI composite plant index. Additionally, costs for process areas in which the production rate differs from that specified by [62] were estimated by scaling the capital costs by the capacity ratio with a 0.6 scaling exponent. The capacity of each subprocess area train was limited to the maximum value listed in [62].

The capital costs for process areas associated with ammonia derivatives production were excluded from the analysis because anhydrous ammonia is the final product of the LWR-HTSE ASU process.

Because all excess steam generated by the ammonia-synthesis process is used for HTSE process-feedwater heating, the non-nuclear steam turbines present in [62] are not included in the LWR-HTSE ASU process design. This eliminates the capital costs associated with this equipment.

Anhydrous ammonia synthesis plant capital costs for a baseline process size corresponding to 852 MW-e of HTSE capacity (3048 tonnes per day NH_3 production) are presented in Table 39. The total capital cost for each process area includes multipliers for engineering (10%) and contingency (18%).

Table 39. LWR-HTSE ASU ammonia production process capital expenses.

Process Area	**Capacity**	**Direct Capital Cost**	**Total Capital Cost**	**Note or Reference**
High-Temperature Steam Electrolysis	852 MWe	$341,890,000	$568,850,000	
H_2 Storage		$0	$0	
H_2 Transport	0.1 mile	$68,000	$68,000	
N_2 Compression	29.7 kg/s	$3,225,000	$3,225,000	
Air Separation Unit	65700 lb/hr O_2	$43,908,000	$56,992,000	Scaled based on values reported in TEV-666
Ammonia Synthesis	3360 ton/day	$253,330,000	$328,830,000	Scaled based on values reported in TEV-666
Steam Turbines (no excess heat from ammonia process is used for power)		$0	$0	Scaled based on values reported in TEV-666*
Cooling Tower	146000 gpm	$5,053,000	$6,559,000	Scaled based on values reported in TEV-666*
Water Systems		$21,463,000	$27,859,000	7.1% of TDCC
Piping		$21,463,000	$27,859,000	7.1% of TDCC
Instrumentation and Control		$7,859,600	$10,202,000	2.6% of TDCC
Electrical Systems		$24,183,000	$31,390,000	8.0% of TDCC
Civil/Structural/Buildings		$27,811,000	$36,099,000	9.2% of TDCC
Ammonia Pipeline		$0	$0	cost of ammonia transport pipeline not included
Total		$750,253,600	$1,097,933,000	

5.3.2.2.5 Operating Costs

LWR-HTSE ASU ammonia-production process's direct and indirect operating costs were calculated using the methodology of TEV-666.[62] Direct operating costs include materials (catalyst replacement, water-treatment chemicals), utilities (electrical power, process, and cooling-water makeup), and labor and maintenance. Indirect operating costs include overhead, insurance, and taxes.

Because no excess steam from the ammonia-production process is available for power generation, all electrical power required to operate the ammonia-production process is obtained from the LWR power plant (this power could also be sourced from the electrical-grid, but would then be associated with the carbon emissions representative of the electrical power-generation technologies used in the regional market).

The cooling-water makeup and treatment costs were adjusted from those reported in [62] to account for the decrease in cooling load associated with the process configuration that produces anhydrous ammonia only (the cooling loads associated with the urea synthesis, nitric-acid synthesis, and ammonium nitrate synthesis process areas were excluded from this analysis).

Anhydrous ammonia synthesis-process operating costs are presented in Table 40. The costs reported correspond to an operating capacity factor of 90.5%, which is consistent with HTSE plant operations. Materials prices were left unchanged from those reported in [62]. A baseline electricity price of $30/MWh-e was selected as the energy cost of power obtained from the LWR. It is assumed that no royalty fees are required for the LWR/HTSE-based anhydrous-ammonia-production process. Anhydrous-ammonia synthesis-process labor and maintenance costs are calculated as 1.15 and 3%, respectively, of the total capital-investment costs. The labor and maintenance costs are based on the TCI for the anhydrous-ammonia process and do not include the costs for the LWR and/or HTSE systems (the O&M costs for the LWR and HTSE systems are accounted for in the hydrogen analysis). Overhead costs are set equal to 65% of the labor and maintenance costs. Insurance and taxes are calculated as 1.5% of the ammonia-synthesis process TCI (excludes LWR and HTSE system costs).

Table 40. LWR-HTSE ASU ammonia process operating expenses.

Direct Costs

Materials	Price	Consumed	Annual Cost	Notes
Makeup H_2O Treatment	0.02 $/k-gal	1072 k-gal/day	$7,000	
Wastewater Treatment	1.32$/k-gal	354.9 k-gal/day	$155,000	
Methanation Catalyst	700 $/ft³	0 ft³/day	$0	
Ammonia Synthesis Catalyst	775 $/ft³	0.029 ft³/day	$7,000	
Utilities				
Electricity	30 $/MW-h	MW-e	$16,557,000	
consumption		MW-e		
generation		MW-e		
Water	0.05 $/k-gal	1072 k-gal/day	$18,000	
Royalties				
O&M, Nuclear				
Labor and Maintenance, non-nuclear			$21,820,000	1.15% of TCI for labor, 3% of TCI for maintenance
Indirect Costs				
Overhead			$14,183,000	65% of labor and maintenance costs
Insurance and Taxes			$7,887,000	1.5% of TCI
HTSE Fixed O&M Costs	18.55 $/kW-yr	852 MWe	$15,809,000	
HTSE Variable O&M Costs (non-energy)	5.06 $/MWh	852 Mwe	$34,201,000	includes stack replacements, process, and cooling water
HTSE Variable O&M Costs (energy)	30 $/MWh	852 Mwe	$209,212,000	Cost includes electrical and thermal power input
Total Manufacturing Costs			$319,856,000	

5.3.2.2.6 Revenues

Oxygen from the ASU is specified to be sold at a fixed price of $45.86/tonne, as reported in [62]. Although the HTSE process also produces oxygen, the HTSE process configuration evaluated previously specifies the use of an air sweep-gas stream that releases an oxygen-enriched air steam. The oxygen-enriched air stream contains approximately 40 mol% oxygen and is therefore not suitable for sale as a final project. The use of an HTSE process configuration that supports the production of a purified oxygen stream would allow this byproduct of the HTSE process to be sold alongside the oxygen product produced by the ASU.

All hydrogen produced by the HTSE plant is used in the production of the anhydrous-ammonia product; thus, the economic analysis does not include revenue from hydrogen-product sales (i.e., the hydrogen is an internal process stream with production costs determined by the HTSE process-capital and operating-cost input specifications). No ammonia-derivative revenue streams are included because urea and ammonium nitrate are not produced by the plant configuration evaluated.

The anhydrous-ammonia selling price was determined using the same methodology as for the conventional gas-to-ammonia case; i.e., for each sensitivity case evaluated, the ammonia selling price was manipulated to the value that resulted in a project NPV equal to zero.

Table 41. LWR-HTSE ASU ammonia-production process, annual revenues.

Product	Price	Produced	Annual Revenue
Oxygen	$45.86 $/tonne	715 tonne/day	$10,840,000
Ammonia	$438.97 $/tonne	3048 tonne/day	$442,106,000
Total			$452,946,000

5.3.3 Process Comparison

A summary comparison of the conventional gas-to-ammonia and LWR-HTSE ASU ammonia-production process is provided in Table 42.

Table 42. Comparison of conventional natural gas-to-NH_3 and LWR-HTSE ASU ammonia-production processes.

	Conventional natural gas-to-Ammonia	**LWR-HTSE ASU Ammonia Production**	**Reference or Note**
Design Production Capacity	3048 tonne NH_3/day	3048 tonne NH_3/day	Based on TEV-666[62]
Natural Gas Consumption	88 MMSCFD	0 MMSCFD	Design point value
Baseline Natural Gas Price	$4.11/MSCF	N/A	Corresponds to modified natural gas pricing case
Net Electrical Power Input	60.6 MW-e	921.6 MW-e net (852 MW-e HTSE; 69.6 MW-e ASU, NH_3 synthesis, etc.)	Design point value
Baseline Electricity Price	$30/MWh-e	$30/MWh-e	
Net Thermal Power Input	0 MW-t	81.5 MW-t supplied by LWR	Design point value
Application For Steam Generated in Ammonia Synthesis Process Operations	Process heat applications (e.g., product purification) with excess steam used for power generation	Process heat applications (e.g., product purification) with excess steam used for HTSE process feed water vaporization	
HTSE Process Thermal Input	N/A	147.5 MW-t net (81.5 MW-t from LWR; 66 MW-t from NH_3 synthesis excess process heat)	
CO_2 Emissions	2.55 kg CO_{2e} per kg NH_3	0.06 kg CO_2 per kg NH_3	
Capacity Factor	92%	90.5%	
Actual Production Rate	2804 tonne NH_3/day	2760 tonne NH_3/day	
Ammonia Production Cost	$296/tonne	$439/tonne	

The sensitivity of the LWR-HTSE ASU ammonia-production cost to several key parameters is shown in Figure 71. This figure indicates the change in ammonia-production cost that could be expected from perturbations to a single input parameter (e.g., the results do not include the compounding or canceling effects that would result from simultaneous perturbation of multiple sensitivity variables). The sensitivity analysis results are presented as a tornado chart in which the variables that have the greatest impact on the ammonia-production cost are listed at the top of the chart. Over the range of input-variable perturbations considered, the price of electricity has the greatest impact on ammonia-production costs, followed by the

HTSE capital cost. Changes to the IRR and ammonia-synthesis plant capital cost may also have a significant impact on ammonia-production costs (changes to either of these variables can affect the NH_3 cost by a value ranging from plus or minus $20–25 per tonne). Changes to the plant capacity or oxygen-byproduct selling price up to 50% different from the base value would be expected to result in a change in ammonia-production cost of less than $10/tonne.

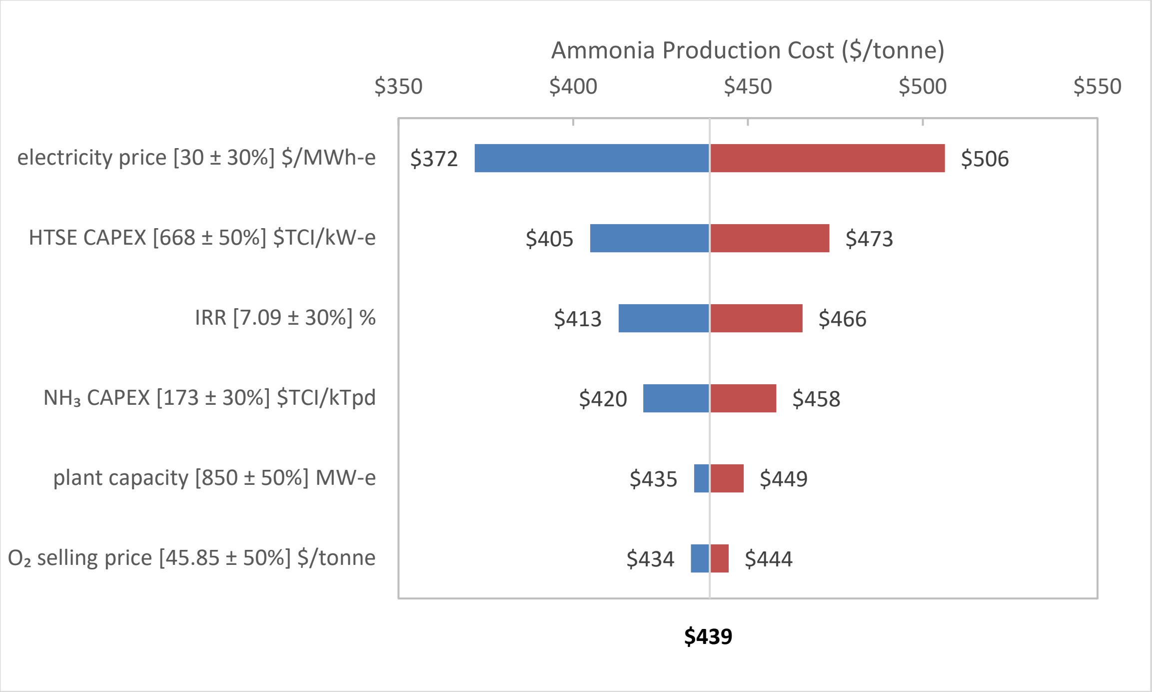

Figure 71. Sensitivity of ammonia-production cost to electricity price, HTSE and NH_3 CAPEX, plant capacity, and IRR. Oxygen sales come from the separated oxygen from the air separation unit as part of the ammonia production process.

Figure 72 is a plot of predicted ammonia-production cost as a function of plant capacity and electricity price. This plot provides additional insight regarding the dependence of the ammonia-production cost to two variables that would be highly likely to inform the decision as to where to site an LWR-HTSE ASU ammonia-production facility. Note that the step changes in the curves are attributed to the step changes in capital cost associated with the requirement to add additional process trains as the plant NH_3 production capacity increases; i.e., production capacity for a single train is maximized (which maximizes the capital-cost economies of scale) at a production capacity of around 1525 tonne/day, and two smaller process trains, each of which has not achieved an optimal economy of scale, are then installed in parallel at a plant capacity greater than 1525 tonne/day. A similar transition from two to three process trains occurs at an NH_3 production capacity of 3050 tonne/day.

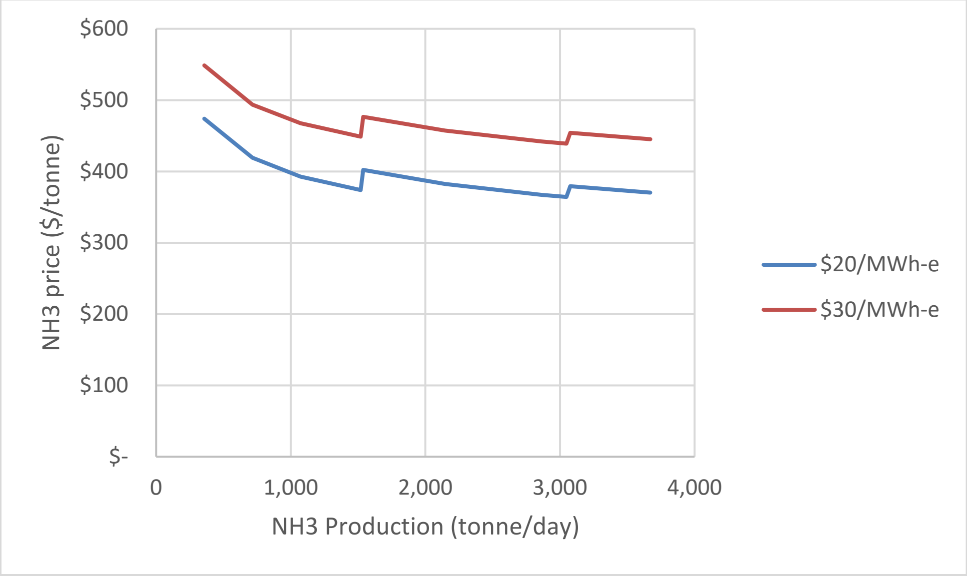

Figure 72. Ammonia production-cost sensitivity to plant capacity and electricity price.

5.3.4 Cost of Avoided CO_2 for Ammonia Production Using H_2 from Nuclear Energy

Ammonia process modeling and financial analysis were carried out in the above section for conventional and alternative ammonia production using NPP-(HTSE & ASU)plant. The ammonia-production rate and price for the alternative process were compared to the price of the conventional production process at different electricity costs. This difference was used to calculate the cost of avoided CO_2 per MT in the equation below.

$$Cost\ of\ avoided\ CO_2 \left(\frac{\$}{MT}\right) = change\ in\ ammonia\ price \left(\frac{\$}{MT}\right) \Big/ change\ of\ CO_2\ emissions \left(\frac{MT\ CO_2}{MT}\right)$$

This will be the cost that would make the NPP-integrated technology on price parity with conventional ammonia production pathway with the emissions of NPP integrated NH_3 production pathways as seen in the Figures below. The cost of avoided CO_2 for nuclear electricity cost of $20/MWh is between $5 and $28/MT of CO_2 and for $30/MWh is between $35 and $58/MT of CO_2, depending on the NH_3 production rate. The life cycle CO_2 emissions for both these pathways are discussed in section 2. The margins for transporting and selling ammonia were not included and could change the cost of CO_2 avoided when compared to the conventional baseline. For utilizing power from a single unit (520 MW) of PI NPP, approximately 1800 MT of NH_3 can be produced using the alternative ammonia synthesis process at a production cost of $570/MT NH_3.

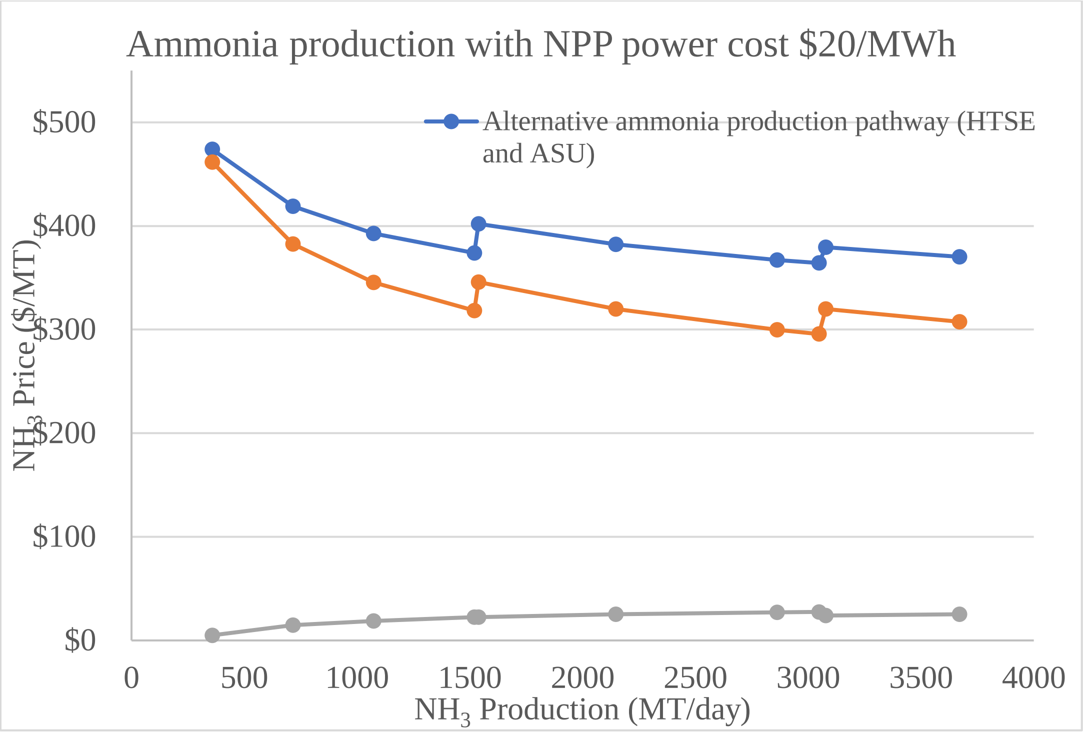

Figure 73. Conventional and alternative ammonia production price and cost of avoided CO_2 as a function of NH_3 plant capacity (based on an electricity price of \$20/MWh).

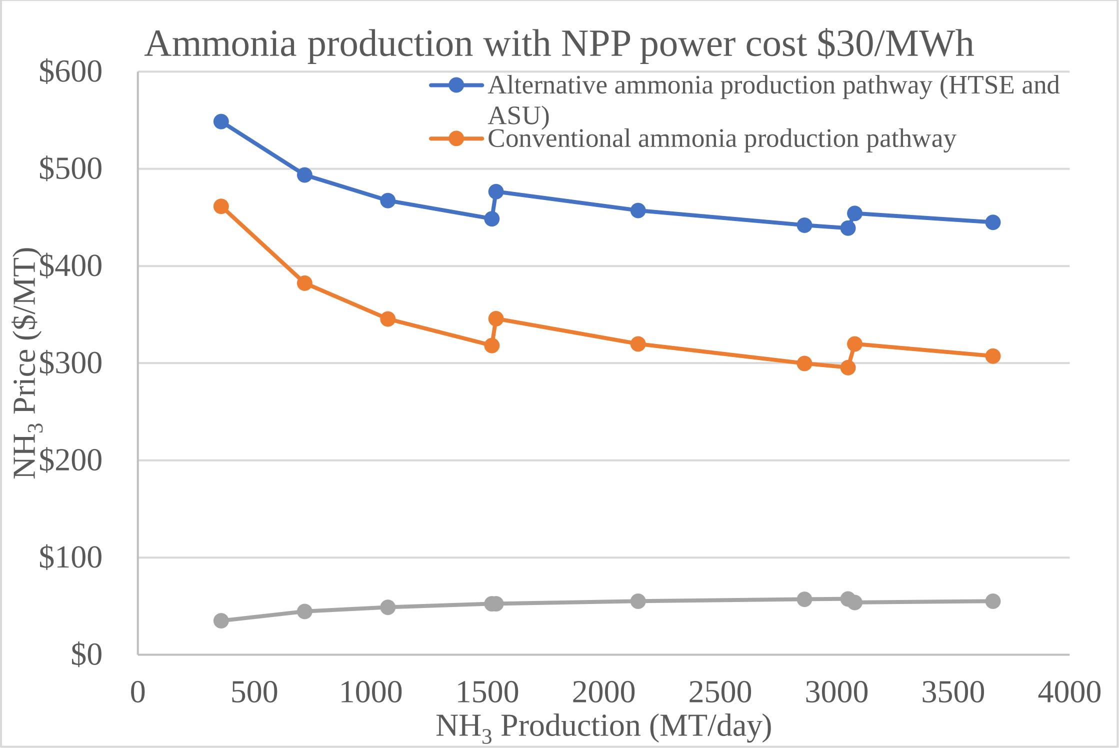

Figure 74. Conventional and alternative ammonia production price and cost of avoided CO_2 as a function of NH_3 plant capacity (based on an electricity price of \$30/MWh).

5.4 Synthetic Fuels

Synfuels and synchemicals are produced by synthesis from chemical building blocks, rather than from conventional petroleum refining. Syngas (carbon monoxide and hydrogen) is a common intermediate building block in the production of synfuels and synchems. Syngas can be produced by many processes, including biomass or fossil-fuel gasification and co-electrolysis. In co-electrolysis, CO_2 is reacted with water to produce syngas. The CO_2 can be sourced from processes that would otherwise eject the CO_2 to the atmosphere, but the cost of CO_2 capture depends on the purity of the source. Ethanol plants, including dozens of large plants in the U.S., ferment corn to make ethanol that is subsequently blended with motor gasoline. Ethanol plants provide a high-purity source of CO_2, they therefore provide a CO_2 source that is low-hanging fruit in terms of cost and ease of capture of large amounts of CO_2. Fossil-fuel processes, such as SMR and natural gas combined cycle (NGCC) power plants, can be other sources of CO_2, but these sources have much lower CO_2 concentrations than ethanol plants; therefore, the CO_2 is more expensive to separate and concentrate. CO_2 is also emitted from biofuels gasification plants.

Conversion of CO_2, which would have otherwise been released to the atmosphere, to synfuels using NE is a potential avenue for adding value to existing LWR facilities while producing lower-carbon-intensity transportation fuels, considered drop-in fuels: those that are compatible with conventional fuels produced via petroleum refining. In the case of CO_2 sourced from ethanol plants, this carbon comes from a biological source (corn), so the transportation fuels thus produced may be able to be considered as biofuels under the existing EPA Renewable Fuel Standard. In the case of CC and usage from fossil-fuel processes, such as SMR and NGCC, the making of transportation fuels using that carbon is giving the carbon a second life and avoiding new carbon release from fossil fuels coming out of the ground; thus, the carbon intensity of the fuels is reduced.

Low-carbon-intensity diesel and jet fuel could be synthesized using electrolysis powered by nuclear power. In the case of jet fuel, the aviation original equipment manufacturers (OEMs) have stringent standards for testing and approval of any fuel produced by a process other than conventional petroleum refining. The path has already been paved for the introduction of percentages of synthetically produced jet fuels with conventional jet fuel in the Specification ASTM D-7566. This specification is in addition to ASTM D-1655 Jet A/A1 Fuel Specification, which covers only jet fuels produced from conventional oil refining. To be allowed to be used to power aircraft, the jet fuel must not only meet the specifications of D-1655, but the process used to make the jet fuel must also be qualified by an ASTM board of aircraft OEMs. The qualification of the synthetic-fuel process is rigorous and analyzes not just the composition of the fuel in detail, but also the process used to make the fuel and any possible contaminants or incompatibilities that could be introduced to the aircraft fuel system which might cause previously unknown issues. After approval, the new synthetic-fuel process can be qualified and written as an annex into Specification D-7566 as an approved alternative jet fuel. Various synthetic-jet-fuel pathways have been approved and included in D-7566, including FT hydroprocessed, synthesized paraffinic kerosene (SPK) and alcohol-to-jet (ATJ) in mixtures of up to 50% with conventionally refined jet fuel. The following sections analyze the ATJ and FT pathways.

5.4.1 Co-electrolysis and Methanol/Ethanol Routes to Synfuels Analysis

A TEA and comparison of two possible synfuel production routes using CO_2 as the feedstock and the co-electrolysis process has been performed in a separately funded effort and is summarized here.[66] The high-level conclusions are summarized for completeness.

Heat from an LWR nuclear plant was integrated to the process wherever possible to positively affect the economics of the LWR by converting power to fuels during times of low grid demand for electricity. Process and economic modeling for a conceptual synfuel production plant co-located with, or in near proximity to an LWR was presented, including the cost of CO_2 captured from an ethanol plant, compressed, and transported to the LWR hybrid plant, co-electrolysis of the CO_2 with water in a SOEC system to produce syngas, and thermocatalytic conversion of the syngas to transportation fuel. The hybrid

LWR/synfuels plant was assumed to be located within 50–150 miles of an ethanol plant (e.g., located in the Midwest region of the U.S.). Performance and NOAK plant economics for the co-electrolysis-based processes were evaluated and compared with biomass-gasification-based technology for the synfuel routes considered. Sensitivity analysis around the price of CO_2 and electricity, two of the major cost drivers, was presented for each case. Consideration of a carbon credit is also included in the sensitivity analysis. The primary results and conclusions of the analysis are the following:

For a plant producing 3,195 barrels per day (BPD) hydrocarbon synfuels via a methanol intermediate with LWR electricity and steam usage of 326 MWe and 133 MWt respectively:

- The modeled minimum fuel selling price (MFSP) of diesel (91%) and gasoline (9%) blend stock with conservative assumptions is \$4.45/gallon for the base case using a CO_2 cost of \$33.3/tonne and an electricity price of \$30/MWh. This is compared to the biomass gasification route to syngas, with its MFSP of \$3.28/gallon. Note that co-electrolysis has a much larger maximum scale of production that can be reached compared to the availability of land competing with food production as in the case of biomass gasification. Also, the scale of the analysis is only about 1/3 of the available energy from a typical LWR but was chosen so that a direct comparison with a biorefinery could be made.
- There are innovative cryogenic carbon capture (CCC) processes that claim to produce CO_2 for a cost as low as \$20–60/tonne CO_2 (SES 2020), which could also have a significant impact on the viability of an LWR/synfuels plant using methanol as the intermediate. Further, the refrigerant used in the CCC process could be produced using LWR energy. The synergies of the LWR with the CCC process and technoeconomic modeling of the CCC process will be explored in detail in future studies.
- Sensitivity analysis (Figure 75) shows that, with optimal CO_2 and electricity prices and inclusion of carbon credits through incentives or mandates, this process could be more cost competitive with petroleum fuels, especially after COVID-19, when oil prices recover somewhat from the current historic lows. With a hypothetical carbon tax of \$100/tonne CO_2, the MFSP is reduced to between roughly \$3.50 and \$3.75/gallon. A renewable-fuel standard (RFS) credit would further aid in the competitiveness of fuels produced via this route. Some states already offer credits for clean fuels, including California and New York. These credits are qualified under the U.S. Environmental Policy Act and are applied to select fuels with the assignment of renewable identification numbers (RINs). Clean-fuel credits in California have ranged upwards from \$0.5 to \$2.5 per gallon of gas equivalent.
- Sensitivity analysis, varying plant scale for the co-electrolysis with a methanol-to-olefins (MTO) fuel process was conducted (Figure 76). At a scale of half the base case (326 MWe; 133 MWt), production cost increases by 9%. At a scale 10 times larger than the base case, the production cost is reduced to about \$3.8/gal. Scaling of the plant up to the entire electrical output of a general 1-GWe LWR of fuel production would result in about 40 cents/gal cost savings. Note that a scaling factor of 1 is assumed for the SOEC stack; therefore, no benefit is gained for this portion of the capital cost.

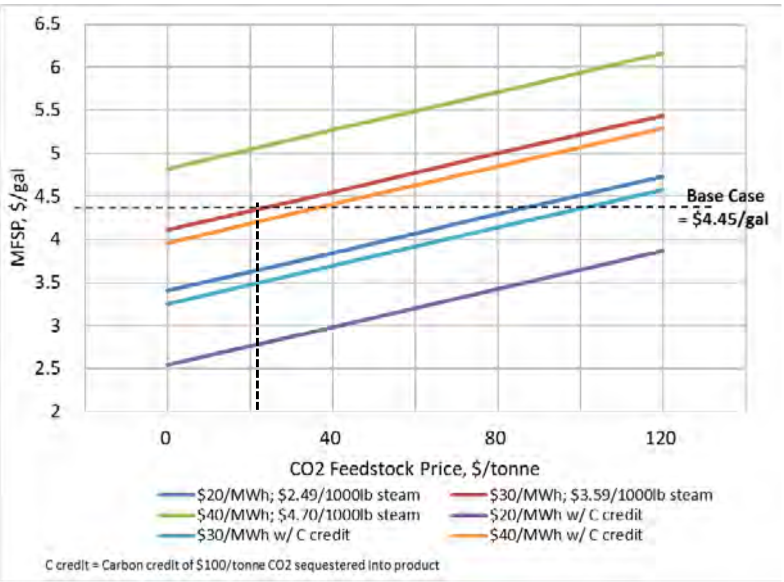

Figure 75. Minimum syn-diesel fuel selling-price sensitivity for fuels via a co-electrolysis-derived syngas to methanol to fuels route. Base-case diesel MFSP $4.45/gal at $33.3/tonne CO_2 feedstock price and $3.59/1000 lb steam price.

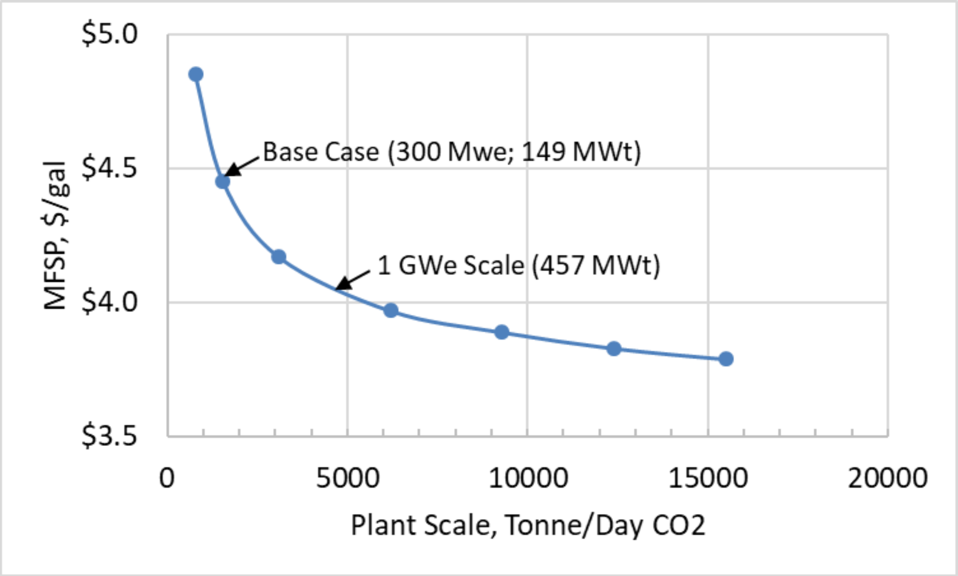

Figure 76. Sensitivity of syn-diesel MFSP to plant scale for the co-electrolysis and MTO fuel process. Electricity (MWe) and thermal power (MWt) were taken from the LWR for use in co-electrolysis and are provided in parentheses.

Co-electrolysis could take all of the energy provided by a single reactor or two reactors to produce the syngas that is converted to methanol. The synfuels could be competitive in price with petroleum fuels if credits for CO_2 emissions reductions reach about $100/tonne CO_2 or when the price of petroleum fuels rises above the current historic lows. The combination of plant scale-up matching the energy produced by an average NPP, together with clean-energy credits, could make synthetic fuels produced by co-electrolysis using LWR energy competitive with petroleum-derived fuels. Together, biomass gasification and nuclear-derived synfuel could feasibly replace a significant volume of U.S. transportation fuels. The nation currently burns 12 million barrels of gasoline and diesel each day. Biomass gasification and co-electrolysis together can feasibly replace over 25% of petroleum fuels.

Future studies should take into account opportunity sources of CO_2, their purity, and location, financial investment terms and options, and clean-energy credits. In addition, synergies between NPPs and the biomass gasification synfuels route should be considered, including drying and torrefaction of biomass feedstock and CO_2 by-product from biomass gasification.

5.4.2 HTSE, RWGS, and Fischer-Tropsch Route to Synfuels Analysis

The following analysis has been completed and represents an alternative to the co-electrolysis route to syngas and synfuels. This analysis produces syngas by using separate hydrogen and carbon dioxide feedstocks. The hydrogen in these scenarios is assumed to come from an HTSE plant integrated with a NPP, the design of which was previously explained in this report. The reverse water gas shift (RWGS) process is a well-known and optimized industrial process in which CO_2 and H_2 are reacted to form CO and water. By combining the HTSE and RWGS processes, the desired concentrations of CO and H_2 (syngas) can be produced for subsequent conversion to synfuels via the FT process.

A technoeconomic analysis of FT liquid-fuel production from H_2 and CO_2 has been reported by Zang et al, with the H_2 recycle pathway having better techno economic results.[67] MFSP was estimated for a broad range of H_2 and CO_2 prices and potential CO_2 credits. H_2 prices were reported to have the largest impact on the MFSP of FT fuel. The analysis showed that a hydrogen price of $0.8/kg was cost competitive with a pretax petroleum-diesel price of $3.1/gal in 2050 (without CO_2 credits). A CO_2 feedstock price of $17.3/metric ton was used in this analysis. When the H_2 price was set to $2.0/kg in the analysis (i.e., at the 2020 DOE target for hydrogen from electrolysis), the minimum selling price of the FT fuel was $5.4–5.9/gal. Figure 77 shows on the left the price of H_2 ($0.8/kg) that would make synthetically produced diesel fuel via the process herein described competitive with conventional diesel fuel at the projected 2050 diesel price. The figure also shows sensitivity studies and the resulting FT fuel price for LCOH of $2.0/kg and $5/kg.

An optimized case in preliminary-analysis phase suggests that the competitive price for H_2 may be well above $0.8/kg. This analysis is discussed in the following section.

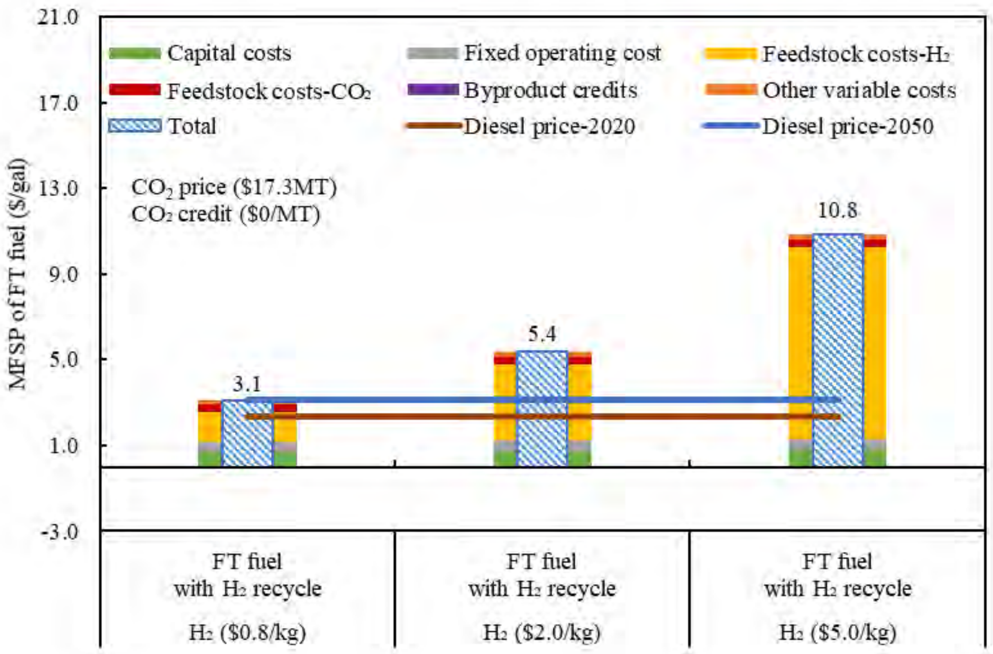

Figure 77. H_2 price effect on minimum fuel selling price of FT fuel production (base case).

Figure 78 shows similar information as is described above, but in trendline form, with the MFSP of diesel contrasted with the cost of hydrogen production.

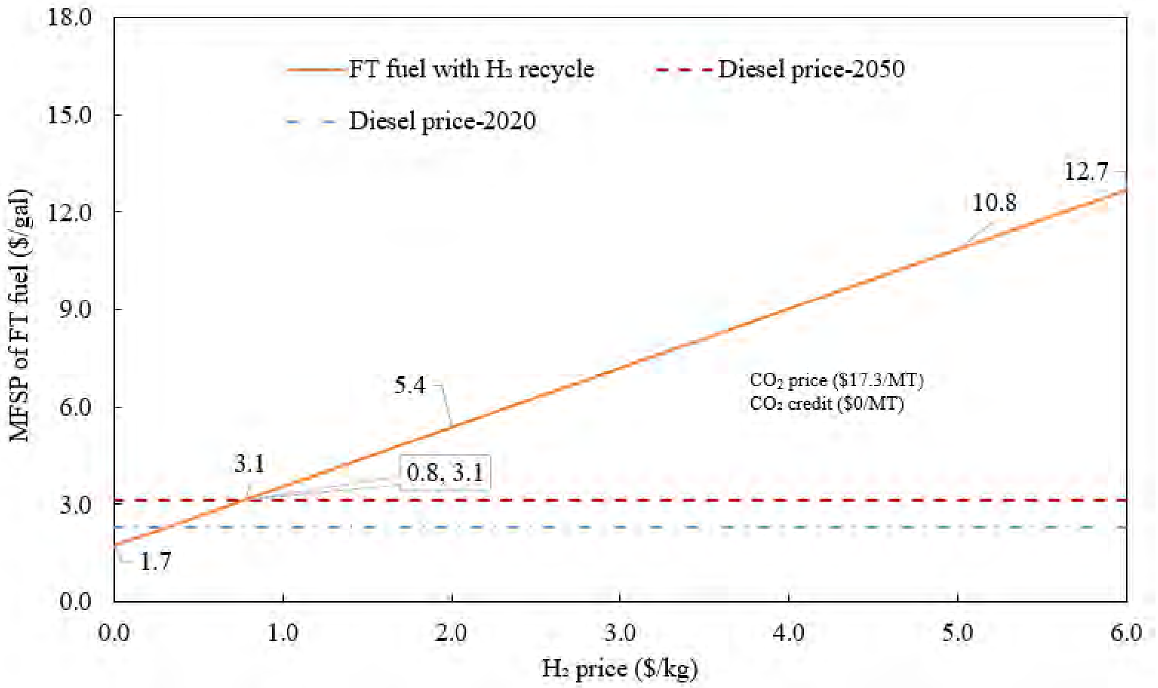

Figure 78. Break-even price for FT fuel production (base case) when compared to diesel prices in 2020 and 2050.

5.4.3 Optimized FT Fuel Production Scenario (Advanced Case)

Preliminary modifications (referred to herein as the advanced case) were implemented to the FT fuel production base case to increase the energy and carbon-conversion efficiency. Modifications included 1) recycling of light-gas from the separator to the FT reactor, 2) capturing pre-combustion CO_2, and 3) modifying the light-gas separation process. These changes increased the carbon-conversion efficiency from 45.5 to between 91.0 and 93.5% and the energy efficiency from 53.3–57.5% to 66.1–67.9% from the base to the advanced case.

The life-cycle GHG emissions are shown in Figure 79 for the advanced case of FT fuel using nuclear H_2, which is about 7 g CO_2 eq./MJ versus 9 g CO_2 eq./MJ for the base case Section 2.

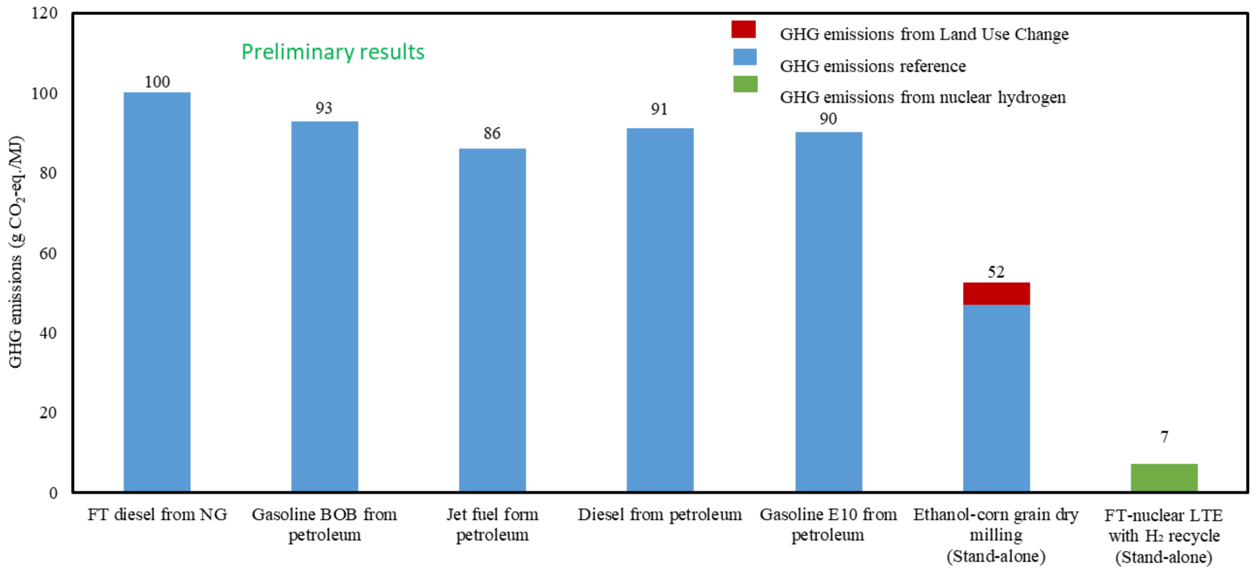

Figure 79. Life-cycle GHG emissions for FT fuel using nuclear H_2 (advanced case).

Due to the above-mentioned improvements to process modeling and the resulting higher energy and carbon-conversion efficiency, the break-even H_2 price, compared with untaxed 2050 diesel, is \$1.14/kg versus \$0.8/kg in the base case. The MFSP of the FT fuel reduced from \$3.1/gal to \$2.5/gal for the advanced case using a H_2 price of \$0.8/kg. Similarly, a hypothetical H_2 price of \$2/kg MFSP for FT fuel reduced from \$5.4/gal for the base case to \$4.2/gal for the advanced case. These results are presented in Figure 80. The CO_2 credit shown in figure below is the amount of credit required to breakeven with the diesel price of 2050, essentially the cost of avoided CO_2.

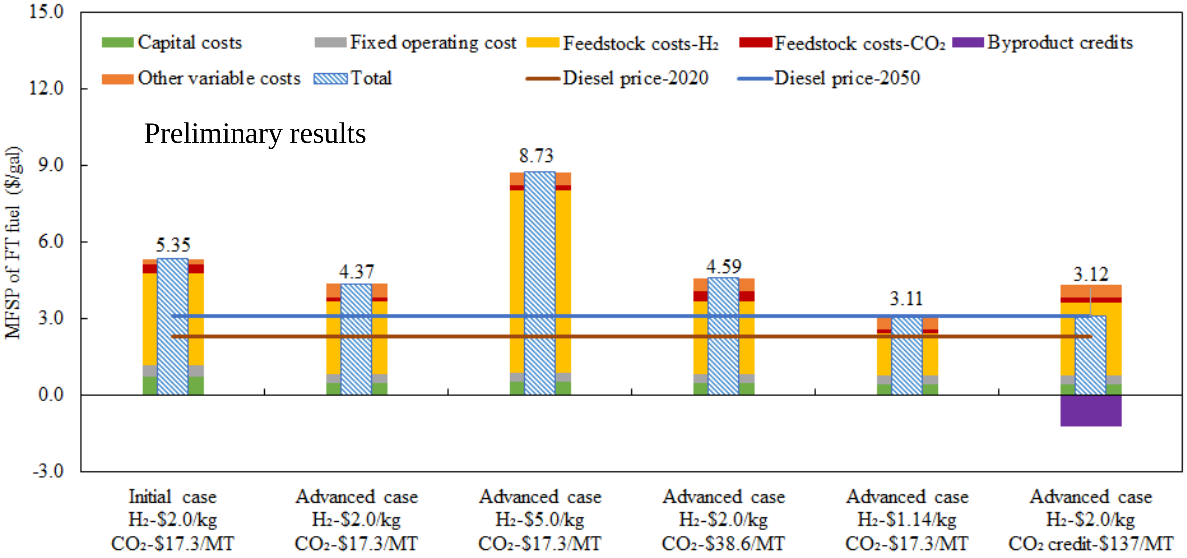

Figure 80. H_2 price effect on minimum fuel selling price of FT fuel production.

The break-even H_2 price for untaxed jet fuel in 2020 is a mere \$0.02/kg due to very low jet-fuel prices while the break-even H_2 price for jet fuel in 2050 is \$0.66/kg, as plotted in Figure A-5. The cost of avoided CO_2 was calculated using the equation below and was compared with untaxed diesel (2050)

\$3.1/gal and untaxed Jet fuel (2050) \$2.6/gal. The cost of avoided CO_2 (using nuclear H_2) for diesel (in the advanced case) in 2050 will be \$137/MT CO_2, and for jet fuel (in the advanced case) in 2050, it would be about \$200/MT CO_2.

$$Cost\ of\ avoided\ CO_2 \left(\frac{\$}{MT}\right) = change\ of\ fuel\ price \left(\frac{\$}{MT}\right) \Big/ change\ of\ CO_2\ emissions \left(\frac{MT\ CO_2}{MT}\right)$$

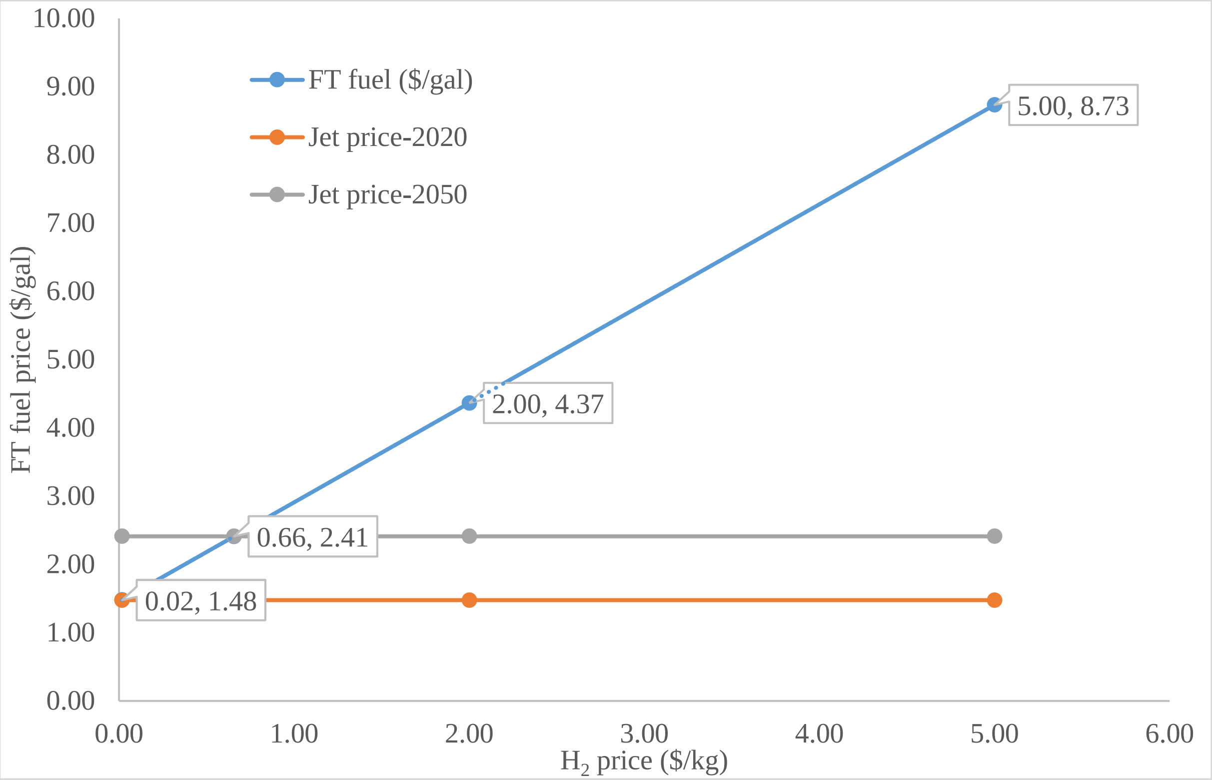

Figure 81. H_2 Break-even price for jet fuel for the advanced FT fuel production case.

5.5 Cryogenic Refrigerant Cycle

A separately funded analysis is in progress to show how NE could be used to power a cryogenic-refrigerant cycle during times of grid overgeneration. This could be thought of as a form of energy storage in the form of a usable cryogenic refrigerant, though the intent is not to go backward to create electricity again, but to store the refrigerant for use onsite as needed or to be transported short distances to point of use. The refrigerant could be used for CCC, hydrogen-compression interstage cooling, hydrogen liquefaction, or air separation for ammonia production.

The captured CO_2 could be used as a feedstock for synthetic fuels production, as has been discussed in this report. It would aid in the decarbonization of fossil-energy sectors.

5.6 Chlor-Alkali

A recent study highlighted the potential integration opportunities of various industries with nuclear reactors, including chlor-alkali.[68] Chlorine gas and sodium hydroxide (i.e., caustic soda or NaOH) are produced electrochemically from NaCl-rich brine, and their production represents around 4% of total industrial CO_2 emissions. The chlor-alkali industry manufactures over 11 MMT[69] of chlorine per year, along with a stoichiometric quantity of NaOH. Chlorine is used directly as a chemical or is incorporated into petrochemical products such as polyvinyl chloride (PVC). Strongly alkaline NaOH is widely applied in industrial processes, and concentration of the NaOH product via evaporation is the major thermal demand in chlor-alkali facilities. Total industry revenue is estimated to be $8 billion/year,[70] and projected compound annual growth rate (CAGR) is strong at more than 4% through 2025.[71] If this growth rate is consistent through 2030, over 6 MMT/year of added capacity will be required, indicating a need for new chlor-alkali facilities. A large chlor-alkali electrolysis unit could be in large demand for carbon-free energy from a new LWR-centered energy industrial park.

Chlor-alkali electrolysis plants show strong technical potential for integration with an NPP for both heat and electricity. The average chlorine facility requires roughly 140 MWe and 25–35 MWth.[72] Modern chlorine-electrolysis cells consume around 2,500 kWh/tonne Cl_2 produced.[73] The heat duty is mainly used to evaporate water and concentrate the NaOH product. Plants producing aqueous NaOH at 20–35% by mass require less heat than do facilities delivering anhydrous NaOH pellets. The representative temperature for concentration process heat is 175°C, which can be readily provided by an LWR. A large facility producing 2,800 thermal power delivery (TPD) Cl_2 (~0.84 MMT/year) would consume over 300 MWe and 75–100 MWth, or potentially more than 1 GWth total of NPP energy. Based on current growth projections, 5–10 plants of this size could be required within the next 10 years. Chlor-alkali production integrates well with LWR coproduction of electricity and heat, and a large facility could serve as a key demand source for an industrial park, consuming >1 GWth of total NPP energy.

Most chlorine is used for industrial processes, including around 40% for PVC. Less than 5% of chlorine is used for water treatment and pharmaceuticals; the remaining fraction is used to synthesize a wide array of other chemicals. A recent analysis found that chlorine plants are well distributed throughout the country, with a concentration in the Gulf Coast region to provide chlorine to industrial processes.[74] Locating a new chlor-alkali facility near industrial demand is therefore likely to be a driving cost consideration. LWRs in the Southeast region (for access to the Gulf Coast) as well as the Midwest (for access to manufacturing) could be strong candidates for integration with a new chlor-alkali plant.

5.7 Formic Acid

Formic acid (FA) can be produced economically using an electrochemical process by baseload low-carbon NPPs using CO_2 from sources such as local ethanol plants and even SMR plants. This has the potential to be game changing for the chemical industry. FA could serve as a durable liquid (at ambient conditions), and energy-dense hydrogen carrier that could be produced by electrolysis (co-electrolysis of CO_2 and water to make FA).[75,76]

5.7.1 Overview of the Current and Near-Future Formic-Acid Market

FA is defined as a high-volume chemical, with global production totaling 1.2 MMT per year.[77] The total market value is estimated to be $1.1 B, indicating a global benchmark price of $1.00/kg; U.S. prices are ~25% higher.[78,79,80] U.S. demand is currently relatively small, around 0.125 MMT/year or ~10% of the global demand, 40% (i.e., 0.05 MMT/y) of which is produced in a single facility in Geismar, Louisiana, operated by Badische Anilin- und SodaFabrik (BASF).[81] The remaining 60% of demand is supplied by imports from China and Germany. Worldwide growth is strong—estimated at nearly 5% CAGR through 2027—and domestic-growth estimates are even more optimistic, ranging from 5–14% CAGR due to emerging applications for FA.[82] Drivers for growth are

1. Current commodity chemical use switching from a U.S. import to a U.S. export
2. Use as a silage preservant to reduce the need for animal antibiotics
3. Fracking/drilling-completion fluids
4. Displacement of phosphoric acid for cleaning and descaling applications[83]
5. Breaking down and hydrogenating carbonaceous (biomass) feedstock into high-value chemicals and fuels

6. Use of a liquid-hydrogen carrier that is easily stored and transported to distributed users.

At this growth rate, the global market will grow to ~2 MMT/year of capacity by 2030.

Conventional processes produce FA via carbonylation of methanol or carbonylation of oxalic acid. Electrochemical catalytic co-electrolysis of CO_2 and water to make FA is a promising emerging process, and one manufacturer, OCO Chemicals, boasts of a 78% efficient process with high selectivity (99%) with their licensed process that reduces CO_2 with *in situ* generated hydrogen from water to FA or formate salts.[84]

Currently, FA is used to make natural and synthetic leathers, textiles, cleaning products, and rubber, and formate salts are used as deicing agents and additives in oil and gas drilling fluids.[85] Abroad, the major use of FA (accounting for 40% of demand) is as an antimicrobial additive in animal feed, but this practice is uncommon in the U.S. due to "generally low commercial availability of formic acid."[86,87,88] Domestic farmers use antibiotics instead, a practice which has come under scrutiny, presenting an opportunity for increased FA production and use.[89] FA is chemically stable and relatively nontoxic (at concentrations below 90%), making it an attractive chemical product for farming applications. The Food and Drug Administration has denoted FA as "generally recognized as safe," and the European Union (EU) has certified it as a permissible additive for both human and animal food.[89,90] The Midwest (particularly Iowa, Minnesota, and Wisconsin) contains clusters of industrial cattle and pig farms while the Southeast is a center for chicken farming.[92] NPPs in these regions making liquid products like FA would have straightforward access to these markets. FA replacement of antibiotics in animal feed, whether motivated by regulation, public scrutiny, or price, would lead to a significant increase in domestic FA demand.

Other possibilities for replacing current chemicals, specifically acetic acid (AA), are also promising, although at a smaller scale than animal feed. Roughly 15 MMT/year of AA are manufactured worldwide, most commonly to produce vinyl acetate monomers, food-grade vinegar, acetic anhydride (an industrial solvent), and acetate salts. FA cannot replace AA as a monomer or in vinegar, but BASF markets FA as a replacement for AA (and acetate salts) in deicing agents and solvents, indicating the possibility of expanded FA demand.[92] In particular, potassium salt deicing agents represent a growth area, as potassium formate has already replaced potassium acetate for deicing at European and North American airports.[89]

In addition to its use by traditional industries, FA also holds substantial promise as a hydrogen carrier for direct or indirect use in fuel-cell technologies. FA is both energy- (1,760 Wh/kg) and hydrogen-dense (53 g/L, 44 g/kg), containing more hydrogen per volume than compressed hydrogen itself (at moderate pressures).[83,85] It is a liquid at ambient temperature, stable, nontoxic, and durable enough for long-term storage. Further, hydrogen release from FA is exergonic (<0 free-energy change) but not exothermic (>0 enthalpy change). This allows hydrogen release to be performed at low temperatures, but perhaps more importantly, at high pressures that may be suitable for storage in fuel-cell vehicles. Other hydrogen carriers (e.g., ammonia and methanol) do not have this property, and it has been estimated to reduce storage, compression, and dispensing costs of a hydrogen refueling station by 60–70%.[94] These properties have led to increased interest in FA as a potential hydrogen carrier.[95,95] Growth of this market, combined with technological advances in co-electrolysis, could see FA become a major industrial chemical in the long term.

5.7.2 Analysis of Nuclear Power Plant Facilities and the Formic Acid Market

With substantial growth in the market, co-electrolysis to generate FA could play a revenue-generating role in a multipurpose NPP-associated facility. A single 1 GW NPP, using currently available low-efficiency co-electrolysis technology, could produce more than the present global demand for FA each year. INL has previously estimated the required energy input for electrolytic FA synthesis at ~4 MWh/1000 kg, assuming a large overpotential (>2 V) to increase cell current densities.[95] One gigawatt of constant electricity input could therefore produce the current annual demand of FA (1.2 MMT) in

~5000 h, or around 7 months. Assuming an electricity price of \$25–40/MWh, the energy input costs are \$0.10–0.22/kg FA. Assuming amortized capital expenditures of \$0.20–0.25/kg as estimated by OCO, the cost of electrochemical FA production is ~\$0.30–0.47/kg.[85] These cost targets would make FA cheaper than many alternative chemicals, opening up new markets, such as silage preservation, cleaning agents, and chemicals processing. To achieve these goals, research and development are needed to increase the efficiency and current density of co-electrolysis cells, which would reduce both operating and capital costs. If cell efficiencies are increased, and the market grows significantly, particularly through the adoption of FA as an H_2-energy carrier, FA production would be very well suited for NPP facility integration, especially when configured as a component of the energy industrial park concept discussed above.

6 SUMMARY OF HYBRID OPTIONS INTEGRATION WITH NPPS

In Table E-1, the hybrid options analyzed in this report are tabulated in order of the cost of avoided carbon from lowest to highest with the TRLs for these options. When carbon reductions are desired, this table provides a useful guide to show those processes that would cost the least, as far as a possible carbon credit is concerned. If there is no credit, then this is the real cost to decarbonize based on the assumptions in this report.

The cost of avoided CO_2 is calculated using the equation below for each application listed in the table. Life cycle CO_2 emissions for all these applications were evaluated in Section 2 wherein NE was considered as an integral power source for all these different pathways. Cost of avoided of CO_2 is very sensitive to underlying assumptions such as natural gas prices, nuclear electricity prices, etc.

$$Cost\ of\ avoided\ CO_2\left(\frac{\$}{MT}\right) = change\ of\ application\ price\ \left(\frac{\$}{MT}\right)\Big/ change\ of\ CO_2\ emissions\ \left(\frac{MT\ CO_2}{MT}\right)$$

Table 43. Hybrid options for integration with LWRs ranked in order or least cost of avoided CO_2 to greatest.

Nuclear-H_2 Applications	Cost of Avoided CO_2 (\$/MT CO_{2e})	TRL (basic = 1, fully commercial = 9)	Notes: Nuclear Electricity Price Assumed to be \$30/MWh, Nuclear-H_2 at \$1.93/kg and natural gas pricing based on Modified 2021 AEO West North Central (WNC) Region Reference Case
Ammonia	\$35–58	8-9	Compared ammonia production facility using nuclear power for air separation unit for N_2 and HTE for H_2 to a conventional ammonia production plant at different production rate. (Section 4.3).
Refineries	\$100	9	Compared nuclear-H_2 to H_2 from natural gas SMR at 1.03/kg.
Synfuels	\$137 (Diesel) \$200 (Jet fuel)	2-3	Compared advanced synfuel production prices to untaxed diesel prices at \$3.1/gal (2050) and untaxed price of jet fuel \$2.6/gal (2050).
natural gas-H2 blending	\$135–172	6-7	Compared nuclear-H_2 to energy equivalent price of natural gas on HHV Btu basis. This cost of avoided CO_2 is for the range of natural gas prices for natural gas electricity generators in the Minnesota's Twin Cities region.
FCEVs	\$55–270	9	Compared H_2 \$5–7/kg (at refueling station), per DOE H_2 fueling cost target, to untaxed gasoline price in 2050 (\$2.96/gal), the cost of avoided carbon is very sensitive to H_2 prices.

Table 44. Hybrid options for integration with LWRs ranked in order or least Cost of Avoided CO_2 to greatest (Advanced HTSE case for H_2 production).

Nuclear-H_2 Applications	Cost of Avoided CO_2 (\$/MT CO_{2e})	TRL (basic = 1, fully commercial = 9)	Notes: Nuclear Electricity Price Assumed to be \$30/MWh, Advanced HTSE Nuclaer-H_2 at \$1.53/kg and natural gas pricing based on Modified 2021 AEO West North Central (WNC) Region Reference Case
Ammonia	\$13–33	8-9	Compared ammonia production facility using nuclear power for air separation unit for N_2 and HTE for H_2 to a conventional ammonia production plant at different production rate. (Section 4.3).
Refineries	\$55	9	Compared nuclear-H_2 to H_2 from natural gas SMR at 1.03/kg.
NG-H_2 blending	\$100–136	6-7	Compared nuclear-H_2 to energy equivalent price of natural gas on HHV Btu basis. This cost of avoided CO_2 is for the range of natural gas prices for natural gas electricity generators in the Minnesota's Twin Cities region.

7 CONCLUSION

Demand analysis forecasts for hydrogen were completed for the U.S. national market as well as the regional market in the greater Minneapolis area. The forecasted U.S. national demand is shown in the table below. The regional market for the greater Minneapolis area around Prairie Island and Monticello NPPs and the associated demand curves were summarized at the end of Section 2. Demand forecast data for this region is found in Appendix F.

Table 45. Summary of forecasted U.S. hydrogen demand potential in 2050.

Application	Potential Hydrogen Consumption[97] [MMT/yr.]
Petroleum Refineries	7.5
Ammonia	2.5
Synfuels from ethanol-CO_2	5
Injection to natural gas infrastructure	16
Iron reduction and steelmaking	8
Light-duty FCEVs (cars and trucks)	21
Medium-duty + heavy-duty FCEVs	8

An LWR-HTSE process evaluation was herein presented. The evaluation determined that an HTSE, scaled to match the energy output of an LWR plant, would require approximately 5% of the LWR steam flow to provide the process-heat input needed to vaporize HTSE-process feedwater. The analysis specified the use of Therminol-66 as the HTF to transfer nuclear process heat a 1 km distance to the HTSE plant. The HTSE plant was determined to have specific energy requirements of 37.4 kWh-e/kg-H_2 and 6.4 kWh-t/kg-H_2. The HTSE plant efficiency was calculated as 88.9% on a HHV basis. Two SOEC technology cases were considered in evaluating HTSE LCOH:

1. The HTSE base case is founded on a stack-cost specification of $155/kW-dc, consistent with that reported for current SOEC technology in the DOE HFTO Hydrogen Production Record.[30] The base case also uses the HFTO Record stack service-life specification of 4 years. This analysis includes annual stack replacements to restore the HTSE plant design-capacity rating at the start of each operating year. The base case NOAK HTSE plant with a hydrogen production design capacity of 383 tonne/day (597 MW-e) has a DCC of $574/kW-ac (including HTSE plant equipment and NPP heat- and power-delivery equipment), and the total capital investment of $742/kW-ac includes project indirect costs in addition to DCC listed above. When energy from the LWR is purchased at a price of $30/MWh-e (the nuclear plant thermal efficiency is used to derive corresponding thermal-energy price), the base case HTSE plant is able to produce hydrogen at an LCOH of $1.93/kg; however, this does not include product storage or transport costs.

2. The HTSE Advanced Case uses a stack module (i.e., stack plus balance-of-module) cost specification of $35/kW as derived from the data presented in [44]. The advanced-case stack service life is specified as 7 years, which is consistent with current best-in-class SOEC technology. The advanced case NOAK HTSE plant with a hydrogen-production design capacity of 383 tonne/day (597 MW-e) has DCC of $345/kW-ac and a total capital investment of $446/kW-ac. When energy from the LWR is purchased at a price of $30/MWh-e, the advanced case HTSE plant is able to produce hydrogen at an LCOH of $1.53/kg (excluding storage and transport costs).

The advanced HTSE case represents SOEC-vendor stack-pricing estimates and best-in-class cell-degradation-rate performance. It is expected that SOEC technology will be aligned with the advanced-HTSE case within the HTSE plant-construction schedule evaluated in this analysis (i.e., start of plant construction in 2026, with plant startup in 2027). The LCOH corresponding to the advanced LCOH is therefore most applicable for the purposes of comparison with natural gas SMR. The combination of decreased stack-module cost and increased stack service life enables the HTSE advanced case to achieve the specified \$0.40/kg reduction in LCOH relative to the HTSE base case across the range of electricity prices evaluated (Figure 53).

A sensitivity analysis evaluated the impact of several key-process and economic parameters on the HTSE LCOH. The upper and lower bounds for each of the input parameters were selected to correspond to expected technology advancement and/or variation in market conditions. Based on the selected range over which the sensitivity variables were perturbed, the parameters that have the greatest impact on LCOH are energy price and SOEC-stack cost. The second set of variables—including the learning rate (for decreases in modular equipment costs as a function of the number of units produced by the equipment manufacturer), stack service life, and IRR—have a medium impact on the LCOH. Once NOAK plant status has been achieved, defined as previous deployment of N = 100 count of 25 MW-e modular blocks, or 2.5 GW-e of production capacity, and a base plant capacity of several hundred MW is considered, perturbations to these variables have a less-pronounced impact on LCOH than the sensitivity variables identified above. Additional results and observations from the sensitivity analysis are listed below:

- Electricity price is major cost driver of HTSE LCOH. A decrease of \$10/MWh-e in the price of the energy obtained from the LWR would result in approximately a \$0.40/kg decrease in the HTSE hydrogen-production cost.
- Stack costs are also a major driver of the HTSE LCOH. The stack costs contribute to the initial plant construction costs as well as the HTSE plant O&M costs (for stack replacement). There is a significant difference between the values of the stack cost specified by DOE HFTO for a “current technology” hydrogen-production cost evaluation[30] versus the stack cost that specific SOEC vendors have reported would be possible using current technology with manufacturing capacity of several hundred megawatts per year.[44] Therefore, a prospective HTSE plant developer could significantly reduce uncertainties in hydrogen production cost by obtaining project-specific stack/system pricing information from SOEC vendors.
- The learning rate affects the HTSE plant modular-equipment capital costs. Variation in the learning rate of ±5% has a moderate impact on LCOH relative to the other sensitivity variables evaluated. Planned expansions in vendor-specific manufacturing capacity could affect the learning rate that is realized as establishment of large-scale SOEC manufacturing capacity continues in the coming years.
- Provided an NOAK HTSE plant is installed at large (i.e., several hundred megawatts) scale, scalable plant components (e.g., nuclear process-heat delivery, electrical-power distribution, utilities) will have achieved sufficient economies of scale, and modular HTSE process components will have obtained cost reductions through economies of mass production. Therefore, there is a relatively minor impact to the LCOH from the HTSE plant-capacity specification over a range from several hundred megawatt up to gigawatt-scale HTSE.

A comparison of LWR-HTSE and natural gas SMR LCOH was performed to identify cases where HTSE could produce hydrogen at a cost competitive with SMR. The SMR LCOH is highly dependent on natural gas pricing. Use of a modified 2021 AEO WNC Region Reference Case natural gas price projection results in an LCOH estimate of \$1.03/kg for an natural gas SMR plant with a design capacity of 380 tonne/day (342 tonne/day actual production rate).

Because hydrogen produced via SMR is associated with significant carbon emissions, it is possible that hydrogen consumers would be willing to pay a price premium for carbon-free green hydrogen or that a price on carbon could increase the effective cost of SMR-derived hydrogen. The natural gas SMR LCOH is increased by approximately \$0.01/kg for every \$1/MT-CO_2 tax that is applied. Specifically, the calculations described in Section 3.3.2 indicate that a carbon tax of \$25/tonne-CO_2 would result in an increase in the natural gas SMR LCOH of \$0.22/kg. In addition to the electricity price and SOEC stack cost/service life, the presence of a CO_2 tax is one of the most significant drivers that could determine the profitability of hydrogen production via HTSE relative to SMR.

The analysis concludes that advanced HTSE technology (e.g., the advanced HTSE case), a low electricity price (e.g., the advanced HTSE case requires an electricity price of \$21.1/MWh-e to compete with natural gas SMR in the absence of a carbon tax), and/or a green-hydrogen production credit or carbon tax on CO_2 emissions from natural gas SMR would be required for HTSE to be cost competitive with SMR.

7.1 HTSE Optimization Summary

The findings of the NPP-HTSE optimization are:

- Adding an HTSE at Prairie Island is competitive at low capital costs. PTC is not needed when CAPEX is below \$300/kw for a few hydrogen delivery levels. CAPEXs above \$300/kW require carbon-free credit or PTC.
- HTSE at Monticello is a more difficult economic proposition. The smaller hydrogen market means that a PTC of at least \$0.6/kg to \$0.7/kg is required to make the system more profitable than BAU.
- The profitability of both systems is highly dependent on the hydrogen market that surrounds the plant. Providing too much hydrogen saturates the market and drops the price of hydrogen, making it difficult for the NPP with HTSE to compete with BAU.
- A matrix of profitable configurations was provided for both PI and Monticello. This matrix provides minimum requirements of HTSE CAPEX, hydrogen demand, and clean-hydrogen credit for the system to be profitable compared to BAU.
- The storage tank sizes for both systems fluctuate between 2–3 hours storage and 5 hours storage, depending on CAPEX. A higher CAPEX depresses storage because more storage requires a larger HTSE. One hour of storage is assumed to cover the required hydrogen demand for 1 hour.
- This analysis is from the perspective of a nuclear-with-HTSE plant. Extending this analysis to a systemwide approach to investigate the value of flexible operation on other generators and their deployment, and overall system cost would be advantageous. This could be done by extending this optimization structure to perform a dispatch with other generators modeled. Another approach would be to use a capacity-expansion model with PI and Monticello IESs included.

7.2 LWR Nuclear-H_2 Utilization Scenarios and Carbon Reduction

- The cost of NH_3 production per the NPP-HTSE-NH_3 plant analyzed in this report assumes an electricity cost of \$30/MWh. The cost of avoided CO_2 is also plotted to show the cost of decarbonization or the hypothetical carbon credit that would make the NPP-HTSE-NH_3 on parity with conventional ammonia production.
- The CO_2 reduction impact of hydrogen blending with natural gas in natural gas power plants was evaluated. A 30 vol% mixture of hydrogen with CO_2 results in just over 10% reduction in CO_2. This is because 30 vol% H_2 with natural gas represents only ~9% blending by energy because the volumetric heating value of hydrogen is approximately 30% of the corresponding heating value of natural gas. Although the potential greenhouse-gas (GHG)-emission reduction for this mixing ratio

appears small, the amount of potential CO_2 abatement is significant due to the large contribution of natural-gas generating plants to the U.S. national GHG-emissions inventory.

- Synthetic fuels production analysis using the HTSE + RWGS + FT pathway was discussed in the above sections. This chart shows that with the advanced synfuels case using the advanced HTSE case inputs, a CO_2 feedstock cost of $17/MT, and the 2050 diesel forecast price, synthetic diesel fuel via this pathway could be competitive with conventional diesel if hydrogen were produced at $1.14/kg.

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APPENDIX A

THERMAL POWER EXTRACTION FROM NUCLEAR POWER PLANTS

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APPENDIX A
THERMAL POWER EXTRACTION FROM NUCLEAR POWER PLANTS

A-1. Thermal Energy Transport Analysis

This appendix discusses the preliminary results of separately completed analysis to understand the costs of thermal-energy transport from an NPP versus heat from a new installation of a commercial natural gas boiler.[98] These analyses are ongoing and fast developing and as such the below summary is not to be considered state of the art. Heat extraction from the NPP and thermal-energy transport are key analyses for understanding the advantages of integration of industrial processes with nuclear power.

The transportation distance of heat via a TDL from the NPP was analyzed and compared to a natural gas boiler, and the break-even distance was found. The break-even distance bounds the approximate limit that thermal energy could be transported from an NPP and still be competitive with the natural gas boiler. Figure A-1 shows the approximate cost of heat from a stationary onsite natural gas boiler versus heat generated from an NPP and transported at various distances. The dashed black line shows the cost of the natural gas boiler heat. The blue, green, and purple lines, respectively, show the cost of NPP heat at $20, $25, and $30/MWhe, transported various distances. This analysis is preliminary and should be taken as a ballpark estimate of costs. A more-involved design analysis would need to be done for any specific project.

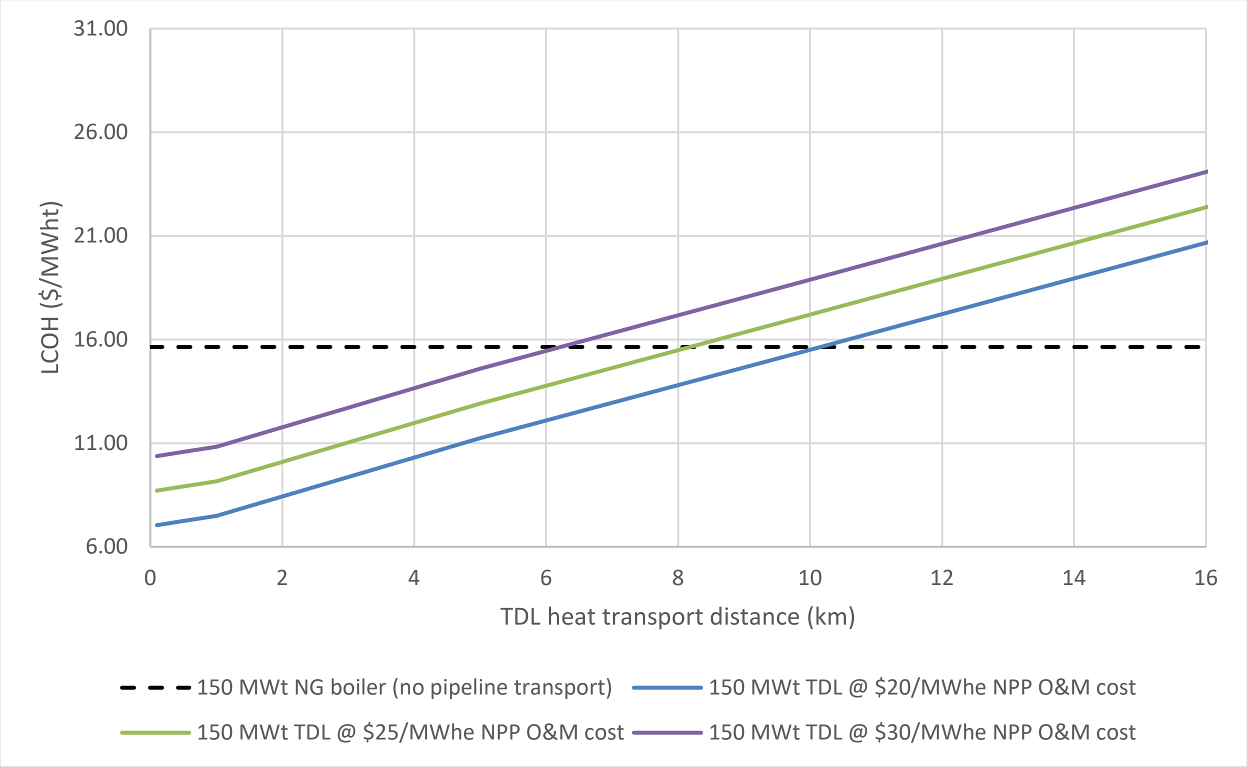

Figure A-1. LCOH versus heat-transport distance and NPP O&M cost. Plotted data points are based on a TDL capacity of 150 MWth and NPP O&M costs ranging from $20 to $30/MWhe. Assumes natural gas is purchased at a price of $4.04/MMBtu.

A-2. TPE Overview

Thermal-hydraulic modeling and NPP simulation have been performed using separate funding within the Light Water Reactor Sustainability (LWRS) Program at INL to support the development of designs for TPE from NPPs. That work includes modifying full-scope, high-fidelity PWR- and BWR-plant simulators to incorporate thermal- and electric power coupling to HTSE plants. It also includes performing hardware-in-the-loop and human-in-the-loop digital real time simulation (DRTS) tests using validated reduced-order NPP simulators, coupled to pilot scale thermal-hydraulic test systems that represent NPP operations and to greater than 50 kW HTSE units. These DRTS tests with robust NPP simulators and both hardware- and human-in-the-loop are necessary to understand the true coupling between NPPs, the bulk electric grid, and HTSE plants. The brief description below only summarizes the initial development of a full-scope, high-fidelity PWR simulator that was modified to incorporate thermal- and electric power coupling to a HTSE plant.

In 2020, the LWRS Program modified a full-scope generic pressurized water reactor (GPWR) simulator from GSE Systems (Sykesville, MD) to include TPE and delivery to an industrial user.[99] The boundary limits of the TPE simulator are shown by the dashed line in Figure A-2. The simulation includes: (1) a TPE line that extracts steam from the main steam line and passes the steam through extraction heat exchangers before returning the steam to the condenser and (2) a TPD loop that circulates synthetic heat-transfer oil between the extraction heat exchangers and a set of heat exchangers at the site of the industrial user (the first user tested will be a hydrogen plant), which may be as far as 1 km from the NPP. Rigorously simulating the modifications needed for electric power switching at the NPP switchyard and also simulating the complex dynamic behavior of the industrial user will be pursued in 2021.

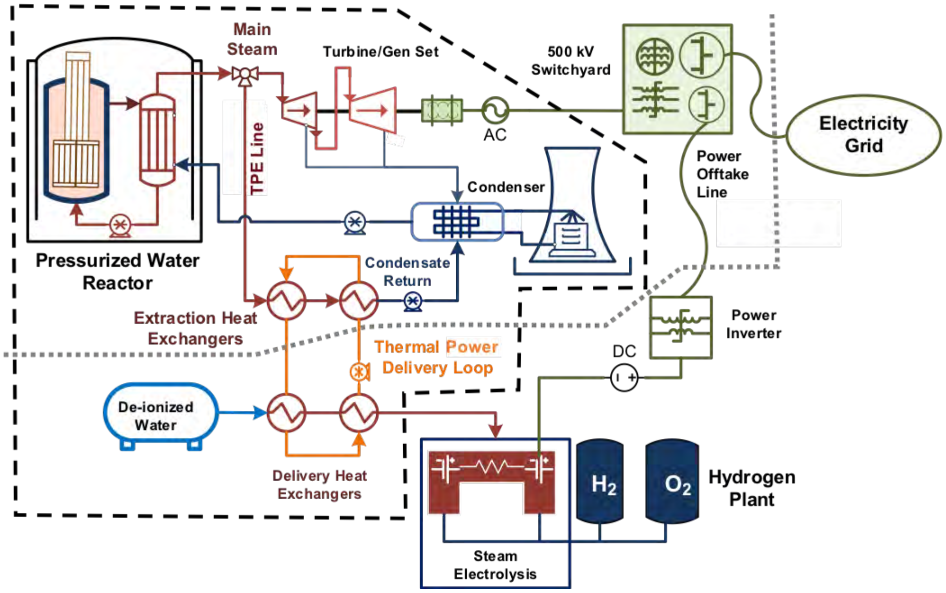

Figure A-2. Boundary limits of the thermal power dispatch GPWR simulator (dashed black line) and site boundary of the NPP (dotted grey line).

A-3. Requirements and Decisions

As noted above, a TPE system must be incorporated into the nuclear plant to transfer thermal power from the PWR to the hydrogen plant. The TPE system includes a TPE line that removes steam from the main steam header, passes this steam through extraction heat exchangers that condense the steam, and then returns the condensate to the NPP condenser. A separate TPD loop transfers the heat from the extraction heat exchangers to the industrial heat user, which may be located a kilometer or more away. Design requirements for the TPE system are summarized in Table A-1.

Table A-1. Design requirements for the TPE System proposed for a PWR.

	Design Requirement
1	Use of the TPE system or any connection to the industrial facility will not impact safety operations at the NPP
2	The nuclear plant operators must have full control of the steam flow in the TPE line with prerogative to completely stop steam flow without the possibility of interference from the industrial facility (such as a hydrogen plant)
3	Changes in the rate of steam diverted to the TPE system from 0 (0 steam flow) to 5% (2.9 105 kg/hr steam) of total thermal power must not cause total NPP thermal power to exceed 100% reactor power (2900 MWth)
4	Use of the TPE system must not adversely affect the existing updated final safety analysis report (UFSAR) design basis accidents (DBA) analyses (specifically, any effects on the step-load decrease transient)
5	The TPE system will be designed to allow switching of at least 90% of power delivery from the industrial facility to the electric grid in less than 10 minutes, such that the power to the industrial facility can act as a dispatchable load; the integrated system shall be capable of cycling power to and from the hydrogen plant at least two times in each 24-hour period

Design decisions that follow from the design requirements include:

I. The TPE System will extract steam from the main steam header (MSH) downstream of the main steam isolation valve (MSIV) so that the extraction point will be outside containment, but prior to the turbine throttle and governor valves, to provide steam with the highest possible temperature (decision to meet Requirement 1).

II. Isolation flow control valves (FCVs) will be installed in the TPE line that will be operable from the main control room to allow NPP operators to immediately stop the flow at any time (decision to meet Requirement 2).

III. Reactor controls will be modified such that the reactor remains between 98 and 100% thermal power while steam flow is increased or decreased in the TPE line, preferably without the use of control rods or adjustments to boron concentration in the reactor coolant (decision to meet Requirement 3).

IV. Steam flow rate in the TPE line is preferred as a control variable because TPE is directly proportional to steam-flow rate (this decision facilitates Requirements 2 and 3).

V. Reactor controls will be modified such that the control rods will not move during normal operations at 100% reactor power (2900 MWth) due to the operation of the TPE system (decision to meet Requirements 1 and 4).

VI. Steam in the TPE line will be fully condensed to liquid water in the extraction heat exchangers.

VII. Condensate from the TPE line will be returned to the condenser. Future work may also consider returning condensate, which has a temperature of approximately 193.3°C, to the feedwater heater system to increase efficiency.

VIII. A closed-loop TPD loop is used to transport heat to the industrial facility to maintain as much flexibility as possible.

IX. The HTF in the TPD loop may be either steam or synthetic oil (see below).

As noted in Design Decision IX, the HTF in the TPD loop may be either superheated steam or synthetic heat-transfer oil. Table A-2 summarizes the relative advantages and disadvantages of these options. Due to the high latent heat of steam, the mass-flow requirement to transport a given quantity of heat is lower for steam than for oil. This is important because the mass momentum of the HTF in a kilometer-long TPD loop will be a limiting factor in starting and stopping flow using valves and pumps. However, if steam is used in the TPD loop, it must be condensed at the delivery heat exchangers to prevent damage to pumps and other equipment in the return condensate line. If steam is used as the HTF, the possibility that an unexpected event could cause a sudden shutdown of the hydrogen plant will require a backup heat sink or heat-storage capacity to mitigate potential damage to the TPD loop pump. Because the vapor pressure of steam at the anticipated operating temperature is much higher than that of synthetic oil, a steam-based TPD loop would have a higher operating pressure. Thermal hydraulic models are developed for both HTF options, and detailed modeling results are presented. A final consideration is that the pump power requirements of a TPD loop with oil are approximately 20 times higher than for a TPD loop with steam, as will be discussed.

Table A-2. Relative advantages and disadvantages of using superheated steam or synthetic oil as the HTF in the TPD loop.

HTF Option	Potential Advantages	Potential Disadvantages
Superheated Steam	Low mass flow required due to the high latent heat High heat-transfer coefficients from phase change allow low approach temperatures Steam is compatible with steam in the main steam line in case of leaks across heat exchangers Low TPD loop pump power requirement (36 kW)	All steam in the TPD loop must be condensed at delivery heat exchangers to avoid damaging TPD loop pump and other equipment Vapor pressure of steam requires high operating pressure
Synthetic Oil	Low vapor pressure of synthetic oil allows low operating pressure Single-phase flow simplifies design and allows greater operational flexibility	High mass flow is required to transport required heat Synthetic oil poses an additional contamination risk if it reaches the condenser due to a leak in the extraction heat exchangers More expensive: the capital cost of the oil is expected to be in the range of $1,000,000 High TPD-loop pump-power requirement (784 kW)

A-4. Thermal Power Dispatch Design for Steam in the TPD Loop

Analysis have been performed for TPE-system designs that employ both synthetic oil and steam in the TPD loop. Only the simulator using steam in the TPD loop is discussed here for brevity. The piping and instrumentation diagram (P&ID) of the TPE line for the model with steam in the TPD loop is shown in Figure A-3.

The steam condition available for extraction at the MSH is saturated steam with a total mass flow rate of 5.8 × 106 kg/hr at 69.5 bar. The extraction heat exchangers required for heat transfer to the TPD loop are located at the NPP site to reduce licensing complications. They are also near the turbine system to reduce losses and to minimize the amount of additional steam inventory that is cycled through the NPP. TPE-1 is the main control valve for the TPE line and the control with the largest effect on reactivity control. During steady-state operations, the flow of steam in the TPD loop should be sufficient to fully condense the steam in the TPE line to avoid sending high-pressure steam to the condenser, which would decrease plant operating efficiency. Steam traps in the main extraction line downstream from TPE-1 remove condensate that forms while saturated steam travels to the extraction heat exchangers. This is a further design choice to maintain dry saturated steam as the heat source to the TPD loop. At a specified condensate level, TPE-3 opens to allow condensate to flow to the TPD-EHX-2. This design ensures that only liquid water can flow to TPD-EHX-2. TPD-EHX-1 has a vent to the condenser for use while the water level is building to the desired level.

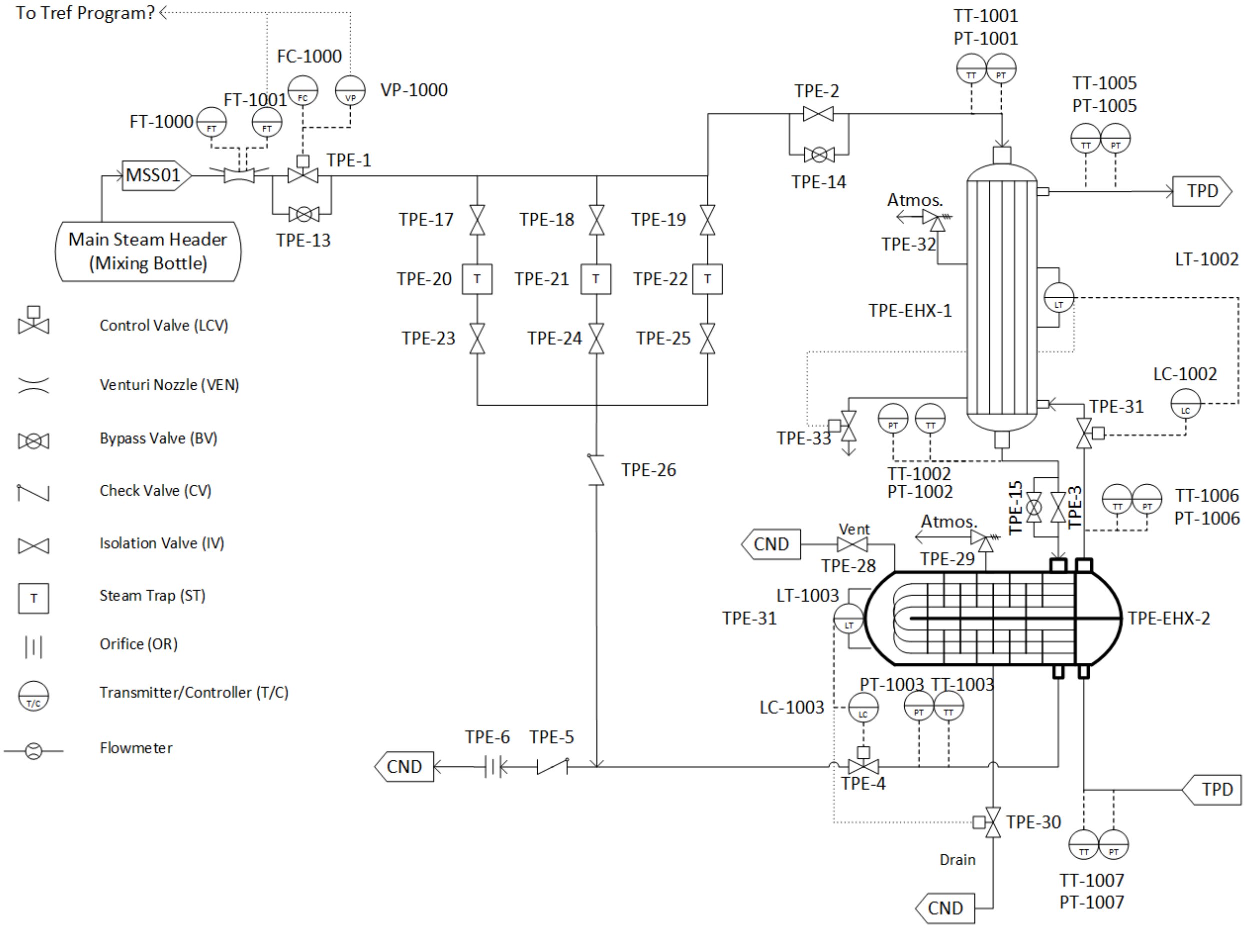

Figure A-3. P&ID for the TPE line for the model with steam in the TPD loop.

For the option in which superheated steam is used in the TPD loop, the extraction heat exchangers comprise a two-stage system because there will be a phase change in both the hot and cold fluids. The first heat exchanger (TPE-EHX-1) is similar to the Babcock and Wilcox once-through steam generator (OTSG). The saturated steam is on the tube side of the heat exchanger, and the delivery steam is evaporated completely and superheated on the shell side. The reason for this design choice is the fact that the OTSG provides slightly superheated steam from a subcooled liquid inlet in a single heat exchanger. This, combined with the vertical nature of the heat exchanger, makes it reasonable to provide the desired heat-transfer and fluid conditions. The TPD loop steam is superheated by about 25°C to assist thermal delivery to the hydrogen plant, approximately one kilometer away, with minimal condensation.

Sending condensate from the TPE line to the condenser is the simplest engineering design for the system for many reasons. This condensate has a temperature of approximately 193.3°C and could alternatively be sent directly to the feedwater heaters to increase operating efficiency. However, doing so would increase design complexity and could introduce a potential process upset that would need to be evaluated. Considering that the total steam extraction is only 5% of total reactor power, the potential gain in efficiency would be small and is not believed to merit the increased design complexity.

In this model, TPD-EHX-2 has a design similar to a feedwater heater. The wet steam from the NPP enters this heat exchanger on the shell side to be condensed and subcooled by the condensate from the TPD loop. The condensate in the TPD loop is preheated in the tube side of the heat exchanger before being fully evaporated and superheated in TPE-EHX-1. The subcooled liquid is designed to exit TPE-EHX-2 at 193.3°C at a high-pressure of 68.3 bar. This liquid is throttled to condenser pressures through an orifice. A check valve prior to the which requires a high differential pressure to open. This helps to ensure that the TPE line remains pressurized in the event of a system malfunction to protect the chemistry of the nuclear steam in the case of a substantial tube leak in either of the extraction heat exchangers.

A-5. Thermal Power Delivery Loop Design with Steam

A P&ID of the proposed TPD loop with steam as the HTF is shown in Figure A-4. As the steam in the TPD loop is pumped through the tubes of TPE-EHX-2, it is preheated to saturation, and then it boils and superheats as it passes through the shell side of TPE-EHX-1. The maximum flow rate of steam exiting the extraction heat exchangers and moving toward the hydrogen plant is 2.715×10^5 kg/hr and the temperature is 252°C. This steam travels approximately 1 km to the hydrogen plant via a pipe that is equipped with steam traps to ensure that dry steam is sent to the hydrogen plant’s steam generator. The delivery-loop heat exchangers (TPD-HX-1 and TPD-HX-2) produce steam for the hydrogen plant by taking in feedwater at saturated conditions (about 5 bars) and producing superheated steam at a rate of 2.751×10^5 kg/hr (approximately the same flow as the steam in the TPD loop noted above). The condensate is then pumped back to the TPE heat exchangers, where it is boiled into steam again.

It is envisioned that a sudden loss of power at the hydrogen plant would trigger an alarm at the nuclear plant to tell operators to reduce steam flow into the TPE line. This alarm setpoint would also control the flow rate of the condensate back to the TPE heat exchangers, which would require either a storage tank or an external discharge. An additional, important consideration is that the pressure of the steam in the TPD loop must be significantly lower than that of the steam in the TPE line to limit contamination across the extraction heat exchangers in the event of leaking tubes.

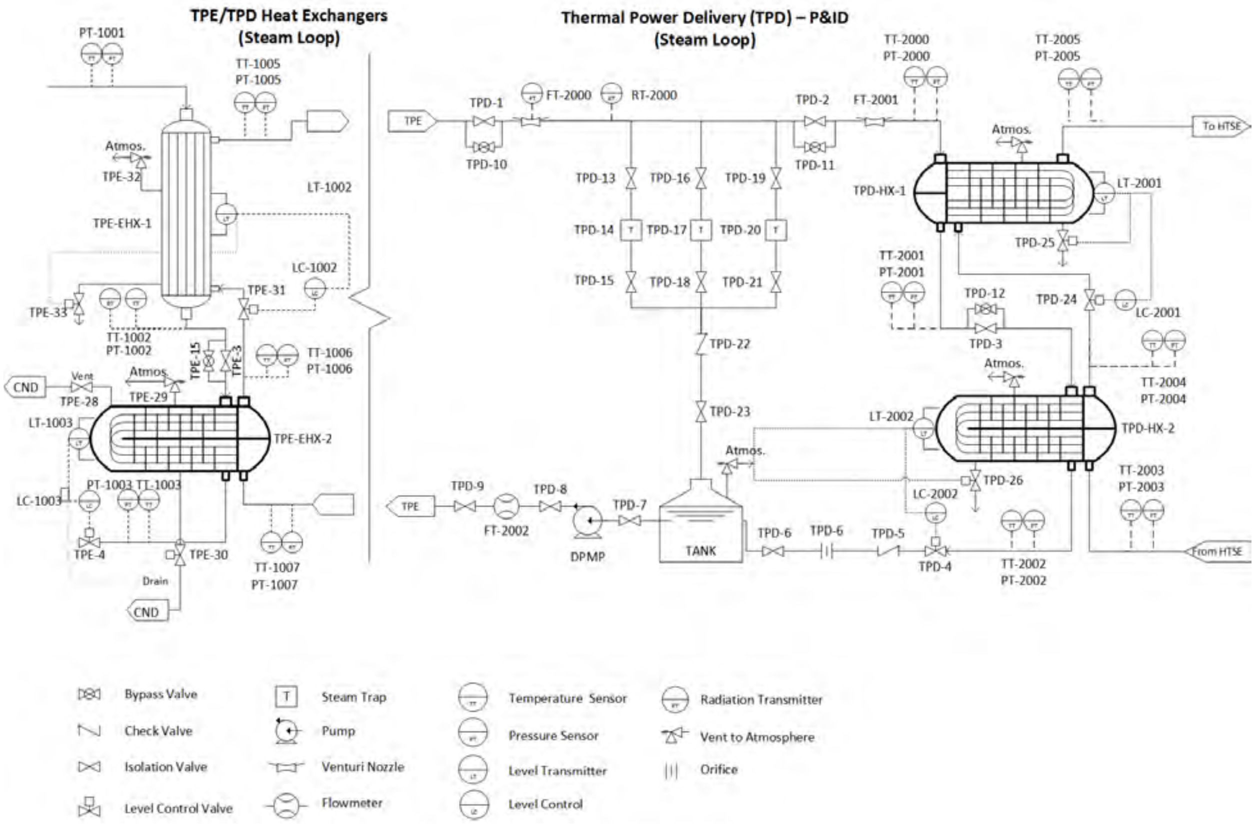

Figure A-4. P&ID for the TPD loop with steam as the HTF.

A-6. Heat-Dispatch Model Descriptions

Multiple thermal-hydraulic models were developed for different scenarios with specific purposes:

1. Provide thermal hydraulic parameters for the heat-dispatch GPWR simulator
2. Gain understanding of scalability of thermal power dispatch from 200 kW to 150 MW and for thermal power dispatch distances from 0.1, 0.5, and 1.0 km.

The scenarios for which the thermal-hydraulic models were developed are summarized in Table A-3. The first scenario featured a thermal-power dispatch of 150 MW and a dispatch distance of 1.0 km. The combined steady-state TPE line and TPD loop thermal hydraulics were modeled using RELAP5-3D, and the steady-state thermal hydraulics of the TPD loop were also modeled using Aspen HYSYS using inputs to replace the TPE line that is consistent with results from the RELAP5-3D model. For the other scenarios, the thermal hydraulics of the TPD loop were only modeled using HYSYS because HYSYS is much more suitable for performing parametric studies. The second and third scenarios that were developed also employed a thermal-power dispatch of 150 MW, but featured dispatch distances of 0.5 and 0.1 km, respectively. For Scenarios 4 and 5, the thermal-power dispatch distance was fixed at 0.1 km while the dispatch power was reduced to 15 MW and 200 kW, respectively, approximately matching Thermal Energy Distribution System (TEDS) at INL. The fluid temperatures and pressure in all cases were the same as those in Scenario 1, and the flow rates were adjusted to achieve the appropriate heat balance. For the first five scenarios, the inside pipe diameter of the TPD loop was 57.45 cm, which corresponds to a 24-inch, schedule 40 NPS pipe. The parameters for Scenario 6 were identical to those of Scenario 5, except that the inside pipe diameter of the TPD loop was reduced to approximately 5.08 cm to match the TED System at INL. Operating conditions of the various simulations are documented separately.

Table A-3. Summary of scenarios for which thermal hydraulic models were developed.

Scenario	Heat dispatch power	Heat dispatch distance	Software and model extent
1	150 MW	1.0 km	RELAP5-3D for TPE line and TPD loop; HYSYS for TPD loop only
2	150 MW	0.5 km	HYSYS for TPD loop only
3	150 MW	0.1 km	HYSYS for TPD loop only
4	15 MW	0.1 km	HYSYS for TPD loop only
5	200 kW	0.1 km	HYSYS for TPD loop only
6*	200 kW	0.1 km	HYSYS for TPD loop only

* For Scenario 6, the inside pipe diameter was decreased to 2 inches to match that of the TEDS at INL.

Figure A-5 shows a block buildup of the RELAP5-3D model of the TPE line and the TPD loop for the model with steam in the TPD loop. For the RELAP5-3D model with steam in the TPD loop, the heat-delivery heat exchangers were modeled as a simple heat sink, as shown by the solid black bar at the far right of Figure A-5. Another addition is a nitrogen surge tank in the TPD loop to regulate pressure in the closed system. This simple heat sink model is fully valid for steady-state flow as long as the magnitude of the heat sink ensures enthalpy balance at the thermal-delivery heat exchangers.

Figure A-6 shows a PFD of the HYSYS model TPD loop, with steam as the heat-transfer media. Similar to the model developed for the oil-based TPD, the stream operating conditions had to be specified along with appropriate pipe dimensions. For the steam-based TPD loop, superheated steam is produced via the TPE heat exchangers, wherein the bypass steam from the MSH condenses on the primary side and feedwater from the hydrogen plant is vaporized and superheated on the secondary side. This superheated steam is transported 1 km to the hydrogen plant, where it is condensed using by the delivery heat exchangers. The condensate is pumped back to the TPE line to complete the loop. Because there is phase change heat transfer in the steam-based TPD, which allows for more heat to be transferred per unit mass of the HTF, less fluid is required in the steam-based TPD. This results in a smaller pipe-diameter requirement for the TPD loop return line from the hydrogen plant, and also less pumping power. Notably, the TPD loop pumping power for oil is more than 20 times larger than that of steam (784 vs 35.6 kW).

The pipe and surrounding material specifications for both the RELAP5-3D and HYSYS models are summarized. Pressure drop in the TPD loop is modeled assuming an internal pipe diameter of 57.45 cm and a surface roughness of 0.05 cm. Heat loss through the walls of the pipe was calculated assuming stainless steel pipe with a wall thickness of 1.75 cm surrounded by 2 inches of urethane foam insulation and buried 1 m underground.

The design of the TPE lines and TPD loops in the models described above meets the design requirements and decisions described above. The TPE line is restricted to the immediate boundaries of the NPP and can be as short as possible to reduce the amount of additional steam that is cycled through the plant secondary system. The controls of the NPP can be designed such that the operation of the TPE line or any connection to the hydrogen plant will not significantly impact safety operations at the NPP. Steam in the TPE line is fully condensed to liquid water in the extraction heat exchangers and is sent directly to the condenser. Future work may also consider returning condensate, which has a temperature of approximately 193.3°C to the feedwater heater system instead of the condenser to increase the efficiency of the power system. A benefit of having a long TPD loop as an intermediary between the NPP and the hydrogen plant is that it provides mass and thermal inertia to smooth fluctuations in the steam flow in the

TPE line, as well as sudden changes that may occur at the hydrogen plant. The oil or steam transport time would provide operators at either plant an additional window to respond to events at the other plant. A long TPD loop will also prolong the time that is required to reach steady state after conditions are altered at either plant.

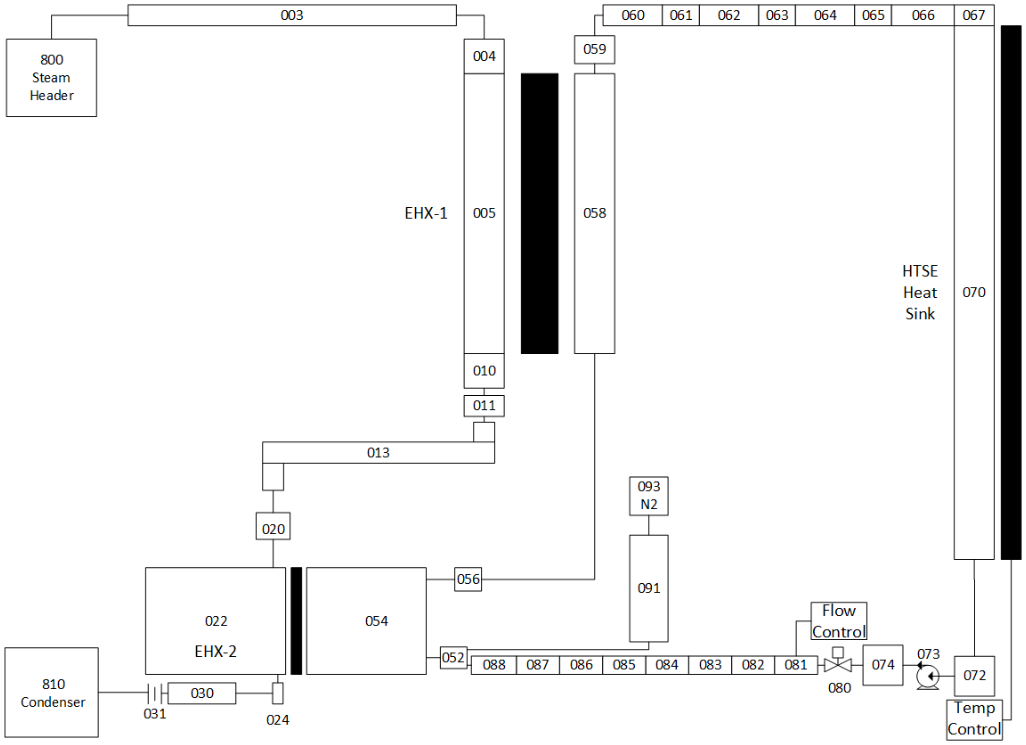

Figure A-5. RELAP5-3D nodalization for both the TPE line and the TPD loop for Scenario 1 with steam in the TPD loop.

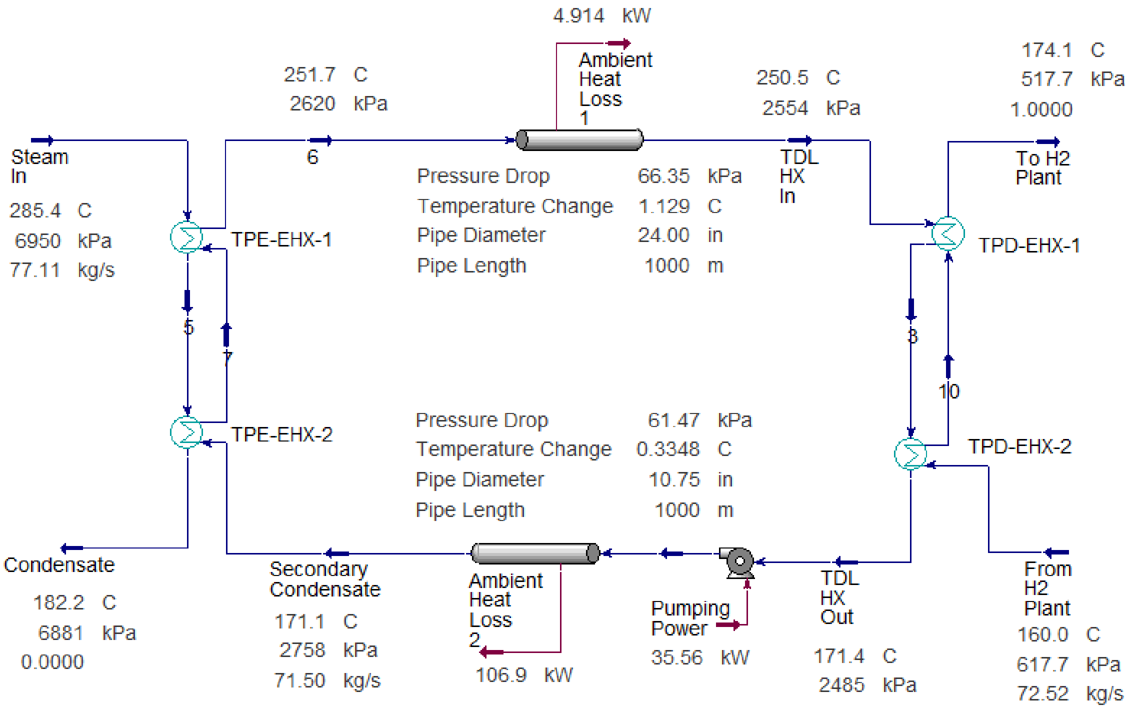

Figure A-6. Aspen HYSYS process model for TPD loop for Scenario 1 with steam in the TPD loop.

A-7. Heat-Dispatch Model Results

Simulation results from Scenario 1 (with a thermal power dispatch of 150 MW and a dispatch distance of 1.0 km) with steam in the TPD loop are summarized in Table A-4. Key parameters for the TPE line include a maximum temperature of 279°C, a mass flow of steam in the TPE line of 80.3 kg/s, and a temperature of condensate that is returned to the condenser of 192.6°C. The pressure in the TPE line decreases from 6.93 MPA to 6.87 MPA at the outlet from TPE-EHX-2, and then to 0.01 MPA at the condenser. Key parameters for the TPD loop are a mass flow rate of thermal oil of 998 kg/s, a maximum temperature of 250.7°C at the outlet of EHX-1, and a minimum temperature of 177.0°C at the inlet to EHX-2. The highest pressure in the TPD loop with steam is 3.049 MPa. For comparison, the simulations using synthetic oil in the TPD loop indicated a maximum pressure of 0.407 MPa. The required flow rates for the simulations with different HTF are also vastly different. The flow rate in the TPD loop with oil is 997.9 kg/s while it is more than an order of magnitude less at 75.26 kg/s for the TPD loop with steam because steam is able to transfer much more heat per unit of mass due to its phase change in the process. The total thermal power dispatch is 156.2 MW, and the total thermal loss in the TPD loop for Scenario 1 is approximately 171 kW (slightly over 1% of the total thermal-power dispatch).

For all scenarios, the fluid temperatures and pressures were consistent with the results from Scenario 1 using RELAP5-3D. These results are important because they confirm that accurate transient models can be developed using shorter thermal-power dispatch distances to reduce computational expense, and then the results can be scaled to any longer-dispatch distance. For Scenario 6, upon maintaining the same operating conditions similar to those of Scenario 5 and simply reducing the pipe diameter, the ΔP increased and the ΔT decreased as expected. Adjustments were made at the hydrogen-plant side to more closely approximate the conditions of the integrated TEDS/HTSE system at INL. The feedwater-inlet temperature and pressure were changed to 20°C and 1 atm, respectively. The flow rate was adjusted such that the outlet conditions of the steam generated was 150°C and 1 atm. Negligible pressure drop across the heat exchanger was assumed based on a design created specifically for this application by an OEM.

Table A-4. Results from RELAP5-3D model for Scenario 1 with steam in the TPD loop.

Node\Description	Pressure, MPa (psia)	Temperature,* °C (°F)	Quality	Mass Flow, kg/s (KPPH)
TPE Line (supersaturated steam to condensate)				
001 (MSH)	6.929 (1004.9)		1	
003 (Pipe to EHX)				80.34 (637.60)
004 (EHX-1 Inlet)	6.880 (998.0)	(Saturated)	1	
010 (EHX-1 Outlet)	6.864 (995.5)	267.7 (513.86)		
020 (EHX-2 Inlet)	6.882 (998.1)	279.44 (535.00)		
024 (EHX-2 Outlet)	6.874 (997.1)	192.59 (378.66)		
810 (Condenser)	0.010 (1.5)			
TPD Loop (supersaturated steam)				
070 (HTSE Outlet)	2.708 (392.81)	176.96 (350.52)		
073 (Pump)				75.26 (597.52)
052 (EHX-2 Inlet)	3.067 (444.89)	177.05 (350.69)		
056 (EHX-2 Outlet/ EHX-1 Inlet)	3.049 (442.19)		0.051	
059 (EHX-1 Outlet)	2.891 (419.25)	254.14 (489.45)		
068 (HTSE Inlet)	2.834 (410.97)	243.44 (470.20)		

* Reported temperatures are based on liquid water, which causes some irregularities. For example, the temperature of the liquid increases slightly from the outlet of EHX-1 to the inlet of EXH-2, presumably because some steam has condensed (quality has decreased).

A-8. Safety Analysis of Thermal-Power Extraction and HTSE

There are two over-arching decisions to be made when considering flexible power operations: the economic case and the safety case. The economic case determines the desire to pursue the design change, and the safety case determines whether the design change will be allowed by the regulator. The safety case also affects cost in determining the lowest-cost design configuration that is acceptable to the regulator.

The safety analysis for HTSE has a natural split between the safety case for the heat-extraction system (HES) in the NPP and the external hazards caused by the HTSE that affect the NPP. The study INL/EXT-20-60104, *Flexible Plant Operation and Generation Probabilistic Risk Assessment of a LWR Coupled with a High-Temperature Electrolysis hydrogen Production Plant*[100] analyzed the safety case for the HES and the external hazards of a high-temperature electrolysis facility (HTEF) by performing a hazard analysis and PRA for both a generic BWR and PWR. The generic BWR was based on a Mark I containment and is applicable to Monticello NPP. The generic PWR model was based on a 2-loop PWR with large dry containment, applicable to the Prairie Island NPP. The internal events for both BWR and PWR generic models match the internal events in the Monticello and Prairie Island NPP licensee and NRC standardized plant analysis risk PRA models. The hazard analysis within[100] identified hazards that were added or modified by the addition of a HES and coupling to a HTEF, recommended design options, and included the minimal safe distance from the HTEF to NPP critical structures. A nominal case where the design assumptions were followed was modeled in a PRA and quantified for the increase in DBA initiating-event frequencies, overall cored damage frequency (CDF), and large early-release frequency (LERF). The DBAs are determined and exhaustively listed and quantified in the existing plant's PRA. They encompass all accidents that can occur within the design basis of the plant's operations. It must be determined what effect any proposed changes have upon the plant's existing DBAs and whether any additional DBAs are introduced. The CDF is the summation of all the sequences of events within the PRA logic model's internal and external events from initiating events (IEs) through a core-damage end state. The LERF applies a set of multipliers (not over 1.0) to those IEs within the PRA which can lead to a large early release to quantify the risk.

The NRC develops various regulatory guides to assist license applicants' implementation of NRC regulations by providing evaluation techniques and data used by the NRC staff. Two distinct pathways through guides and codes of federal regulations (CFRs) are identified for use in the proposed LWR plant-configuration change approval.

One pathway uses 10 CFR 50.59, "Changes Tests and Experiments,"[101] to review the effects on frequencies of DBAs, amendment of the UFSAR, and determination of whether a licensing amendment review (LAR) is required. This pathway is dependent on the IE frequency, which is on the front end of the PRA.

A supporting pathway uses RG-1.174, "An Approach for Using Probabilistic Risk Assessment in Risk-Informed Decisions on Plant-Specific Changes to the Licensing Basis,"[102] through use of risk-informed metrics to approve a plant-configuration change based on the effect on the overall CDF of an approved PRA. This pathway is dependent on the tail end, the CDF-resulting metric of the PRA.

The final pathway is the LAR process, which would use PRA results as well; however, the process uses 10 CFR 50.90, "Application for amendment of license or construction permit at the request of the holder,"[103] and should be avoided if possible due to the lengthy review and monetary burden.

Eight criteria are required to be met for licensing through 10 CFR 50.59[101]:

1. Result in more than a minimal increase in the frequency of occurrence of an accident previously evaluated in the final safety analysis report (as updated)
2. Result in more than a minimal increase in the likelihood of occurrence of a malfunction of a structure, system, or component important to safety previously evaluated in the final safety analysis report (as updated)
3. Result in more than a minimal increase in the consequences of an accident previously evaluated in the final safety analysis report (as updated)
4. Result in more than a minimal increase in the consequences of a malfunction of an SSC important to safety previously evaluated in the final safety analysis report (as updated)
5. Create a possibility for an accident of a different type than any previously evaluated in the final safety analysis report (as updated)

6. Create a possibility for a malfunction of a system, structure, or component (SSC) important to safety with a different result than any previously evaluated in the final safety analysis report (as updated)
7. Result in a design basis limit for a fission product barrier as described in the Final Safety Analysis Report (FSAR) (as updated) being exceeded or altered
8. Result in a departure from a method of evaluation described in the FSAR (as updated) used in establishing the design bases or in the safety analyses.

The bounding criterion of 10 CFR 50.59 is the “result in more than a minimal increase in the frequency of occurrence of an accident previously evaluated in the FSAR (as updated).” A “minimal increase in the frequency of occurrence” is understood to mean <15%. A preliminary PWR PRA[104] used a conservative estimated increase of over 15% and still met the RG-1.174 criteria. It was the task of the final PRA[100] not only to include a BWR model, but to also remove the conservatisms of the preliminary PRA through increased design knowledge and a thorough hazard analysis. This final PRA presented a realistic set of DBA frequencies for both BWR and PWR NPPs.

The results of [100] were compared to the criteria in U.S, NRC, 10 CFR 50.59[101] and RG-1.174.[102] The conclusion was that none of the DBA IE frequencies do not increase by more than 5.6% for any DBA, which is well within the criteria proposed in 10 CFR 50.59.[101] The report also conclude that both the CDF and the LERF increases were within Region III of RG-1.174,[102] providing further support for risk-informed decision making on design changes.

The generic cases for BWR and PWR NPPs set forth in [100] can be used as a roadmap for the site-specific safety cases at Monticello and Prairie Island. Site-specific data that will be required to use PRA for licensing include:

1. Specific design of the HTEF that affects the external overpressure event on the NPP
   a. H_2 production capacities
   b. On-site storage and piping or shipping process
   c. Presence of dedicated ceiling ventilation and other options shown in [100]
2. Specific design of the HES
   a. Isolation valves configuration, number of bypass trains, and other options discussed in [100]
3. Site specifications
   a. Natural and man-made barriers to blast overpressure
   b. Other considerations and sensitivities to potential overpressure event
      i. Pipelines in close proximity to HTEF
      ii. Significant power lines in close proximity to HTEF
   c. Other hazards identified as local to the site that can be affected by the addition of HES and HTEF

Safety analysis has only been done for the HES and HTEF thus far. Other possible hybrid integrations, such as an ammonia plant adjacent to an NPP, will require a hazard analysis specific to an ammonia-production facility and modification of the PRA to account for the identified hazards.

APPENDIX B
ALTERNATIVE MARKET ANALYSIS APPROACH

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APPENDIX B
ALTERNATIVE MARKET ANALYSIS APPROACH

Dispatch and optimization analysis developed in this report was from the viewpoint of the nuclear plant and coupled HTSE system. While this method is valuable, it does not provide insight into the perspective of a grid operator in a regulated system.

In a regulated market, the capacity and the electricity-generating assets are an acquired right of the ratepayer. The production of hydrogen is not currently part of the services provided to the ratepayer; therefore, its future introduction will have to be negotiated between regulators and utilities as to whether electricity sourced for the production of hydrogen could be acquired at the wholesale cost of production (behind the meter) or at the retail price (electricity from the grid).

Production of hydrogen using existing electricity-generating facilities, such as NPPs, is a new paradigm that may require new negotiation and out-of-the-box thinking in order to reach a novel, mutually beneficial state for all parties involved. Various alternative approaches may be found to framing the benefit of hydrogen co-generation and shifting an asset from baseload generation to a responsive one. The following describes one potential approach to framing this concept for a grid operator. The analysis intends to provide insight on valuing the production of hydrogen as a grid service.

B-1. Alternative Economic Approach

One possible demonstration of an IES's value to the grid would be to classify the system as a load response. The HTSE procures electricity at retail price in times of low electricity demand to produce hydrogen and use that storage during peak demand, turning off the HTSE load. Because this load response also acts in a similar fashion to a battery, it provides inherent value to the grid. This could be paid out by a grid operator or independent system operator. A differential cash-flow analysis will be investigated to show some cases that compare IES flexibility to other load-response measures.

Figure B-1 provides an illustration of how the HTSE could be operated in an optimal fashion to minimize the cost of hydrogen and impact (increase in cost) to the grid as a system. During nominal conditions (i.e., charging), the NPP provides electricity to the grid and purchases it back to generate H_2 at the HTSE ($e^- + \Delta e^-$). A fraction of the NPP steam is diverted to the HTSE. This feed allows for both enough hydrogen for the market (H_2) and enough for storage (ΔH_2). During high-demand conditions (i.e., discharging), only the hot-standby capacity is taken from the grid (e^-_{HS}). This can be equated to "roundtrip" efficiency losses. The accumulated ΔH_2 in storage is then discharged to the market (at the nominal H_2 rate).

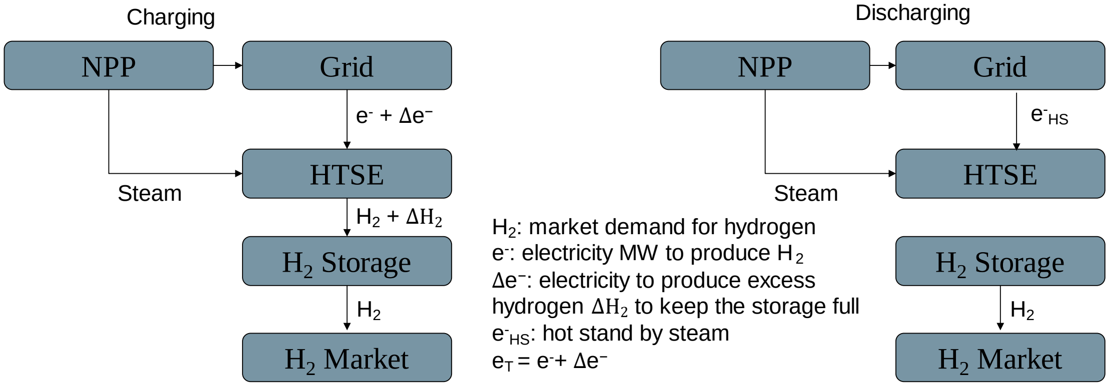

Figure B-1. Block flow diagram of the HTSE charge and discharge modes of operation.

Two main analogies can be made toward already-existing grid-asset types and contractual agreements: load-response or battery-like behavior. These analogies can help determine the value of the IES asset to the grid by giving meaningful comparisons. For example, if the IES system can provide the same service to the grid as a battery, then the economics could be compared to a battery to understand the comparability of the configuration.

Each of those analogies starts with the assumption that the electricity and needed steam are bought by the HTSE as an external rate payer. The HTSE load would shift the generator bid stack to the right by the amount of energy required to produce hydrogen for storage and sale. This shift is depicted in Figure B-2. Similarly, by load-following, or operating as a battery, the return of electricity to the grid during peak demand will result in a shift back to the left of the stack during these instances. These dynamics will drive how a regulator would price the types of electricity consumed by the HTSE (e^- vs. Δe^-) based on how they impact the stack. The main assumption from a grid-benefit standpoint is that the additional demand for storage during nominal load will cause a minimal shift in the stack; freeing additional capacity during peak demand is likely to shift overall prices down for the ratepayer by decreasing the clearing price.

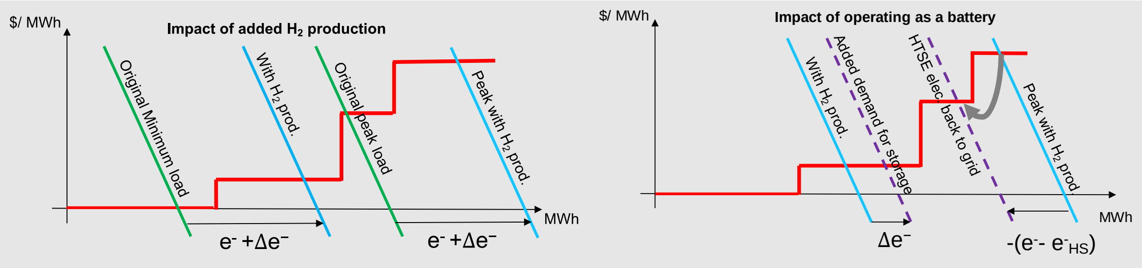

Figure B-2. Generator bid stack and effect of adding HTSE load to the system. The HTSE load would effectively move the clearing price up the stack by the amount of electric load required for hydrogen production and storage, but down by the amount of electricity returned to the grid during peak demand.

B-1.1 Load Response

The HTSE operation is analogous to a load-response system or a load-modifying resource (LMR) or demand-response resource as defined by the Midcontinent Independent System Operator (MISO). MISO defines these two load-response categories as:

- Demand Response Resource (DRR) refers to a resource type: one that provides service to the energy and ancillary services market.
- LMR refers to the use of a demand resource toward meeting planning reserve margin requirement.

In either classification, the HTSE would be entitled to a capacity payment. Capacity payments in MISO have been low in recent years. For 2020 and 2021, the average clearing price in MISO was $5/MW-day with respect to a CONE of around $90,000/MW-year (i.e., ~$246/MW-day). The capacity market should, in theory, account for the benefit of not needing a new CONE by providing the equivalent capacity payment to the HTSE system (in terms of electricity returned back to the grid). The CONE payment in this case could be equivalent to the CAPEX payment for a gas-turbine of around $750/kW.

The requirements for LMR and DRR in MISO have recently changed, but it seems reasonable that, given the ramp rate of the HTSE, it would qualify as an LMR and could possibly be classified as a DRR, Type II. As a reference, the requirements for qualification as an LMR are summarized below:

- May be claimed by only one market participant
- ≥100 kW (grouping allowed)
- Schedulable within 12 hours (startup time ≤12 hours)

- Able to achieve the target level provided during registration
- Maintain target level for four continuous hours
- Able to respond at least five times per year
- Response is an obligation during emergencies
- Submits monthly availability in MWs and notification time for the upcoming planning year.

The amount of power that can be committed from an LMR or DRR is equal to the lowest amount of curtailable consumption which is always available. This electricity returned to the system would correspond to the energy consumed minus hot-standby needs.

As already mentioned, the MISO capacity payment in the last few years (except in Zone 7, which cleared at CONE) has been very low, but considering the long-term investment outlook for the project (25 years) it is likely that most of the CONE revenue could be recovered. Recovering about 60% of the CONE would cover nearly all the capital investment of the HTSE at $400/kW installed.

B-1.2 Battery Equivalence

Intuitively, one could also argue that the HTSE system behaves more closely to a battery, rather than a load response because it consumes excess electricity for storage and discharges it during peak times. The main difference is that hydrogen is physically stored, instead of electricity, and the HTSE system essentially stops consuming electricity during peak demand instead of actually dumping electricity back to the grid.

As a starting point, an HTSE plant providing no load following would purchase a certain amount of electricity (e^-) at the retail rate to generate H_2 for its customer (this includes the steam-equivalent consumption as well). However, if this HTSE plant purchased a slight excess of electricity (Δe^-) to generate an excess of hydrogen that was subsequently stored in order to allow flexible operations during peak demand, then it is also providing a service to the grid, not dissimilar to a battery. This mode of operation can be compared to charging a battery. While the electricity purchased to generate sold hydrogen (e^-) is purchased at the retail rate, this excess electricity is assumed to be equivalent to the variable battery O&M costs. Following consultation with Xcel, a value of -$5/MWh was deemed representative for variable battery costs. The negative value reflects the difference between the low electricity purchase price during the day and the higher sales price during peaks.

During periods of high demand, electricity from the HTSE is diverted back to the grid, with some roundtrip efficiency losses. In this context, these losses equate to the hot-standby requirements (e^-_{HS}) of the HTSE plant. The resulting electricity sales ($e^- - e^-_{HS}$) can be returned to the grid at a similar arbitrage rate as a battery system. Conservatively, the equivalent battery capacity (with corresponding CAPEX, O&M, etc.) for such as system would correspond to the net power returned to the grid (i.e., $e^- - e^-_{HS}$), ignoring roundtrip efficiencies for the battery. For instance, a 300 MW_e HTSE plant ($e^- + \Delta e^-$), would be equated to something closer to a 220 MW_e ($e^- - e^-_{HS}$) battery in order to account for these “grid-level losses.”

From the perspective of the H_2 market, no change in operation is observed: excess hydrogen produced (from Δe^-) and stored is discharged from the tanks at the nominal rate and provided to the end user. The system will also encounter flat or idle days—i.e., times at which H_2 storage capacity is full and no discharge is required. Under those circumstances, the system only purchases from the grid the electricity needed for nominal hydrogen production. No variable costs or sales occur during those instances. This would be equivalent to an HTSE system that does not load follow.

The different modes of operations are illustrated in Figure B-3. During flat days, the system behaves as a typical grid customer. During all other days, the system is a hybrid: it partially operates as a typical end user in regards to equivalent electricity purchased to produce sold hydrogen (e^-), and partially as a battery charging at a given rate (Δe^-) and discharging at another ($e^- – e^-_{HS}$).

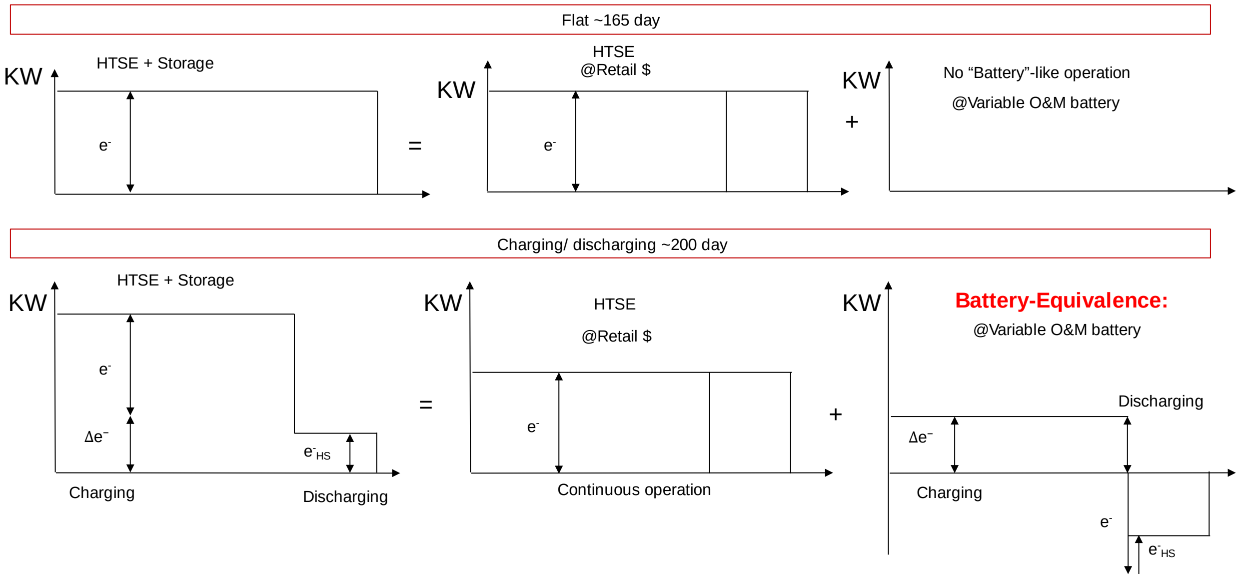

Figure B-3. Battery-equivalence model under charging and discharging conditions, as opposed to flat days under Option 1.

This alternative approach assumes neither that the HTSE can purchase electricity at the NPP O&M cost nor that it receives a specified price at the node; rather, it assumes electricity is bought at the retail price, and steam is paid for at the converted equivalent-electricity retail value. This approach also does not attempt to quantify the value to grid as a whole. Instead, it conducts a differential analysis, comparing this solution to an equivalent battery system. Based on feedback provided by Xcel Energy, a standard battery was assumed to operate under the following conditions:

- Hold storage capacity for roughly 4 hours
- Discharge capacity about 200 times in a year
- Recharge to full capacity within 24 hours.

The fixed battery O&M is specified at 2% of CAPEX. Battery CAPEX will be detailed in a later section, but prices are assumed to decrease between 2020 and 2030. Similarly, a battery lifetime of 10 years is considered for 2020 technology, and 15 years for 2030 technology. As a result, the analysis is conducted for a 25-year timeframe. The HTSE lifetime is assumed to last 25 years with its fixed O&M accounting for stack replacement (leading to their being significantly higher than those of a battery). The remaining parameters are the variable and O&M prices of both systems: i.e., electricity prices. Sensitivity analyses on these parameters will be conducted in Section B-3.

B-2. Alternative Approach Limitations

While these comparisons provide simplified comparison points between very different technologies, the approaches are not without limitations. In the case of the load-response analogy, the main concerns are:

- Ability to be classified as DRR, which depends on ramp rates and the market requirements/constraints
- Predicting MISO capacity payments as a reasonable fraction of the CONE in the future.

For the battery alternative, the concerns surround:

- Ramp rate and capabilities to meet the requirements for participating to the ancillary-services market
- Having a "charging time" over 20 hours, which is not in line with a battery profile.

The prolonged charging time concern has been investigated in other projects. At first level of approximation, a recharging time of 18 hours was found not to alter system cost significantly. Another limitation in both analogies is that they do not capture potential increases in system costs stemming from HTSE hydrogen demand. Currently, this demand is captured via heat sources outside the grid (i.e., SMR). The main justification for neglecting these prices rests on the assumption that the HTSE will be charged at retail price for the cost of electricity used in hydrogen sales. In doing so, the grid is responsible for setting the retail price balancing these system costs.

B-3. Battery Equivalency Results

The battery-equivalence approach is investigated further in this section, with an attempt to quantify the break-even prices of hydrogen and purchased retail electricity that enable the HTSE cogeneration system to be competitive to an equivalent battery. Both high and low battery-cost predictions are considered, based on conversations with Xcel. Because the battery will need to be replaced during the analysis timeframe, the battery CAPEX estimates are adjusted for 2020 and 2030 values. The study does not account for the necessary capacity expansion as load changes through the years or how the stack would be affected by such expansion.

The analysis compares the cost of flexibly operating an HTSE coupled to a nuclear plant with a battery system. By framing the HTSE as a battery-equivalent service, the economics can be directly compared. Break-even levels essentially determine conditions under which the HTSE system is economically equivalent to a battery in terms of providing the same load shifting to the grid. The arbitrage price is defined as the difference in the electricity prices at procurement (charging) and sale (discharging). This price essentially amounts to the net revenue a storage system generates for each MWh.

Looking at the LCOH—i.e. the break-even price of H_2 for $\Delta NPV = 0$ relative to the cost of an equivalent battery—a wide range of economically viable cogeneration configurations are possible at different negotiated retail prices (Table B-1). For instance, if the electricity purchased to sell hydrogen stands at \$35/MWh, and the plant is able to generate \$10/MWh in load-shifting arbitrage revenue, the break-even price for hydrogen could be as low as \$1.56/kg-H_2. Under these conditions, the HTSE would be competitive with a battery system under high-cost assumptions. If, on the other hand, the negotiated electricity price is \$45/MWh, and the arbitrage gains are only \$5/MWh, the break-even point would increase to \$1.98/kg-H_2. The hydrogen prices required to break even are slightly higher if a lower battery CAPEX is assumed in the analysis.

Table B-1. LCOH under different retail prices, arbitrage returns, and battery CAPEX for Option 1.

		Battery Costs	**High**	**Low**
		2020 CAPEX	$350.00/kWh	$250.00/kWh
		2030 CAPEX	$202.50/kWh	$177.50/kWh
Retail Price ($/MWh)	**Arbitrage ($/MWh)**		**LCOH ($/kg)**	**LCOH ($/kg)**
35	-10		1.56	1.72
40	-10		1.76	1.92
45	-10		1.96	2.13
35	-5		1.57	1.74
40	-5		1.77	1.94
45	-5		1.98	2.14

Similarly, Table B-2 flips the analysis around by fixing the H_2 price and arbitrage gains to determine what is an affordable retail electricity price. For instance, if the price of hydrogen is in the \$1.66/kg-H_2 range, and the arbitrage rate is \$10/MWh (approximately the current battery gains), then the break-even electricity purchase price for the HTSE would be \$37.5/MWh. In the case of lower battery costs, the break-even retail price of electricity would need to be lower than \$33/MWh. These values are expected to be close to the market rates in light of the known NPP O&M costs at this node.

Table B-2. Affordable retail electricity price under different hydrogen sales prices, arbitrage returns, and battery CAPEX for Option 1.

		Battery Costs	**High**	**Low**
		2020 CAPEX	$350.00/kWh	$250.00/kWh
		2030 CAPEX	$202.50/kWh	$177.50/kWh
H_2 price ($/kg)	**Aribitrage ($/MWh)**		**Affordable retail price ($/MWh)**	**Affordable retail price ($/MWh)**
1.33	-10		29	25
1.66	-10		37.5	33
2.00	-10		46	42
1.33	-5		29	25
1.66	-5		37	33
2.00	-5		45.5	41

The differential analysis shows the range of conditions that allow the co-generation option to be competitive with a battery system. These estimates essentially account for the system-level costs associated with typical forms of storage. They therefore demonstrate how cogeneration at an NPP could be an attractive value proposition for Minnesota ratepayers.

APPENDIX C
ENERGY STORAGE OPTIONS EVALUATION

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ENERGY STORAGE OPTIONS EVALUATION

This section is a survey of energy storage options that could be integrated with NPPs as part of a larger solution of providing NE to a hydrogen plant or other hybrid industrial plants integrated with the NPP. The possible options were considered based on their availability, cost, efficiency, lifespan, capacity, TRL, feasibility, and risks (Figure C-1). Due to their low energy capacity, flywheels, ultracapacitors, and superconducting magnetic energy-storage were not considered in this analysis. Other options are discussed briefly, but not in detail because they are judged not to be feasible at this time for utility-scale energy storage. All options were analyzed holistically, considering key metrics, risk factors, expected research progress, and threats to the technology. Key metrics included: 1) installation cost, 2) round-trip efficiency, 3) cycle life, 4) lifespan, 5) self-discharge rates, and 6) rates of degradation.

The options were sorted into various storage classes such as batteries, chemical, mechanical, and thermal. Each option was analyzed both holistically and numerically using data to generate a levelized cost of energy (LCOE) for the storage medium for various scenarios. The options with the lowest LCOE were compared directly against one another. These options are lead-acid batteries, lithium-ion batteries, sodium sulfur batteries, above-ground compressed-air energy storage (CAES), liquid-air energy storage (LAES), and molten-salt energy storage.

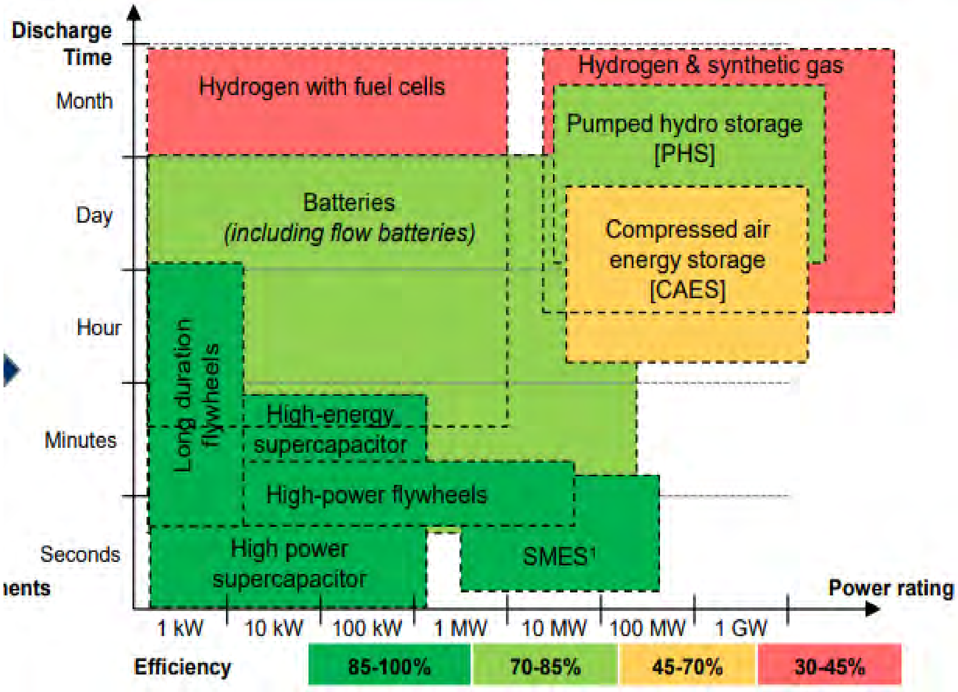


Figure C-1. Projected discharge times of various energy-storage options with their approximate applicable scales.

C-1. Methodology

The primary metric of comparison used between different energy-storage options is the LCOE. LCOE is the total cost (including initial CAPEX and ongoing OPEX) of operating divided by the total energy over the assumed lifetime of the plant. For this analysis an electricity cost of $25 per MWh is assumed.

A scenario-dependent degradation schedule was built that factors in both calendric and cyclical degradation in whichever cycling scenario was selected. An end-o-f-life battery health of 50% was used, and calendar life, cycle life, number of cycles annually, scenario length, round-trip efficiency, and self-discharge rate were inputted. Cycle efficiency is the base loss of power in each cycle due to inherent system inefficiencies. Self-discharge rate is the rate at which the energy storage discharges when left completely idle. This model was then used to calculate the total quantity of energy discharged by the energy-storage medium based on these factors. Based on the cycling rate, the losses due to self-discharge were calculated, and the losses due to cycle efficiency were also calculated. These losses were then accounted for in the model to give a final total energy discharge over the storage system's lifetime.

Costs are accounted for with four metrics. First, the capital cost of each energy-storage system is accounted for with a cost per unit power ($/kw) and cost per unit energy ($/kWh). These two factors of capital cost are not separate ways of measuring capital costs; rather, they reflect the costs of different parts of a system. The $/kWh factor measures the incremental cost per additional unit of energy capacity. $/kW, on the other hand, measures the incremental cost per unit of discharge power. When multiplied by the power output and energy capacity of the system, respectively, they can be added together for a total system cost. Once the capital cost has been accounted for, ongoing expenses must also be, so a maintenance cost in $/kWh-yr is included. Finally, decommissioning costs were included. Many more-mature storage technologies, such as battery installations, have a standard for manufacturer removal and disposal that is functionally included in the installation capital costs, but for less-mature technologies, such costs must be estimated and counted.

C-2. Batteries

Flow Batteries

Flow batteries (Figure C-2) use chemical agents dissolved in liquid to store energy. Each of the liquids is stored in a separate container, one serving as the anode, and the other as the cathode. When the battery is charged or discharged, the liquids are pumped into a third container, where ion exchange occurs across a separating membrane to transfer electricity. Vanadium and zinc-bromide are two types of flow batteries currently on the market. Both have very similar performances. These batteries tend to be large and not very energy-dense. They function using many mechanical elements to pump liquid from one section to another.

Flow batteries have much larger operating-temperature ranges than other battery technologies, and they have an extremely low risk of fire. They also see extremely low levels of capacity degradation, only around 0.4% each year, making them much longer-lasting batteries than other available options. They have cycle lives of well over 10,000 cycles as well.

Despite their long lives, flow batteries are more expensive than most comparable batteries due to their low energy density and the auxiliary equipment required. Prices usually sit above $200 per kW and $700 per kWh installed. They are also known to be unreliable because a large number of mechanical components wear down and break. This causes a significant increase in maintenance costs. Additionally, flow batteries see low efficiency when compared to other batteries, with roundtrip efficiencies of around 70%.

R&D of flow batteries is ongoing in the areas of more-efficient scale-up of flow geometries, better state-of-charge sensors, and less-expensive membranes. Membrane improvements show promise to significantly improve the technology, but the batteries are not among the best near-term options for an NPP energy-storage system until significant advancements can be made.

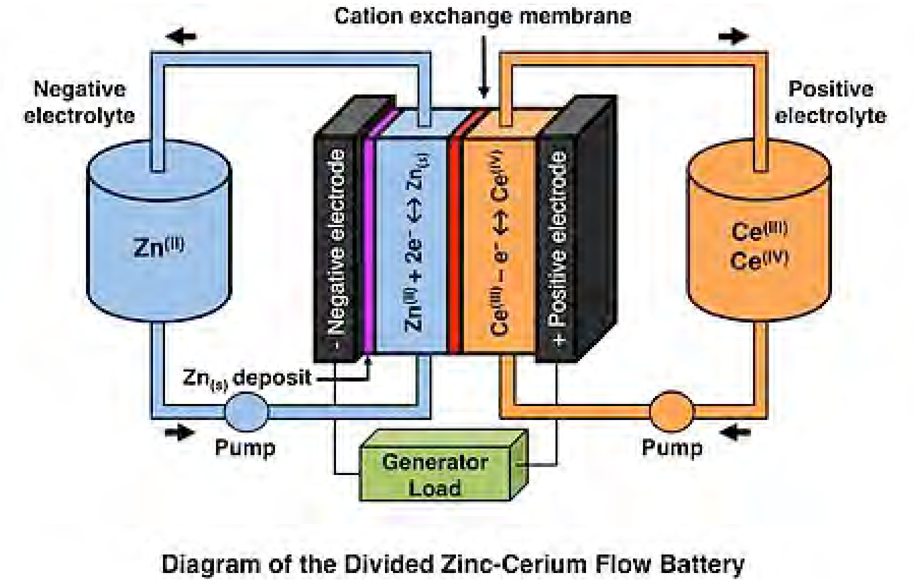


Figure C-2. Diagram of a divided zinc-cerium flow battery.

Sodium Sulfur Batteries

Sodium sulfur batteries (Figure C-3) consist of a molten sodium positive electrode and a molten sulfur negative electrode that are kept above 350°C. The two liquids are separated by a ceramic barrier that only sodium ions flow through. The ions travel across the barrier and store energy by creating Na_2S_4.[114]

Sodium sulfur batteries have efficiencies of around 80%, squarely in the middle of battery options. The batteries are almost exclusively made by NGK Insulators, a Japanese company with a history of excellent service and support throughout the battery's lifetime.[115,116] Sodium sulfur batteries do not self-discharge, unlike other types of batteries. Their typical cycle life is 4,000 cycles, and they suffer an average of just 0.34% annual degradation.[117]

Figure C-3. Photograph of an installed NGK sodium sulfur battery module.

Though these batteries do not self-discharge, they do require a heat source to keep their internal components at the proper temperature, and this function can consume over 20% of the battery's charge each day, a significant maintenance expense. In the case of a pairing with an NPP, the power plant's waste heat could be used to provide this heating. The batteries' insulation must be designed well due to the batteries' high internal temperature, and any defect in the insulation can lead to large fires. NGK has improved its technology to protect against this scenario, but such fires are still a possibility.[118] Any installation would need to be placed at least 200 yards from sensitive structures.

Lead-Acid Batteries

Lead-acid batteries store energy using the complex series of chemical reactions that occur when diodes made of lead (Pb) and lead dioxide (PbO_2) are submerged in sulfuric acid (H_2SO_4)[119] as shown in Figure C-4. This chemistry degrades much more quickly than other battery types, giving this class of battery cycle life of around 2,500 cycles.[117] However, this degradation is experienced primarily at lower levels of charge, so if the batteries were infrequently cycled and discharged close to an NPP flex, they would see very little degradation.

Lead-acid batteries are easily the least-expensive class of large batteries, costing an average of only $250 per installed kWh.[115] This cost is split, with the batteries only costing $200 per kWh, installed capacity, and $80 per kW, power. This low cost is due to inexpensive battery chemistry as well as the age and mature nature of the technology. These batteries have been used for decades, and they are the primary type of battery used in the automotive industry. They are widely available because of this fact. Lead-acid batteries' calendric degradation slows dramatically when held at above 80% charge. This fact gives them potential to be very effective as an option for weekly to monthly cycling scenarios in which the batteries could rest charged.[115]

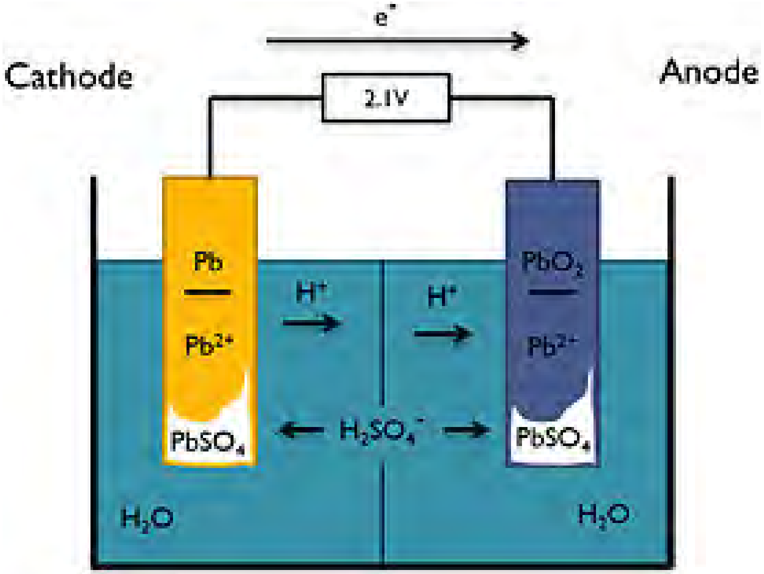


Figure C-4. Diagram displaying a common lead-acid battery chemistry.

Lead-acid batteries degrade at an average rate of 5.4% annually; this degradation is much faster than any of the other types of batteries discussed in this report.[117] This degradation accelerates rapidly after reaching 80% battery health. Additionally, they have a self-discharge rate of roughly 2% each month. This self-discharge rate decreases efficiency when energy is held for longer periods of time, potentially harming the technology's competitiveness in the long-term storage space. Upon disposal, the toxic chemicals in these batteries must be handled with care, and the batteries are usually recycled by companies that specialize in their disposal.

Very little research is being done on lead-acid batteries, and they are not expected to advance significantly in the foreseeable future due to the technology's mature state.

Lithium-Ion Batteries

Lithium-ion (Li-ion) batteries (Figure C-5) use lithium as the key component of their electrochemistry. In these batteries, lithium ions travel across the electrolyte.[120] These chemistries, which often rely on cobalt, allow them to be much more energy-dense than other batteries. Already heavily used in automotive and consumer-goods industries, lithium-ion batteries have rapidly penetrated the market to become the primary battery choice of utilities across the U.S., reaching 90% of the new market share in 2017.[117]

These batteries are the most energy-dense battery option available, both volumetrically and by weight, as shown in Figure C-6.[4] They are also among the cheapest and most readily available, with an installation cost that is now close to $400/kWh according to Indie Energy's Evan Ture.[115] Maintenance expense is only anticipated to be $8/kWh-yr. Additionally, lithium-ion batteries have an exceptionally low self-discharge rate of only 1–2% per month.[121] This rate allows them to hold energy for much longer than other technologies with minimal losses.

Figure C-5. Tesla 80 MWh lithium-ion battery installation in Mira Loma, CA.

Lithium-ion batteries can spontaneously combust if they overheat or are put under high levels of pressure.[122] These fires burn at temperatures well in excess of 500°C and are extremely difficult to extinguish due to the battery's internal chemistry[123]; these fires have the potential to cause severe damage if the batteries are proximate to other flammable structures. Any pairing with an NPP would need to place the installation a safe distance from sensitive structures.

Materials used in the production of Li-ion batteries are rare and could continue to have geo-political and supply-chain constraints, causing some uncertainty as to the maximum large-scale deployment of these battery systems.

Lithium-ion batteries are one of the most-researched energy-storage options, with technology advancements coming rapidly over the past decade. This progress is expected to continue for the next decade, with prices dropping, cycle lives increasing, and hazards being mitigated. Most predictions foresee prices below $200 per kWh by 2030, cycle lives of at least 9,000, and the introduction of highly efficient lithium chemistries that exclude cobalt.[124] Additionally, new observation software and stricter construction and fire codes are steadily decreasing the risks that these batteries pose.

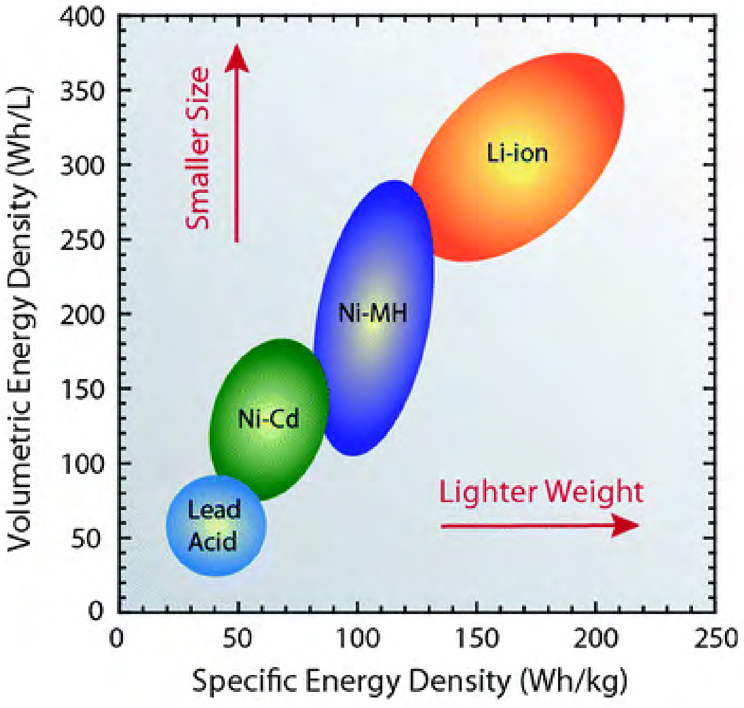


Figure C-6. Volumetric and specific energy densities for various classes of battery.[4]

Summary—Battery Energy Storage

Financial analysis was conducted on each battery-storage option to assess its viability to pair with an NPP. The inputs for that analysis are displayed in Table C-1. Data on battery-storage options are readily available, but the rapid technological progress and price changes across the industry render even last year's data on battery installations unrepresentative of the market as it currently stands. Information on battery pricing, efficiency, and lifetime are sourced from the 2020 NREL Cost of Battery Storage Update, and these figures were verified directly by an industry source and are reflective of the state of the technology and markets in 2020.[115,125] These data were supplemented with data from the 2019 PNNL Energy Storage and Technology Cost Characterization Report on these metrics, as well as cycle life and degradation. Finally, data on self-discharge rates were sourced from the University of Washington's Clean Energy Institute, and degradation was assumed to be half calendric and half cyclical per the recommendation.[115,120] An electricity cost of $25 per MWh is assumed for all modeling.

Table C-1. Summary of key metrics for different battery energy-storage options.Table D-1

Battery storage option	Installation cost* ($/kW)	Installation cost* ($/kWh)	Maintenance cost* ($/kWh)	Round-trip efficiency* (%)	Cycle life*	Lifetime (years)*	Self-discharge rate (%/month)*
Flow	200	700 465-700	8	70 70-75	10,000	20	26
Lead Acid	80	200 160-250	8	80 79-84	2,500	15	2
Li-ion	120	340	8	85 90-94	7,500	20	1
Sodium Sulfate	200 500-1000	700 500-1000	24	80 77-83	4,000	15	0

* Approximated.

Many factors are involved in a company’s decisions about the installation of an energy-storage system, but the most important by far is the system’s profitability. Below, each of the battery storage options are analyzed, using the data from Table C-1 to output a levelized cost of storage (LCOS) and a LCOE for a 20-year period, where LCOS ($/MWh) is the cost of the storage system, including both CAPEX and OPEX, per MWh of electricity stored and LCOE ($/MWh) is the total break-even price of the electricity retrieved from storage per MWh discharged (Table C-2 through Table C-10). While LCOE is a far better metric for overall system evaluation, the inclusion of LCOS in the results should provide a window into the specific effects that system costs, round-trip storage efficiency, and self-discharge have on the cost of the storage system.

The two primary sources of uncertainty surrounding the cost of storage are the frequency with which nuclear plants will be asked to flex and the amount of time that they will move electricity off of the grid in each instance. To account for this uncertainty, nine scenarios are modeled accounting for each of these factors. Four-hour and 12-hour storage are modeled across low-cycle, medium-cycle, high-cycle, and daily-cycle scenarios. Additionally, a 100-hour storage option that completes two full cycles each year is modeled to show the possibility that stored energy is held for seasonal use.

Table C-2. Financial analysis of battery storage options for 4-hour storage cycled 12 times annually.

Battery Storage Option	LCOS ($/MWh)	LCOE ($/MWh)
Flow Batteries	4456.99	60,461,560.48
Sodium Sulfur Batteries	6,371.44	7,989.30
Lead Acid Batteries	1,979.56	4,559.59
Lithium-ion Batteries	2,599.17	4,176.24

Table C-3. Financial analysis of battery storage options for 4-hour storage cycled 60 times annually.

Battery Storage Option	LCOS ($/MWh)	LCOE ($/MWh)
Flow Batteries	905.27	8100.80
Sodium Sulfur Batteries	1,323.08	1,678.85
Lead Acid Batteries	420.27	619.04
Lithium-ion Batteries	530.63	688.63

Table C-4.Financial analysis of battery storage options for 4-hour storage cycled 200 times annually.

Battery Storage Option	LCOS ($/MWh)	LCOE ($/MWh)
Flow Batteries	283.94	727.73
Sodium Sulfur Batteries	441.50	576.87
Lead Acid Batteries	202.39	287.39
Lithium-ion Batteries	168.87	227.35

Table C-5. Financial analysis of battery storage options for 4-hour storage cycled daily.

Battery Storage Option	LCOS ($/MWh)	LCOE ($/MWh)
Flow Batteries	163.81	341.23
Sodium Sulfur Batteries	402.98	528.73
Lead Acid Batteries	190.90	268.50
Lithium-ion Batteries	99.03	142.68

Table C-6. Financial analysis of battery storage options for 12-hour storage cycled 12 times annually.

Battery Storage Option	LCOS ($/MWh)	LCOE ($/MWh)
Flow Batteries	4456.99	60,461,560.48
Sodium Sulfur Batteries	6,198.78	7,773.47
Lead Acid Batteries	1,910.10	4439.07
Lithium-ion Batteries	2,501.09	4,019.59

Table C-7. Financial analysis of battery storage options for 12-hour storage cycled 60 times annually.

Battery Storage Option	LCOS ($/MWh)	LCOE ($/MWh)
Flow Batteries	905.27	8,100.80
Sodium Sulfur Batteries	1,287.22	1,634.03
Lead Acid Batteries	405.53	598.20
Lithium-ion Batteries	510.61	663.59

Table C-8. Financial analysis of battery storage options for 12-hour storage cycled 200 times annually.

Battery Storage Option	LCOS ($/MWh)	LCOE ($/MWh)
Flow Batteries	283.94	727.73
Sodium Sulfur Batteries	429.53	561.92
Lead Acid Batteries	195.29	278.28
Lithium-ion Batteries	162.50	219.71

Table C-9. Financial analysis of battery storage options for 12-hour storage cycled daily.

Battery Storage Option	LCOS ($/MWh)	LCOE ($/MWh)
Flow Batteries	163.81	341.23
Sodium Sulfur Batteries	392.06	515.07
Lead Acid Batteries	184.21	259.96
Lithium-ion Batteries	95.29	130.65

Table C-10. Financial analysis of battery storage options for 100-hour storage cycled bi-annually.

Battery Storage Option	LCOS ($/MWh)	LCOE ($/MWh)
Flow Batteries	n/a	n/a
Sodium Sulfur Batteries	36,447.82	45,584.78
Lead Acid Batteries	11,135.81	555,766.53
Lithium-ion Batteries	14,684.27	108,170.96

* The option with the lowest LCOE has all of its data listed in bold.

These analyses show that flow batteries are not currently economically competitive with other battery options. Sodium sulfur batteries perform similarly, with the LCOE coming in above Li-ion in almost all scenarios. However, in the long-duration scenario (100 hr of storage cycled bi-annually), Na-S batteries come in with the lowest LCOE.

Lead-acid batteries price competitively in all but the 100-hour storage scenario. Its low CAPEX requirement of only $200 per kWh installed, $140 per kWh less expensive than lithium-ion, gives it an advantage in lower cycle scenarios because its low cycle life of 2,500 cycles, 1,500 less than sodium sulfur and 5,000 less than lithium-ion, does not hamper its performance. Its self-discharge rate of 2% per month, higher than lithium's 1%, provides it with a slightly higher LCOE in the 12-cycle scenarios.

Finally, lithium-ion batteries perform competitively in all scenarios, and they produce the lowest LCOE in all four higher-cycle scenarios, as well as in the seasonal storage scenario. The technology's combination of relatively low cost ($340 per kWh installed) and relatively long cycle life (7,500 cycles) make it significantly less expensive than other options when cycled frequently. Its low cost gives it an advantage over higher-cost options such as flow batteries and sodium sulfur batteries while its long cycle life allows the cost to be distributed over far more cycles than lead acid in a higher-cycle scenario. Lithium-ion batteries also provide the lowest LCOE for the 100-hour seasonal-storage scenario. Lithium-ion batteries have very low self-discharge rates (roughly 1–2% per month). Flow batteries and lead-acid batteries both suffer from upwards of 10% self-discharge. This fact makes flow and lead-acid batteries incapable of holding the charge for long enough periods to provide seasonal storage. Although lithium-ion does not reach the 0% self-discharge rate of sodium sulfur technology, its much lower capital and maintenance costs cause it to have a much lower LCOE.

C-3. Hydrogen Energy Storage

Water electrolysis, the splitting of water into hydrogen and oxygen, has been understood for over a century. However, it is traditionally high-level of the expense compared to other energy sources prevented significant development of the technology. The production of H_2 out of H_2O using electrolysis received significant research attention over the past decade, and the element's potential use as energy storage has played a key part in that attention. In this section, H_2 will be discussed specifically as a means of storing energy to be later converted back to grid electricity at utility scale. This discussion and analysis does not apply to the case of H_2 as a value-added product used in downstream industries such as ammonia production, refineries, etc.

Such a storage system uses excess electricity to create H_2 through the electrolysis of H_2O. The H_2 is then stored in large, highly pressurized vessels. This H_2 is sent through a fuel cell to make electricity when needed, with H_2O as the only byproduct. For a more in-depth discussion on electrolysis, see Section 0 of this report.

Summary—Hydrogen Energy Storage

Reversible SOFC/EC electrolyzers that can provide hydrogen energy storage are currently expensive compared to other energy-storage options, with a cost of over $2400 per kW. Efficiencies are around 35%, dramatically lower than competing storage options (Table C-11). Ongoing research on SOFCs and SOECs has the potential to improve both areas, with capital costs expected to come down to around $600 per kWh by 2030 and efficiency expected to improve to nearly 60% in the same time frame.

Table C-11. Summary of key metrics for hydrogen energy-storage.

Hydrogen Storage Option	Installation Cost ($/kWh)	Round Trip Efficiency (%)	Cycle Life	Self-Discharge Rate (%/Month)
SOE (2020)	2400	30-40	10,000	0.1
SOE (2030)	600	40-60	10,000	0.1

C-4. Mechanical

Mechanical energy storage refers to the use of mechanical processes, such as changes in pressure, gravity, or rotational inertia, to store energy for later use. There are four classes of mechanical-energy storage that might, at present, successfully pair with an NPP: 1) CAES, 2) pumped hydro storage (PHS), 3) solid mass gravitational energy storage, and 4) LAES. The concepts needed to implement these methods have been familiar for a long time, and mechanical-energy storage is by far the most used energy-storage category, both in the U.S. and globally. Due to a lack of requisite geologic structures and terrain in many locations, CAES and PHS are often not options; LAES removes this geographical constraint.

Compressed Air Energy Storage

CAES is the storage of energy by compressing air that can later be released through turbines to produce electricity. This compression can be done either in above-ground tanks or in certain types of basalt subterranean geologic structures, the latter being more economical, but geography dictates whether this option is available. Two large operational below-ground CAES facilities, which take advantage of local underground geologic formations, include one 110 MWe facility in Alabama and one 270 MWe facility in Germany.[126] These facilities see round-trip efficiencies of over 50%.[117] In the Minnesota Twin Cities region, there are no known geologic formations where CAES could be employed; thus, below ground CAES is likely not an option.

There are several smaller above-ground CAES facilities; these facilities usually cost roughly $400 per kWh. Like underground CAES, round-trip efficiency is usually around 50%.[117]

Pumped Hydro Storage

PHS is the most commonly used energy-storage method globally, by far—making up 99% of all global energy storage with over 9,000 GWh installed.[127] PHS is simply the act of storing energy as gravitational potential energy. This is usually done by using electricity to pump water uphill, generally into a lake or pond, and then allowing the water to flow downhill through a turbine when the electricity is needed. Modern PHS systems achieve round-trip efficiencies of around 80%. These systems have a wide range of capital costs, with a high end of over $3,000 per kW, a low end of $386 per kW, and a median of $920 per kW in 2020 dollars.[128] However, PHS is also geographically limited due to the required elevation changes. The lack of substantial changes of elevation in the Twin Cities region makes PHS an unlikely option. Some PHS systems involve underground water storage, but like CAES, these require specific geologic formations.

Solid Mass Gravitational Energy Storage

Solid mass gravitational energy storage (SMGES) is another option that has received significant interest over the past several years. This method can take many forms, with some of the most popular being advanced rail energy storage (ARES), using cranes to lift large objects, and using old oil wells as tunnels along which weights can be lifted and dropped. Though this technology is proven and can often achieve efficiencies of over 80%,[129] its massive weight and space requirements provide a significant challenge to its use for grid-scale energy storage. A simple calculation finds that moving 1 million pounds 1 mile would store roughly 1.9885 MWh of gravitational potential energy.

ARES is the primary SMGES method in use, with a large facility under construction near Pahrump, Nevada. The rail system has an altitude change of 3000 ft over 5.5 miles and can deliver 50 MW of electricity to the grid. It has a storage capacity of 12.5 MWh. This project cost $55 million total, an average of $4,400 per kWh. It is expected to be operational for 40 years.[129]

Liquid Air Energy Storage (LAES)

LAES is a new concept based on CAES and thermal-storage concepts. The United Kingdom’s Highview Power pioneered a LAES concept under the name of CRYOBattery, shown in Figure C-7.[130] In this storage system, the air is cooled to -196°C, its liquefaction point, and stored in insulated, low-pressure vessels. When electricity is needed, exposure to ambient temperatures causes rapid regasification and a 700-fold expansion in volume, which is then used to drive a turbine and create electricity. Highview modularized their technology and is in the process of constructing its first commercial installation, a 50 MW, 250 MWh CRYOBattery in Manchester, United Kingdom. They believe that after the completion of this project, they will be in a position to deploy gigawatt-scale facilities.[131]

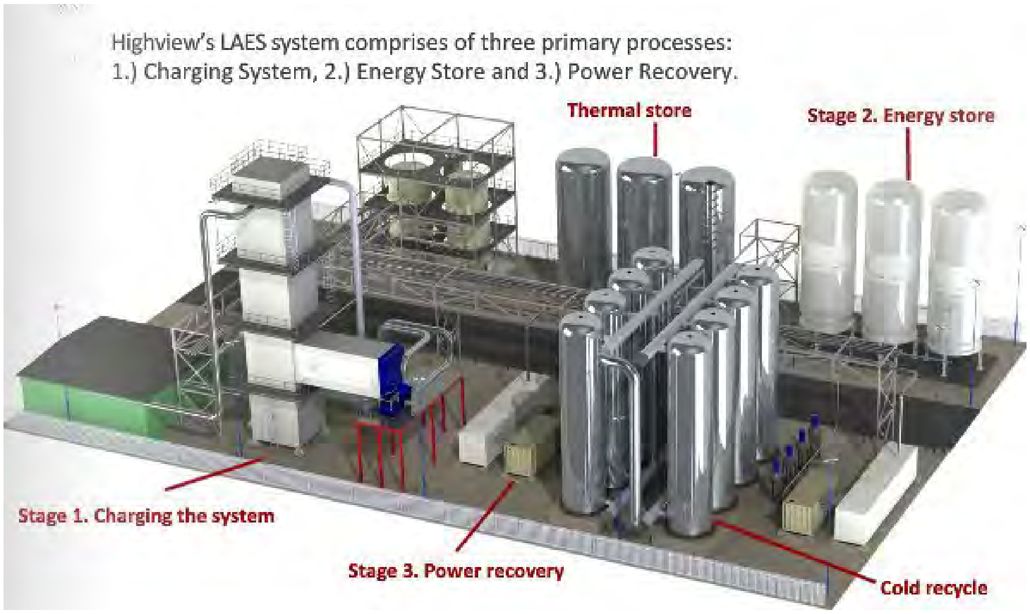


Figure C-7. Concept image of a Highview Power CRYOBattery, their new LAirES technology.[130]

The technology has been successful thus far, achieving round trip efficiencies above 70% when a source of waste heat is present.[130] The CRYOBattery has a high cost on a per-unit power basis of roughly $1000 per kW, but the system has a very low cost of additional capacity.[132] Above 8 hours of duration, it becomes extremely cost-effective due to the small marginal cost of additional capacity. Additionally, unlike some other storage options, there are no particularly hazardous or exotically expensive materials or components used in these facilities.

All of the construction and processes used by the CRYOBattery have been well understood for some time. Because LAES builds on established and mature technologies of air liquefaction and vaporization and uses primarily off-the-shelf parts, its future development is likely limited, even though Highview is working to make incremental improvements. Despite this, as the technology gains market share, its costs will likely come down.

Summary—Mechanical Energy Storage

Financial analysis was conducted on each mechanical energy-storage option across all nine of the established scenarios to assess viability. The inputs to the analysis are summarized in Table C-12, with more detailed data regarding cycling regimes in Table C-13 through Table C-21. Data on mechanical energy-storage options is sporadically available depending on the type of storage. Degradation is assumed to be 1/5 that of batteries for all options in this category due to the lack of chemical degradation. This degradation is assumed to be half calendric and half cyclical. An electrical generation cost of $25 per MWh is assumed for all modeling.

Table C-12. Summary of key metrics for different mechanical energy-storage options.

Mechanical Storage Option	Installation Cost ($/kW)	Installation Cost ($/kWh)	Round-Trip Efficiency (%)	Cycle Life	Lifetime	Self-Discharge Rate (%/month)
CAES (Above Ground)	1,500 1050-2544	165 355-490	52 50-55	10,000	25	0
CAES (Below Ground)	1,500 1050-2544	0	52 50-55	15,000	40	0
PHS	920	0	80	10,000	30	1.5
ARES	200	3,400	78	10,000	40	0
LAES	1,000	200	70 60-70	15,000	30	1

Table C-13. Financial analysis of mechanical energy-storage options for 4-hour storage cycled 12 times annually.

Mechanical Energy Storage Option	LCOS ($/MWh)	LCOE ($/MWh)
CAES (Above Ground)	3,168.36	6,118.00
CAES (Below Ground)	2,262.01	4,375.02
PHS	1,657.10	2,096.38
ARES	15,338.84	19,690.18
LAES	2,591.27	5,050.48

Table C-14. Financial analysis of mechanical energy storage options for 4-hour storage cycled 60 times annually.

Mechanical Energy Storage Option	LCOS ($/MWh)	LCOE ($/MWh)
CAES (Above Ground)	640.44	1,256.51
CAES (Below Ground)	453.26	896.66
PHS	332.37	440.46
ARES	3,076.51	3,969.25
LAES	519.24	813.54

Table C-15. Financial analysis of mechanical energy storage options for 4-hour storage cycled 200 times annually.

Mechanical Energy Storage Option	LCOS ($/MWh)	LCOE ($/MWh)
CAES (Above Ground)	198.13	406.02
CAES (Below Ground)	136.73	287.95
PHS	100.54	150.67
ARES	930.64	1,218.12
LAES	156.64	252.91

Table C-16. Financial analysis of mechanical energy storage options for 4-hour storage cycled daily.

Mechanical Energy Storage Option	LCOS ($/MWh)	LCOE ($/MWh)
CAES (Above Ground)	112.52	241.38
CAES (Below Ground)	75.41	170.02
PHS	55.63	95.60
ARES	514.93	685.16
LAES	86.39	149.66

Table C-17. Financial analysis of mechanical energy storage options for 12-hour storage cycled 12 times annually.

Mechanical Energy Storage Option	LCOS ($/MWh)	LCOE ($/MWh)
CAES (Above Ground)	2,036.80	3,941.93
CAES (Below Ground)	1,205.00	2,342.30
PHS	1,005.59	2,015.58
ARES	15,197.21	19,508.60
LAES	1,883.27	3,677.40

Table C-18. Financial analysis of mechanical energy storage options for 12-hour storage cycled 60 times annually.

Mechanical Energy Storage Option	LCOS ($/MWh)	LCOE ($/MWh)
CAES (Above Ground)	411.71	816.75
CAES (Below Ground)	241.46	489.34
PHS	201.69	301.39
ARES	3,048.11	3,932.83
LAES	377.37	598.09

Table C-19. Financial analysis of mechanical energy storage options for 12-hour storage cycled 200 times annually.

Mechanical Energy Storage Option	LCOS ($/MWh)	LCOE ($/MWh)
CAES (Above Ground)	127.37	269.94
CAES (Below Ground)	72.84	165.08
PHS	61.01	103.40
ARES	922.04	1,207.11
LAES	113.84	190.64

Table C-20. Financial analysis of mechanical energy storage options for 12-hour storage cycled daily.

Mechanical Energy Storage Option	LCOS ($/MWh)	LCOE ($/MWh)
CAES (Above Ground)	72.33	164.10
CAES (Below Ground)	40.17	102.26
PHS	33.76	67.84
ARES	510.17	679.07
LAES	62.78	115.60

Table C-21. Financial analysis of mechanical energy storage options for 100-hour storage cycled bi-annually.

Mechanical Energy Storage Option	LCOS ($/MWh)	LCOE ($/MWh)
CAES (Above Ground)	9,213.05	17,742.41
CAES (Below Ground)	4,437.70	8,559.05
PHS	4,311.01	85,016.82
ARES	90,755.46	116,378.15
LAES	9,426.79	84,327.95

PHS emerges from this analysis as the clear winner, yielding the lowest LCOE in all but the long-duration storage scenario. Its low cost per capacity and relatively low cost per unit power compared to other comparable storage options make it remarkably inexpensive, with LCOE in the 12-hour cycled daily scenario reaching just $42.95 per MWh. The only aspect of PHS that makes it less competitive in the long-duration scenario is the inevitable evaporation of water from the storage pool, which depletes the storage pool at an average rate of 1–2% monthly. Despite its economic competitiveness, the lack of elevation change in the Twin Cities region makes PHS an unlikely option for this region of study.

Below ground, CAES is also shown to be very competitive, with the second-lowest LCOE in each of the 12-hour scenarios as well as the 4-hour cycled daily. Additionally, below ground CAES easily yields the lowest LCOE in the long-duration scenario. Its 0% self-discharge rate results in an advantage in the long duration scenario. As mentioned, due to geography, this is not an option for the Twin Cities region.

Above ground, CAES is shown to be competitive with LAES for the lowest LCOE out of the geographically feasible options. It has a higher per unit power cost than LAES, but it also has a lower capacity cost due to the decreased complexity of storing compressed air compared to cryogenically liquified air. This combination makes it competitive in LCOE in all but the 12 cycles and 60 cycle 4-hour storage options. However, it only comes in with a lower LCOE than LAES in the long-duration scenario.

LAES seems to be the most economically efficient of the geographically feasible options in eight of the nine scenarios. It's cost per unit power of roughly $1000 per kWh and high efficiency of 70% compared to above-ground CAES’s figures of $1500 and 52% respectively give LAES an advantage.

This analysis shows that ARES is the least-competitive of the mechanical storage options. Though the technology has no self-discharge, the massive capital requirement of laying rails over long distances is responsible for the bulk of the LCOE in this case. This capital cost is primarily weighted toward capacity instead of power due to the reliance on the cost of rail, making ARES increasingly uncompetitive as scenario duration increases. Additionally, the significant elevation change required for ARES makes it an unlikely option in the Twin Cities region.

C-5. Thermal

Interest in the use of advanced materials to store heat in an insulated environment for thermal energy storage (TES) has increased significantly over the past several years. When the stored energy is required, the heated material is commonly used to heat water to create steam, which is then sent to a turbine-generator to make electricity. There are three different types of media that will be discussed in this section: 1) phase change (PC-TES), 2) solid state (SS-TES), and 3) liquid state (LS-TES). Phase-change TES uses a material's phase change to store large amounts of latent heat. These systems are not analyzed in this report, but current research and potential breakthroughs in phase-change TES are discussed. Solid-state TES uses material that remains solid throughout the process, such as concrete, to store energy as sensible heat. Liquid state TES uses liquid to store energy as sensible heat. The most common liquid state TES medium is molten salt.

Phase Change

The use of phase change materials for TES is a relatively simple, yet undeveloped concept. This concept uses the isothermal phase change of materials to store energy as latent heat in addition to that stored as sensible heat as the material heats up. The large amount of heat required to transition a material from one phase to another allows for a very high energy-storage density and allows the use of much less material per unit of energy stored. Though this concept has existed for a long time, very few products have been developed using it, and those that have been primarily aimed at structural heating applications instead of storage for electrical generation. The technology's primary challenge is recovering the latent heat stored in the phase change. The isothermal nature of phase changes causes a large energy loss to entropy. Further research in the area is underway and may find an effective solution.

Molten Salt

Molten salt generally refers to some combination of potassium nitrate, sodium nitrite, or sodium nitrate. Other salts have been used and experimented with, but these three make up almost the entirety of commercial molten-salt applications.[133] In the case of an NPP, steam would be diverted prior to going through the generator turbine and sent through a heat exchanger where the heat would be transferred to the molten salt.

Molten salts are efficient thermal sinks due to their high boiling points, high heat density, and low vapor pressure. Molten salt TES also has relatively low storage capital costs, with many systems starting close to $30/kWh.[134] However, they do have significant costs in terms of cost per unit power, with the associated steam turbines, generators, and other equipment costing in the neighborhood of $500 per kW.[23] Though they have historically been used primarily in concentrated solar plants and for experimental nuclear systems, these salts perform well for energy-storage applications across the board. Round-trip energy efficiency is roughly 70%.[133]

Although molten salts are still being researched as a storage option and are constantly improving, their temperature profile struggles to meet that of most LWR plants, the generating facilities of which operate around 300°C. The melting point of molten salts is nearly always in excess of 265°C. This decreases the available temperatures by over 82% and leaves a very narrow range of temperatures for heating, dramatically decreasing the energy density of the system and significantly increasing the cost. With this change, costs are around $175 per kWh. Unlike other, more commercially developed energy storage options that have manufacturers which include the cost of removal in the purchase price, molten salt is a newer technology, and any such installation would have to be removed by its user. Thus, end-of-life costs must be accounted for. This change could bring the cost per kW to $750 and the cost per kWh to $262.5.

High-Temperature Concrete

Concrete can be used to store energy by heating large concrete blocks to high-temperatures. The heat is used to either boil water or heat air which, in turn, spins a turbine to generate electricity when needed. Research has shown that concrete blocks have been able to cycle to 500°C approximately 2,500 times without unacceptable levels of material degradation.[125] This result indicates the technology's viability as an energy-storage option.

High-temperature concrete also has the potential to be inexpensive, with costs around $1 per kWh. Costs per unit power are greater due to the cost associated with steam turbines, generators, and other equipment coming out to around $500 per kW. As with molten-salt storage, the temperature at which a nuclear plant operates is significantly lower than the maximum temperature for high-temperature concrete, shrinking the range of temperatures available for storage by 60%. This change brings the cost per kWh to $2.50.

Despite these very promising numbers, the technology is still nascent, and it is not feasible to deploy at scale. Significant developments need to be made to increase its heat-transfer rate, and the technology must be demonstrated at scale. Research is funded and ongoing. Should the technology develop, overcome material heat-transfer limitations, and undergo a successful commercial trial, it would be transformative for energy storage.

Summary—Thermal Energy Storage

Key metrics for TES methods are displayed in Table C-22, and more-detailed data on cycling regimes are provided in Table C-23 through Table C-31.

Table C-22. Summary of key metrics for different thermal energy storage options.

Thermal Storage Option	Installation Cost ($/kW)	Installation Cost ($/kWh)	Round-Trip Efficiency (%)	Cycle Life	Lifetime	Self-Discharge Rate (%/month)
Molten Salt	750	267.5	70	10,000	25	26
High-Temperature Concrete	500	2.50	70	2,500	30	26

Table C-23. Financial analysis of thermal energy storage options for 4-hour storage cycled 12 times annually.

Thermal Energy Storage Option	LCOS ($/MWh)	LCOE ($/MWh)
Molten Salt	2,601.63	35,292,550.76
High-Temperature Concrete	1490.31	20,216,901

Table C-24. Financial analysis of thermal energy storage options for 4-hour storage cycled 60 times annually.

Thermal Energy Storage Option	LCOS ($/MWh)	LCOE ($/MWh)
Molten Salt	521.81	4,680.00

High-Temperature Concrete	301.47	2,714.37

Table C-25. Financial analysis of thermal energy storage options for 4-hour storage cycled 200 times annually.

Thermal Energy Storage Option	LCOS ($/MWh)	LCOE ($/MWh)
Molten Salt	157.84	415.64
High-Temperature Concrete	145.86	320.00

Table C-26. Financial analysis of thermal energy storage options for 4-hour storage cycled daily.

Thermal Energy Storage Option	LCOS ($/MWh)	LCOE ($/MWh)
Molten Salt	87.33	193.60
High-Temperature Concrete	145.06	305.03

Table C-27. Financial analysis of thermal energy storage options for 12-hour storage cycled 12 times annually.

Thermal Energy Storage Option	LCOS ($/MWh)	LCOE ($/MWh)
Molten Salt	2,068.51	28,060,475.81
High-Temperature Concrete	961.07	13,037,432.03

Table C-28. Financial analysis of thermal energy storage options for 12-hour storage cycled 60 times annually.

Thermal Energy Storage Option	LCOS ($/MWh)	LCOE ($/MWh)
Molten Salt	414.88	3,726.10
High-Temperature Concrete	204.55	1,849.74

Table C-29. Financial analysis of thermal energy storage options for 12-hour storage cycled 200 times annually.

Thermal Energy Storage Option	LCOS ($/MWh)	LCOE ($/MWh)
Molten Salt	125.50	335.59
High-Temperature Concrete	102.49	278.66

Table C-30. Financial analysis of thermal energy storage options for 12-hour storage cycled daily.

Thermal Energy Storage Option	LCOS ($/MWh)	LCOE ($/MWh)
Molten Salt	69.44	159.05
High-Temperature Concrete	99.48	217.05

Table C-31. Financial analysis of thermal energy storage options for 100-hour storage cycled bi-annually.

Thermal Energy Storage Option	LCOS ($/MWh)	LCOE ($/MWh)
Molten Salt	10,997.10	n/a

High-Temperature Concrete	4,669.07	n/a

This analysis shows that if high-temperature concrete develops as a technology, it will be a very efficient option. This performance is due to its much lower cost per capacity than molten salt combined with lows in other key areas, such as cost per unit power, efficiency, and self-discharge rate. The relatively low cycle life of high-temperature concrete of 2,500 cycles does cause it to falter against molten salt at higher cycling rates because it is not able to cycle as many times. Despite its economic efficiency in certain scenarios, high-temperature concrete cannot be seriously considered for use at this time due to its current lack of technological development.

Molten salt has a lower LCOE than high-temperature concrete in only the two daily-cycling scenarios, but its much more-progressed technological development makes it an option that can be seriously considered.

C-6. Summary of Energy Storage Options

The energy storage options that held the lowest LCOE in scenarios for any storage category are considered further in the following final comparison. Sodium-sulfur, lead-acid, and lithium-ion batteries, LAES, above-ground CAES, and molten salt are compared and contrasted in summary in Table C-32 and by cycling regime in Table C-33 through Table C-41.

Table C-32. Summary of key metrics for different energy storage options.

Energy Storage Option	Installation Cost ($/kW)	Installation Cost ($/kWh)	Round-Trip Efficiency (%)	Cycle Life	Lifetime	Self-Discharge Rate (%/month)
Lead Acid Battery	80	200 160-250	80	2500 79-84	15	2
Lithium-ion Battery	120	340	85	7500 90-94	20	1
LAES	1,000	200	70 60-70	15,000	30	1–2
CAES (Above Ground)	1,500 1050-2544	165 355-490	52 50-55	10,000	25	0
Molten Salt	500	267.5	70	10,000	25	26

Table C-33. Financial analysis of energy storage options for 4-hour storage cycled 12 times annually.

Energy Storage Option	LCOS ($/MWh)	LCOE ($/MWh)
Sodium Sulfur Batteries	6,371.44	7,989.30
Lead Acid Batteries	1,979.56	4,559.59
Lithium-ion Batteries	2,599.17	4,176.24
LAES	2,591.27	5,050.48
CAES (Above Ground)	3,168.36	6,118.00
Molten Salt	2,601.63	35,292,550.76

Table C-34. Financial analysis of energy storage options for 4-hour storage cycled 60 times annually.

Energy Storage Option	LCOS ($/MWh)	LCOE ($/MWh)
Sodium Sulfur Batteries	1,323.08	1,678.85
Lead Acid Batteries	420.27	619.04
Lithium-ion Batteries	530.63	688.63
LAES	519.24	813.54
CAES (Above Ground)	640.44	1,256.51
Molten Salt	521.81	4,680.00

Table C-35. Financial analysis of energy storage options for 4-hour storage cycled 200 times annually.

Energy Storage Option	LCOS ($/MWh)	LCOE ($/MWh)
Sodium Sulfur Batteries	441.50	576.87
Lead Acid Batteries	202.39	287.39
Lithium-ion Batteries	168.87	227.35
LAES	156.64	252.91
CAES (Above Ground)	198.13	406.02
Molten Salt	157.84	415.64

Table C-36. Financial analysis of energy storage options for 4-hour storage cycled daily.

Energy Storage Option	LCOS ($/MWh)	LCOE ($/MWh)
Sodium Sulfur Batteries	402.98	528.73
Lead Acid Batteries	190.90	268.50
Lithium-ion Batteries	99.03	142.68
LAES	86.39	149.66
CAES (Above Ground)	112.52	241.38
Molten Salt	87.33	193.60

Table C-37. Financial analysis of mechanical energy storage options for 12-hour storage cycled 12 times annually.

Energy Storage Option	LCOS ($/MWh)	LCOE ($/MWh)
Sodium Sulfur Batteries	6,198.78	7,773.47
Lead Acid Batteries	1,910.10	4439.07
Lithium-ion Batteries	2,501.09	4,019.59
LAES	1,883.27	3,677.40
CAES (Above Ground)	2,036.80	3,941.93
Molten Salt	2,068.51	28,060,475.81

Table C-38. Financial analysis of mechanical energy storage options for 12-hour storage cycled 60 times annually.

Energy Storage Option	LCOS ($/MWh)	LCOE ($/MWh)
Sodium Sulfur Batteries	1,287.22	1,634.03
Lead Acid Batteries	405.53	598.20
Lithium-ion Batteries	510.61	663.59
LAES	377.37	598.09
CAES (Above Ground)	411.71	816.75
Molten Salt	414.88	3,726.10

Table C-39. Financial analysis of mechanical energy storage options for 12-hour storage cycled 200 times annually.

Energy Storage Option	LCOS ($/MWh)	LCOE ($/MWh)
Sodium Sulfur Batteries	429.53	561.92
Lead Acid Batteries	195.29	278.28
Lithium-ion Batteries	162.50	219.71
LAES	113.84	190.64
CAES (Above Ground)	127.37	269.94
Molten Salt	125.50	335.59

Table C-40. Financial analysis of energy storage options for 12-hour storage cycled daily.

Energy Storage Option	LCOS ($/MWh)	LCOE ($/MWh)
Sodium Sulfur Batteries	392.06	515.07
Lead Acid Batteries	184.21	259.96
Lithium-ion Batteries	95.29	130.65
LAES	62.78	115.60
CAES (Above Ground)	72.33	164.10
Molten Salt	69.44	159.05

Table C-41. Financial analysis of energy storage options for 100-hour storage cycled bi-annually.

Energy Storage Option	LCOS ($/MWh)	LCOE ($/MWh)
Sodium Sulfur Batteries	36,447.82	45,584.78
Lead Acid Batteries	11,135.81	555,766.53
Lithium-ion Batteries	14,684.27	108,170.96
LAES	9,426.79	84,327.95
CAES (Above Ground)	9,213.05	17,742.41
Molten Salt	10,997.10	n/a

When these options are compared directly, four stand out as the most economically efficient option in different scenarios—lead-acid and Li-ion batteries, LAES, and above-ground CAES.

Lead-acid batteries have the lowest LCOE in the 60 cycle, 4-hour scenario. This advantage exists due to their low cost and the inability of their limited cycle lives to impact their performance significantly in low-cycle scenarios. This scenario highlights the low levels of degradation when kept at high levels of charge. Their closest competitor in this scenario is Li-ion batteries, which has an LCOE of $688.63, compared to lead acid's $619.04. Lead-acid batteries are also the second lowest LCOE option in the 12 cycle, 4-hour scenario with an LCOE of $4,559.59, slightly higher than lithium-ion's $4,176.24. They are held back in this scenario by their slightly higher self-discharge rate: 2% per month to lithium-ion's 1% per month. Self-discharge begins to be a significant factor in lower-cycle scenarios due to the increased spans of time between charging and discharging.

Lithium-ion batteries stand out in the 12-cycle, 200-cycle, and daily cycle 4-hour scenarios. The low cost per unit power of $120 per kW, moderate cost per kWh capacity of $320 per kWh, and relatively high cycle life of 7,500 cycles give it an advantage over other options in these categories. In the 12-cycle category, it is trailed by lead-acid batteries for the reasons discussed above. In the other two, it is trailed closely by LAES. Highview Power's new CRYOBattery LAES system has a relatively high efficiency of 70%, with a very low cost per kWh capacity of just $200. Ultimately, it is LAES's lower efficiency—70% compared to lithium-ion's 85%—and its much-higher cost per unit power of $1,000 per kW that result in a higher LCOE in these lower duration scenarios.

As duration increases, LAES's low cost of $200 per kW capacity becomes an increasing advantage. It has the lowest LCOE in all four 12-hour scenarios. In the 12-cycle scenario, it is followed by above-ground CAES, which is lifted above other technologies in this low-cycle scenario by its lack of any self-discharge. In the 60-cycle scenario, lead acid batteries come in with a $0.11 higher LCOE. It is competitive in this scenario due to its low cost and this scenario's provides an ideal number of cycles to minimize the technology's flaws, like the 4-hour 60-cycle scenario. In the 200 and daily cycle 12-hour scenarios, LAES is followed by lithium-ion batteries, which are competitive for the same reasons they are in a 4-hour scenario. However, in these longer-duration scenarios, LAES's low cost per kWh gives it an edge.

In the long duration scenario, above-ground CAES easily has the lowest LCOE. In this extremely low-cycle scenario, self-discharge rate becomes easily the most important factor.

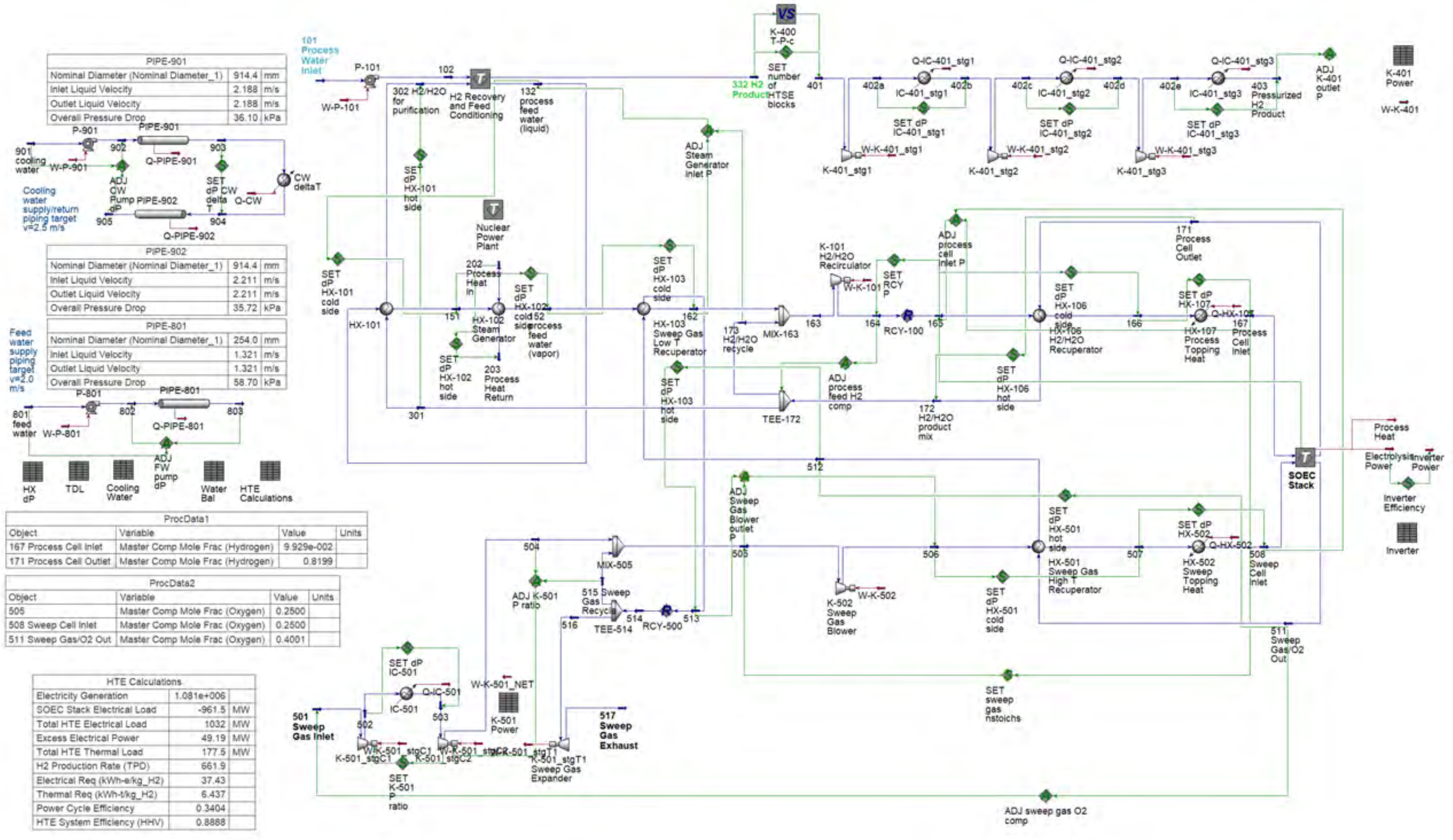
Above-ground CAES's 0% self-discharge rate allows it to overcome factors such as expensive cost per unit power and lower efficiency. The closest competitor in this scenario is sodium sulfur batteries which also have 0% self-discharge, though the batteries' high maintenance costs and high cost per kWh make it nearly three times as expensive as above-ground CAES.

Energy storage is an important and rapidly developing field in the move toward cleaner energy production. The massive field of technologies has the potential to successfully store energy integrated with NPPs as they begin to flex increasingly due to increased renewables penetration. Though there are hundreds of different developed and potential technologies for energy storage, only a select few standout. While a daily use scenario—such as the 4-hour daily scenario in which lithium-ion comes in at $142.68 per MWh—might be competitive at MISO wholesale electricity prices, a less frequently cycling scenario, such as the 4-hour 60 annual cycle scenario in which lead-acid batteries come in at $619.04 per MWh, is less likely to be profitable. Seasonal storage scenarios, such as the 12-hour 12 annual cycle scenario in which LAES comes in at $3,677.40 or the 100-hour two annual cycle scenario in which above-ground CAES comes in at $17,742.4, are likely to be compared with the cost of natural gas peaker plants. Decisions on profitability must be made, but if they are successfully achieved, energy storage technologies present a proven and reliable option for the use of excess nuclear electricity.

APPENDIX D
HTSE PROCESS ANALYSIS SUPPLEMENTAL INFORMATION

APPENDIX D
HTSE PROCESS ANALYSIS SUPPLEMENTAL INFORMATION

D-1. HTSE Detailed Process Flow Diagrams



PIPE-901		
Nominal Diameter (Nominal Diameter_1)	914.4	mm
Inlet Liquid Velocity	2.188	m/s
Outlet Liquid Velocity	2.188	m/s
Overall Pressure Drop	36.10	kPa

PIPE-902		
Nominal Diameter (Nominal Diameter_1)	914.4	mm
Inlet Liquid Velocity	2.211	m/s
Outlet Liquid Velocity	2.211	m/s
Overall Pressure Drop	35.72	kPa

PIPE-801		
Nominal Diameter (Nominal Diameter_1)	254.0	mm
Inlet Liquid Velocity	1.321	m/s
Outlet Liquid Velocity	1.321	m/s
Overall Pressure Drop	58.70	kPa

ProcData1

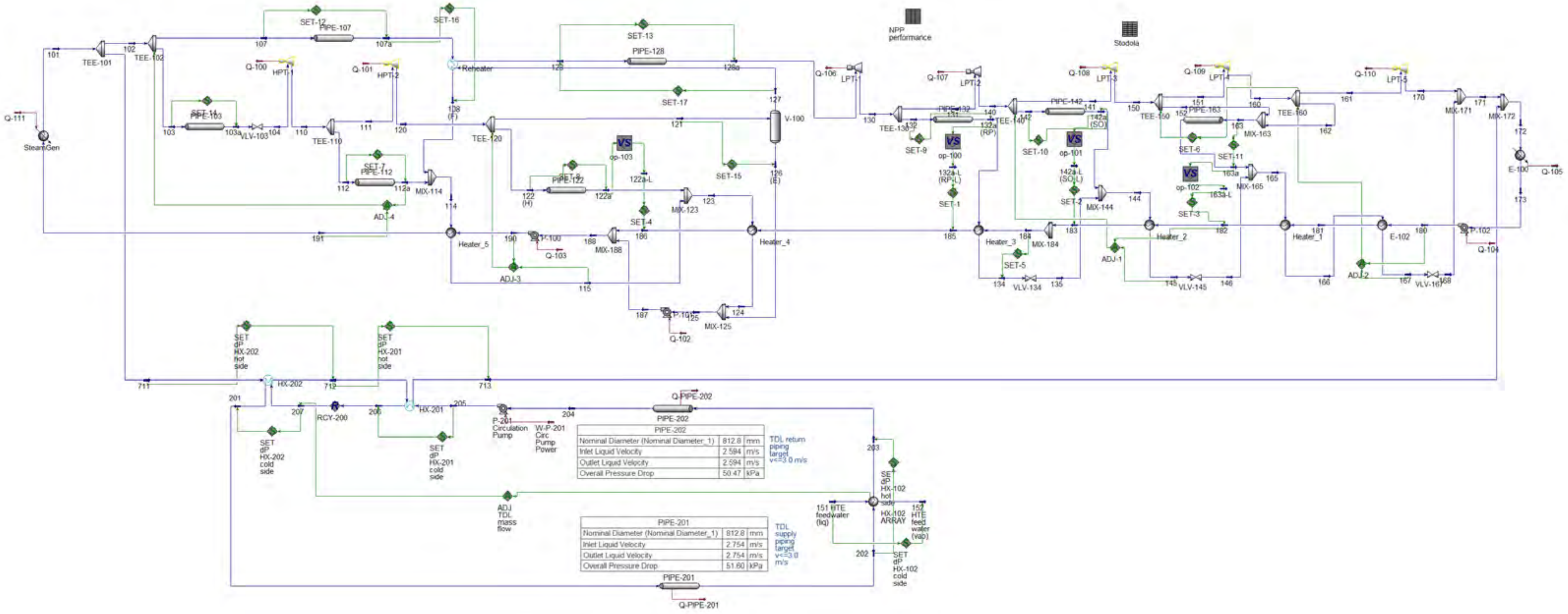
Object	Variable	Value	Units
167 Process Cell Inlet	Master Comp Mole Frac (Hydrogen)	9.929e-002	
171 Process Cell Outlet	Master Comp Mole Frac (Hydrogen)	0.8199	

ProcData2

Object	Variable	Value	Units
505	Master Comp Mole Frac (Oxygen)	0.2500	
508 Sweep Cell Inlet	Master Comp Mole Frac (Oxygen)	0.2500	
511 Sweep Gas/O2 Out	Master Comp Mole Frac (Oxygen)	0.4001	

HTE Calculations		
Electricity Generation	1.081e+006	
SOEC Stack Electrical Load	-961.5	MW
Total HTE Electrical Load	1032	MW
Excess Electrical Power	49.19	MW
Total HTE Thermal Load	177.5	MW
H2 Production Rate (TPD)	661.9	
Electrical Req (kWh-e/kg_H2)	37.43	
Thermal Req (kWh-t/kg_H2)	6.437	
Power Cycle Efficiency	0.3404	
HTE System Efficiency (HHV)	0.8888	

Figure D-1. Process Flow Diagram 1.



PIPE-202		
Nominal Diameter (Nominal Diameter_1)	812.8	mm
Inlet Liquid Velocity	2.594	m/s
Outlet Liquid Velocity	2.594	m/s
Overall Pressure Drop	50.47	kPa

PIPE-201		
Nominal Diameter (Nominal Diameter_1)	812.8	mm
Inlet Liquid Velocity	2.754	m/s
Outlet Liquid Velocity	2.754	m/s
Overall Pressure Drop	51.60	kPa

Figure D-2. Process Flow Diagram 2.

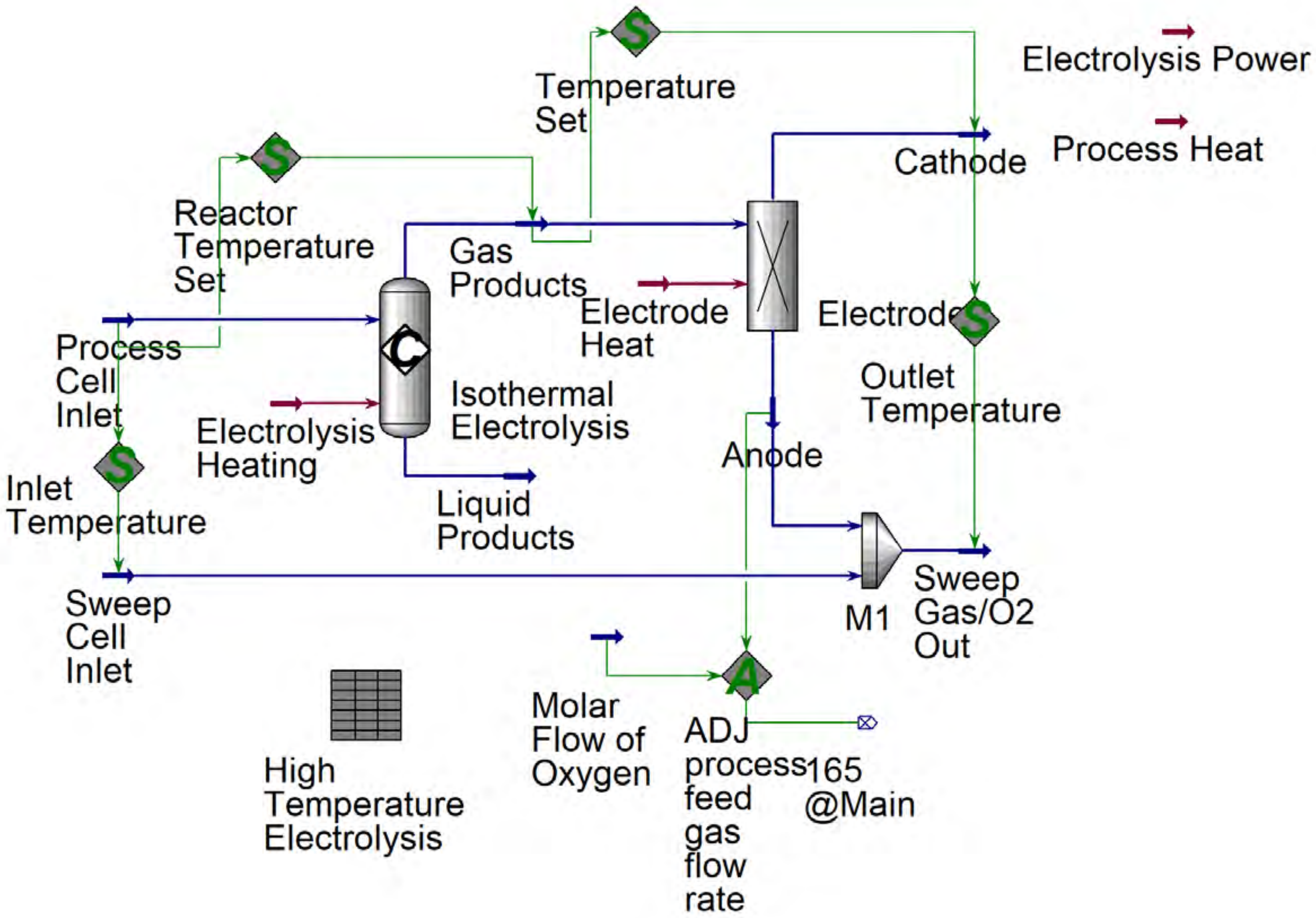


Figure D-3. Process Flow Diagram 3.

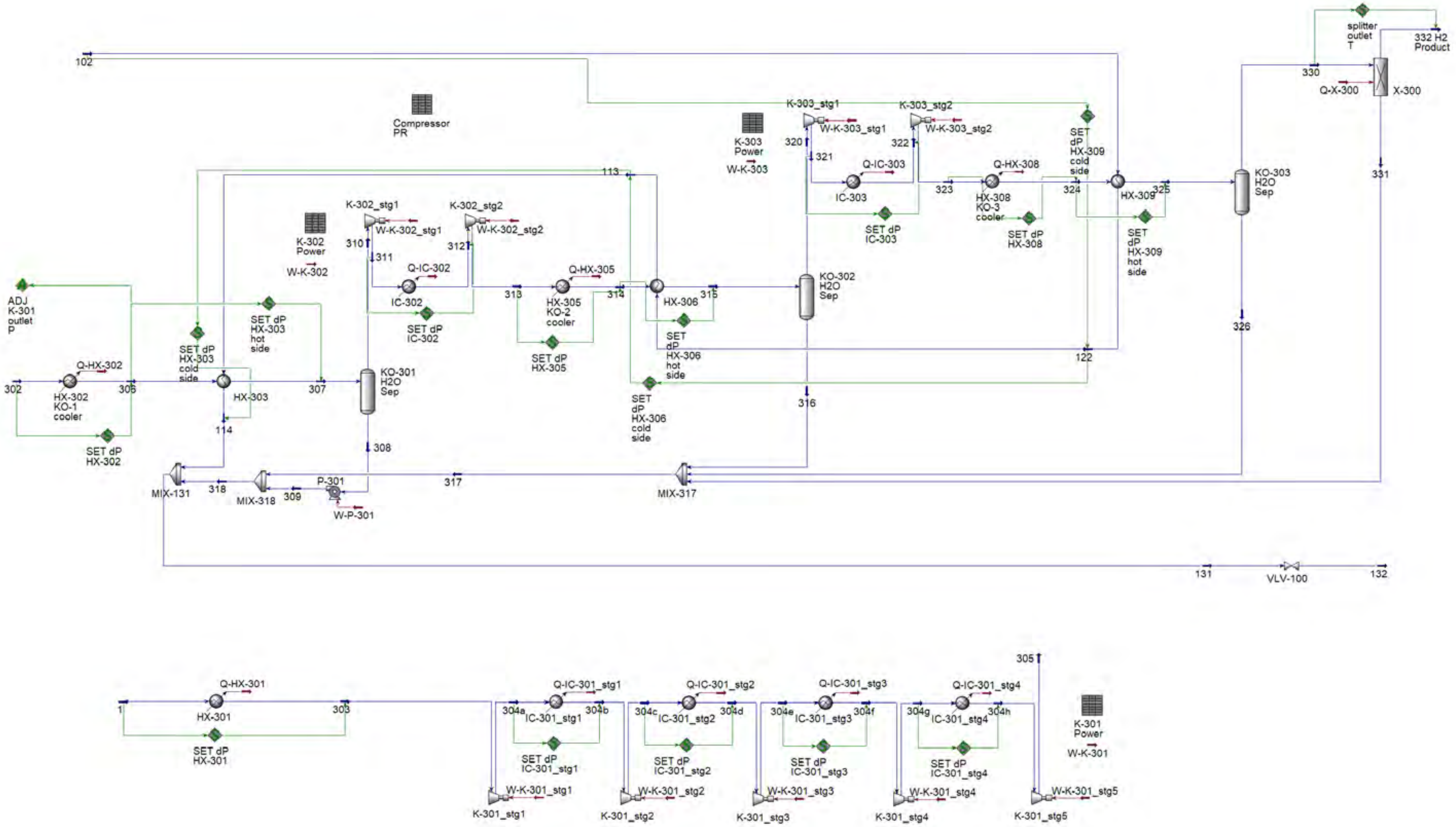


Figure D-4. Process Flow Diagram 4.

HTSE Process Flow Diagrams

Subprocess area boundaries are included; flowsheet controls and energy streams are not shown.

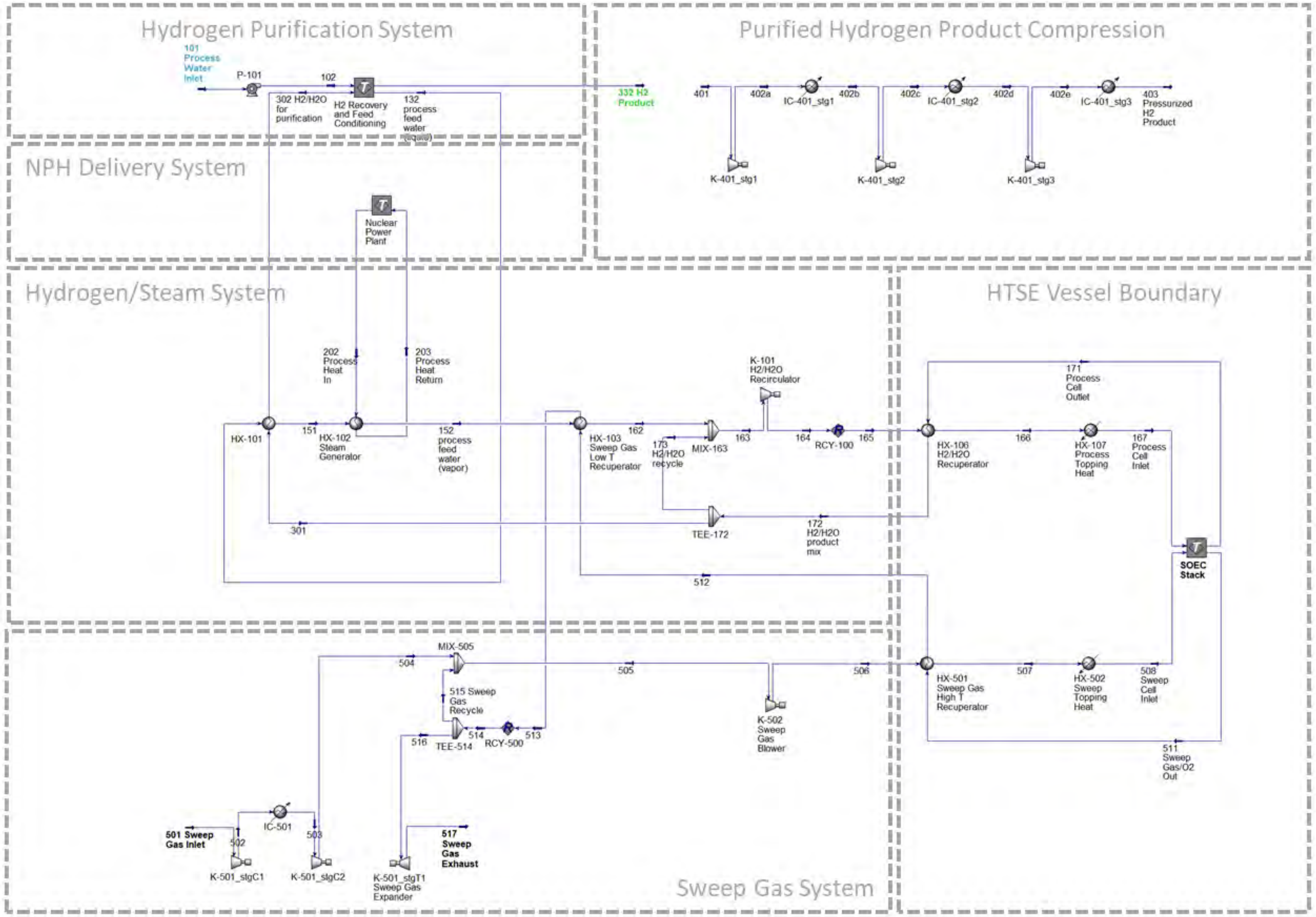


Figure D-5. Process Flow Diagram 5.

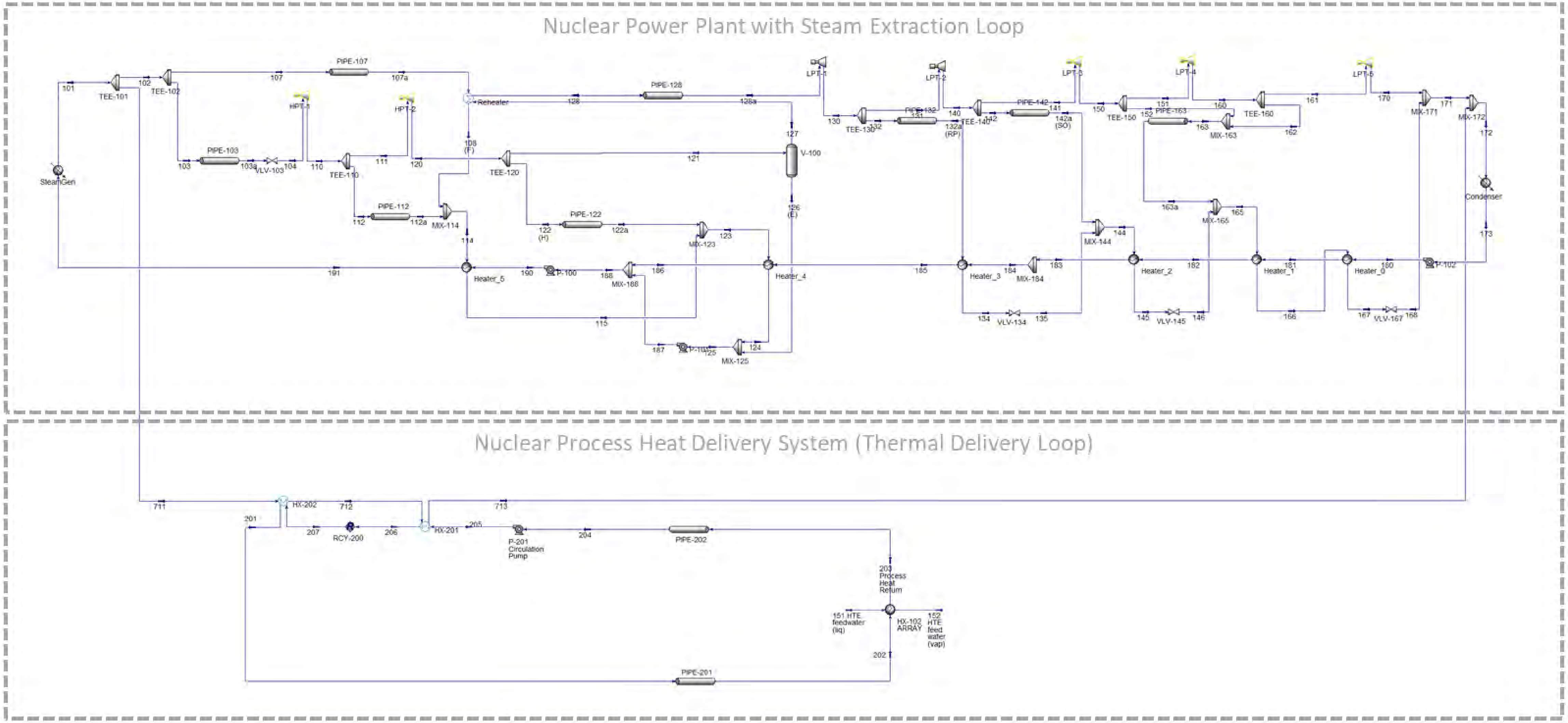


Figure D-6. Process Flow Diagram 6.

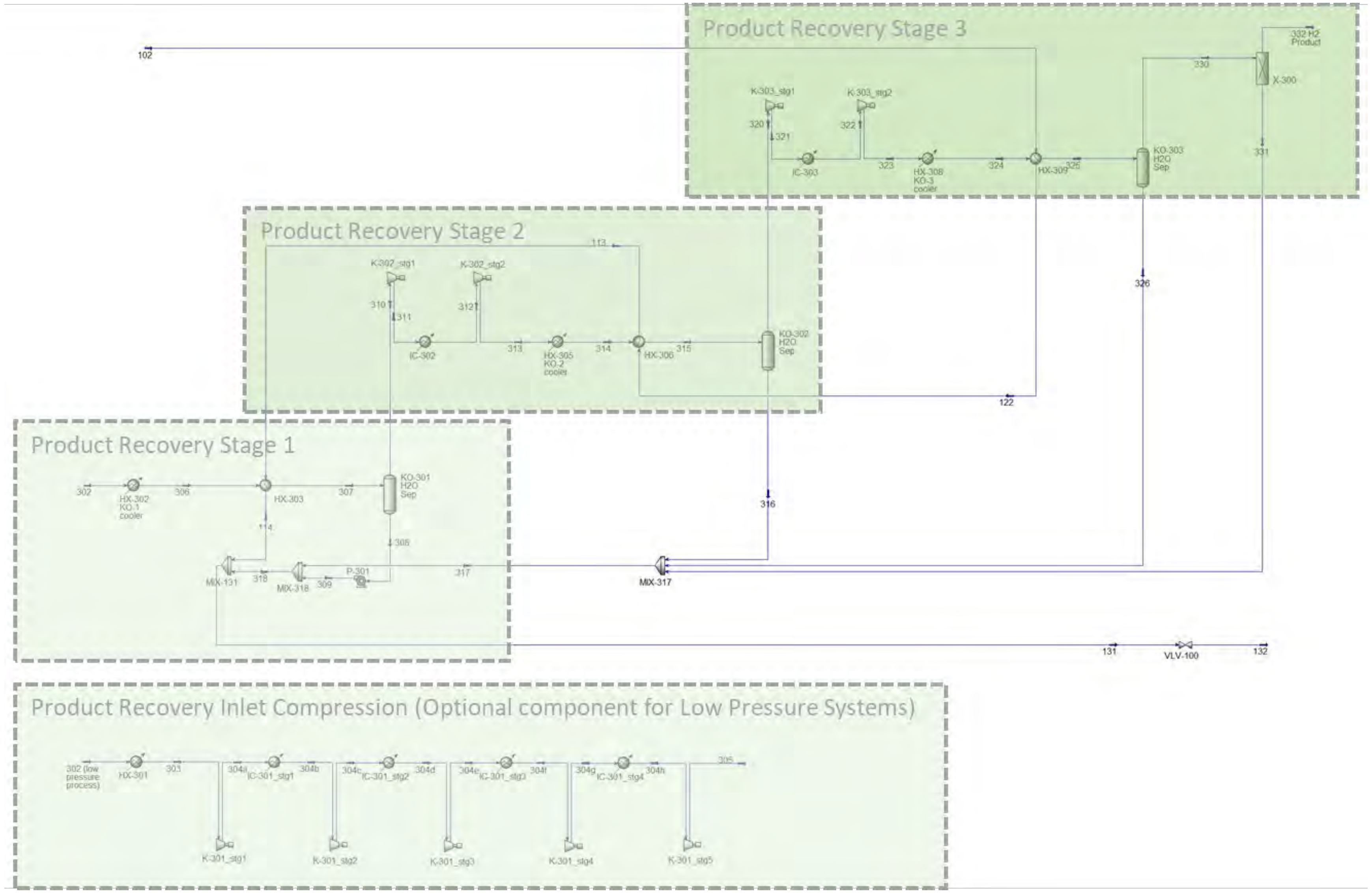


Figure D-7. Process Flow Diagram 7.

Prairie Island LWR-HTSE Capital Costs

Table D-1. PI NPP-HTSE capital-cost breakdown.

Equipment		Process System	Scale of supported system (TPD H_2)	Equipment Capacity	Unit	Equipment Description	APEA model	Equipment Costs (uninstalled costs of equipment in one HTE block)	Quantity	Scaling Factor [or Progress Rate]	Uninstalled Cost (scaled, FOAK plant, 2020$)	Uninstalled Cost (scaled, NOAK plant, 2020$)	Installation Cost Factor	Installed Cost (scaled, FOAK plant, 2020$)	Installed Cost (scaled, NOAK plant, 2020$)
HTSE Module (stacks + balance-of-module)	modular	HTSE system	17,416	25,300	kW-dc	SOEC Module (electrolysis stacks, module assembly, high & low temp recuperators, topping heaters)		$0	38	[1.000]	$0	$0	1.00	$0	$0
HTSE Vessel Shell	modular	HTSE system	17,416	1	unit	Horizontal drum (L 84" × D 42", 500°C, 500 kPa)	HT HORIZ DRUM	$360,978	38	[0.936]	$10,867,822	$9,755,814	5.13	$55,781,743	$50,074,088
HTSE Vessel Isolation Valves	modular	HTSE system	17,416	1	unit	Isolation Valves, 4 per module		$135,040	38	[0.936]	$4,065,596	$3,649,599	4.00	$16,262,384	$14,598,397
SOE Cells	modular	HTSE system	17,416	25,300	kWdc			$4,297,787	38	[1.000]	$163,315,924	$163,315,924	1.00	$163,315,924	$163,315,924
SOEC Module Assembly	modular	HTSE system	17,416	195.84	m² (cell area)			$425,161	38	[0.936]	$12,800,133	$11,490,408	1.00	$12,800,133	$11,490,408
SOEC Electrical Connector Assemblies	modular	HTSE system	17,416	2.516544	MWe			$36,079	38	[0.936]	$1,086,207	$975,065	3.00	$3,258,620	$2,925,194
Sleeved Process Connections	modular	HTSE system	17,416	1	unit			$394,796	38	[0.936]	$11,885,957	$10,669,771	1.00	$11,885,957	$10,669,771
HX-501 Sweep Gas High-Temperature Recuperator	modular	HTSE system	17,416	382.7	kW			$188,161	38	[0.936]	$5,664,891	$5,085,252	1.00	$5,664,891	$5,085,252
HX-106 H_2/H_2O Recuperator	modular	HTSE system	17,416	200.9	kW			$217,395	38	[0.936]	$6,545,031	$5,875,336	1.00	$6,545,031	$5,875,336
HX-502 Sweep Gas Topping Heater	modular	HTSE system	17,416	10	kW			$15,925	38	[0.936]	$479,435	$430,379	5.77	$2,767,107	$2,483,973
HX-107 H_2/H_2O Topping Heater	modular	HTSE system	17,416	51.9	kW			$51,223	38	[0.936]	$1,542,154	$1,384,359	5.77	$8,900,694	$7,989,965
HTSE Block Container (shipping container)	modular	HTSE system	17,416	1	unit			$9,000	38	[0.936]	$270,959	$243,234	1.00	$270,959	$243,234
grid interconnect	modular	HTSE system	17,416	17,416	kg/day H_2			$0	38	[0.936]	$0	$0	1.00	$0	$0
control module	modular	HTSE system	17,416	17,416	kg/day H_2			$0	38	[0.936]	$0	$0	1.00	$0	$0
thermocouples	modular	HTSE system	17,416	17,416	kg/day H_2			$0	38	[0.936]	$0	$0	1.00	$0	$0
pressure sensors	modular	HTSE system	17,416	17,416	kg/day H_2			$0	38	[0.936]	$0	$0	1.00	$0	$0
plumbing and fittings	modular	HTSE system	17,416	17,416	kg/day H_2			$0	38	[0.936]	$0	$0	1.00	$0	$0
hardware	modular	HTSE system	17,416	17,416	kg/day H_2			$0	38	[0.936]	$0	$0	1.00	$0	$0
frame/housing	modular	HTSE system	17,416	17,416	kg/day H_2			$45,955	38	[0.936]	$1,383,544	$1,241,978	1.00	$1,383,544	$1,241,978
Rectifier/Power Supply	modular	Feed & Utility System	17,416	25	MVA	One rectifier unit per HTSE block		$3,879,606	38	[0.936]	$116,801,682	$104,850,393	1.00	$116,801,682	$104,850,393
Disconnect Switch	modular	Feed & Utility System	17,416	20	kV	20 kV	BELSDISCNCT SW	$4,116	38	[0.936]	$123,919	$111,240	1.00	$123,919	$111,240
Transformer	modular	Feed & Utility System	17,416	25.2	MVA	25 MVA load, 20 kV primary voltage	BELSTRANSFORM	$458,557	38	[0.936]	$13,805,594	$12,392,989	1.00	$13,805,594	$12,392,989
Switch Board	modular	Feed & Utility System	17,416	25.2	MVA	25 MVA	BELSSWITCH BRD	$71,387	38	[0.936]	$2,149,223	$1,929,311	1.00	$2,149,223	$1,929,311
DC Bus Power Distribution	modular	Feed & Utility System	17,416	25.2	MVA	25 MVA, 20 m	BELSBUS DUCT	$137,875	38	[0.936]	$4,150,949	$3,726,219	1.00	$4,150,949	$3,726,219
DC Bus Power Distribution	modular	Feed & Utility System	17,416	5760	A	one inlet and one outlet DC bus bar per HTSE block; DC power distribution to each of the four stack columns within each SOEC module		$0	38	[0.936]	$0	$0	3.00	$0	$0

Equipment		Process System	Scale of supported system (TPD H_2)	Equipment Capacity	Unit	Equipment Description	APEA model	Equipment Costs (uninstalled costs of equipment in one HTE block)	Quantity	Scaling Factor [or Progress Rate]	Uninstalled Cost (scaled, FOAK plant, 2020$)	Uninstalled Cost (scaled, NOAK plant, 2020$)	Installation Cost Factor	Installed Cost (scaled, FOAK plant, 2020$)	Installed Cost (scaled, NOAK plant, 2020$)
Power Pole Lines	conventional	Feed & Utility System	661,792	1000	m	1000 MVA, 1 km	BELSPOLE LINE		1	[0.000]	$928,884	$928,884	1.00	$928,884	$928,884
Purified Water Storage Tank	modular	Feed & Utility System	17,416	311.3	m^3	48 hrs storage (Shaw Report, 2009) @ 1.8 kg/s	VT STORAGE	$97,482	38	[0.936]	$2,934,847	$2,634,550	2.13	$6,257,799	$5,617,493
PIPE-801 Feed Water Supply Piping	conventional	Feed & Utility System	661,792	250	mm	1000 m length; nominal pipe diameter selected to achieve ~2.0 m/s average fluid velocity	BPIPIPE		1	0.325	$0	$0	4.00	$0	$0
P-801 Feed Water Supply Pump	conventional	Feed & Utility System	661,792	67.2	L/s	7.5 kW driver	CP CENTRIF		1	0.217	$0	$0	5.15	$0	$0
Water Pretreatment Filter/Softener System	conventional	Feed & Utility System	661,792						1	0.600	$2,241,667	$2,241,667	1.00	$2,241,667	$2,241,667
Water Treatment RO/EDI System	conventional	Feed & Utility System	661,792						1	0.600	$9,881,831	$9,881,831	1.00	$9,881,831	$9,881,831
PIPE-901 Cooling Water Supply Piping	conventional	Feed & Utility System	661,792	900	mm	1000 m length; nominal pipe diameter selected to achieve ~2.5 m/s average fluid velocity	BPIPIPE		1	0.690	$0	$0	4.00	$0	$0
PIPE-902 Cooling Water Return Piping	conventional	Feed & Utility System	661,792	900	mm	1000 m length; nominal pipe diameter selected to achieve ~2.5 m/s average fluid velocity	BPIPIPE		1	0.690	$0	$0	4.00	$0	$0
P-901 Cooling Water Recirculation Pump	conventional	Feed & Utility System	661,792	1319.2	L/s	1000 kW driver	CP CENTRIF		1	0.625	$0	$0	2.41	$0	$0
CT-901 Cooling Tower	conventional	Feed & Utility System	661,792	1319.2	L/s	14°C range, 4°C approach, 20°C wet bulb	CTWCOOLING		1	0.572	$1,037,055	$1,037,055	1.47	$1,524,397	$1,524,397
Air Filter	modular	Air Sweep Gas System	17,416	17,416	kg/day H_2			$2,298	38	[0.936]	$69,177	$62,099	1.00	$69,177	$62,099
K-501C Sweep Gas Compressor	modular	Air Sweep Gas System	17,416	13647.4	m^3/hr	800 kW driver; 2 stage intercooled centrifugal air compressor	AC CENTRIF M	$666,684	38	[0.936]	$20,071,566	$18,017,819	1.33	$26,726,165	$23,991,512
K-501T Sweep Gas Exhaust Turbine	modular	Air Sweep Gas System	17,416	6865	m^3/hr	880 kW power output	TURTURBOEXP	$470,594	38	[0.936]	$14,167,966	$12,718,283	1.95	$27,685,963	$24,853,102
K-502 Sweep Gas Blower	modular	Air Sweep Gas System	17,416	5675.4	m^3/hr	107 kW driver	FN PROPELLER	$1,342	38	[0.936]	$40,413	$36,277	24.67	$996,843	$894,845
P-101 Water Pump	modular	Hydrogen/Steam System	17,416	1.8	L/s	2.2 kW driver	CP CENTRIF	$5,145	38	[0.936]	$154,891	$139,043	6.70	$1,037,098	$930,981
F-101 Water Filter	modular	Hydrogen/Steam System	17,416	1.8	L/s		F CARTRIDGE	$5,812	38	[0.936]	$174,970	$157,067	2.36	$412,665	$370,440
DI Polisher	modular	Hydrogen/Steam System	17,416	17,416	kg/day H_2			$6,893	38	[0.936]	$207,532	$186,297	1.00	$207,532	$186,297
Water Flow Meter	modular	Hydrogen/Steam System	17,416	17,416	kg/day H_2			$11,489	38	[0.936]	$345,886	$310,495	1.00	$345,886	$310,495
HX-101 Condenser & Water Preheater	modular	Hydrogen/Steam System	17,416	15.8	m^2	BEM exchanger, 1 × 300 mm dia shell, 67 tubes/shell × 25 mm × 3 m	HE TEMA EXCH	$27,381	38	[0.936]	$824,338	$739,991	4.70	$3,877,916	$3,481,123
HX-102 Feed Water Vaporizer	modular	Hydrogen/Steam System	17,416	165.9	m^2	BEM exchanger, 1 × 650 mm dia shell, 352 tubes/shell × 25 mm × 6 m	HE TEMA EXCH	$38,568	38	[0.936]	$1,161,140	$1,042,331	3.61	$4,189,085	$3,760,453
HX-103 Sweep Gas Low Temp Recuperator	modular	Hydrogen/Steam System	17,416	163.5	m^2	BEM exchanger, 1 × 650 mm dia shell, 347 tubes/shell × 25 mm × 6 m	HE TEMA EXCH	$38,461	38	[0.936]	$1,157,922	$1,039,442	5.99	$6,937,909	$6,228,014
K-101 hydrogen Recycle Blower	modular	Hydrogen/Steam System	17,416	4973.9	m^3/hr	52 kW driver	FN PROPELLER	$1,342	38	[0.936]	$40,413	$36,277	22.92	$926,121	$831,359
HX-303 Feedwater Heater #1	modular	Hydrogen Purification System	17,416	8.5	m^2	BEM exchanger, 1 × 250 mm dia shell, 36 tubes/shell × 25 mm × 3 m	HE TEMA EXCH	$12,252	38	[0.936]	$368,867	$331,124	5.30	$1,953,392	$1,753,519
HX-306 Feedwater Heater #2	modular	Hydrogen Purification System	17,416	7.5	m^2	BEM exchanger, 1 × 200 mm dia shell, 32 tubes/shell × 25 mm × 3 m	HE TEMA EXCH	$11,080	38	[0.936]	$333,584	$299,451	5.60	$1,866,788	$1,675,776
HX-309 Feedwater Heater #3	modular	Hydrogen Purification System	17,416	15.6	m^2	BEM exchanger, 1 × 300 mm dia shell, 66 tubes/shell × 25 mm × 3 m	HE TEMA EXCH	$14,702	38	[0.936]	$442,641	$397,349	5.01	$2,219,618	$1,992,504

Equipment		Process System	Scale of supported system (TPD H_2)	Equipment Capacity	Unit	Equipment Description	APEA model	Equipment Costs (uninstalled costs of equipment in one HTE block)	Quantity	Scaling Factor [or Progress Rate]	Uninstalled Cost (scaled, FOAK plant, 2020$)	Uninstalled Cost (scaled, NOAK plant, 2020$)	Installation Cost Factor	Installed Cost (scaled, FOAK plant, 2020$)	Installed Cost (scaled, NOAK plant, 2020$)
HX-301 Product Purification Area Inlet Stream Cooler	modular	Hydrogen Purification System	17,416	20.3	m²	BEM exchanger, 1 × mm dia shell, 43 tubes/shell × 25 mm × 6 m	HE TEMA EXCH	$0	38	[0.936]	$0	$0	5.59	$0	$0
HX-302 Separation Vessel #1 Precooler	modular	Hydrogen Purification System	17,416	8	m²	BEM exchanger, 1 × 150 mm dia shell, 17 tubes/shell × 25 mm × 6 m	HE TEMA EXCH	$11,932	38	[0.936]	$359,244	$322,486	5.56	$1,998,298	$1,793,829
HX-305 Separation Vessel #2 Precooler	modular	Hydrogen Purification System	17,416	3.3	m²	BEM exchanger, 1 × 100 mm dia shell, 7 tubes/shell × 25 mm × 6 m	HE TEMA EXCH	$10,760	38	[0.936]	$323,962	$290,813	6.05	$1,959,807	$1,759,277
HX-308 Separation Vessel #3 Precooler	modular	Hydrogen Purification System	17,416	44.3	m²	BEM exchanger, 1 × 350 mm dia shell, 94 tubes/shell × 25 mm × 6 m	HE TEMA EXCH	$11,187	38	[0.936]	$336,792	$302,331	5.90	$1,985,467	$1,782,312
P-301 KO-1 Outlet Pump	modular	Hydrogen Purification System	17,416	0.2	L/s	0.2 kW driver	CP CENTRIF	$4,921	38	[0.936]	$148,157	$132,997	6.48	$959,652	$861,459
K-301 H_2 Purification Multistage Compressor #1	modular	Hydrogen Purification System	17,416	0	m³/hr	900 kW driver; 3 stage reciprocating gas compressor	GC RECIP MOTR	$0	38	[0.936]	$0	$0	1.09	$0	$0
K-302 H_2 Purification Multistage Compressor #2	modular	Hydrogen Purification System	17,416	2509.1	m³/hr	300 kW driver; 2 stage reciprocating gas compressor	GC RECIP MOTR	$869,261	38	[0.936]	$26,170,493	$23,492,696	1.13	$29,683,017	$26,645,815
K-303 H_2 Purification Multistage Compressor #3	modular	Hydrogen Purification System	17,416	1027.6	m³/hr	250 kW driver; 2 stage reciprocating gas compressor	GC RECIP MOTR	$724,179	38	[0.936]	$21,802,570	$19,571,705	1.14	$24,887,394	$22,340,886
KO-301 H_2 Separation Vessel #1	modular	Hydrogen Purification System	17,416	3.4	m³	L/D = 3	VT CYLINDER	$19,603	38	[0.936]	$590,177	$529,789	5.50	$3,245,973	$2,913,841
KO-302 H_2 Separation Vessel #2	modular	Hydrogen Purification System	17,416	3.4	m³	L/D = 3	VT CYLINDER	$20,029	38	[0.936]	$603,007	$541,306	5.20	$3,136,918	$2,815,945
KO-303 H_2 Separation Vessel #3	modular	Hydrogen Purification System	17,416	3.4	m³	L/D = 3	VT CYLINDER	$23,438	38	[0.936]	$705,646	$633,444	4.38	$3,088,806	$2,772,756
Hydrogen H_2O Adsorber Regen Cooler	modular	Hydrogen Purification System	17,416	17,416	kg/day H_2			$0	38	[0.936]	$0	$0	2.91	$0	$0
Regen Water Knockout Drum	modular	Hydrogen Purification System	17,416	17,416	kg/day H_2			$0	38	[0.936]	$0	$0	1.41	$0	$0
X-300 hydrogen H_2O Adsorbers	modular	Hydrogen Purification System	17,416	17,416	kg/day H_2			$0	38	[0.936]	$0	$0	3.59	$0	$0
Hydrogen H_2O Adsorber Regen Heater	modular	Hydrogen Purification System	17,416	17,416	kg/day H_2			$0	38	[0.936]	$0	$0	2.91	$0	$0
Backup Electric Boiler	modular	NPH Delivery System	17,416	17,416	kg/day H_2			$241,530	38	[0.936]	$7,271,650	$6,527,606	1.00	$7,271,650	$6,527,606
Backup natural gas Boiler	conventional	NPH Delivery System	661,792						1	[1.000]	$0	$0	1.31	$0	$0
PIPE-201 Nuclear Process Heat Piping (supply)	conventional	NPH Delivery System	661,792	800	mm	1000 m length; nominal pipe diameter selected to achieve <30 m/s average vapor or <3.0 m/s average liquid velocity	BPIPIPE		1	0.483	$1,224,911	$1,224,911	4.00	$4,899,643	$4,899,643
PIPE-202 Nuclear Process Heat Piping (return)	conventional	NPH Delivery System	661,792	800	mm	1000 m length; nominal pipe diameter selected to achieve <3.0 m/s average fluid velocity	BPIPIPE		1	0.483	$1,224,911	$1,224,911	4.00	$4,899,643	$4,899,643
P-201 Nuclear Process Heat Circulation Pump	conventional	NPH Delivery System	661,792	88.7	L/s	80 kW driver	CP CENTRIF		1	0.285	$282,257	$282,257	2.72	$767,403	$767,403

Equipment		Process System	Scale of supported system (TPD H_2)	Equipment Capacity	Unit	Equipment Description	APEA model	Equipment Costs (uninstalled costs of equipment in one HTE block)	Quantity	Scaling Factor [or Progress Rate]	Uninstalled Cost (scaled, FOAK plant, 2020$)	Uninstalled Cost (scaled, NOAK plant, 2020$)	Installation Cost Factor	Installed Cost (scaled, FOAK plant, 2020$)	Installed Cost (scaled, NOAK plant, 2020$)
HX-201 Nuclear Process Heat TDL HX	conventional	NPH Delivery System	661,792	237	m^2	BEM exchanger, 1 × 775 mm dia shell, 503 tubes/shell × 25 mm × 6 m	HE TEMA EXCH		1	0.700	$174,881	$174,881	2.20	$384,483	$384,483
HX-202 Nuclear Process Heat TDL HX	conventional	NPH Delivery System	661,792	4122.4	m^2	BEM exchanger, 4 × 1600 mm dia shell, 2187 tubes/shell × 25 mm × 6 m	HE TEMA EXCH		1	0.700	$1,030,967	$1,030,967	1.53	$1,573,716	$1,573,716
Therminol-66 HTF	conventional	NPH Delivery System	661,792	700,000	kg	2 × 1000 m × 700 mm × 900 kg/m^3			1	1.000	$2,681,383	$2,681,383	1.00	$2,681,383	$2,681,383
K-401 High-Pressure Multistage Reciprocating Compressor	conventional	Product Compression System	661,792	17,650	m^3/hr	18,000 kW driver; 3 stage intercooled reciprocating gas compressor	GC RECIP MOTR		1	0.602	$4,104,907	$4,104,907	1.30	$5,336,379	$5,336,379
H_2 Product Storage	conventional	Product Compression System	661,792						1	0.570	$0	$0	1.41	$0	$0
CB-101 Control Building	conventional	Control System	661,792			15 m × 25 m	BCIVBUILDING		1	0.000	$498,879	$498,879	1.00	$498,879	$498,879
OC-101 Operator Center	conventional	Control System	661,792			5 operator display units	BINSOPER CENT		1	0.000	$329,225	$329,225	1.00	$329,225	$329,225
											$483,388,625	$453,262,096		$625,742,826	$582,104,942

Monticello LWR-HTSE Capital Costs

Table D-2. Monticello NPP-HTSE capital-cost breakdown.

Equipment		Process System	Scale of supported system (TPD H_2)	Equipment Capacity	Unit	Equipment Description	APEA Model	Equipment Costs (uninstalled costs of equipment in one HTE block)	Quantity	Scaling Factor [or Progress Rate]	Uninstalled Cost (scaled, FOAK plant, 2020$)	Uninstalled Cost (scaled, NOAK plant, 2020$)	Installation Cost Factor	Installed Cost (scaled, FOAK plant, 2020$)	Installed Cost (scaled, NOAK plant, 2020$)
HTSE Module (stacks + balance-of-module)	modular	HTSE system	17,416	25,300	kW-dc	SOEC Module (electrolysis stacks, module assembly, high & low temp recuperators, topping heaters)		$0	22	[1.000]	$0	$0	1.00	$0	$0
HTSE Vessel Shell	modular	HTSE system	17,416	1	unit	Horizontal drum (L 84" x D 42", 500°C, 500 kPa)	HT HORIZ DRUM	$360,978	22	[0.936]	$6,515,911	$5,648,103	5.13	$33,444,497	$28,990,261
HTSE Vessel Isolation Valves	modular	HTSE system	17,416	1	unit	Isolation Valves, 4 per module		$135,040	22	[0.936]	$2,437,568	$2,112,926	4.00	$9,750,273	$8,451,703
SOE Cells	modular	HTSE system	17,416	25,300	kWdc			$4,297,787	22	[1.000]	$94,551,325	$94,551,325	1.00	$94,551,325	$94,551,325
SOEC Module Assembly	modular	HTSE system	17,416	195.84	m² (cell area)			$425,161	22	[0.936]	$7,674,446	$6,652,341	1.00	$7,674,446	$6,652,341
SOEC Electrical Connector Assemblies	modular	HTSE system	17,416	2.516544	MWe			$36,079	22	[0.936]	$651,246	$564,511	3.00	$1,953,738	$1,693,533
Sleeved Process Connections	modular	HTSE system	17,416	1	unit			$394,796	22	[0.936]	$7,126,343	$6,177,236	1.00	$7,126,343	$6,177,236
HX-501 Sweep Gas High-Temperature Recuperator	modular	HTSE system	17,416	382.7	kW			$188,161	22	[0.936]	$3,396,441	$2,944,093	1.00	$3,396,441	$2,944,093
HX-106 H_2/H_2O Recuperator	modular	HTSE system	17,416	200.9	kW			$217,395	22	[0.936]	$3,924,138	$3,401,510	1.00	$3,924,138	$3,401,510
HX-502 Sweep Gas Topping Heater	modular	HTSE system	17,416	10	kW			$15,925	22	[0.936]	$287,450	$249,167	5.77	$1,659,046	$1,438,090
HX-107 H_2/H_2O Topping Heater	modular	HTSE system	17,416	51.9	kW			$51,223	22	[0.936]	$924,614	$801,471	5.77	$5,336,499	$4,625,769
HTSE Block Container (shipping container)	modular	HTSE system	17,416	1	unit			$9,000	22	[0.936]	$162,456	$140,820	1.00	$162,456	$140,820
grid interconnect	modular	HTSE system	17,416	17,416	kg/day H_2			$0	22	[0.936]	$0	$0	1.00	$0	$0
control module	modular	HTSE system	17,416	17,416	kg/day H_2			$0	22	[0.936]	$0	$0	1.00	$0	$0

Equipment		Process System	Scale of supported system (TPD H_2)	Equipment Capacity	Unit	Equipment Description	APEA Model	Equipment Costs (uninstalled costs of equipment in one HTE block)	Quantity	Scaling Factor [or Progress Rate]	Uninstalled Cost (scaled, FOAK plant, 2020$)	Uninstalled Cost (scaled, NOAK plant, 2020$)	Installation Cost Factor	Installed Cost (scaled, FOAK plant, 2020$)	Installed Cost (scaled, NOAK plant, 2020$)
thermocouples	modular	HTSE system	17,416	17,416	kg/day H_2			$0	22	[0.936]	$0	$0	1.00	$0	$0
pressure sensors	modular	HTSE system	17,416	17,416	kg/day H_2			$0	22	[0.936]	$0	$0	1.00	$0	$0
plumbing and fittings	modular	HTSE system	17,416	17,416	kg/day H_2			$0	22	[0.936]	$0	$0	1.00	$0	$0
hardware	modular	HTSE system	17,416	17,416	kg/day H_2			$0	22	[0.936]	$0	$0	1.00	$0	$0
frame/housing	modular	HTSE system	17,416	17,416	kg/day H_2			$45,955	22	[0.936]	$829,518	$719,040	1.00	$829,518	$719,040
Rectifier/Power Supply	modular	Feed & Utility System	17,416	25	MVA	One rectifier unit per HTSE block		$3,879,606	22	[0.936]	$70,029,606	$60,702,859	1.00	$70,029,606	$60,702,859
Disconnect Switch	modular	Feed & Utility System	17,416	20	kV	20 kV	BELSDISCNCT SW	$4,116	22	[0.936]	$74,297	$64,402	1.00	$74,297	$64,402
Transformer	modular	Feed & Utility System	17,416	25.2	MVA	25 MVA load, 20 kV primary voltage	BELSTRANSFORM	$458,557	22	[0.936]	$8,277,281	$7,174,888	1.00	$8,277,281	$7,174,888
Switch Board	modular	Feed & Utility System	17,416	25.2	MVA	25 MVA	BELSSWITCH BRD	$71,387	22	[0.936]	$1,288,588	$1,116,970	1.00	$1,288,588	$1,116,970
DC Bus Power Distribution	modular	Feed & Utility System	17,416	25.2	MVA	25 MVA, 20 m	BELSBUS DUCT	$137,875	22	[0.936]	$2,488,743	$2,157,285	1.00	$2,488,743	$2,157,285
DC Bus Power Distribution	modular	Feed & Utility System	17,416	5760	A	one inlet and one outlet DC bus bar per HTSE block; DC power distribution to each of the four stack columns within each SOEC module		$0	22	[0.936]	$0	$0	3.00	$0	$0
Power Pole Lines	conventional	Feed & Utility System	383,143	1000	m	1000 MVA, 1 km	BELSPOLE LINE		1	[0.000]	$928,884	$928,884	1.00	$928,884	$928,884
Purified Water Storage Tank	modular	Feed & Utility System	17,416	311.3	m³	48 hrs storage (Shaw Report, 2009) @ 1.8 kg/s	VT STORAGE	$97,482	22	[0.936]	$1,759,616	$1,525,266	2.13	$3,751,925	$3,252,233
PIPE-801 Feed Water Supply Piping	conventional	Feed & Utility System	383,143	250	mm	1000 m length; nominal pipe diameter selected to achieve ~2.0 m/s average fluid velocity	BPIPIPE		1	0.325	$0	$0	4.00	$0	$0
P-801 Feed Water Supply Pump	conventional	Feed & Utility System	383,143	38.9	L/s	7.5 kW driver	CP CENTRIF		1	0.217	$0	$0	5.15	$0	$0

Equipment		Process System	Scale of supported system (TPD H_2)	Equipment Capacity	Unit	Equipment Description	APEA Model	Equipment Costs (uninstalled costs of equipment in one HTE block)	Quantity	Scaling Factor [or Progress Rate]	Uninstalled Cost (scaled, FOAK plant, 2020$)	Uninstalled Cost (scaled, NOAK plant, 2020$)	Installation Cost Factor	Installed Cost (scaled, FOAK plant, 2020$)	Installed Cost (scaled, NOAK plant, 2020$)
Water Pretreatment Filter/Softener System	conventional	Feed & Utility System	383,143						1	0.600	$1,614,933	$1,614,933	1.00	$1,614,933	$1,614,933
Water Treatment RO/EDI System	conventional	Feed & Utility System	383,143						1	0.600	$7,119,031	$7,119,031	1.00	$7,119,031	$7,119,031
PIPE-901 Cooling Water Supply Piping	conventional	Feed & Utility System	383,143	900	mm	1000 m length; nominal pipe diameter selected to achieve ~2.5 m/s average fluid velocity	BPIPIPE		1	0.690	$0	$0	4.00	$0	$0
PIPE-902 Cooling Water Return Piping	conventional	Feed & Utility System	383,143	900	mm	1000 m length; nominal pipe diameter selected to achieve ~2.5 m/s average fluid velocity	BPIPIPE		1	0.690	$0	$0	4.00	$0	$0
P-901 Cooling Water Recirculation Pump	conventional	Feed & Utility System	383,143	763.7	L/s	1000 kW driver	CP CENTRIF		1	0.625	$0	$0	2.41	$0	$0
CT-901 Cooling Tower	conventional	Feed & Utility System	383,143	763.7	L/s	14°C range, 4°C approach, 20°C wet bulb	CTWCOOLING		1	0.572	$758,726	$758,726	1.47	$1,115,274	$1,115,274
Air Filter	modular	Air Sweep Gas System	17,416	17,416	kg/day H_2			$2,298	22	[0.936]	$41,476	$35,952	1.00	$41,476	$35,952
K-501C Sweep Gas Compressor	modular	Air Sweep Gas System	17,416	13647.4	m^3/hr	800 kW driver; 2 stage intercooled centrifugal air compressor	AC CENTRIFM	$666,684	22	[0.936]	$12,034,106	$10,431,369	1.33	$16,023,937	$13,889,823
K-501T Sweep Gas Exhaust Turbine	modular	Air Sweep Gas System	17,416	6865	m^3/hr	880 kW power output	TURTURBOEXP	$470,594	22	[0.936]	$8,494,544	$7,363,216	1.95	$16,599,394	$14,388,638
K-502 Sweep Gas Blower	modular	Air Sweep Gas System	17,416	5675.4	m^3/hr	107 kW driver	FN PROPELLER	$1,342	22	[0.936]	$24,230	$21,003	24.67	$597,667	$518,068
P-101 Water Pump	modular	Hydrogen/Steam System	17,416	1.8	L/s	2.2 kW driver	CP CENTRIF	$5,145	22	[0.936]	$92,867	$80,498	6.70	$621,802	$538,989
F-101 Water Filter	modular	Hydrogen/Steam System	17,416	1.8	L/s		F CARTRIDGE	$5,812	22	[0.936]	$104,905	$90,933	2.36	$247,417	$214,465
DI Polisher	modular	Hydrogen/Steam System	17,416	17,416	kg/day H_2			$6,893	22	[0.936]	$124,428	$107,856	1.00	$124,428	$107,856

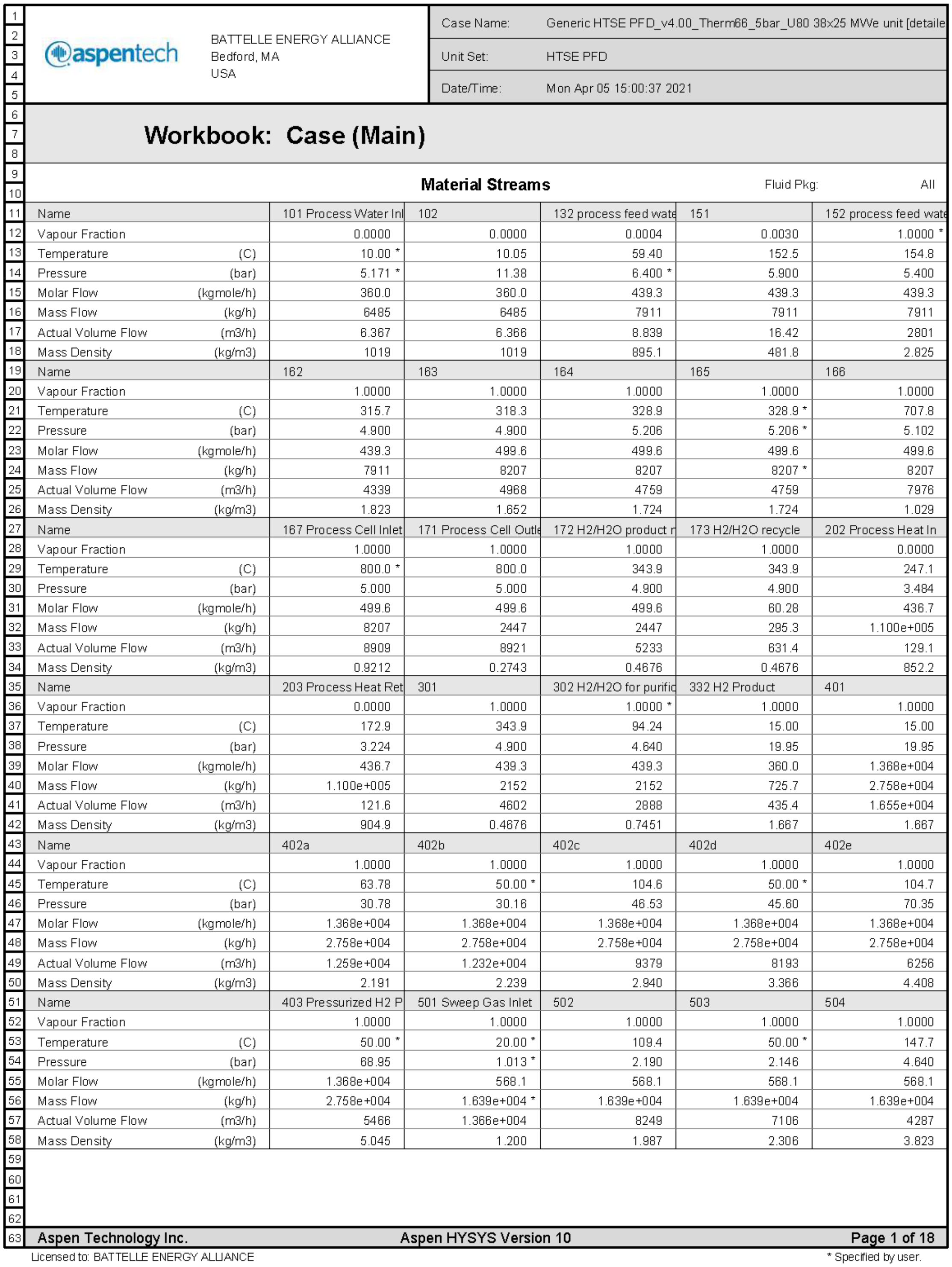
Equipment		Process System	Scale of supported system (TPD H_2)	Equipment Capacity	Unit	Equipment Description	APEA Model	Equipment Costs (uninstalled costs of equipment in one HTE block)	Quantity	Scaling Factor [or Progress Rate]	Uninstalled Cost (scaled, FOAK plant, 2020$)	Uninstalled Cost (scaled, NOAK plant, 2020$)	Installation Cost Factor	Installed Cost (scaled, FOAK plant, 2020$)	Installed Cost (scaled, NOAK plant, 2020$)
Water Flow Meter	modular	Hydrogen/Steam System	17,416	17,416	kg/day H_2			$11,489	22	[0.936]	$207,379	$179,760	1.00	$207,379	$179,760
HX-101 Condenser & Water Preheater	modular	Hydrogen/Steam System	17,416	15.8	m^2	BEM exchanger, 1 × 300 mm dia shell, 67 tubes/shell × 25 mm × 3 m	HE TEMA EXCH	$27,381	22	[0.936]	$494,240	$428,416	4.70	$2,325,043	$2,015,387
HX-102 Feed Water Vaporizer	modular	Hydrogen/Steam System	17,416	165.9	m^2	BEM exchanger, 1 × 650 mm dia shell, 352 tubes/shell × 25 mm × 6 m	HE TEMA EXCH	$38,568	22	[0.936]	$696,173	$603,455	3.61	$2,511,607	$2,177,104
HX-103 Sweep Gas Low Temp Recuperator	modular	Hydrogen/Steam System	17,416	163.5	m^2	BEM exchanger, 1 × 650 mm dia shell, 347 tubes/shell × 25 mm × 6 m	HE TEMA EXCH	$38,461	22	[0.936]	$694,244	$601,782	5.99	$4,159,692	$3,605,692
K-101 hydrogen Recycle Blower	modular	Hydrogen/Steam System	17,416	4973.9	m^3/hr	52 kW driver	FN PROPELLER	$1,342	22	[0.936]	$24,230	$21,003	22.92	$555,265	$481,313
HX-303 Feedwater Heater #1	modular	Hydrogen Purification System	17,416	8.5	m^2	BEM exchanger, 1 × 250 mm dia shell, 36 tubes/shell × 25 mm × 3 m	HE TEMA EXCH	$12,252	22	[0.936]	$221,158	$191,703	5.30	$1,171,175	$1,015,195
HX-306 Feedwater Heater #2	modular	Hydrogen Purification System	17,416	7.5	m^2	BEM exchanger, 1 × 200 mm dia shell, 32 tubes/shell × 25 mm × 3 m	HE TEMA EXCH	$11,080	22	[0.936]	$200,004	$173,367	5.60	$1,119,251	$970,186
HX-309 Feedwater Heater #3	modular	Hydrogen Purification System	17,416	15.6	m^2	BEM exchanger, 1 × 300 mm dia shell, 66 tubes/shell × 25 mm × 3 m	HE TEMA EXCH	$14,702	22	[0.936]	$265,390	$230,044	5.01	$1,330,794	$1,153,555
HX-301 Product Purification Area Inlet Stream Cooler	modular	Hydrogen Purification System	17,416	20.3	m^2	BEM exchanger, 1 × mm dia shell, 43 tubes/shell × 25 mm × 6 m	HE TEMA EXCH	$0	22	[0.936]	$0	$0	5.59	$0	$0
HX-302 Separation Vessel #1 Precooler	modular	Hydrogen Purification System	17,416	8	m^2	BEM exchanger, 1 × 150 mm dia shell, 17 tubes/shell × 25 mm × 6 m	HE TEMA EXCH	$11,932	22	[0.936]	$215,389	$186,703	5.56	$1,198,099	$1,038,533
HX-305 Separation Vessel #2 Precooler	modular	Hydrogen Purification System	17,416	3.3	m^2	BEM exchanger, 1 × 100 mm dia shell, 7 tubes/shell × 25 mm × 6 m	HE TEMA EXCH	$10,760	22	[0.936]	$194,234	$168,366	6.05	$1,175,022	$1,018,529
HX-308 Separation Vessel #3 Precooler	modular	Hydrogen Purification System	17,416	44.3	m^2	BEM exchanger, 1 × 350 mm dia shell, 94 tubes/shell × 25 mm × 6 m	HE TEMA EXCH	$11,187	22	[0.936]	$201,927	$175,034	5.90	$1,190,407	$1,031,865
P-301 KO-1 Outlet Pump	modular	Hydrogen Purification System	17,416	0.2	L/s	0.2 kW driver	CP CENTRIF	$4,921	22	[0.936]	$88,829	$76,998	6.48	$575,369	$498,740
K-301 H_2 Purification Multistage Compressor #1	modular	Hydrogen Purification System	17,416	0	m^3/hr	900 kW driver; 3 stage reciprocating gas compressor	GC RECIP MOTR	$0	22	[0.936]	$0	$0	1.09	$0	$0

Equipment		Process System	Scale of supported system (TPD H_2)	Equipment Capacity	Unit	Equipment Description	APEA Model	Equipment Costs (uninstalled costs of equipment in one HTE block)	Quantity	Scaling Factor [or Progress Rate]	Uninstalled Cost (scaled, FOAK plant, 2020$)	Uninstalled Cost (scaled, NOAK plant, 2020$)	Installation Cost Factor	Installed Cost (scaled, FOAK plant, 2020$)	Installed Cost (scaled, NOAK plant, 2020$)
K-302 H_2 Purification Multistage Compressor #2	modular	Hydrogen Purification System	17,416	2509.1	m³/hr	300 kW driver; 2 stage reciprocating gas compressor	GC RECIP MOTR	$869,261	22	[0.936]	$15,690,778	$13,601,035	1.13	$17,796,747	$15,426,524
K-303 H_2 Purification Multistage Compressor #3	modular	Hydrogen Purification System	17,416	1027.6	m³/hr	250 kW driver; 2 stage reciprocating gas compressor	GC RECIP MOTR	$724,179	22	[0.936]	$13,071,947	$11,330,987	1.14	$14,921,484	$12,934,197
KO-301 H_2 Separation Vessel #1	modular	Hydrogen Purification System	17,416	3.4	m³	L/D = 3	VT CYLINDER	$19,603	22	[0.936]	$353,846	$306,720	5.50	$1,946,155	$1,686,960
KO-302 H_2 Separation Vessel #2	modular	Hydrogen Purification System	17,416	3.4	m³	L/D = 3	VT CYLINDER	$20,029	22	[0.936]	$361,539	$313,388	5.20	$1,880,770	$1,630,284
KO-303 H_2 Separation Vessel #3	modular	Hydrogen Purification System	17,416	3.4	m³	L/D = 3	VT CYLINDER	$23,438	22	[0.936]	$423,077	$366,731	4.38	$1,851,924	$1,605,280
Hydrogen H_2O Adsorber Regen Cooler	modular	Hydrogen Purification System	17,416	17,416	kg/day H_2			$0	22	[0.936]	$0	$0	2.91	$0	$0
Regen Water Knockout Drum	modular	Hydrogen Purification System	17,416	17,416	kg/day H_2			$0	22	[0.936]	$0	$0	1.41	$0	$0
X-300 hydrogen H_2O Adsorbers	modular	Hydrogen Purification System	17,416	17,416	kg/day H_2			$0	22	[0.936]	$0	$0	3.59	$0	$0
Hydrogen H_2O Adsorber Regen Heater	modular	Hydrogen Purification System	17,416	17,416	kg/day H_2			$0	22	[0.936]	$0	$0	2.91	$0	$0
Backup Electric Boiler	modular	NPH Delivery System	17,416	17,416	kg/day H_2			$241,530	22	[0.936]	$4,359,790	$3,779,140	1.00	$4,359,790	$3,779,140
Backup natural gas Boiler	conventional	NPH Delivery System	383,143						1	[1.000]	$0	$0	1.31	$0	$0
PIPE-201 Nuclear Process Heat Piping (supply)	conventional	NPH Delivery System	383,143	800	mm	1000 m length; nominal pipe diameter selected to achieve <30 m/s average vapor or <3.0 m/s average liquid velocity	BPIPIPE		1	0.483	$940,838	$940,838	4.00	$3,763,351	$3,763,351
PIPE-202 Nuclear Process Heat Piping (return)	conventional	NPH Delivery System	383,143	800	mm	1000 m length; nominal pipe diameter selected to achieve <3.0 m/s average fluid velocity	BPIPIPE		1	0.483	$940,838	$940,838	4.00	$3,763,351	$3,763,351
P-201 Nuclear Process Heat Circulation Pump	conventional	NPH Delivery System	383,143	88.7	L/s	80 kW driver	CP CENTRIF		1	0.285	$241,600	$241,600	2.72	$656,864	$656,864

Equipment		Process System	Scale of supported system (TPD H_2)	Equipment Capacity	Unit	Equipment Description	APEA Model	Equipment Costs (uninstalled costs of equipment in one HTE block)	Quantity	Scaling Factor [or Progress Rate]	Uninstalled Cost (scaled, FOAK plant, 2020$)	Uninstalled Cost (scaled, NOAK plant, 2020$)	Installation Cost Factor	Installed Cost (scaled, FOAK plant, 2020$)	Installed Cost (scaled, NOAK plant, 2020$)
HX-201 Nuclear Process Heat TDL HX	conventional	NPH Delivery System	383,143	237	m²	BEM exchanger, 1 × 775 mm dia shell, 503 tubes/shell × 25 mm × 6 m	HE TEMA EXCH		1	0.700	$119,260	$119,260	2.20	$262,197	$262,197
HX-202 Nuclear Process Heat TDL HX	conventional	NPH Delivery System	383,143	4122.4	m²	BEM exchanger, 4 × 1600 mm dia shell, 2187 tubes/shell × 25 mm × 6 m	HE TEMA EXCH		1	0.700	$703,067	$703,067	1.53	$1,073,194	$1,073,194
Therminol-66 HTF	conventional	NPH Delivery System	383,143	700,000	kg	2 × 1000 m × 700 mm × 900 kg/m³			1	1.000	$1,552,380	$1,552,380	1.00	$1,552,380	$1,552,380
K-401 High-Pressure Multistage Reciprocating Compressor	conventional	Product Compression System	383,143	17,650	m³/hr	18,000 kW driver; 3 stage intercooled reciprocating gas compressor	GC RECIP MOTR		1	0.602	$2,953,435	$2,953,435	1.30	$3,839,465	$3,839,465
H_2 Product Storage	conventional	Product Compression System	383,143						1	0.570	$0	$0	1.41	$0	$0
CB-101 Control Building	conventional	Control System	383,143			15 m × 25 m	BCIVBUILDING		1	0.000	$498,879	$498,879	1.00	$498,879	$498,879
OC-101 Operator Center	conventional	Control System	383,143			5 operator display units	BINSOPER CENT		1	0.000	$329,225	$329,225	1.00	$329,225	$329,225
											$289,781,408	$266,270,764		$376,768,283	$342,713,423

HTSE Process Model Stream Tables

Table D-3. Aspentech HTSE process model.



BATTELLE ENERGY ALLIANCE
Bedford, MA
USA

Case Name: Generic HTSE PFD_v4.00_Therm66_5bar_U80 38x25 MWe unit [detaile

Unit Set: HTSE PFD

Date/Time: Mon Apr 05 15:00:37 2021

Workbook: Case (Main)

Material Streams

Fluid Pkg: All

Name		101 Process Water Inl	102	132 process feed wate	151	152 process feed wate
Vapour Fraction		0.0000	0.0000	0.0004	0.0030	1.0000 *
Temperature	(C)	10.00 *	10.05	59.40	152.5	154.8
Pressure	(bar)	5.171 *	11.38	6.400 *	5.900	5.400
Molar Flow	(kgmole/h)	360.0	360.0	439.3	439.3	439.3
Mass Flow	(kg/h)	6485	6485	7911	7911	7911
Actual Volume Flow	(m3/h)	6.367	6.366	8.839	16.42	2801
Mass Density	(kg/m3)	1019	1019	895.1	481.8	2.825

Name		162	163	164	165	166
Vapour Fraction		1.0000	1.0000	1.0000	1.0000	1.0000
Temperature	(C)	315.7	318.3	328.9	328.9 *	707.8
Pressure	(bar)	4.900	4.900	5.206	5.206 *	5.102
Molar Flow	(kgmole/h)	439.3	499.6	499.6	499.6	499.6
Mass Flow	(kg/h)	7911	8207	8207	8207 *	8207
Actual Volume Flow	(m3/h)	4339	4968	4759	4759	7976
Mass Density	(kg/m3)	1.823	1.652	1.724	1.724	1.029

Name		167 Process Cell Inlet	171 Process Cell Outle	172 H2/H2O product r	173 H2/H2O recycle	202 Process Heat In
Vapour Fraction		1.0000	1.0000	1.0000	1.0000	0.0000
Temperature	(C)	800.0 *	800.0	343.9	343.9	247.1
Pressure	(bar)	5.000	5.000	4.900	4.900	3.484
Molar Flow	(kgmole/h)	499.6	499.6	499.6	60.28	436.7
Mass Flow	(kg/h)	8207	2447	2447	295.3	1.100e+005
Actual Volume Flow	(m3/h)	8909	8921	5233	631.4	129.1
Mass Density	(kg/m3)	0.9212	0.2743	0.4676	0.4676	852.2

Name		203 Process Heat Ret	301	302 H2/H2O for purific	332 H2 Product	401
Vapour Fraction		0.0000	1.0000	1.0000 *	1.0000	1.0000
Temperature	(C)	172.9	343.9	94.24	15.00	15.00
Pressure	(bar)	3.224	4.900	4.640	19.95	19.95
Molar Flow	(kgmole/h)	436.7	439.3	439.3	360.0	1.368e+004
Mass Flow	(kg/h)	1.100e+005	2152	2152	725.7	2.758e+004
Actual Volume Flow	(m3/h)	121.6	4602	2888	435.4	1.655e+004
Mass Density	(kg/m3)	904.9	0.4676	0.7451	1.667	1.667

Name		402a	402b	402c	402d	402e
Vapour Fraction		1.0000	1.0000	1.0000	1.0000	1.0000
Temperature	(C)	63.78	50.00 *	104.6	50.00 *	104.7
Pressure	(bar)	30.78	30.16	46.53	45.60	70.35
Molar Flow	(kgmole/h)	1.368e+004	1.368e+004	1.368e+004	1.368e+004	1.368e+004
Mass Flow	(kg/h)	2.758e+004	2.758e+004	2.758e+004	2.758e+004	2.758e+004
Actual Volume Flow	(m3/h)	1.259e+004	1.232e+004	9379	8193	6256
Mass Density	(kg/m3)	2.191	2.239	2.940	3.366	4.408

Name		403 Pressurized H2 P	501 Sweep Gas Inlet	502	503	504
Vapour Fraction		1.0000	1.0000	1.0000	1.0000	1.0000
Temperature	(C)	50.00 *	20.00 *	109.4	50.00 *	147.7
Pressure	(bar)	68.95	1.013 *	2.190	2.146	4.640
Molar Flow	(kgmole/h)	1.368e+004	568.1	568.1	568.1	568.1
Mass Flow	(kg/h)	2.758e+004	1.639e+004 *	1.639e+004	1.639e+004	1.639e+004
Actual Volume Flow	(m3/h)	5466	1.366e+004	8249	7106	4287
Mass Density	(kg/m3)	5.045	1.200	1.987	2.306	3.823

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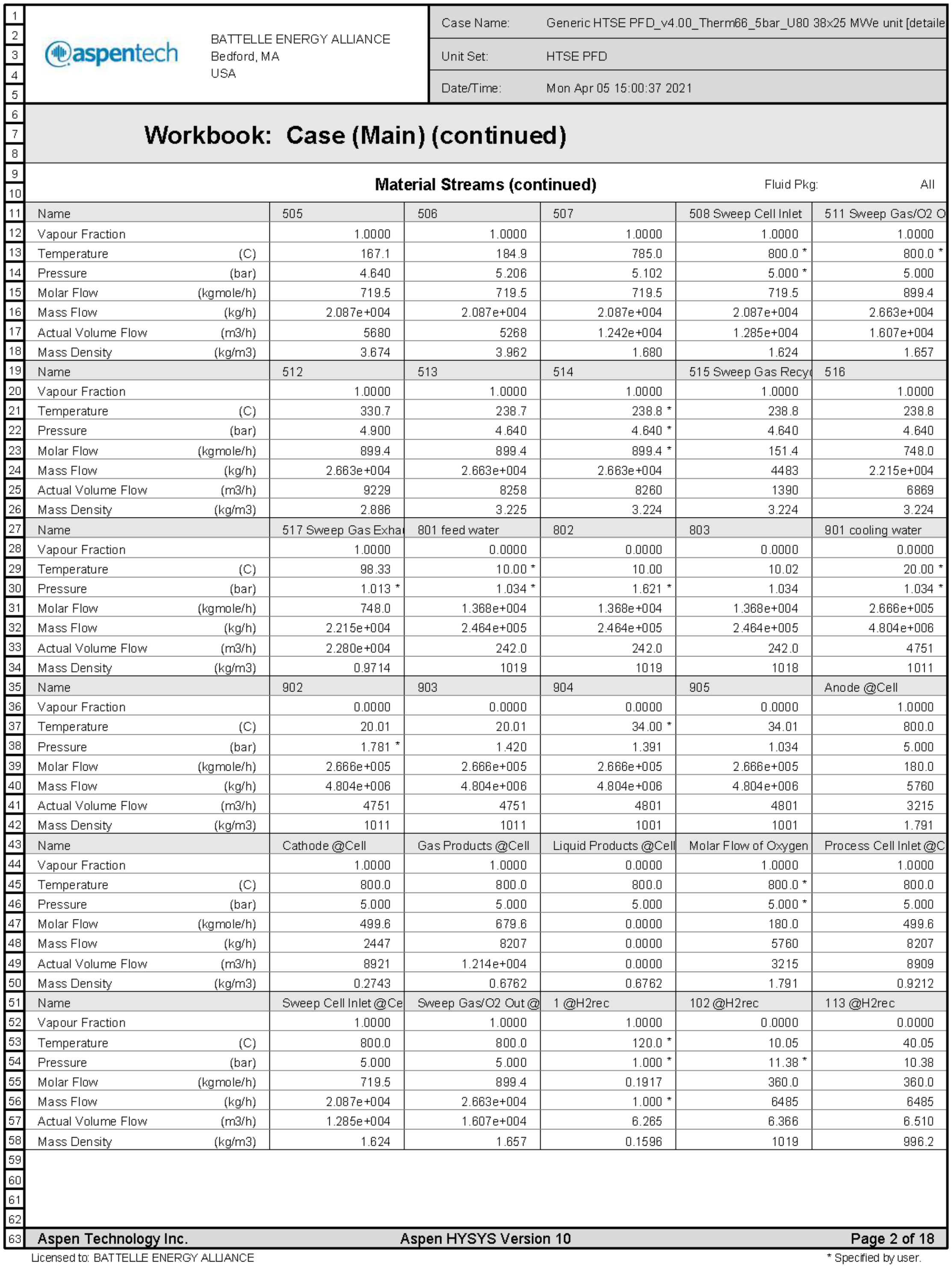

BATTELLE ENERGY ALLIANCE
Bedford, MA
USA

Case Name:	Generic HTSE PFD_v4.00_Therm66_5bar_U80 38x25 MWe unit [detaile
Unit Set:	HTSE PFD
Date/Time:	Mon Apr 05 15:00:37 2021

Workbook: Case (Main) (continued)

Material Streams (continued)

Fluid Pkg: All

Name	505	506	507	508 Sweep Cell Inlet	511 Sweep Gas/O2 O
Vapour Fraction	1.0000	1.0000	1.0000	1.0000	1.0000
Temperature (C)	167.1	184.9	785.0	800.0 *	800.0 *
Pressure (bar)	4.640	5.206	5.102	5.000 *	5.000
Molar Flow (kgmole/h)	719.5	719.5	719.5	719.5	899.4
Mass Flow (kg/h)	2.087e+004	2.087e+004	2.087e+004	2.087e+004	2.663e+004
Actual Volume Flow (m3/h)	5680	5268	1.242e+004	1.285e+004	1.607e+004
Mass Density (kg/m3)	3.674	3.962	1.680	1.624	1.657

Name	512	513	514	515 Sweep Gas Recy	516
Vapour Fraction	1.0000	1.0000	1.0000	1.0000	1.0000
Temperature (C)	330.7	238.7	238.8 *	238.8	238.8
Pressure (bar)	4.900	4.640	4.640 *	4.640	4.640
Molar Flow (kgmole/h)	899.4	899.4	899.4 *	151.4	748.0
Mass Flow (kg/h)	2.663e+004	2.663e+004	2.663e+004	4483	2.215e+004
Actual Volume Flow (m3/h)	9229	8258	8260	1390	6869
Mass Density (kg/m3)	2.886	3.225	3.224	3.224	3.224

Name	517 Sweep Gas Exha	801 feed water	802	803	901 cooling water
Vapour Fraction	1.0000	0.0000	0.0000	0.0000	0.0000
Temperature (C)	98.33	10.00 *	10.00	10.02	20.00 *
Pressure (bar)	1.013 *	1.034 *	1.621 *	1.034	1.034 *
Molar Flow (kgmole/h)	748.0	1.368e+004	1.368e+004	1.368e+004	2.666e+005
Mass Flow (kg/h)	2.215e+004	2.464e+005	2.464e+005	2.464e+005	4.804e+006
Actual Volume Flow (m3/h)	2.280e+004	242.0	242.0	242.0	4751
Mass Density (kg/m3)	0.9714	1019	1019	1018	1011

Name	902	903	904	905	Anode @Cell
Vapour Fraction	0.0000	0.0000	0.0000	0.0000	1.0000
Temperature (C)	20.01	20.01	34.00 *	34.01	800.0
Pressure (bar)	1.781 *	1.420	1.391	1.034	5.000
Molar Flow (kgmole/h)	2.666e+005	2.666e+005	2.666e+005	2.666e+005	180.0
Mass Flow (kg/h)	4.804e+006	4.804e+006	4.804e+006	4.804e+006	5760
Actual Volume Flow (m3/h)	4751	4751	4801	4801	3215
Mass Density (kg/m3)	1011	1011	1001	1001	1.791

Name	Cathode @Cell	Gas Products @Cell	Liquid Products @Cell	Molar Flow of Oxygen	Process Cell Inlet @C
Vapour Fraction	1.0000	1.0000	0.0000	1.0000	1.0000
Temperature (C)	800.0	800.0	800.0	800.0 *	800.0
Pressure (bar)	5.000	5.000	5.000	5.000 *	5.000
Molar Flow (kgmole/h)	499.6	679.6	0.0000	180.0	499.6
Mass Flow (kg/h)	2447	8207	0.0000	5760	8207
Actual Volume Flow (m3/h)	8921	1.214e+004	0.0000	3215	8909
Mass Density (kg/m3)	0.2743	0.6762	0.6762	1.791	0.9212

Name	Sweep Cell Inlet @Ce	Sweep Gas/O2 Out @	1 @H2rec	102 @H2rec	113 @H2rec
Vapour Fraction	1.0000	1.0000	1.0000	0.0000	0.0000
Temperature (C)	800.0	800.0	120.0 *	10.05	40.05
Pressure (bar)	5.000	5.000	1.000 *	11.38 *	10.38
Molar Flow (kgmole/h)	719.5	899.4	0.1917	360.0	360.0
Mass Flow (kg/h)	2.087e+004	2.663e+004	1.000 *	6485	6485
Actual Volume Flow (m3/h)	1.285e+004	1.607e+004	6.265	6.366	6.510
Mass Density (kg/m3)	1.624	1.657	0.1596	1019	996.2

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Workbook: Case (Main) (continued)

Material Streams (continued)

Fluid Pkg: All

Name		114 @H2rec	122 @H2rec	131 @H2rec	132 @H2rec	302 @H2rec
Vapour Fraction		0.0000	0.0000	0.0004	0.0004	1.0000
Temperature	(C)	60.38	20.05	59.34	59.40	94.25
Pressure	(bar)	9.876	10.88	9.189	6.400	4.640
Molar Flow	(kgmole/h)	360.0	360.0	439.3	439.3	439.3
Mass Flow	(kg/h)	6485	6485	7911	7911	2152
Actual Volume Flow	(m3/h)	6.615	6.413	8.588	8.839	2888
Mass Density	(kg/m3)	980.4	1011	921.3	895.1	0.7451
Name		303 @H2rec	304a @H2rec	304b @H2rec	304c @H2rec	304d @H2rec
Vapour Fraction		1.0000 *	1.0000	1.0000 *	1.0000	1.0000 *
Temperature	(C)	59.66	59.67	59.23	59.23	58.80
Pressure	(bar)	0.9800	0.9800	0.9604	0.9604	0.9412
Molar Flow	(kgmole/h)	0.1917	0.1917	0.1917	0.1917	0.1917
Mass Flow	(kg/h)	1.000	1.000	1.000	1.000	1.000
Actual Volume Flow	(m3/h)	5.410	5.410	5.513	5.513	5.619
Mass Density	(kg/m3)	0.1848	0.1848	0.1814	0.1814	0.1780
Name		304e @H2rec	304f @H2rec	304g @H2rec	304h @H2rec	305 @H2rec
Vapour Fraction		1.0000	1.0000 *	1.0000	1.0000 *	1.0000
Temperature	(C)	58.81	58.37	58.38	57.94	57.95
Pressure	(bar)	0.9412	0.9224	0.9224	0.9039	0.9039
Molar Flow	(kgmole/h)	0.1917	0.1917	0.1917	0.1917	0.1917
Mass Flow	(kg/h)	1.000	1.000	1.000	1.000	1.000
Actual Volume Flow	(m3/h)	5.619	5.726	5.726	5.835	5.836
Mass Density	(kg/m3)	0.1780	0.1746	0.1746	0.1714	0.1714
Name		306 @H2rec	307 @H2rec	308 @H2rec	309 @H2rec	310 @H2rec
Vapour Fraction		0.8822	0.8603	0.0000	0.0000	1.0000
Temperature	(C)	70.38	60.00 *	60.00	60.05	60.00
Pressure	(bar)	4.547	4.287	4.287	9.189	4.287
Molar Flow	(kgmole/h)	439.3	439.3	61.37	61.37	377.9
Mass Flow	(kg/h)	2152	2152	1106	1106	1046
Actual Volume Flow	(m3/h)	2436	2445	1.128	1.127	2444
Mass Density	(kg/m3)	0.8832	0.8801	980.5	980.7	0.4281
Name		311 @H2rec	312 @H2rec	313 @H2rec	314 @H2rec	315 @H2rec
Vapour Fraction		1.0000	1.0000 *	1.0000	0.9785	0.9609
Temperature	(C)	113.0	68.60	123.0	64.00	40.00 *
Pressure	(bar)	6.495	6.365	9.642	9.449	9.189
Molar Flow	(kgmole/h)	377.9	377.9	377.9	377.9	377.9
Mass Flow	(kg/h)	1046	1046	1046	1046	1046
Actual Volume Flow	(m3/h)	1871	1689	1294	1100	1032
Mass Density	(kg/m3)	0.5591	0.6193	0.8087	0.9511	1.013
Name		316 @H2rec	317 @H2rec	318 @H2rec	320 @H2rec	321 @H2rec
Vapour Fraction		0.0000	0.0101	0.0023	1.0000	1.0000
Temperature	(C)	40.00	35.71	54.56	40.00	90.37
Pressure	(bar)	9.189	9.189	9.189	9.189	13.92
Molar Flow	(kgmole/h)	14.77	17.95	79.32	363.2	363.2
Mass Flow	(kg/h)	266.1	320.5	1426	780.1	780.1
Actual Volume Flow	(m3/h)	0.2671	0.8285	1.990	1032	792.1
Mass Density	(kg/m3)	996.2	386.8	716.6	0.7559	0.9849

* Specified by user.

Workbook: Case (Main) (continued)

Material Streams (continued)

Fluid Pkg: All

Name		322 @H2rec	323 @H2rec	324 @H2rec	325 @H2rec	326 @H2rec
Vapour Fraction		1.0000 *	1.0000	0.9952	0.9927	0.0000
Temperature	(C)	47.25	98.78	37.79	15.00 *	15.00
Pressure	(bar)	13.64	20.66	20.25	19.95	19.95
Molar Flow	(kgmole/h)	363.2	363.2	363.2	363.2	2.665
Mass Flow	(kg/h)	780.1	780.1	780.1	780.1	48.00
Actual Volume Flow	(m3/h)	712.3	547.0	464.6	436.0	4.728e-002
Mass Density	(kg/m3)	1.095	1.426	1.679	1.789	1015
Name		330 @H2rec	331 @H2rec	332 H2 Product @H2r	101 @NPP	102 @NPP
Vapour Fraction		1.0000	0.3511	1.0000	1.0000 *	1.0000
Temperature	(C)	15.00	15.00	15.00	267.1	267.1
Pressure	(bar)	19.95	19.95	19.95	52.54 *	52.54
Molar Flow	(kgmole/h)	360.5	0.5134	360.0	3.663e+005	3.483e+005
Mass Flow	(kg/h)	732.1	6.367	725.7	6.599e+006 *	6.275e+006
Actual Volume Flow	(m3/h)	436.0	0.2239	435.4	2.469e+005	2.348e+005
Mass Density	(kg/m3)	1.679	28.44	1.667	26.72	26.72
Name		103 @NPP	103a @NPP	104 @NPP	107 @NPP	107a @NPP
Vapour Fraction		1.0000	0.9991	0.9979	1.0000	0.9989
Temperature	(C)	267.1	265.2	262.0	267.1	264.5
Pressure	(bar)	52.54	50.96	48.46	52.54	50.44
Molar Flow	(kgmole/h)	3.230e+005	3.230e+005	3.230e+005	2.534e+004	2.534e+004
Mass Flow	(kg/h)	5.818e+006	5.818e+006	5.818e+006	4.566e+005	4.566e+005
Actual Volume Flow	(m3/h)	2.177e+005	2.247e+005	2.367e+005	1.709e+004	1.782e+004
Mass Density	(kg/m3)	26.72	25.89	24.58	26.72	25.62
Name		108 (F) @NPP	110 @NPP	111 @NPP	112 @NPP	112a @NPP
Vapour Fraction		0.0236	0.9387	0.9387	0.9387	0.9391
Temperature	(C)	263.9	223.6	223.6	223.6	222.0
Pressure	(bar)	49.93	24.83	24.83	24.83	24.08
Molar Flow	(kgmole/h)	2.534e+004	3.230e+005	2.952e+005	2.778e+004	2.778e+004
Mass Flow	(kg/h)	4.566e+005	5.818e+006	5.318e+006	5.004e+005	5.004e+005
Actual Volume Flow	(m3/h)	999.1	4.401e+005	4.022e+005	3.785e+004	3.902e+004
Mass Density	(kg/m3)	457.0	13.22	13.22	13.22	12.82
Name		114 @NPP	115 @NPP	120 @NPP	121 @NPP	122 (H) @NPP
Vapour Fraction		0.5529	0.0000	0.8948	0.8948	0.8948
Temperature	(C)	222.0	186.9	183.4	183.4	183.4
Pressure	(bar)	24.08	24.08	10.84	10.84	10.84
Molar Flow	(kgmole/h)	5.312e+004	5.312e+004	2.952e+005	2.676e+005	2.759e+004
Mass Flow	(kg/h)	9.569e+005	9.569e+005	5.318e+006	4.821e+006	4.971e+005
Actual Volume Flow	(m3/h)	4.441e+004	1087	8.572e+005	7.771e+005	8.013e+004
Mass Density	(kg/m3)	21.55	880.5	6.204	6.204	6.204
Name		122a @NPP	122a-L @NPP	123 @NPP	124 @NPP	125 @NPP
Vapour Fraction		0.8961	0.0000 *	0.3148	0.0000	0.0000 *
Temperature	(C)	181.2	181.2	181.2	181.0	181.2
Pressure	(bar)	10.30	10.30	10.30	10.30	10.30
Molar Flow	(kgmole/h)	2.759e+004	2.759e+004	8.071e+004	8.071e+004	1.086e+005
Mass Flow	(kg/h)	4.971e+005	4.971e+005	1.454e+006	1.454e+006	1.956e+006
Actual Volume Flow	(m3/h)	8.428e+004	561.2	8.766e+004	1641	2208
Mass Density	(kg/m3)	5.898	885.8	16.59	886.0	885.8

* Specified by user.

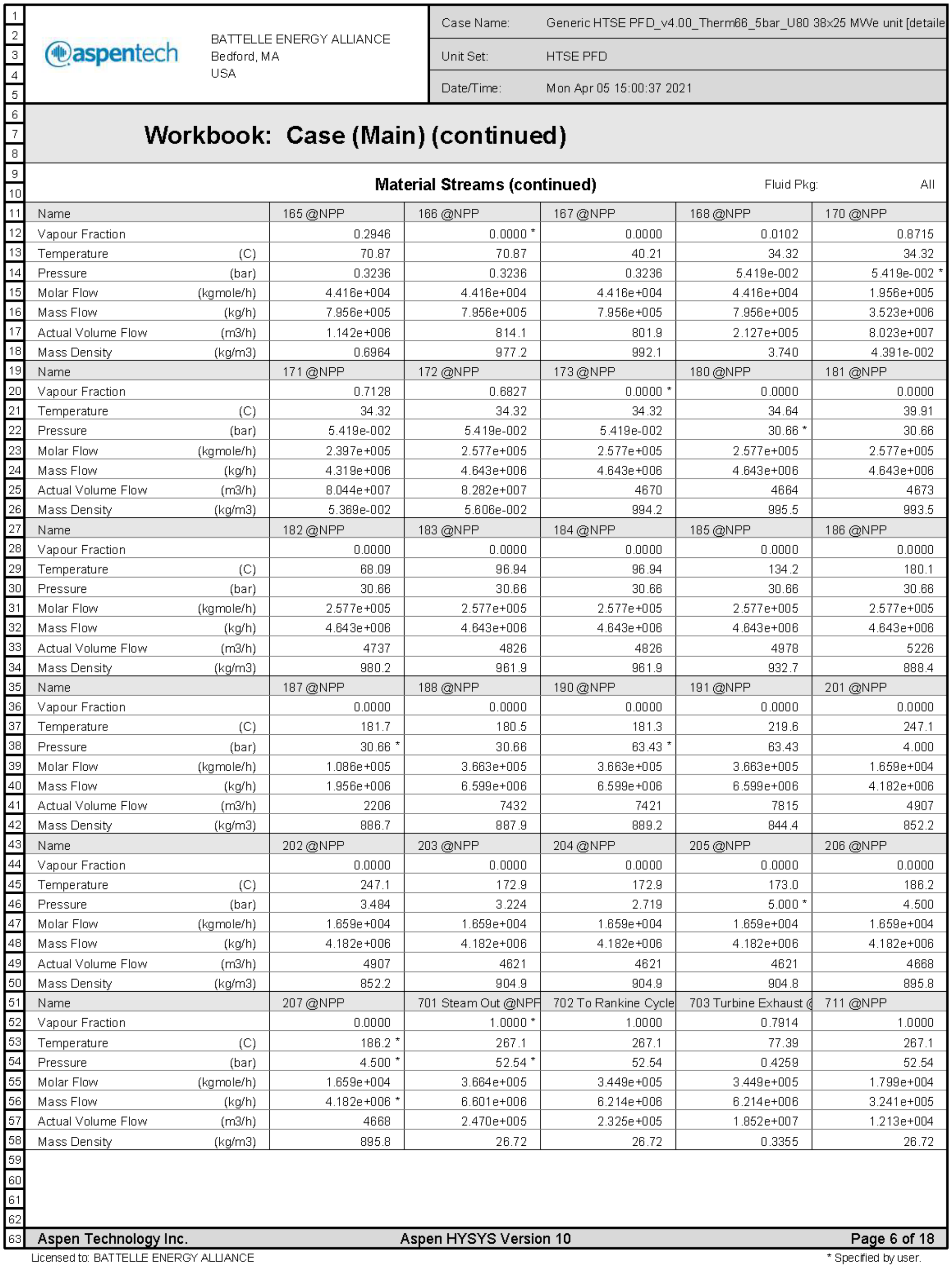
Workbook: Case (Main) (continued)

Material Streams (continued)

Fluid Pkg: All

Name		126 (E) @NPP	127 @NPP	128 @NPP	128a @NPP	130 @NPP
Vapour Fraction		0.0000	1.0000	1.0000	1.0000	1.0000
Temperature	(C)	181.6	181.6	252.9	252.5	147.2
Pressure	(bar)	10.40	10.40	9.988	9.788	3.488
Molar Flow	(kgmole/h)	2.787e+004	2.397e+005	2.397e+005	2.397e+005	2.397e+005
Mass Flow	(kg/h)	5.021e+005	4.319e+006	4.319e+006	4.319e+006	4.319e+006
Actual Volume Flow	(m3/h)	567.2	8.084e+005	1.012e+006	1.033e+006	2.326e+006
Mass Density	(kg/m3)	885.3	5.343	4.266	4.180	1.857
Name		131 @NPP	132 @NPP	132a (RP) @NPP	132a-L (RP-L) @NPP	134 @NPP
Vapour Fraction		1.0000	1.0000	1.0000	0.0000 *	0.0000
Temperature	(C)	147.2	147.2	146.5	137.0	102.5
Pressure	(bar)	3.488	3.488	3.314	3.314	3.314
Molar Flow	(kgmole/h)	2.222e+005	1.752e+004	1.752e+004	1.752e+004	1.752e+004
Mass Flow	(kg/h)	4.003e+006	3.157e+005	3.157e+005	3.157e+005	3.157e+005
Actual Volume Flow	(m3/h)	2.156e+006	1.700e+005	1.789e+005	339.8	330.0
Mass Density	(kg/m3)	1.857	1.857	1.764	928.9	956.7
Name		135 @NPP	140 @NPP	141 @NPP	142 @NPP	142a (SO) @NPP
Vapour Fraction		0.0053	0.9594	0.9594	0.9594	0.9605
Temperature	(C)	99.71	101.2	101.2	101.2	99.71
Pressure	(bar)	1.003	1.056	1.056	1.056	1.003
Molar Flow	(kgmole/h)	1.752e+004	2.222e+005	2.095e+005	1.274e+004	1.274e+004
Mass Flow	(kg/h)	3.157e+005	4.003e+006	3.773e+006	2.296e+005	2.296e+005
Actual Volume Flow	(m3/h)	3139	6.186e+006	5.831e+006	3.548e+005	3.726e+005
Mass Density	(kg/m3)	100.6	0.6471	0.6471	0.6471	0.6161
Name		142a-L (SO-L) @NPP	144 @NPP	145 @NPP	146 @NPP	150 @NPP
Vapour Fraction		0.0000 *	0.4075	0.0000	0.0050	0.9417
Temperature	(C)	99.71	99.71	73.65	70.87	87.46
Pressure	(bar)	1.003	1.003	1.003	0.3236	0.6362
Molar Flow	(kgmole/h)	1.274e+004	3.027e+004	3.027e+004	3.027e+004	2.095e+005
Mass Flow	(kg/h)	2.296e+005	5.453e+005	5.453e+005	5.453e+005	3.773e+006
Actual Volume Flow	(m3/h)	239.5	3.758e+005	558.9	1.389e+004	9.192e+006
Mass Density	(kg/m3)	958.6	1.451	975.7	39.25	0.4105
Name		151 @NPP	151 HTE feedwater (lic	152 @NPP	152 HTE feed water (v	160 @NPP
Vapour Fraction		0.9417	0.0026	0.9417	1.0000	0.9216
Temperature	(C)	87.46	152.9	87.46	155.3	72.07
Pressure	(bar)	0.6362	5.900	0.6362	5.400	0.3407
Molar Flow	(kgmole/h)	2.082e+005	1.669e+004	1270	1.669e+004	2.082e+005
Mass Flow	(kg/h)	3.751e+006	3.006e+005	2.288e+004	3.006e+005	3.751e+006
Actual Volume Flow	(m3/h)	9.136e+006	584.5	5.574e+004	1.066e+005	1.605e+007
Mass Density	(kg/m3)	0.4105	514.4	0.4105	2.821	0.2337
Name		161 @NPP	162 @NPP	163 @NPP	163a @NPP	163a-L @NPP
Vapour Fraction		0.9216	0.9216	0.9245	0.9255	0.0000 *
Temperature	(C)	72.07	72.07	72.06	70.87	70.87
Pressure	(bar)	0.3407	0.3407	0.3407	0.3236	0.3236
Molar Flow	(kgmole/h)	1.956e+005	1.262e+004	1.389e+004	1.389e+004	1.389e+004
Mass Flow	(kg/h)	3.523e+006	2.274e+005	2.503e+005	2.503e+005	2.503e+005
Actual Volume Flow	(m3/h)	1.507e+007	9.730e+005	1.074e+006	1.128e+006	256.1
Mass Density	(kg/m3)	0.2337	0.2337	0.2330	0.2218	977.2

* Specified by user.



BATTELLE ENERGY ALLIANCE
Bedford, MA
USA

Case Name: Generic HTSE PFD_v4.00_Therm66_5bar_U80 38x25 MWe unit [detaile
Unit Set: HTSE PFD
Date/Time: Mon Apr 05 15:00:37 2021

Workbook: Case (Main) (continued)

Material Streams (continued)

Fluid Pkg: All

Name		165 @NPP	166 @NPP	167 @NPP	168 @NPP	170 @NPP
Vapour Fraction		0.2946	0.0000 *	0.0000	0.0102	0.8715
Temperature	(C)	70.87	70.87	40.21	34.32	34.32
Pressure	(bar)	0.3236	0.3236	0.3236	5.419e-002	5.419e-002 *
Molar Flow	(kgmole/h)	4.416e+004	4.416e+004	4.416e+004	4.416e+004	1.956e+005
Mass Flow	(kg/h)	7.956e+005	7.956e+005	7.956e+005	7.956e+005	3.523e+006
Actual Volume Flow	(m3/h)	1.142e+006	814.1	801.9	2.127e+005	8.023e+007
Mass Density	(kg/m3)	0.6964	977.2	992.1	3.740	4.391e-002
Name		171 @NPP	172 @NPP	173 @NPP	180 @NPP	181 @NPP
Vapour Fraction		0.7128	0.6827	0.0000 *	0.0000	0.0000
Temperature	(C)	34.32	34.32	34.32	34.64	39.91
Pressure	(bar)	5.419e-002	5.419e-002	5.419e-002	30.66 *	30.66
Molar Flow	(kgmole/h)	2.397e+005	2.577e+005	2.577e+005	2.577e+005	2.577e+005
Mass Flow	(kg/h)	4.319e+006	4.643e+006	4.643e+006	4.643e+006	4.643e+006
Actual Volume Flow	(m3/h)	8.044e+007	8.282e+007	4670	4664	4673
Mass Density	(kg/m3)	5.369e-002	5.606e-002	994.2	995.5	993.5
Name		182 @NPP	183 @NPP	184 @NPP	185 @NPP	186 @NPP
Vapour Fraction		0.0000	0.0000	0.0000	0.0000	0.0000
Temperature	(C)	68.09	96.94	96.94	134.2	180.1
Pressure	(bar)	30.66	30.66	30.66	30.66	30.66
Molar Flow	(kgmole/h)	2.577e+005	2.577e+005	2.577e+005	2.577e+005	2.577e+005
Mass Flow	(kg/h)	4.643e+006	4.643e+006	4.643e+006	4.643e+006	4.643e+006
Actual Volume Flow	(m3/h)	4737	4826	4826	4978	5226
Mass Density	(kg/m3)	980.2	961.9	961.9	932.7	888.4
Name		187 @NPP	188 @NPP	190 @NPP	191 @NPP	201 @NPP
Vapour Fraction		0.0000	0.0000	0.0000	0.0000	0.0000
Temperature	(C)	181.7	180.5	181.3	219.6	247.1
Pressure	(bar)	30.66 *	30.66	63.43 *	63.43	4.000
Molar Flow	(kgmole/h)	1.086e+005	3.663e+005	3.663e+005	3.663e+005	1.659e+004
Mass Flow	(kg/h)	1.956e+006	6.599e+006	6.599e+006	6.599e+006	4.182e+006
Actual Volume Flow	(m3/h)	2206	7432	7421	7815	4907
Mass Density	(kg/m3)	886.7	887.9	889.2	844.4	852.2
Name		202 @NPP	203 @NPP	204 @NPP	205 @NPP	206 @NPP
Vapour Fraction		0.0000	0.0000	0.0000	0.0000	0.0000
Temperature	(C)	247.1	172.9	172.9	173.0	186.2
Pressure	(bar)	3.484	3.224	2.719	5.000 *	4.500
Molar Flow	(kgmole/h)	1.659e+004	1.659e+004	1.659e+004	1.659e+004	1.659e+004
Mass Flow	(kg/h)	4.182e+006	4.182e+006	4.182e+006	4.182e+006	4.182e+006
Actual Volume Flow	(m3/h)	4907	4621	4621	4621	4668
Mass Density	(kg/m3)	852.2	904.9	904.9	904.8	895.8
Name		207 @NPP	701 Steam Out @NPF	702 To Rankine Cycle	703 Turbine Exhaust @	711 @NPP
Vapour Fraction		0.0000	1.0000 *	1.0000	0.7914	1.0000
Temperature	(C)	186.2 *	267.1	267.1	77.39	267.1
Pressure	(bar)	4.500 *	52.54 *	52.54	0.4259	52.54
Molar Flow	(kgmole/h)	1.659e+004	3.664e+005	3.449e+005	3.449e+005	1.799e+004
Mass Flow	(kg/h)	4.182e+006 *	6.601e+006	6.214e+006	6.214e+006	3.241e+005
Actual Volume Flow	(m3/h)	4668	2.470e+005	2.325e+005	1.852e+007	1.213e+004
Mass Density	(kg/m3)	895.8	26.72	26.72	0.3355	26.72

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* Specified by user.

Workbook: Case (Main) (continued)

Material Streams (continued)

Fluid Pkg: All

Name		711 SEL Inlet @NPP	712 @NPP	713 @NPP	713 SEL Condensate (	721 NPP Condensate
Vapour Fraction		1.0000	0.0000	0.0000	0.0000	0.0000 *
Temperature	(C)	267.1	264.3	193.0	193.0 *	76.90
Pressure	(bar)	52.54	51.91	51.29	52.04	0.4174
Molar Flow	(kgmole/h)	2.148e+004	1.799e+004	1.799e+004	2.148e+004	3.449e+005
Mass Flow	(kg/h)	3.869e+005	3.241e+005	3.241e+005	3.869e+005	6.214e+006
Actual Volume Flow	(m3/h)	1.448e+004	417.0	370.2	441.9	6382
Mass Density	(kg/m3)	26.72	777.2	875.5	875.6	973.7

Name		722 @NPP	723 @NPP	724 @NPP	725 @NPP	726 Heated Feedwate
Vapour Fraction		0.0000	0.0000	0.0000	0.0000	0.0000
Temperature	(C)	77.31	84.26	84.74	222.2 *	222.2 *
Pressure	(bar)	30.66 *	30.66	64.73	63.43 *	63.43 *
Molar Flow	(kgmole/h)	3.449e+005	3.664e+005	3.664e+005	3.664e+005	3.664e+005
Mass Flow	(kg/h)	6.214e+006	6.601e+006	6.601e+006	6.601e+006	6.601e+006 *
Actual Volume Flow	(m3/h)	6374	6802	6793	7848	7848
Mass Density	(kg/m3)	974.8	970.5	971.7	841.1	841.1

Compositions

Fluid Pkg: All

Name	101 Process Water Inl	102	132 process feed wate	151	152 process feed wate
Comp Mole Frac (H2O)	1.0000 *	1.0000	0.9996	0.9996	0.9996
Comp Mole Frac (Hydrogen)	0.0000 *	0.0000	0.0004	0.0004	0.0004
Comp Mole Frac (Oxygen)	0.0000 *	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.0000 *	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO2)	0.0000 *	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	0.0000 *	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***

Name	162	163	164	165	166
Comp Mole Frac (H2O)	0.9996	0.9007	0.9007	0.9007 *	0.9007
Comp Mole Frac (Hydrogen)	0.0004	0.0993	0.0993	0.0993 *	0.0993
Comp Mole Frac (Oxygen)	0.0000	0.0000	0.0000	0.0000 *	0.0000
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000 *	0.0000
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000 *	0.0000
Comp Mole Frac (CO)	0.0000	0.0000	0.0000	0.0000 *	0.0000
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***

Name	167 Process Cell Inlet	171 Process Cell Outle	172 H2/H2O product r	173 H2/H2O recycle	202 Process Heat In
Comp Mole Frac (H2O)	0.9007	0.1801	0.1801	0.1801	***
Comp Mole Frac (Hydrogen)	0.0993	0.8199	0.8199	0.8199	***
Comp Mole Frac (Oxygen)	0.0000	0.0000	0.0000	0.0000	***
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000	***
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	***
Comp Mole Frac (CO)	0.0000	0.0000	0.0000	0.0000	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	1.0000 *

* Specified by user.

Workbook: Case (Main) (continued)

Compositions (continued)

Fluid Pkg: All

Name	203 Process Heat Ret	301	302 H2/H2O for purific	332 H2 Product	401
Comp Mole Frac (H2O)	***	0.1801	0.1801	0.0000	0.0000
Comp Mole Frac (Hydrogen)	***	0.8199	0.8199	1.0000	1.0000
Comp Mole Frac (Oxygen)	***	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Nitrogen)	***	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO2)	***	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	***	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	1.0000	***	***	***	***
Name	402a	402b	402c	402d	402e
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Hydrogen)	1.0000	1.0000	1.0000	1.0000	1.0000
Comp Mole Frac (Oxygen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Name	403 Pressurized H2 P	501 Sweep Gas Inlet	502	503	504
Comp Mole Frac (H2O)	0.0000	0.0000 *	0.0000	0.0000	0.0000
Comp Mole Frac (Hydrogen)	1.0000	0.0000 *	0.0000	0.0000	0.0000
Comp Mole Frac (Oxygen)	0.0000	0.2100 *	0.2100	0.2100	0.2100
Comp Mole Frac (Nitrogen)	0.0000	0.7900 *	0.7900	0.7900	0.7900
Comp Mole Frac (CO2)	0.0000	0.0000 *	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	0.0000	0.0000 *	0.0000	0.0000	0.0000
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Name	505	506	507	508 Sweep Cell Inlet	511 Sweep Gas/O2 O
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000	0.0000 *	0.0000
Comp Mole Frac (Hydrogen)	0.0000	0.0000	0.0000	0.0000 *	0.0000
Comp Mole Frac (Oxygen)	0.2500	0.2500	0.2500	0.2500 *	0.4001
Comp Mole Frac (Nitrogen)	0.7500	0.7500	0.7500	0.7500 *	0.5999
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000 *	0.0000
Comp Mole Frac (CO)	0.0000	0.0000	0.0000	0.0000 *	0.0000
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Name	512	513	514	515 Sweep Gas Recy	516
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000 *	0.0000	0.0000
Comp Mole Frac (Hydrogen)	0.0000	0.0000	0.0000 *	0.0000	0.0000
Comp Mole Frac (Oxygen)	0.4001	0.4001	0.4001 *	0.4001	0.4001
Comp Mole Frac (Nitrogen)	0.5999	0.5999	0.5999 *	0.5999	0.5999
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000 *	0.0000	0.0000
Comp Mole Frac (CO)	0.0000	0.0000	0.0000 *	0.0000	0.0000
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***

* Specified by user.

Workbook: Case (Main) (continued)

Compositions (continued)

Fluid Pkg: All

Name	517 Sweep Gas Exha	801 feed water	802	803	901 cooling water
Comp Mole Frac (H2O)	0.0000	1.0000 *	1.0000	1.0000	1.0000 *
Comp Mole Frac (Hydrogen)	0.0000	0.0000 *	0.0000	0.0000	0.0000 *
Comp Mole Frac (Oxygen)	0.4001	0.0000 *	0.0000	0.0000	0.0000 *
Comp Mole Frac (Nitrogen)	0.5999	0.0000 *	0.0000	0.0000	0.0000 *
Comp Mole Frac (CO2)	0.0000	0.0000 *	0.0000	0.0000	0.0000 *
Comp Mole Frac (CO)	0.0000	0.0000 *	0.0000	0.0000	0.0000 *
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***

Name	902	903	904	905	Anode @Cell
Comp Mole Frac (H2O)	1.0000	1.0000	1.0000	1.0000	0.0000
Comp Mole Frac (Hydrogen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Oxygen)	0.0000	0.0000	0.0000	0.0000	1.0000
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***

Name	Cathode @Cell	Gas Products @Cell	Liquid Products @Cell	Molar Flow of Oxygen	Process Cell Inlet @C
Comp Mole Frac (H2O)	0.1801	0.1324	0.1324	0.0000 *	0.9007
Comp Mole Frac (Hydrogen)	0.8199	0.6027	0.6027	0.0000 *	0.0993
Comp Mole Frac (Oxygen)	0.0000	0.2648	0.2648	1.0000 *	0.0000
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000 *	0.0000
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000 *	0.0000
Comp Mole Frac (CO)	0.0000	0.0000	0.0000	0.0000 *	0.0000
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***

Name	Sweep Cell Inlet @Ce	Sweep Gas/O2 Out @	1 @H2rec	102 @H2rec	113 @H2rec
Comp Mole Frac (H2O)	0.0000	0.0000	0.2000 *	1.0000	1.0000
Comp Mole Frac (Hydrogen)	0.0000	0.0000	0.8000 *	0.0000	0.0000
Comp Mole Frac (Oxygen)	0.2500	0.4001	0.0000 *	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.7500	0.5999	0.0000 *	0.0000	0.0000
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000 *	0.0000	0.0000
Comp Mole Frac (CO)	0.0000	0.0000	0.0000 *	0.0000	0.0000
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***

Name	114 @H2rec	122 @H2rec	131 @H2rec	132 @H2rec	302 @H2rec
Comp Mole Frac (H2O)	1.0000	1.0000	0.9996	0.9996	0.1801
Comp Mole Frac (Hydrogen)	0.0000	0.0000	0.0004	0.0004	0.8199
Comp Mole Frac (Oxygen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***

* Specified by user.

Workbook: Case (Main) (continued)

Compositions (continued)

Fluid Pkg: All

Name	303 @H2rec	304a @H2rec	304b @H2rec	304c @H2rec	304d @H2rec
Comp Mole Frac (H2O)	0.2000	0.2000	0.2000	0.2000	0.2000
Comp Mole Frac (Hydrogen)	0.8000	0.8000	0.8000	0.8000	0.8000
Comp Mole Frac (Oxygen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***

Name	304e @H2rec	304f @H2rec	304g @H2rec	304h @H2rec	305 @H2rec
Comp Mole Frac (H2O)	0.2000	0.2000	0.2000	0.2000	0.2000
Comp Mole Frac (Hydrogen)	0.8000	0.8000	0.8000	0.8000	0.8000
Comp Mole Frac (Oxygen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***

Name	306 @H2rec	307 @H2rec	308 @H2rec	309 @H2rec	310 @H2rec
Comp Mole Frac (H2O)	0.1801	0.1801	1.0000	1.0000	0.0470
Comp Mole Frac (Hydrogen)	0.8199	0.8199	0.0000	0.0000	0.9530
Comp Mole Frac (Oxygen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***

Name	311 @H2rec	312 @H2rec	313 @H2rec	314 @H2rec	315 @H2rec
Comp Mole Frac (H2O)	0.0470	0.0470	0.0470	0.0470	0.0470
Comp Mole Frac (Hydrogen)	0.9530	0.9530	0.9530	0.9530	0.9530
Comp Mole Frac (Oxygen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***

Name	316 @H2rec	317 @H2rec	318 @H2rec	320 @H2rec	321 @H2rec
Comp Mole Frac (H2O)	1.0000	0.9900	0.9977	0.0083	0.0083
Comp Mole Frac (Hydrogen)	0.0000	0.0100	0.0023	0.9917	0.9917
Comp Mole Frac (Oxygen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***

Workbook: Case (Main) (continued)

Compositions (continued)

Fluid Pkg: All

Name	322 @H2rec	323 @H2rec	324 @H2rec	325 @H2rec	326 @H2rec
Comp Mole Frac (H2O)	0.0083	0.0083	0.0083	0.0083	1.0000
Comp Mole Frac (Hydrogen)	0.9917	0.9917	0.9917	0.9917	0.0000
Comp Mole Frac (Oxygen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Name	330 @H2rec	331 @H2rec	332 H2 Product @H2r	101 @NPP	102 @NPP
Comp Mole Frac (H2O)	0.0009	0.6492	0.0000	1.0000 *	1.0000
Comp Mole Frac (Hydrogen)	0.9991	0.3508	1.0000	***	***
Comp Mole Frac (Oxygen)	0.0000	0.0000	0.0000	***	***
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	***	***
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	***	***
Comp Mole Frac (CO)	0.0000	0.0000	0.0000	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Name	103 @NPP	103a @NPP	104 @NPP	107 @NPP	107a @NPP
Comp Mole Frac (H2O)	1.0000	1.0000	1.0000	1.0000	1.0000
Comp Mole Frac (Hydrogen)	***	***	***	***	***
Comp Mole Frac (Oxygen)	***	***	***	***	***
Comp Mole Frac (Nitrogen)	***	***	***	***	***
Comp Mole Frac (CO2)	***	***	***	***	***
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Name	108 (F) @NPP	110 @NPP	111 @NPP	112 @NPP	112a @NPP
Comp Mole Frac (H2O)	1.0000	1.0000	1.0000	1.0000	1.0000
Comp Mole Frac (Hydrogen)	***	***	***	***	***
Comp Mole Frac (Oxygen)	***	***	***	***	***
Comp Mole Frac (Nitrogen)	***	***	***	***	***
Comp Mole Frac (CO2)	***	***	***	***	***
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Name	114 @NPP	115 @NPP	120 @NPP	121 @NPP	122 (H) @NPP
Comp Mole Frac (H2O)	1.0000	1.0000	1.0000	1.0000	1.0000
Comp Mole Frac (Hydrogen)	***	***	***	***	***
Comp Mole Frac (Oxygen)	***	***	***	***	***
Comp Mole Frac (Nitrogen)	***	***	***	***	***
Comp Mole Frac (CO2)	***	***	***	***	***
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***

* Specified by user.

Workbook: Case (Main) (continued)

Compositions (continued)

Fluid Pkg: All

Name	122a @NPP	122a-L @NPP	123 @NPP	124 @NPP	125 @NPP
Comp Mole Frac (H2O)	1.0000	1.0000	1.0000	1.0000	1.0000
Comp Mole Frac (Hydrogen)	***	***	***	***	***
Comp Mole Frac (Oxygen)	***	***	***	***	***
Comp Mole Frac (Nitrogen)	***	***	***	***	***
Comp Mole Frac (CO2)	***	***	***	***	***
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***

Name	126 (E) @NPP	127 @NPP	128 @NPP	128a @NPP	130 @NPP
Comp Mole Frac (H2O)	1.0000	1.0000	1.0000	1.0000	1.0000
Comp Mole Frac (Hydrogen)	***	***	***	***	***
Comp Mole Frac (Oxygen)	***	***	***	***	***
Comp Mole Frac (Nitrogen)	***	***	***	***	***
Comp Mole Frac (CO2)	***	***	***	***	***
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***

Name	131 @NPP	132 @NPP	132a (RP) @NPP	132a-L (RP-L) @NPP	134 @NPP
Comp Mole Frac (H2O)	1.0000	1.0000	1.0000	1.0000	1.0000
Comp Mole Frac (Hydrogen)	***	***	***	***	***
Comp Mole Frac (Oxygen)	***	***	***	***	***
Comp Mole Frac (Nitrogen)	***	***	***	***	***
Comp Mole Frac (CO2)	***	***	***	***	***
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***

Name	135 @NPP	140 @NPP	141 @NPP	142 @NPP	142a (SO) @NPP
Comp Mole Frac (H2O)	1.0000	1.0000	1.0000	1.0000	1.0000
Comp Mole Frac (Hydrogen)	***	***	***	***	***
Comp Mole Frac (Oxygen)	***	***	***	***	***
Comp Mole Frac (Nitrogen)	***	***	***	***	***
Comp Mole Frac (CO2)	***	***	***	***	***
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***

Name	142a-L (SO-L) @NPP	144 @NPP	145 @NPP	146 @NPP	150 @NPP
Comp Mole Frac (H2O)	1.0000	1.0000	1.0000	1.0000	1.0000
Comp Mole Frac (Hydrogen)	***	***	***	***	***
Comp Mole Frac (Oxygen)	***	***	***	***	***
Comp Mole Frac (Nitrogen)	***	***	***	***	***
Comp Mole Frac (CO2)	***	***	***	***	***
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***

Workbook: Case (Main) (continued)

Compositions (continued)

Fluid Pkg: All

Name	151 @NPP	151 HTE feedwater (li	152 @NPP	152 HTE feed water (v	160 @NPP
Comp Mole Frac (H2O)	1.0000	0.9997	1.0000	0.9997	1.0000
Comp Mole Frac (Hydrogen)	***	0.0003	***	0.0003	***
Comp Mole Frac (Oxygen)	***	0.0000	***	0.0000	***
Comp Mole Frac (Nitrogen)	***	0.0000	***	0.0000	***
Comp Mole Frac (CO2)	***	0.0000	***	0.0000	***
Comp Mole Frac (CO)	***	0.0000	***	0.0000	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***

Name	161 @NPP	162 @NPP	163 @NPP	163a @NPP	163a-L @NPP
Comp Mole Frac (H2O)	1.0000	1.0000	1.0000	1.0000	1.0000
Comp Mole Frac (Hydrogen)	***	***	***	***	***
Comp Mole Frac (Oxygen)	***	***	***	***	***
Comp Mole Frac (Nitrogen)	***	***	***	***	***
Comp Mole Frac (CO2)	***	***	***	***	***
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***

Name	165 @NPP	166 @NPP	167 @NPP	168 @NPP	170 @NPP
Comp Mole Frac (H2O)	1.0000	1.0000	1.0000	1.0000	1.0000
Comp Mole Frac (Hydrogen)	***	***	***	***	***
Comp Mole Frac (Oxygen)	***	***	***	***	***
Comp Mole Frac (Nitrogen)	***	***	***	***	***
Comp Mole Frac (CO2)	***	***	***	***	***
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***

Name	171 @NPP	172 @NPP	173 @NPP	180 @NPP	181 @NPP
Comp Mole Frac (H2O)	1.0000	1.0000	1.0000	1.0000	1.0000
Comp Mole Frac (Hydrogen)	***	***	***	***	***
Comp Mole Frac (Oxygen)	***	***	***	***	***
Comp Mole Frac (Nitrogen)	***	***	***	***	***
Comp Mole Frac (CO2)	***	***	***	***	***
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***

Name	182 @NPP	183 @NPP	184 @NPP	185 @NPP	186 @NPP
Comp Mole Frac (H2O)	1.0000	1.0000	1.0000	1.0000	1.0000
Comp Mole Frac (Hydrogen)	***	***	***	***	***
Comp Mole Frac (Oxygen)	***	***	***	***	***
Comp Mole Frac (Nitrogen)	***	***	***	***	***
Comp Mole Frac (CO2)	***	***	***	***	***
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***

* Specified by user.

Workbook: Case (Main) (continued)

Compositions (continued)

Fluid Pkg: All

Name	187 @NPP	188 @NPP	190 @NPP	191 @NPP	201 @NPP
Comp Mole Frac (H2O)	1.0000	1.0000	1.0000	1.0000	***
Comp Mole Frac (Hydrogen)	***	***	***	***	***
Comp Mole Frac (Oxygen)	***	***	***	***	***
Comp Mole Frac (Nitrogen)	***	***	***	***	***
Comp Mole Frac (CO2)	***	***	***	***	***
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	1.0000

Name	202 @NPP	203 @NPP	204 @NPP	205 @NPP	206 @NPP
Comp Mole Frac (H2O)	***	***	***	***	***
Comp Mole Frac (Hydrogen)	***	***	***	***	***
Comp Mole Frac (Oxygen)	***	***	***	***	***
Comp Mole Frac (Nitrogen)	***	***	***	***	***
Comp Mole Frac (CO2)	***	***	***	***	***
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	1.0000	1.0000	1.0000	1.0000	1.0000

Name	207 @NPP	701 Steam Out @NPP	702 To Rankine Cycle	703 Turbine Exhaust @	711 @NPP
Comp Mole Frac (H2O)	***	1.0000	1.0000	1.0000	1.0000
Comp Mole Frac (Hydrogen)	***	***	***	***	***
Comp Mole Frac (Oxygen)	***	***	***	***	***
Comp Mole Frac (Nitrogen)	***	***	***	***	***
Comp Mole Frac (CO2)	***	***	***	***	***
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	1.0000 *	***	***	***	***

Name	711 SEL Inlet @NPP	712 @NPP	713 @NPP	713 SEL Condensate (	721 NPP Condensate
Comp Mole Frac (H2O)	1.0000	1.0000	1.0000	1.0000	1.0000
Comp Mole Frac (Hydrogen)	***	***	***	***	***
Comp Mole Frac (Oxygen)	***	***	***	***	***
Comp Mole Frac (Nitrogen)	***	***	***	***	***
Comp Mole Frac (CO2)	***	***	***	***	***
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***

Name	722 @NPP	723 @NPP	724 @NPP	725 @NPP	726 Heated Feedwate
Comp Mole Frac (H2O)	1.0000	1.0000	1.0000	1.0000	1.0000 *
Comp Mole Frac (Hydrogen)	***	***	***	***	***
Comp Mole Frac (Oxygen)	***	***	***	***	***
Comp Mole Frac (Nitrogen)	***	***	***	***	***
Comp Mole Frac (CO2)	***	***	***	***	***
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***

Energy Streams

Fluid Pkg: All

Name		Electrolysis Power	Inverter Power	Process Heat	Q-CW	Q-HX-107
Heat Flow	(MW)	-24.80	-25.30	2.755e-002	80.50	0.5193
Mass Flow	(kg/h)	---	---	---	---	---

* Specified by user.

Workbook: Case (Main) (continued)

Energy Streams (continued)

Fluid Pkg: All

Name		Q-HX-502	Q-IC-501	Q-IC-401_stg1	Q-IC-401_stg2	Q-IC-401_stg3
Heat Flow	(MW)	0.1003	0.2769	1.502	5.966	5.997
Mass Flow	(kg/h)	---	1.702e+004	9.231e+004	3.668e+005	3.687e+005
Name		Q-PIPE-801	Q-PIPE-901	Q-PIPE-902	W-K-101	W-K-401
Heat Flow	(MW)	1.639e-006	7.534e-003	8.385e-003	5.159e-002	17.30
Mass Flow	(kg/h)	---	---	---	---	---
Name		W-K-502	W-K-401_stg1	W-K-401_stg2	W-K-401_stg3	W-K-501_NET
Heat Flow	(MW)	0.1071	5.308	5.977	6.018	-9.231e-003
Mass Flow	(kg/h)	---	---	---	---	---
Name		W-K-501_stgC1	W-K-501_stgC2	W-K-501_stgT1	W-P-101	W-P-801
Heat Flow	(MW)	0.4142	0.4563	0.8798	1.463e-003	5.261e-003
Mass Flow	(kg/h)	---	---	---	---	---
Name		W-P-901	Electrode Heat @Cell	Electrolysis Heating @	Electrolysis Power @C	Process Heat @Cell
Heat Flow	(MW)	0.1314	-1.626e-005	24.82	-24.80	2.755e-002
Mass Flow	(kg/h)	---	---	---	---	---
Name		Q-HX-301 @H2rec	Q-HX-302 @H2rec	Q-HX-305 @H2rec	Q-HX-308 @H2rec	Q-IC-302 @H2rec
Heat Flow	(MW)	9.542e-005	0.6934	0.2747	0.1975	0.1348
Mass Flow	(kg/h)	5.866	4.262e+004	1.689e+004	1.214e+004	8286
Name		Q-IC-303 @H2rec	Q-IC-301_stg1 @H2re	Q-IC-301_stg2 @H2re	Q-IC-301_stg3 @H2re	Q-IC-301_stg4 @H2re
Heat Flow	(MW)	0.1246	6.886e-007	6.846e-007	6.882e-007	6.864e-007
Mass Flow	(kg/h)	7662	4.233e-002	4.208e-002	4.230e-002	4.219e-002
Name		Q-X-300 @H2rec	W-K-301 @H2rec	W-K-302 @H2rec	W-K-303 @H2rec	W-K-301_stg1 @H2re
Heat Flow	(MW)	-4.093e-003	9.210e-008	0.3254	0.2946	2.003e-008
Mass Flow	(kg/h)	---	---	---	---	---
Name		W-K-301_stg2 @H2re	W-K-301_stg3 @H2re	W-K-301_stg4 @H2re	W-K-301_stg5 @H2re	W-K-302_stg1 @H2re
Heat Flow	(MW)	1.136e-008	2.017e-008	2.024e-008	2.031e-008	0.1606
Mass Flow	(kg/h)	---	---	---	---	---
Name		W-K-302_stg2 @H2re	W-K-303_stg1 @H2re	W-K-303_stg2 @H2re	W-P-301 @H2rec	Excess Electricity @N
Heat Flow	(MW)	0.1648	0.1455	0.1491	2.047e-004	49.19
Mass Flow	(kg/h)	---	---	---	---	---
Name		Q-100 @NPP	Q-101 @NPP	Q-102 @NPP	Q-103 @NPP	Q-104 @NPP
Heat Flow	(MW)	165.3	176.1	1.666	9.021	5.294
Mass Flow	(kg/h)	---	---	---	---	---
Name		Q-105 @NPP	Q-106 @NPP	Q-107 @NPP	Q-108 @NPP	Q-109 @NPP
Heat Flow	(MW)	2130	236.9	183.1	67.27	77.88
Mass Flow	(kg/h)	---	---	---	---	---
Name		Q-110 @NPP	Q-111 @NPP	Q-COND-100 @NPP	Q-FWH-700 @NPP	Q-PIPE-201 @NPP
Heat Flow	(MW)	190.7	3389	3165	1090	7.491e-002
Mass Flow	(kg/h)	---	---	---	---	---
Name		Q-PIPE-202 @NPP	Q-SG-100 Thermal En	Q-TDL @NPP	Rankine Energy @NP	W-K-700 Electricity G
Heat Flow	(MW)	5.148e-002	3368	211.6 *	3171	1097
Mass Flow	(kg/h)	---	---	---	---	---
Name		W-P-201 Circ Pump P	W-P-701 Condensate	W-P-702 Feedwater P		
Heat Flow	(MW)	0.3903	7.148	8.581		
Mass Flow	(kg/h)	---	---	---		

Unit Ops

Operation Name	Operation Type	Feeds	Products	Ignored	Calc Level
TEE-172	Tee	172 H2/H2O product mix	173 H2/H2O recycle 301	No	500.0 *
TEE-514	Tee	514	515 Sweep Gas Recycle 516	No	500.0 *

* Specified by user.

BATTELLE ENERGY ALLIANCE
Bedford, MA
USA

Case Name: Generic HTSE PFD_v4.00_Therm66_5bar_U80 38x25 MWe unit [detaile
Unit Set: HTSE PFD
Date/Time: Mon Apr 05 15:00:37 2021

Workbook: Case (Main) (continued)

Unit Ops (continued)

Operation Name	Operation Type	Feeds	Products	Ignored	Calc Level
HX-106 H2/H2O Recuperator	Heat Exchanger	171 Process Cell Outlet 165	172 H2/H2O product mix 166	No	500.0 *
HX-103 Sweep Gas Low T Re	Heat Exchanger	512 152 process feed water (vapo	513 162	No	500.0 *
HX-501 Sweep Gas High T R	Heat Exchanger	511 Sweep Gas/O2 Out 506	512 507	No	500.0 *
HX-102 Steam Generator	Heat Exchanger	202 Process Heat In 151	203 Process Heat Return 152 process feed water (vapo	No	500.0 *
HX-101	Heat Exchanger	301 132 process feed water (liquid	302 H2/H2O for purification 151	No	500.0 *
MIX-163	Mixer	173 H2/H2O recycle 162	163	No	500.0 *
MIX-505	Mixer	504 515 Sweep Gas Recycle	505	No	500.0 *
K-502 Sweep Gas Blower	Compressor	505 W-K-502	506	No	500.0 *
K-101 H2/H2O Recirculator	Compressor	163 W-K-101	164	No	500.0 *
K-401_stg2	Compressor	402b W-K-401_stg2	402c	No	500.0 *
K-401_stg3	Compressor	402d W-K-401_stg3	402e	No	500.0 *
K-401_stg1	Compressor	401 W-K-401_stg1	402a	No	500.0 *
K-501_stgC1	Compressor	501 Sweep Gas Inlet W-K-501_stgC1	502	No	500.0 *
K-501_stgC2	Compressor	503 W-K-501_stgC2	504	No	500.0 *
HX-107 Process Topping Hea	Heater	166 Q-HX-107	167 Process Cell Inlet	No	500.0 *
HX-502 Sweep Topping Heat	Heater	507 Q-HX-502	508 Sweep Cell Inlet	No	500.0 *
CW deltaT	Heater	903 Q-CW	904	No	500.0 *
SOEC Stack	Standard Sub-Flowsheet	167 Process Cell Inlet 508 Sweep Cell Inlet	171 Process Cell Outlet 511 Sweep Gas/O2 Out Electrolysis Power Process Heat	No	2500 *
H2 Recovery and Feed Condi	Standard Sub-Flowsheet	302 H2/H2O for purification 102	132 process feed water (liquid 332 H2 Product	No	2500 *
Nuclear Power Plant	Standard Sub-Flowsheet			No	3600 *
ADJ process feed H2 comp	Adjust			No	3500 *
ADJ CW Pump dP	Adjust			No	3500 *
ADJ FW pump dP	Adjust			No	3500 *
ADJ process cell inlet P	Adjust			No	3500 *
ADJ K-401 outlet P	Adjust			No	3500 *
ADJ Steam Generator inlet P	Adjust			No	3500 *
ADJ Sweep Gas Blower outlet	Adjust			Yes	3500 *
ADJ K-501 P ratio	Adjust			No	3500 *
ADJ sweep gas O2 comp	Adjust			No	3500 *
RCY-100	Recycle	164	165	No	3700 *

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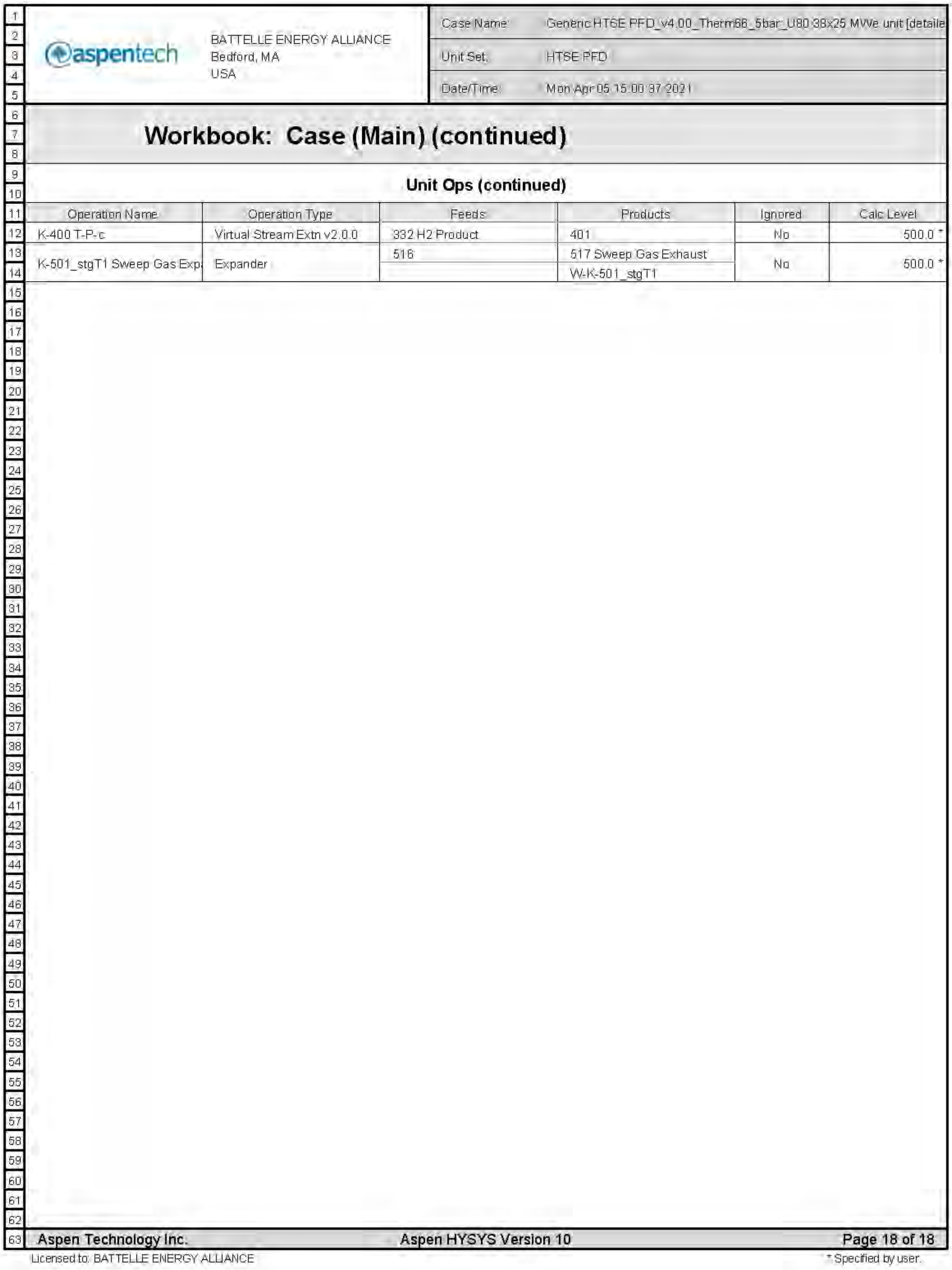
* Specified by user.

Workbook: Case (Main) (continued)

Unit Ops (continued)

Operation Name	Operation Type	Feeds	Products	Ignored	Calc Level
RCY-500	Recycle	513	514	No	3800 *
HTE Calculations	Spreadsheet			No	500.0 *
Cooling Water	Spreadsheet			No	500.0 *
K-401 Power	Spreadsheet			No	500.0 *
TDL	Spreadsheet			No	500.0 *
Water Bal	Spreadsheet			No	500.0 *
Inverter	Spreadsheet			No	500.0 *
HX dP	Spreadsheet			No	500.0 *
K-501 Power	Spreadsheet			No	500.0 *
PIPE-801	Pipe Segment	802	803 Q-PIPE-801	No	500.0 *
PIPE-901	Pipe Segment	902	903 Q-PIPE-801	No	500.0 *
PIPE-902	Pipe Segment	904	905 Q-PIPE-902	No	500.0 *
P-101	Pump	101 Process Water Inlet W-P-101	102	No	500.0 *
P-801	Pump	801 feed water W-P-801	802	No	500.0 *
P-901	Pump	901 cooling water W-P-901	902	No	500.0 *
SET RCY P	Set			No	500.0 *
SET number of HTSE blocks	Set			No	500.0 *
Inverter Efficiency	Set			No	500.0 *
SET dP IC-401_stg1	Set			No	500.0 *
SET dP IC-401_stg2	Set			No	500.0 *
SET dP IC-401_stg3	Set			No	500.0 *
SET sweep gas nstoichs	Set			Yes	500.0 *
SET dP HX-103 cold side	Set			No	500.0 *
SET dP HX-103 hot side	Set			No	500.0 *
SET dP HX-106 cold side	Set			No	500.0 *
SET dP HX-106 hot side	Set			No	500.0 *
SET dP HX-107	Set			No	500.0 *
SET dP HX-501 cold side	Set			No	500.0 *
SET dP HX-501 hot side	Set			No	500.0 *
SET dP HX-502	Set			No	500.0 *
SET dP CW delta T	Set			No	500.0 *
SET dP HX-101 cold side	Set			No	500.0 *
SET dP HX-101 hot side	Set			No	500.0 *
SET dP HX-102 cold side	Set			No	500.0 *
SET dP HX-102 hot side	Set			No	500.0 *
SET K-501 P ratio	Set			No	500.0 *
SET dP IC-501	Set			No	500.0 *
IC-401_stg2	Cooler	402c	402d Q-IC-401_stg2	No	500.0 *
IC-401_stg1	Cooler	402a	402b Q-IC-401_stg1	No	500.0 *
IC-401_stg3	Cooler	402e	403 Pressurized H2 Product Q-IC-401_stg3	No	500.0 *
IC-501	Cooler	502	503 Q-IC-501	No	500.0 *

* Specified by user.



BATTELLE ENERGY ALLIANCE
Bedford, MA
USA

Case Name:	Generic HTSE PFD_v4.00_Therm66_5bar_U80 38x25 MWe unit [detaile
Unit Set:	HTSE PFD
Date/Time:	Mon Apr 05 15:00:37 2021

Workbook: Case (Main) (continued)

Unit Ops (continued)

Operation Name	Operation Type	Feeds	Products	Ignored	Calc Level
K-400 T-P-c	Virtual Stream Extn v2.0.0	332 H2 Product	401	No	500.0 *
K-501_stgT1 Sweep Gas Exp:	Expander	516	517 Sweep Gas Exhaust W-K-501_stgT1	No	500.0 *

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Comparison of HTSE and SMR LCOH

The SMR LCOH is plotted for each of the natural gas price cases with zero, $25/tonne, and $100/tonne costs of CO_2 in Figure F-8, Figure F-9, and Figure F-10, respectively. The LCOH for both HTSE cases is plotted in each of these figures as a function of the electricity cost (the sensitivity variable with the greatest impact on HTSE LCOH). Hydrogen transportation costs of $0.16/kg are included in the HTSE LCOH shown in these figures (no delivery costs are included in the SMR LCOH under the assumption that the SMR plant would be co-located with the end-use industrial hydrogen customer).

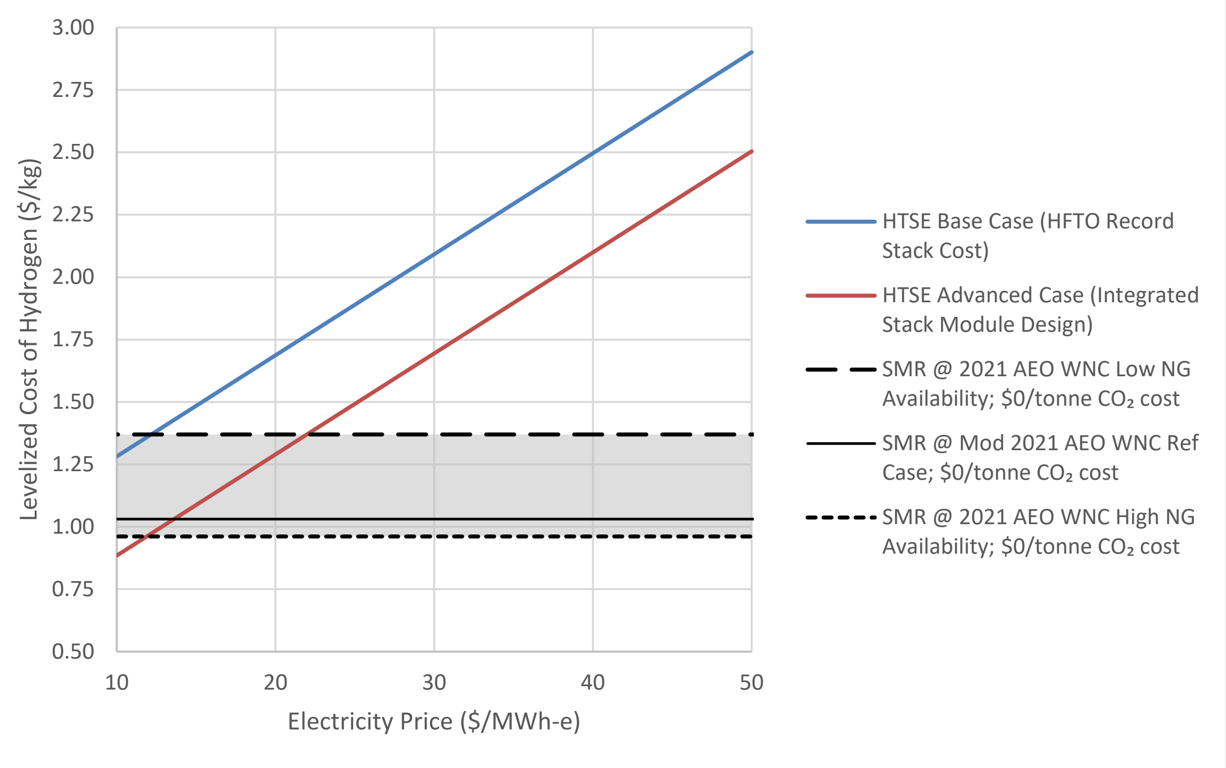

Figure D-8. LCOH of 347 tonne/day HTSE Base and Advanced Cases versus 342 tonne/day SMR with zero cost of CO_2. Economic model input parameters for SMR and HTSE LCOH calculations provided in Table 24 and Table 25, respectively. A hydrogen transportation cost of $0.16/kg is included in the HTSE LCOH (the SMR LCOH does not include hydrogen transportation costs).

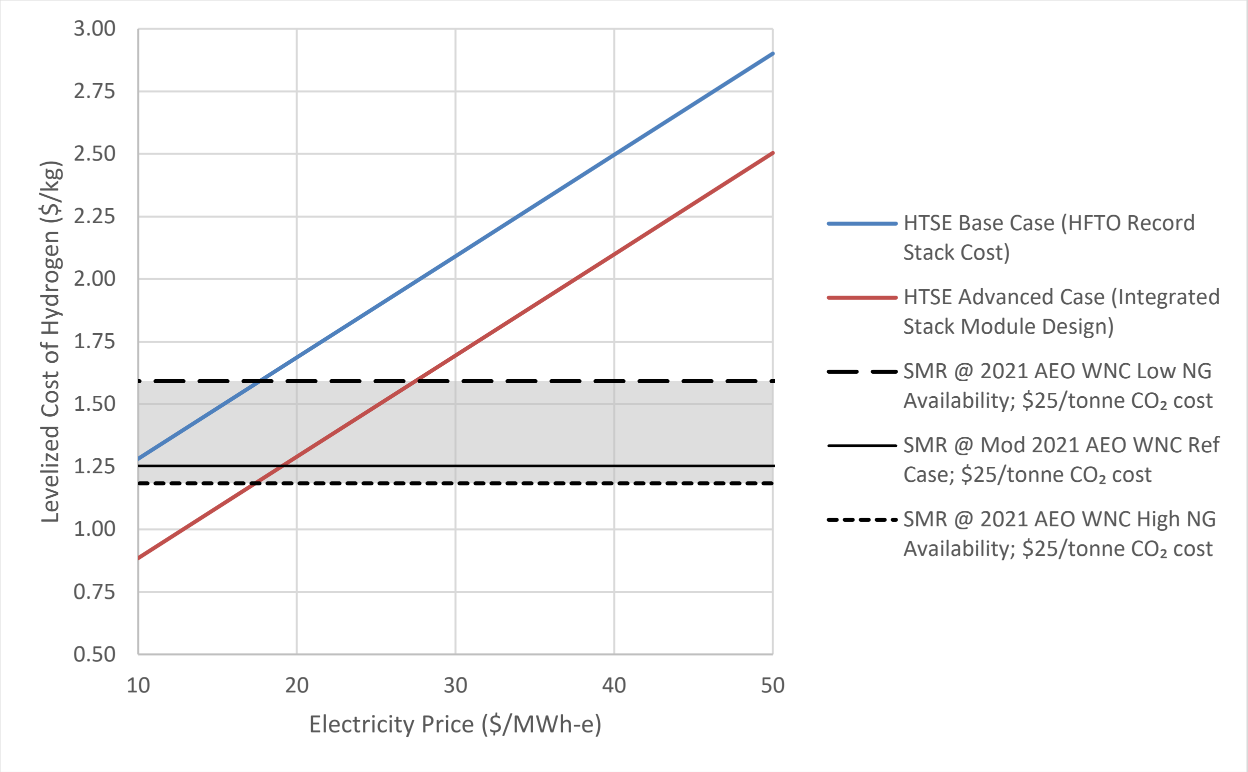

Figure D-9. LCOH of 347 tonne/day HTSE Base and Advanced Cases versus 342 tonne/day SMR with $25/tonne cost of CO2. Economic model input parameters for SMR and HTSE LCOH calculations provided in Table 24 and Table 25, respectively. A hydrogen transportation cost of $0.16/kg is included in the HTSE LCOH (the SMR LCOH does not include hydrogen transportation costs).

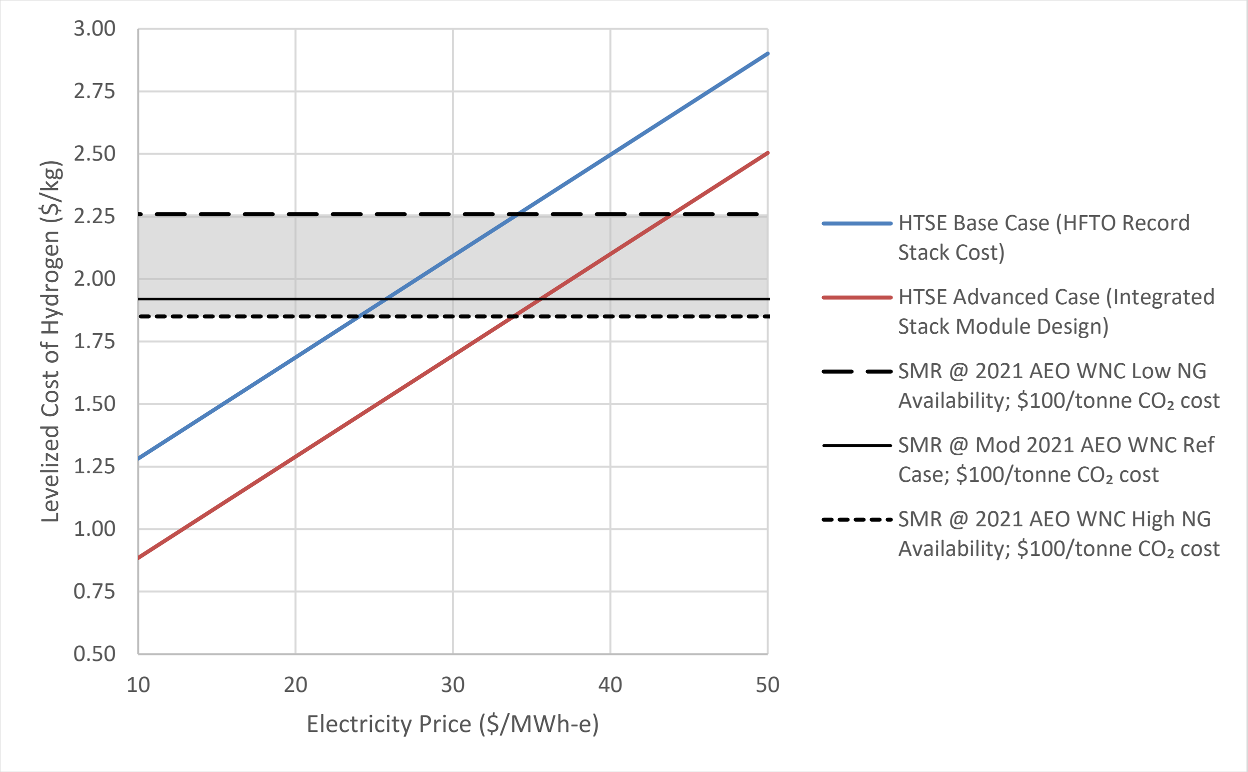

Figure D-10. LCOH of 347 tonne/day HTSE Base and Advanced Cases versus 342 tonne/day SMR with $100/tonne cost of CO2. Economic model input parameters for SMR and HTSE LCOH calculations provided in Table 24 and Table 25, respectively. A hydrogen transportation cost of $0.16/kg is included in the HTSE LCOH (the SMR LCOH does not include hydrogen transportation costs).

APPENDIX E
LWR-HTSE ASU AMMONIA SYNTHESIS PROCESS FLOW DIAGRAMS

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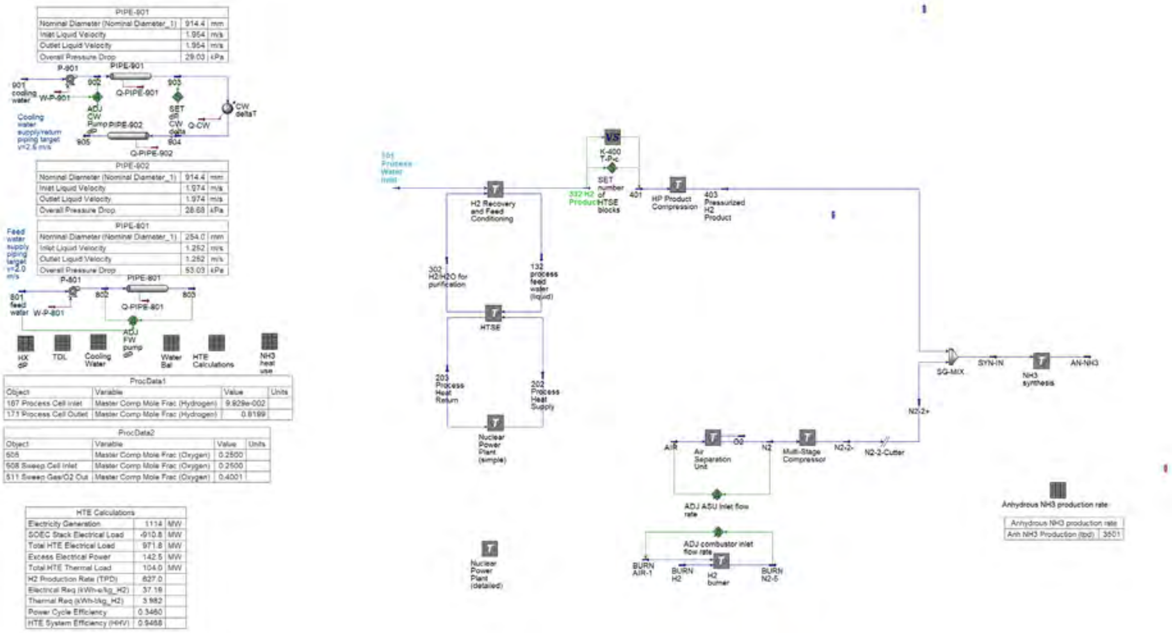
APPENDIX E
LWR-HTSE ASU AMMONIA SYNTHESIS PROCESS FLOW DIAGRAMS

E-1. Main Process



PIPE-901		
Nominal Diameter (Nominal Diameter_1)	914.4	mm
Inlet Liquid Velocity	1.954	m/s
Outlet Liquid Velocity	1.954	m/s
Overall Pressure Drop	29.03	kPa

PIPE-902		
Nominal Diameter (Nominal Diameter_1)	914.4	mm
Inlet Liquid Velocity	1.974	m/s
Outlet Liquid Velocity	1.974	m/s
Overall Pressure Drop	28.68	kPa

PIPE-801		
Nominal Diameter (Nominal Diameter_1)	254.0	mm
Inlet Liquid Velocity	1.252	m/s
Outlet Liquid Velocity	1.252	m/s
Overall Pressure Drop	53.03	kPa

ProcData1

Object	Variable	Value	Units
167 Process Cell Inlet	Master Comp Mole Frac (Hydrogen)	9.929e-002	
171 Process Cell Outlet	Master Comp Mole Frac (Hydrogen)	0.8199	

ProcData2

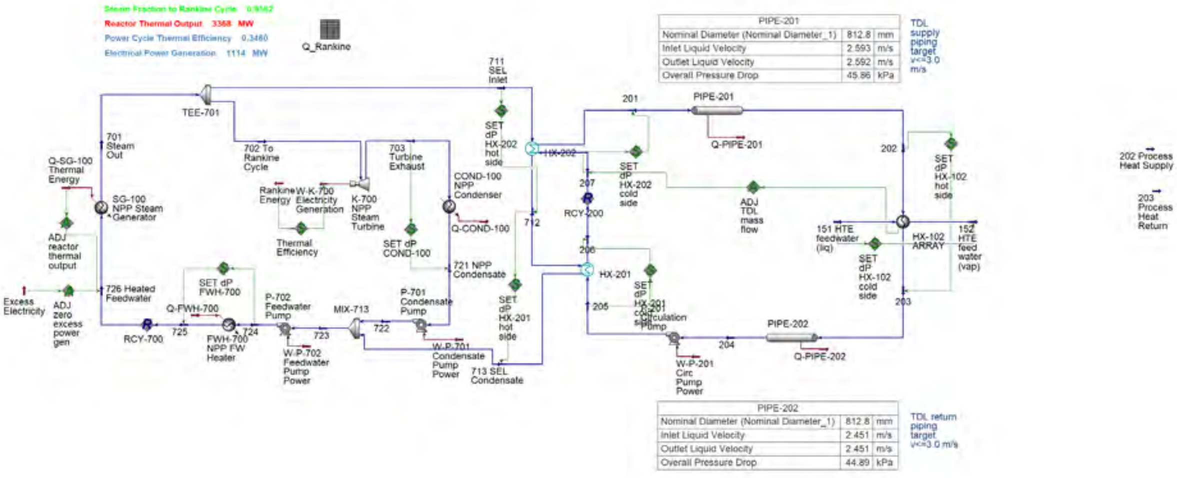
Object	Variable	Value	Units
505	Master Comp Mole Frac (Oxygen)	0.2500	
508 Sweep Cell Inlet	Master Comp Mole Frac (Oxygen)	0.2500	
511 Sweep Gas/O2 Out	Master Comp Mole Frac (Oxygen)	0.4001	

HTE Calculations		
Electricity Generation	1114	MW
SOEC Stack Electrical Load	-910.8	MW
Total HTE Electrical Load	971.8	MW
Excess Electrical Power	142.5	MW
Total HTE Thermal Load	104.0	MW
H2 Production Rate (TPD)	627.0	
Electrical Req (kWh-e/kg_H2)	37.19	
Thermal Req (kWh-t/kg_H2)	3.982	
Power Cycle Efficiency	0.3460	
HTE System Efficiency (HHV)	0.9468	

Anhydrous NH3 production rate	
Anh NH3 Production (tpd)	3501

Figure E-1. Process flow diagram for NPP-HTSE ammonia synthesis, main process.

LWR/TDL



PIPE-201		
Nominal Diameter (Nominal Diameter_1)	812.8	mm
Inlet Liquid Velocity	2.593	m/s
Outlet Liquid Velocity	2.592	m/s
Overall Pressure Drop	45.86	kPa

PIPE-202		
Nominal Diameter (Nominal Diameter_1)	812.8	mm
Inlet Liquid Velocity	2.451	m/s
Outlet Liquid Velocity	2.451	m/s
Overall Pressure Drop	44.89	kPa

Figure E-2. Process flow diagram for NPP-HTSE ammonia synthesis, reactor with TDL.

High-Temperature Steam Electrolysis

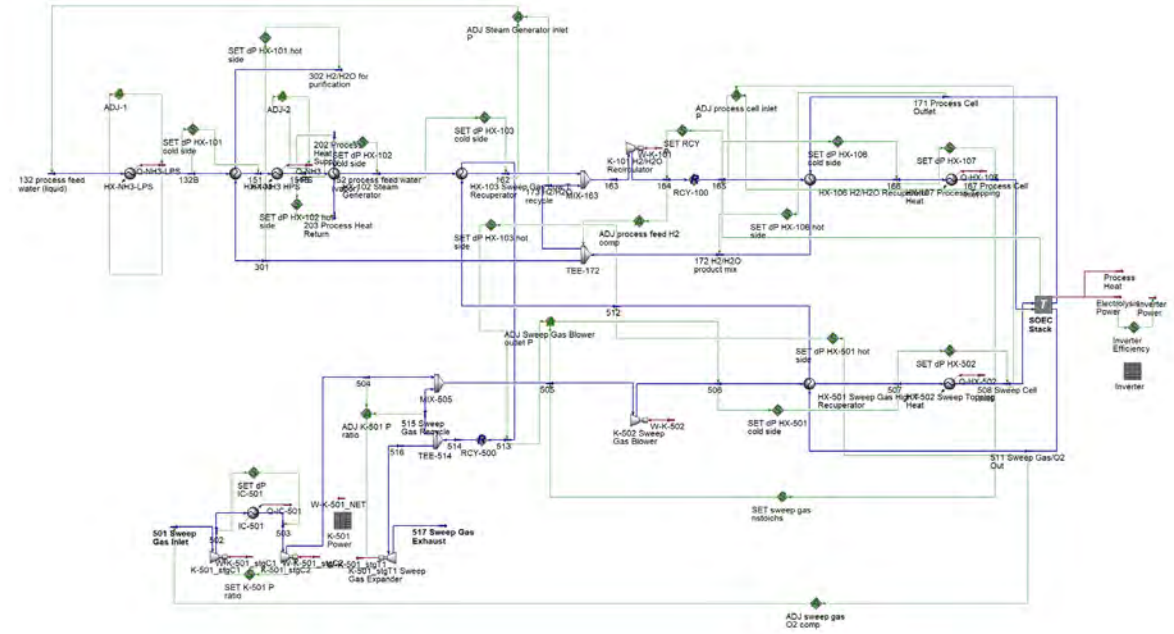


Figure E-3. Process flow diagram for NPP-HTSE ammonia synthesis, HTSE.

H_2 Recovery

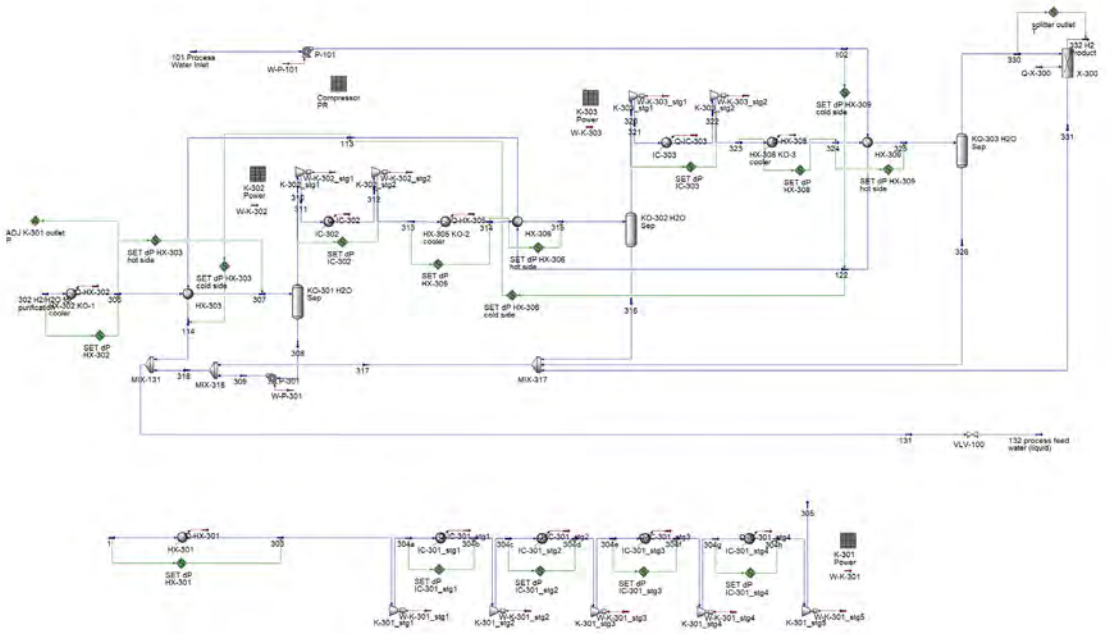


Figure E-4. Process flow diagram for NPP-HTSE ammonia synthesis, hydrogen recovery.

H_2 Product Compression

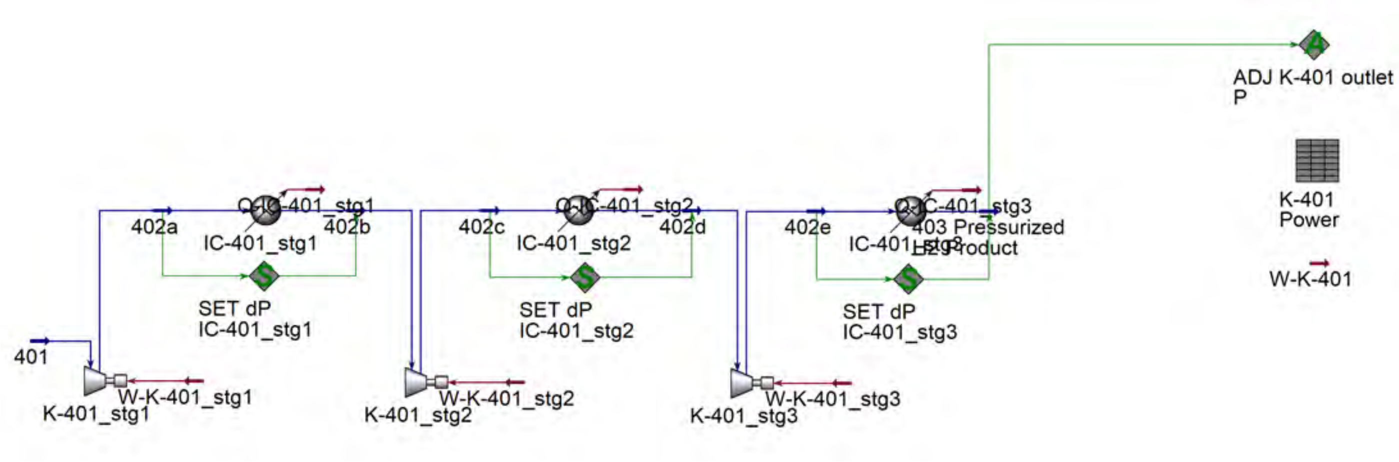


Figure E-5. Process flow diagram for NPP-HTSE ammonia synthesis, hydrogen-product compression.

Air Separation Unit

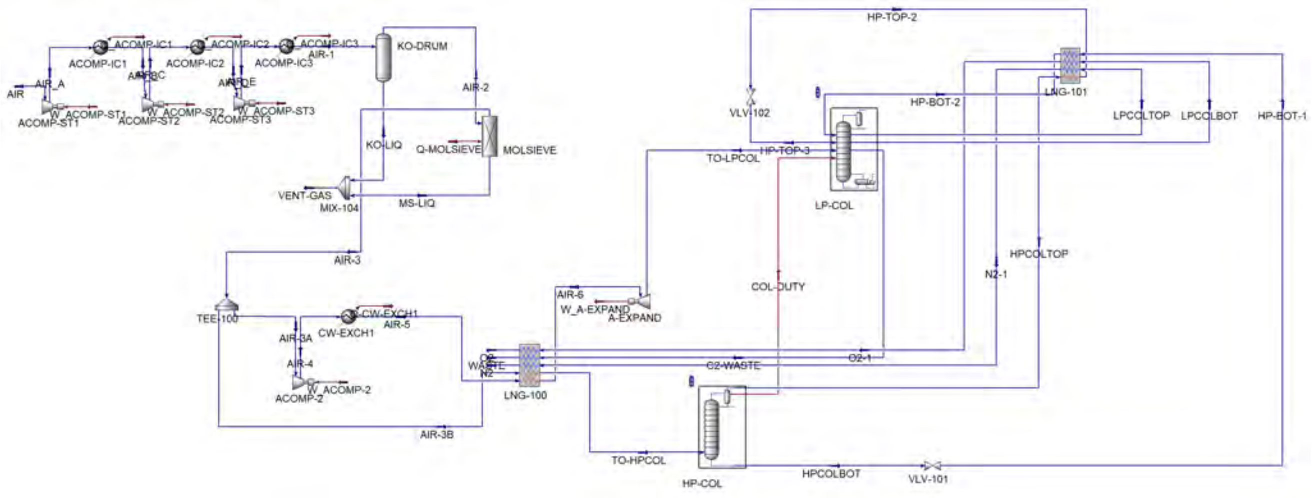


Figure E-6. Process flow diagram for NPP-HTSE ammonia synthesis, ASU.

N_2 Compression

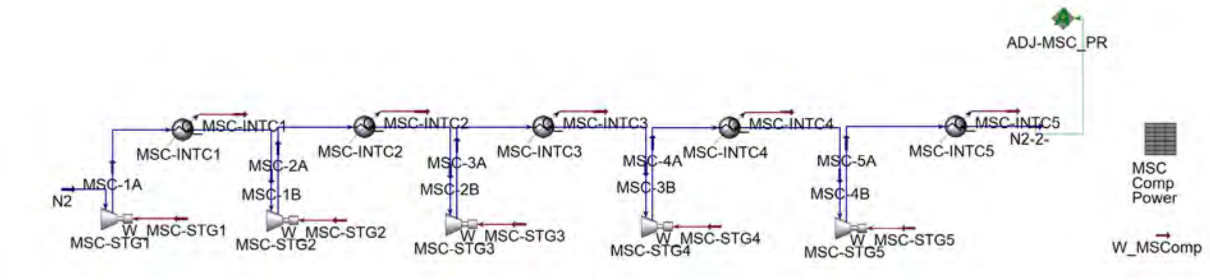
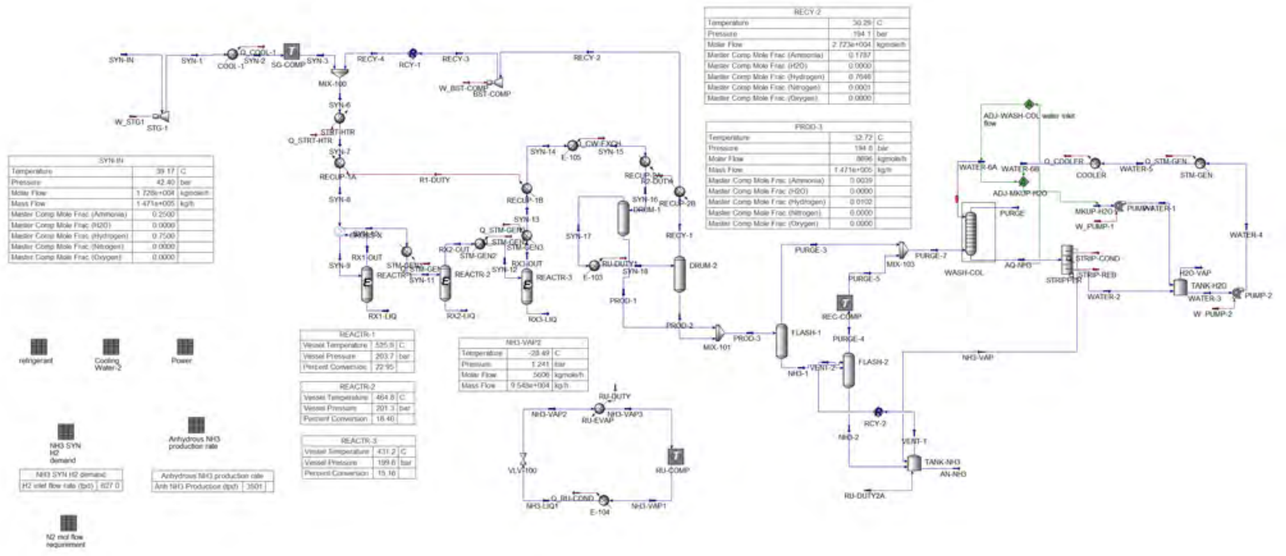


Figure E-7. Process flow diagram for NPP-HTSE ammonia synthesis, nitrogen compression.

NH₃ Synthesis



RECY-2		
Temperature	30.29	C
Pressure	194.1	bar
Molar Flow	2.723e+004	kgmole/h
Master Comp Mole Frac (Ammonia)	0.1797	
Master Comp Mole Frac (H2O)	0.0000	
Master Comp Mole Frac (Hydrogen)	0.7646	
Master Comp Mole Frac (Nitrogen)	0.0001	
Master Comp Mole Frac (Oxygen)	0.0000	

PROD-3		
Temperature	32.72	C
Pressure	194.6	bar
Molar Flow	8696	kgmole/h
Mass Flow	1.471e+005	kg/h
Master Comp Mole Frac (Ammonia)	0.0039	
Master Comp Mole Frac (H2O)	0.0000	
Master Comp Mole Frac (Hydrogen)	0.0102	
Master Comp Mole Frac (Nitrogen)	0.0000	
Master Comp Mole Frac (Oxygen)	0.0000	

SYN-IN		
Temperature	39.17	C
Pressure	42.40	bar
Molar Flow	1.726e+004	kgmole/h
Mass Flow	1.471e+005	kg/h
Master Comp Mole Frac (Ammonia)	0.2500	
Master Comp Mole Frac (H2O)	0.0000	
Master Comp Mole Frac (Hydrogen)	0.7500	
Master Comp Mole Frac (Nitrogen)	0.0000	
Master Comp Mole Frac (Oxygen)	0.0000	

REACTR-1		
Vessel Temperature	525.9	C
Vessel Pressure	203.7	bar
Percent Conversion	22.95	

REACTR-2		
Vessel Temperature	464.8	C
Vessel Pressure	201.3	bar
Percent Conversion	18.40	

REACTR-3		
Vessel Temperature	431.2	C
Vessel Pressure	199.6	bar
Percent Conversion	15.16	

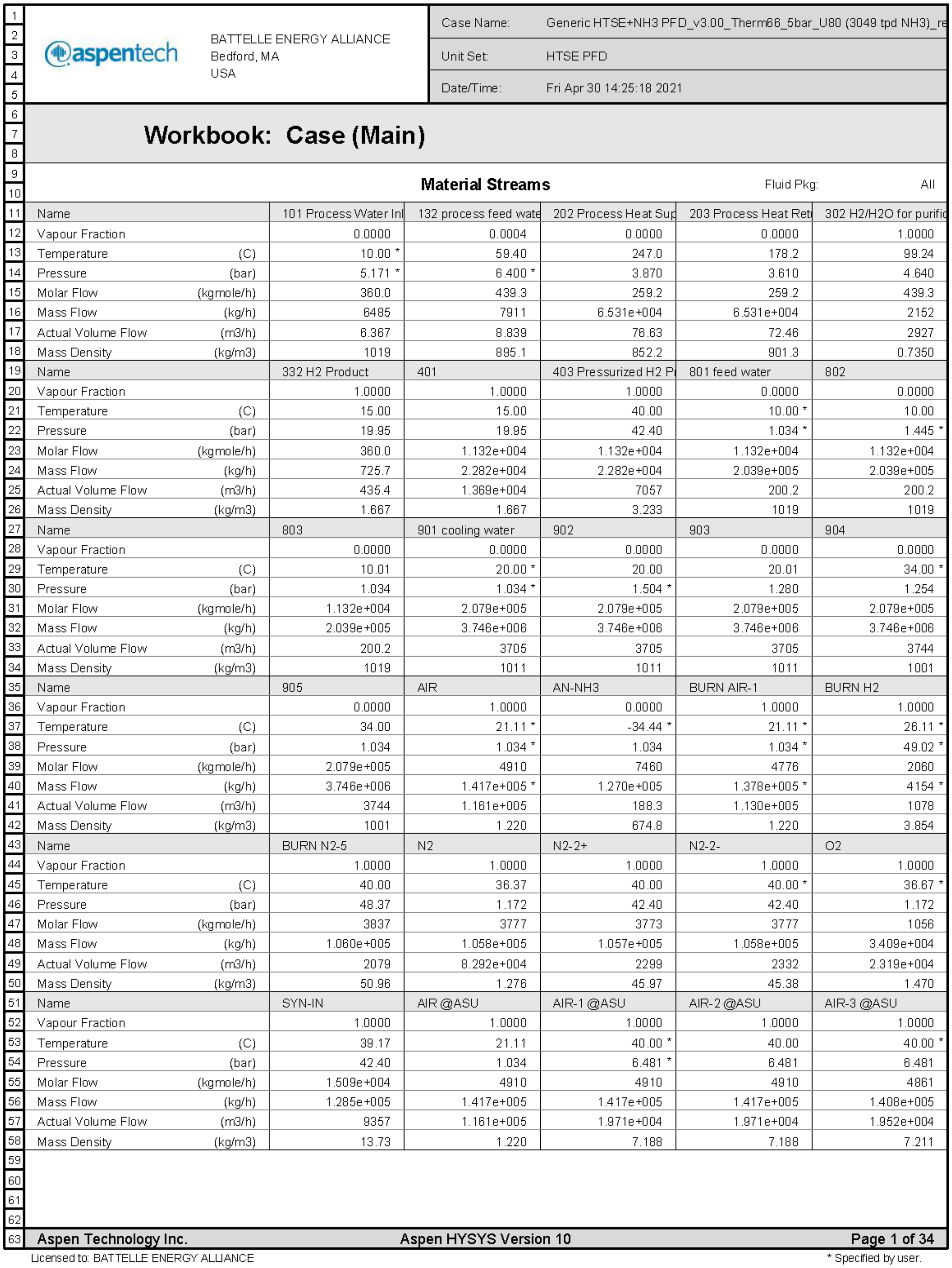
NH3-VAP2		
Temperature	-28.49	C
Pressure	1.241	bar
Molar Flow	5606	kgmole/h
Mass Flow	9.548e+004	kg/h

NH3 SYN H2 demand	
H2 inlet flow rate (tpd)	627.0

Anhydrous NH3 production rate		
Anh NH3 Production (tpd)	3501	

Figure E-8. Process flow diagram for NPP-HTSE ammonia synthesis.

LWR-HTSE ASU Process Model Stream Tables



BATTELLE ENERGY ALLIANCE
Bedford, MA
USA

Case Name: Generic HTSE+NH3 PFD_v3.00_Therm66_5bar_U80 (3049 tpd NH3)_re

Unit Set: HTSE PFD

Date/Time: Fri Apr 30 14:25:18 2021

Workbook: Case (Main)

Material Streams Fluid Pkg: All

Name		101 Process Water Inl	132 process feed wate	202 Process Heat Sup	203 Process Heat Ret	302 H2/H2O for purific
Vapour Fraction		0.0000	0.0004	0.0000	0.0000	1.0000
Temperature	(C)	10.00 *	59.40	247.0	178.2	99.24
Pressure	(bar)	5.171 *	6.400 *	3.870	3.610	4.640
Molar Flow	(kgmole/h)	360.0	439.3	259.2	259.2	439.3
Mass Flow	(kg/h)	6485	7911	6.531e+004	6.531e+004	2152
Actual Volume Flow	(m3/h)	6.367	8.839	76.63	72.46	2927
Mass Density	(kg/m3)	1019	895.1	852.2	901.3	0.7350
Name		332 H2 Product	401	403 Pressurized H2 P	801 feed water	802
Vapour Fraction		1.0000	1.0000	1.0000	0.0000	0.0000
Temperature	(C)	15.00	15.00	40.00	10.00 *	10.00
Pressure	(bar)	19.95	19.95	42.40	1.034 *	1.445 *
Molar Flow	(kgmole/h)	360.0	1.132e+004	1.132e+004	1.132e+004	1.132e+004
Mass Flow	(kg/h)	725.7	2.282e+004	2.282e+004	2.039e+005	2.039e+005
Actual Volume Flow	(m3/h)	435.4	1.369e+004	7057	200.2	200.2
Mass Density	(kg/m3)	1.667	1.667	3.233	1019	1019
Name		803	901 cooling water	902	903	904
Vapour Fraction		0.0000	0.0000	0.0000	0.0000	0.0000
Temperature	(C)	10.01	20.00 *	20.00	20.01	34.00 *
Pressure	(bar)	1.034	1.034 *	1.504 *	1.280	1.254
Molar Flow	(kgmole/h)	1.132e+004	2.079e+005	2.079e+005	2.079e+005	2.079e+005
Mass Flow	(kg/h)	2.039e+005	3.746e+006	3.746e+006	3.746e+006	3.746e+006
Actual Volume Flow	(m3/h)	200.2	3705	3705	3705	3744
Mass Density	(kg/m3)	1019	1011	1011	1011	1001
Name		905	AIR	AN-NH3	BURN AIR-1	BURN H2
Vapour Fraction		0.0000	1.0000	0.0000	1.0000	1.0000
Temperature	(C)	34.00	21.11 *	-34.44 *	21.11 *	26.11 *
Pressure	(bar)	1.034	1.034 *	1.034	1.034 *	49.02 *
Molar Flow	(kgmole/h)	2.079e+005	4910	7460	4776	2060
Mass Flow	(kg/h)	3.746e+006	1.417e+005 *	1.270e+005	1.378e+005 *	4154 *
Actual Volume Flow	(m3/h)	3744	1.161e+005	188.3	1.130e+005	1078
Mass Density	(kg/m3)	1001	1.220	674.8	1.220	3.854
Name		BURN N2-5	N2	N2-2+	N2-2-	O2
Vapour Fraction		1.0000	1.0000	1.0000	1.0000	1.0000
Temperature	(C)	40.00	36.37	40.00	40.00 *	36.67 *
Pressure	(bar)	48.37	1.172	42.40	42.40	1.172
Molar Flow	(kgmole/h)	3837	3777	3773	3777	1056
Mass Flow	(kg/h)	1.060e+005	1.058e+005	1.057e+005	1.058e+005	3.409e+004
Actual Volume Flow	(m3/h)	2079	8.292e+004	2299	2332	2.319e+004
Mass Density	(kg/m3)	50.96	1.276	45.97	45.38	1.470
Name		SYN-IN	AIR @ASU	AIR-1 @ASU	AIR-2 @ASU	AIR-3 @ASU
Vapour Fraction		1.0000	1.0000	1.0000	1.0000	1.0000
Temperature	(C)	39.17	21.11	40.00 *	40.00	40.00 *
Pressure	(bar)	42.40	1.034	6.481 *	6.481	6.481
Molar Flow	(kgmole/h)	1.509e+004	4910	4910	4910	4861
Mass Flow	(kg/h)	1.285e+005	1.417e+005	1.417e+005	1.417e+005	1.408e+005
Actual Volume Flow	(m3/h)	9357	1.161e+005	1.971e+004	1.971e+004	1.952e+004
Mass Density	(kg/m3)	13.73	1.220	7.188	7.188	7.211

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Workbook: Case (Main) (continued)

Material Streams (continued)

Fluid Pkg: All

Name		AIR-4 @ASU	AIR-5 @ASU	AIR-6 @ASU	AIR-3A @ASU	AIR-3B @ASU
Vapour Fraction		1.0000	1.0000	1.0000	1.0000	1.0000
Temperature	(C)	57.97	40.00 *	-170.0 *	40.00	40.00
Pressure	(bar)	7.515 *	7.515	7.033	6.481	6.481
Molar Flow	(kgmole/h)	364.6	364.6	364.6	364.6	4496
Mass Flow	(kg/h)	1.056e+004	1.056e+004	1.056e+004	1.056e+004	1.302e+005
Actual Volume Flow	(m3/h)	1336	1263	379.7	1464	1.806e+004
Mass Density	(kg/m3)	7.901	8.361	27.81	7.211	7.211
Name		AIR_A @ASU	AIR_B @ASU	AIR_C @ASU	AIR_D @ASU	AIR_E @ASU
Vapour Fraction		1.0000	1.0000	1.0000	1.0000	1.0000
Temperature	(C)	95.91	40.00 *	119.2	40.00 *	117.8
Pressure	(bar)	1.913	1.913	3.540	3.540	6.481
Molar Flow	(kgmole/h)	4910	4910	4910	4910	4910
Mass Flow	(kg/h)	1.417e+005	1.417e+005	1.417e+005	1.417e+005	1.417e+005
Actual Volume Flow	(m3/h)	7.877e+004	6.680e+004	4.530e+004	3.610e+004	2.467e+004
Mass Density	(kg/m3)	1.798	2.121	3.127	3.924	5.743
Name		C2-WASTE @ASU	HP-BOT-1 @ASU	HP-BOT-2 @ASU	HP-TOP-2 @ASU	HP-TOP-3 @ASU
Vapour Fraction		1.0000	0.1846	0.4058	0.0000	0.0358
Temperature	(C)	-190.9	-190.8	-190.0 *	-191.1 *	-194.5
Pressure	(bar)	1.172	1.172	1.172	5.998	1.172
Molar Flow	(kgmole/h)	28.72	2429	2429	2067	2067
Mass Flow	(kg/h)	849.6	7.228e+004	7.228e+004	5.794e+004	5.794e+004
Actual Volume Flow	(m3/h)	160.9	2578	5633	73.78	464.9
Mass Density	(kg/m3)	5.280	28.03	12.83	785.3	124.6
Name		HPCOLBOT @ASU	HPCOLTOP @ASU	KO-LIQ @ASU	LPCOLBOT @ASU	LPCOLTOP @ASU
Vapour Fraction		0.0000	1.0000	0.0000	0.0000	1.0000
Temperature	(C)	-172.6	-176.9	40.00	-181.6	-194.5
Pressure	(bar)	5.998	5.998	6.481	1.172	1.172
Molar Flow	(kgmole/h)	2429	2067	0.0000	1056	3777
Mass Flow	(kg/h)	7.228e+004	5.794e+004	0.0000	3.409e+004	1.058e+005
Actual Volume Flow	(m3/h)	86.41	2359	0.0000	29.71	2.015e+004
Mass Density	(kg/m3)	836.4	24.56	746.9	1147	5.252
Name		MS-LIQ @ASU	N2 @ASU	N2-1 @ASU	O2 @ASU	O2-1 @ASU
Vapour Fraction		0.0000	1.0000	1.0000	1.0000	1.0000 *
Temperature	(C)	40.00 *	36.37	-183.5	36.67	-181.6
Pressure	(bar)	6.481	1.172	1.172	1.172	1.172
Molar Flow	(kgmole/h)	49.10	3777	3777	1056	1056
Mass Flow	(kg/h)	884.6	1.058e+005	1.058e+005	3.409e+004	3.409e+004
Actual Volume Flow	(m3/h)	0.8880	8.292e+004	2.326e+004	2.319e+004	6634
Mass Density	(kg/m3)	996.1	1.276	4.550	1.470	5.138
Name		TO-HPCOL @ASU	TO-LPCOL @ASU	VENT-GAS @ASU	WASTE @ASU	Anode @Cell
Vapour Fraction		1.0000	0.8938	0.0000	1.0000	1.0000
Temperature	(C)	-172.2 *	-190.5	40.00	36.67 *	800.0
Pressure	(bar)	5.998	1.172 *	6.481	1.172	5.000
Molar Flow	(kgmole/h)	4496	364.6	49.10	28.72	180.0
Mass Flow	(kg/h)	1.302e+005	1.056e+004	884.6	849.6	5760
Actual Volume Flow	(m3/h)	5471	1835	0.8880	631.1	3215
Mass Density	(kg/m3)	23.80	5.752	996.1	1.346	1.791

Workbook: Case (Main) (continued)

Material Streams (continued)

Fluid Pkg: All

Name		Cathode @Cell	Gas Products @Cell	Liquid Products @Cell	Molar Flow of Oxygen	Process Cell Inlet @C
Vapour Fraction		1.0000	1.0000	0.0000	1.0000	1.0000
Temperature	(C)	800.0	800.0	800.0	800.0 *	800.0
Pressure	(bar)	5.000	5.000	5.000	5.000 *	5.000
Molar Flow	(kgmole/h)	499.6	679.6	0.0000	180.0	499.6
Mass Flow	(kg/h)	2447	8207	0.0000	5760	8207
Actual Volume Flow	(m3/h)	8921	1.214e+004	0.0000	3215	8909
Mass Density	(kg/m3)	0.2743	0.6762	0.6762	1.791	0.9212

Name		Sweep Cell Inlet @Ce	Sweep Gas/O2 Out @	AQ-NH3 @COL4	PURGE @COL4	PURGE-7 @COL4
Vapour Fraction		1.0000	1.0000	0.0000	1.0000	0.2239
Temperature	(C)	800.0	800.0	74.57	40.00	32.21
Pressure	(bar)	5.000	5.000	17.24	17.24	17.24
Molar Flow	(kgmole/h)	719.5	899.4	8668	107.6	1913
Mass Flow	(kg/h)	2.087e+004	2.663e+004	1.544e+005	980.8	3.173e+004
Actual Volume Flow	(m3/h)	1.285e+004	1.607e+004	179.8	164.2	627.3
Mass Density	(kg/m3)	1.624	1.657	858.5	5.974	50.58

Name		WATER-6A @COL4	HPCOLBOT @COL5	HPCOLTOP @COL5	Reflux @COL5	To Condenser @COL
Vapour Fraction		0.0000	0.0000	1.0000	0.0000	1.0000
Temperature	(C)	40.00	-172.6	-176.9	-176.9	-176.9
Pressure	(bar)	17.24	5.998	5.998	5.998	5.998
Molar Flow	(kgmole/h)	6863	2429	2067	2778	4845
Mass Flow	(kg/h)	1.236e+005	7.228e+004	5.794e+004	7.789e+004	1.358e+005
Actual Volume Flow	(m3/h)	124.1	86.41	2359	109.4	5529
Mass Density	(kg/m3)	996.5	836.4	24.56	711.7	24.57

Name		TO-HPCOL @COL5	Boilup @COL6	C2-WASTE @COL6	FlshV @COL6	HP-BOT-2 @COL6
Vapour Fraction		1.0000	1.0000	1.0000	1.0000	0.4058
Temperature	(C)	-172.2	-181.6	-190.9	-181.6	-190.0
Pressure	(bar)	5.998	1.172	1.172	1.172	1.172
Molar Flow	(kgmole/h)	4496	1939	28.72	1939	2429
Mass Flow	(kg/h)	1.302e+005	6.264e+004	849.6	6.264e+004	7.228e+004
Actual Volume Flow	(m3/h)	5471	1.218e+004	160.9	1.218e+004	5633
Mass Density	(kg/m3)	23.80	5.142	5.280	5.142	12.83

Name		HP-TOP-3 @COL6	LPCOLBOT @COL6	LPCOLTOP @COL6	SumpV @COL6	To Reboiler @COL6
Vapour Fraction		0.0358	0.0000	1.0000	1.0000	0.0000
Temperature	(C)	-194.5	-181.6	-194.5	-181.6	-181.7
Pressure	(bar)	1.172	1.172	1.172	1.172	1.172
Molar Flow	(kgmole/h)	2067	1056	3777	8.859e-004	2994
Mass Flow	(kg/h)	5.794e+004	3.409e+004	1.058e+005	2.874e-002	9.672e+004
Actual Volume Flow	(m3/h)	464.9	29.71	2.015e+004	5.562e-003	84.26
Mass Density	(kg/m3)	124.6	1147	5.252	5.167	1148

Name		TO-LPCOL @COL6	ToFlsh @COL6	ToReb @COL6	ToSump @COL6	ToTee @COL6
Vapour Fraction		0.8938	0.7870	0.0000	0.0000	0.0000
Temperature	(C)	-190.5	-181.6	-181.6	-181.6	-181.6
Pressure	(bar)	1.172	1.172	1.172	1.172 *	1.172 *
Molar Flow	(kgmole/h)	364.6	2463	2463	524.8	3519
Mass Flow	(kg/h)	1.056e+004	7.953e+004	7.953e+004	1.690e+004	1.136e+005
Actual Volume Flow	(m3/h)	1835	1.220e+004	69.32	14.76	99.03
Mass Density	(kg/m3)	5.752	6.521	1147	1144	1147

* Specified by user.

Workbook: Case (Main) (continued)

Material Streams (continued)

Fluid Pkg: All

Name		1 @H2burn	2 @H2burn	3 @H2burn	4 @H2burn	5 @H2burn
Vapour Fraction		0.0000	0.0000	0.0000	1.0000 *	1.0000
Temperature	(C)	40.00	40.00	40.01	265.6	265.6 *
Pressure	(bar)	48.37	48.37	49.57 *	49.37	49.37 *
Molar Flow	(kgmole/h)	1997	1.030e+004	1.030e+004	1.030e+004	1.030e+004
Mass Flow	(kg/h)	3.598e+004	1.856e+005	1.856e+005	1.856e+005 *	1.856e+005 *
Actual Volume Flow	(m3/h)	36.08	186.1	186.1	7812	7812
Mass Density	(kg/m3)	997.3	997.3	997.3	23.75	23.75
Name		6 @H2burn	BURN AIR-1 @H2burn	BURN AIR-2 @H2burn	BURN AIR-3 @H2burn	BURN H2 @H2burn
Vapour Fraction		0.2378	1.0000	1.0000	1.0000	1.0000
Temperature	(C)	40.00 *	21.11	745.6	160.0 *	26.11
Pressure	(bar)	48.37	1.034	49.64 *	49.64	49.02
Molar Flow	(kgmole/h)	1.613e+004	4776	4776	4776	2060
Mass Flow	(kg/h)	3.275e+005	1.378e+005	1.378e+005	1.378e+005	4154
Actual Volume Flow	(m3/h)	2301	1.130e+005	8272	3534	1078
Mass Density	(kg/m3)	142.3	1.220	16.66	39.00	3.854
Name		BURN N2-1 @H2burn	BURN N2-2 @H2burn	BURN N2-3 @H2burn	BURN N2-4 @H2burn	BURN N2-5 @H2burn
Vapour Fraction		1.0000	0.9304	0.3474	0.2550	1.0000
Temperature	(C)	981.4	243.6	195.0 *	131.1 *	40.00
Pressure	(bar)	49.02	48.81	48.61	48.40	48.37
Molar Flow	(kgmole/h)	1.613e+004	1.613e+004	1.613e+004	1.613e+004	3837
Mass Flow	(kg/h)	3.275e+005	3.275e+005	3.275e+005	3.275e+005	1.060e+005
Actual Volume Flow	(m3/h)	3.444e+004	1.196e+004	4614	3115	2079
Mass Density	(kg/m3)	9.511	27.38	70.98	105.1	50.96
Name		BURNER LIQ @H2bu	WATER @H2burn	1 @H2rec	101 Process Water Inl	102 @H2rec
Vapour Fraction		0.0000	0.0000	1.0000	0.0000	0.0000
Temperature	(C)	981.4	40.00	120.0 *	10.00	10.05
Pressure	(bar)	49.02	48.37	1.000 *	5.171	11.38 *
Molar Flow	(kgmole/h)	0.0000	1.230e+004	0.1917	360.0	360.0
Mass Flow	(kg/h)	0.0000	2.215e+005	1.000 *	6485	6485
Actual Volume Flow	(m3/h)	0.0000	222.1	6.265	6.367	6.366
Mass Density	(kg/m3)	9.511	997.3	0.1596	1019	1019
Name		113 @H2rec	114 @H2rec	122 @H2rec	131 @H2rec	132 process feed wate
Vapour Fraction		0.0000	0.0000	0.0000	0.0004	0.0004
Temperature	(C)	40.05	60.38	20.05	59.34	59.40
Pressure	(bar)	10.38	9.876	10.88	9.189	6.400
Molar Flow	(kgmole/h)	360.0	360.0	360.0	439.3	439.3
Mass Flow	(kg/h)	6485	6485	6485	7911	7911
Actual Volume Flow	(m3/h)	6.510	6.615	6.413	8.588	8.839
Mass Density	(kg/m3)	996.2	980.4	1011	921.3	895.1
Name		302 H2/H2O for purific	303 @H2rec	304a @H2rec	304b @H2rec	304c @H2rec
Vapour Fraction		1.0000	1.0000 *	1.0000	1.0000 *	1.0000
Temperature	(C)	99.24	59.66	59.67	59.23	59.23
Pressure	(bar)	4.640	0.9800	0.9800	0.9604	0.9604
Molar Flow	(kgmole/h)	439.3	0.1917	0.1917	0.1917	0.1917
Mass Flow	(kg/h)	2152	1.000	1.000	1.000	1.000
Actual Volume Flow	(m3/h)	2927	5.410	5.410	5.513	5.513
Mass Density	(kg/m3)	0.7350	0.1848	0.1848	0.1814	0.1814

* Specified by user.

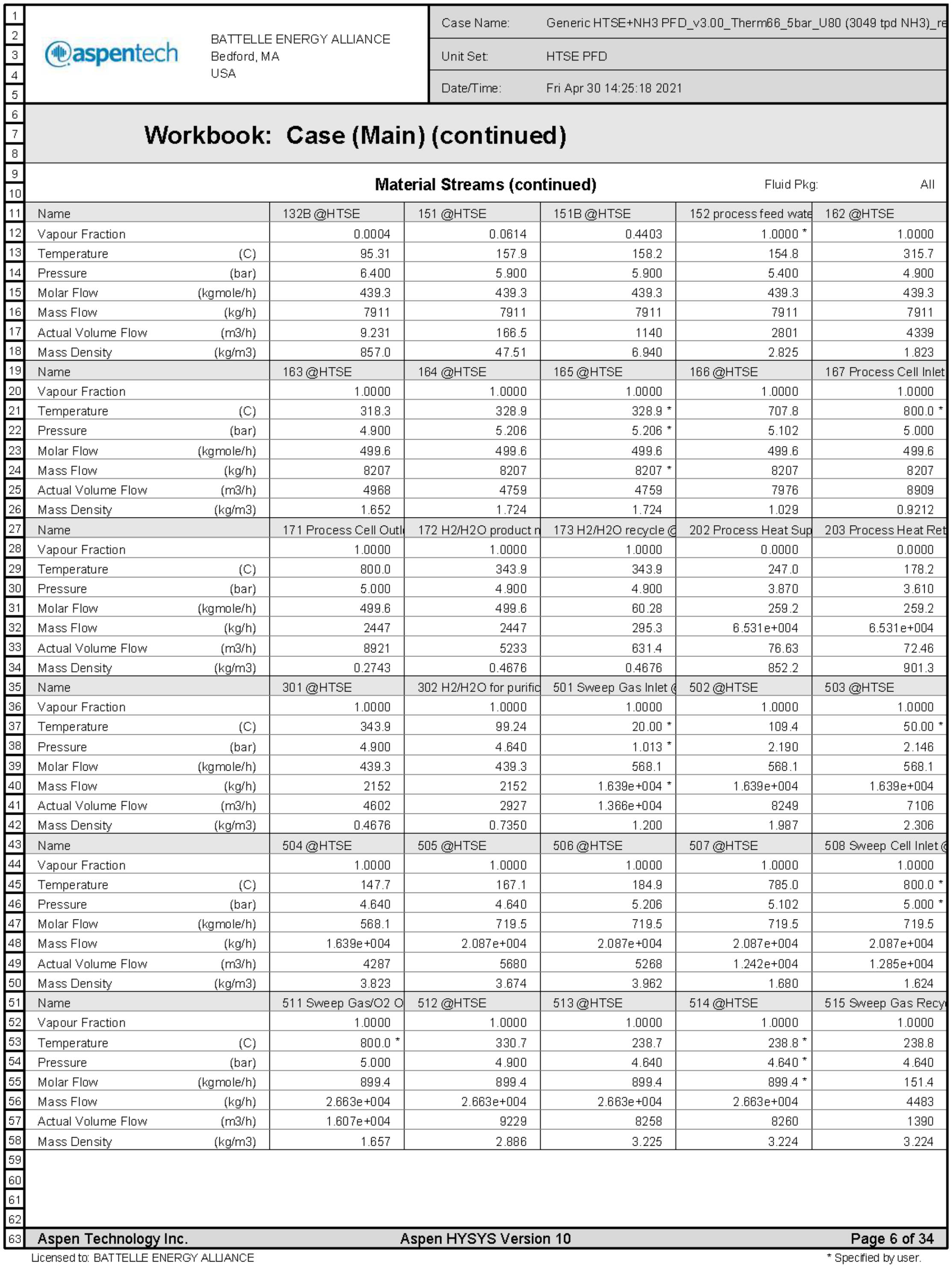
Workbook: Case (Main) (continued)

Material Streams (continued)

Fluid Pkg: All

Name		304d @H2rec	304e @H2rec	304f @H2rec	304g @H2rec	304h @H2rec
Vapour Fraction		1.0000 *	1.0000	1.0000 *	1.0000	1.0000 *
Temperature	(C)	58.80	58.81	58.37	58.38	57.94
Pressure	(bar)	0.9412	0.9412	0.9224	0.9224	0.9039
Molar Flow	(kgmole/h)	0.1917	0.1917	0.1917	0.1917	0.1917
Mass Flow	(kg/h)	1.000	1.000	1.000	1.000	1.000
Actual Volume Flow	(m3/h)	5.619	5.619	5.726	5.726	5.835
Mass Density	(kg/m3)	0.1780	0.1780	0.1746	0.1746	0.1714
Name		305 @H2rec	306 @H2rec	307 @H2rec	308 @H2rec	309 @H2rec
Vapour Fraction		1.0000	0.8822	0.8603	0.0000	0.0000
Temperature	(C)	57.95	70.38	60.00 *	60.00	60.05
Pressure	(bar)	0.9039	4.547	4.287	4.287	9.189
Molar Flow	(kgmole/h)	0.1917	439.3	439.3	61.37	61.37
Mass Flow	(kg/h)	1.000	2152	2152	1106	1106
Actual Volume Flow	(m3/h)	5.836	2436	2445	1.128	1.127
Mass Density	(kg/m3)	0.1714	0.8832	0.8801	980.5	980.7
Name		310 @H2rec	311 @H2rec	312 @H2rec	313 @H2rec	314 @H2rec
Vapour Fraction		1.0000	1.0000	1.0000 *	1.0000	0.9785
Temperature	(C)	60.00	113.0	68.60	123.0	64.00
Pressure	(bar)	4.287	6.495	6.365	9.642	9.449
Molar Flow	(kgmole/h)	377.9	377.9	377.9	377.9	377.9
Mass Flow	(kg/h)	1046	1046	1046	1046	1046
Actual Volume Flow	(m3/h)	2444	1871	1689	1294	1100
Mass Density	(kg/m3)	0.4281	0.5591	0.6193	0.8087	0.9511
Name		315 @H2rec	316 @H2rec	317 @H2rec	318 @H2rec	320 @H2rec
Vapour Fraction		0.9609	0.0000	0.0101	0.0023	1.0000
Temperature	(C)	40.00 *	40.00	35.71	54.56	40.00
Pressure	(bar)	9.189	9.189	9.189	9.189	9.189
Molar Flow	(kgmole/h)	377.9	14.77	17.95	79.32	363.2
Mass Flow	(kg/h)	1046	266.1	320.5	1426	780.1
Actual Volume Flow	(m3/h)	1032	0.2671	0.8285	1.990	1032
Mass Density	(kg/m3)	1.013	996.2	386.8	716.6	0.7559
Name		321 @H2rec	322 @H2rec	323 @H2rec	324 @H2rec	325 @H2rec
Vapour Fraction		1.0000	1.0000 *	1.0000	0.9952	0.9927
Temperature	(C)	90.37	47.25	98.78	37.79	15.00 *
Pressure	(bar)	13.92	13.64	20.66	20.25	19.95
Molar Flow	(kgmole/h)	363.2	363.2	363.2	363.2	363.2
Mass Flow	(kg/h)	780.1	780.1	780.1	780.1	780.1
Actual Volume Flow	(m3/h)	792.1	712.3	547.0	464.6	436.0
Mass Density	(kg/m3)	0.9849	1.095	1.426	1.679	1.789
Name		326 @H2rec	330 @H2rec	331 @H2rec	332 H2 Product @H2r	132 process feed wate
Vapour Fraction		0.0000	1.0000	0.3511	1.0000	0.0004
Temperature	(C)	15.00	15.00	15.00	15.00	59.40
Pressure	(bar)	19.95	19.95	19.95	19.95	6.400
Molar Flow	(kgmole/h)	2.665	360.5	0.5134	360.0	439.3
Mass Flow	(kg/h)	48.00	732.1	6.367	725.7	7911
Actual Volume Flow	(m3/h)	4.728e-002	436.0	0.2239	435.4	8.839
Mass Density	(kg/m3)	1015	1.679	28.44	1.667	895.1

* Specified by user.



BATTELLE ENERGY ALLIANCE
Bedford, MA
USA

Case Name: Generic HTSE+NH3 PFD_v3.00_Therm66_5bar_U80 (3049 tpd NH3)_re

Unit Set: HTSE PFD

Date/Time: Fri Apr 30 14:25:18 2021

Workbook: Case (Main) (continued)

Material Streams (continued)

Fluid Pkg: All

Name		132B @HTSE	151 @HTSE	151B @HTSE	152 process feed wate	162 @HTSE
Vapour Fraction		0.0004	0.0614	0.4403	1.0000 *	1.0000
Temperature	(C)	95.31	157.9	158.2	154.8	315.7
Pressure	(bar)	6.400	5.900	5.900	5.400	4.900
Molar Flow	(kgmole/h)	439.3	439.3	439.3	439.3	439.3
Mass Flow	(kg/h)	7911	7911	7911	7911	7911
Actual Volume Flow	(m3/h)	9.231	166.5	1140	2801	4339
Mass Density	(kg/m3)	857.0	47.51	6.940	2.825	1.823
Name		163 @HTSE	164 @HTSE	165 @HTSE	166 @HTSE	167 Process Cell Inlet
Vapour Fraction		1.0000	1.0000	1.0000	1.0000	1.0000
Temperature	(C)	318.3	328.9	328.9 *	707.8	800.0 *
Pressure	(bar)	4.900	5.206	5.206 *	5.102	5.000
Molar Flow	(kgmole/h)	499.6	499.6	499.6	499.6	499.6
Mass Flow	(kg/h)	8207	8207	8207 *	8207	8207
Actual Volume Flow	(m3/h)	4968	4759	4759	7976	8909
Mass Density	(kg/m3)	1.652	1.724	1.724	1.029	0.9212
Name		171 Process Cell Outl	172 H2/H2O product n	173 H2/H2O recycle @	202 Process Heat Sup	203 Process Heat Ret
Vapour Fraction		1.0000	1.0000	1.0000	0.0000	0.0000
Temperature	(C)	800.0	343.9	343.9	247.0	178.2
Pressure	(bar)	5.000	4.900	4.900	3.870	3.610
Molar Flow	(kgmole/h)	499.6	499.6	60.28	259.2	259.2
Mass Flow	(kg/h)	2447	2447	295.3	6.531e+004	6.531e+004
Actual Volume Flow	(m3/h)	8921	5233	631.4	76.63	72.46
Mass Density	(kg/m3)	0.2743	0.4676	0.4676	852.2	901.3
Name		301 @HTSE	302 H2/H2O for purific	501 Sweep Gas Inlet @	502 @HTSE	503 @HTSE
Vapour Fraction		1.0000	1.0000	1.0000	1.0000	1.0000
Temperature	(C)	343.9	99.24	20.00 *	109.4	50.00 *
Pressure	(bar)	4.900	4.640	1.013 *	2.190	2.146
Molar Flow	(kgmole/h)	439.3	439.3	568.1	568.1	568.1
Mass Flow	(kg/h)	2152	2152	1.639e+004 *	1.639e+004	1.639e+004
Actual Volume Flow	(m3/h)	4602	2927	1.366e+004	8249	7106
Mass Density	(kg/m3)	0.4676	0.7350	1.200	1.987	2.306
Name		504 @HTSE	505 @HTSE	506 @HTSE	507 @HTSE	508 Sweep Cell Inlet @
Vapour Fraction		1.0000	1.0000	1.0000	1.0000	1.0000
Temperature	(C)	147.7	167.1	184.9	785.0	800.0 *
Pressure	(bar)	4.640	4.640	5.206	5.102	5.000 *
Molar Flow	(kgmole/h)	568.1	719.5	719.5	719.5	719.5
Mass Flow	(kg/h)	1.639e+004	2.087e+004	2.087e+004	2.087e+004	2.087e+004
Actual Volume Flow	(m3/h)	4287	5680	5268	1.242e+004	1.285e+004
Mass Density	(kg/m3)	3.823	3.674	3.962	1.680	1.624
Name		511 Sweep Gas/O2 O	512 @HTSE	513 @HTSE	514 @HTSE	515 Sweep Gas Recy
Vapour Fraction		1.0000	1.0000	1.0000	1.0000	1.0000
Temperature	(C)	800.0 *	330.7	238.7	238.8 *	238.8
Pressure	(bar)	5.000	4.900	4.640	4.640 *	4.640
Molar Flow	(kgmole/h)	899.4	899.4	899.4	899.4 *	151.4
Mass Flow	(kg/h)	2.663e+004	2.663e+004	2.663e+004	2.663e+004	4483
Actual Volume Flow	(m3/h)	1.607e+004	9229	8258	8260	1390
Mass Density	(kg/m3)	1.657	2.886	3.225	3.224	3.224

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* Specified by user.

Workbook: Case (Main) (continued)

Material Streams (continued)

Fluid Pkg: All

Name		516 @HTSE	517 Sweep Gas Exhau	401 @KhpH2	402a @KhpH2	402b @KhpH2
Vapour Fraction		1.0000	1.0000	1.0000	1.0000	1.0000
Temperature	(C)	238.8	98.33	15.00	46.83	40.00 *
Pressure	(bar)	4.640	1.013 *	19.95	26.17	25.65
Molar Flow	(kgmole/h)	748.0	748.0	1.132e+004	1.132e+004	1.132e+004
Mass Flow	(kg/h)	2.215e+004	2.215e+004	2.282e+004	2.282e+004	2.282e+004
Actual Volume Flow	(m3/h)	6869	2.280e+004	1.369e+004	1.161e+004	1.159e+004
Mass Density	(kg/m3)	3.224	0.9714	1.667	1.965	1.968
Name		402c @KhpH2	402d @KhpH2	402e @KhpH2	403 Pressurized H2 Pr	PURGE-4 @Krecov
Vapour Fraction		1.0000	1.0000	1.0000	1.0000	1.0000
Temperature	(C)	74.53	40.00 *	74.56	40.00 *	-32.19
Pressure	(bar)	33.65	32.98	43.27	42.40	1.034
Molar Flow	(kgmole/h)	1.132e+004	1.132e+004	1.132e+004	1.132e+004	1577
Mass Flow	(kg/h)	2.282e+004	2.282e+004	2.282e+004	2.282e+004	2.684e+004
Actual Volume Flow	(m3/h)	9837	9042	7678	7057	3.010e+004
Mass Density	(kg/m3)	2.319	2.524	2.972	3.233	0.8917
Name		PURGE-5 @Krecov	PURGE-4A @Krecov	PURGE-4B @Krecov	PURGE-4C @Krecov	PURGE-4D @Krecov
Vapour Fraction		0.0246	1.0000	1.0000	1.0000	1.0000
Temperature	(C)	40.00	26.71	26.71	96.86	40.00 *
Pressure	(bar)	17.56	2.099	2.099	4.262	4.262
Molar Flow	(kgmole/h)	1577	1577	1577	1577	1577
Mass Flow	(kg/h)	2.684e+004	2.684e+004	2.684e+004	2.684e+004	2.684e+004
Actual Volume Flow	(m3/h)	96.84	1.842e+004	1.842e+004	1.119e+004	9352
Mass Density	(kg/m3)	277.2	1.457	1.457	2.400	2.870
Name		PURGE-4E @Krecov	PURGE-4F @Krecov	PURGE-4G @Krecov	NH3-VAP1 @KRU	NH3-VAP3 @KRU
Vapour Fraction		1.0000	1.0000	1.0000	1.0000	1.0000
Temperature	(C)	112.7	40.00 *	113.3	88.33	-30.86
Pressure	(bar)	8.652	8.652	17.56	15.77	1.103
Molar Flow	(kgmole/h)	1577	1577	1577	---	---
Mass Flow	(kg/h)	2.684e+004	2.684e+004	2.684e+004	---	---
Actual Volume Flow	(m3/h)	5666	4455	2698	---	---
Mass Density	(kg/m3)	4.738	6.025	9.948	9.643	0.9474
Name		NH3-VAP3A @KRU	NH3-VAP3B @KRU	NH3-VAP3C @KRU	NH3-VAP3D @KRU	NH3-VAP3E @KRU
Vapour Fraction		1.0000	---	---	---	---
Temperature	(C)	24.21	---	---	---	---
Pressure	(bar)	2.139	2.139	---	---	---
Molar Flow	(kgmole/h)	---	---	---	---	---
Mass Flow	(kg/h)	---	---	---	---	---
Actual Volume Flow	(m3/h)	---	---	---	---	---
Mass Density	(kg/m3)	1.499	---	---	---	---
Name		NH3-VAP3F @KRU	NH3-VAP3G @KRU	SYN-2 @Ksyn	SYN-3 @Ksyn	SYN-2A @Ksyn
Vapour Fraction		---	---	1.0000	1.0000	1.0000
Temperature	(C)	88.33 *	---	40.00	86.11	104.8
Pressure	(bar)	---	15.77	42.40	206.8	71.91
Molar Flow	(kgmole/h)	---	---	1.509e+004	1.509e+004	1.509e+004
Mass Flow	(kg/h)	---	---	1.285e+005	1.285e+005	1.285e+005
Actual Volume Flow	(m3/h)	---	---	9508	2442	6857
Mass Density	(kg/m3)	---	---	13.52	52.62	18.74

Workbook: Case (Main) (continued)

Material Streams (continued)

Fluid Pkg: All

Name		SYN-2B @Ksyn	SYN-2C @Ksyn	SYN-2D @Ksyn	SYN-2E @Ksyn	MSC-1A @MSC
Vapour Fraction		1.0000	1.0000	1.0000	1.0000	1.0000
Temperature	(C)	86.11 *	160.2	86.11 *	160.4	129.2
Pressure	(bar)	71.91	122.0	122.0	206.8	2.402
Molar Flow	(kgmole/h)	1.509e+004	1.509e+004	1.509e+004	1.509e+004	3777
Mass Flow	(kg/h)	1.285e+005	1.285e+005	1.285e+005	1.285e+005	1.058e+005
Actual Volume Flow	(m3/h)	6527	4726	3957	2899	5.263e+004
Mass Density	(kg/m3)	19.69	27.19	32.48	44.33	2.011
Name		MSC-1B @MSC	MSC-2A @MSC	MSC-2B @MSC	MSC-3A @MSC	MSC-3B @MSC
Vapour Fraction		1.0000	1.0000	1.0000	1.0000	1.0000
Temperature	(C)	40.00 *	133.8	40.00 *	133.8	40.00 *
Pressure	(bar)	2.402	4.924	4.924	10.09	10.09
Molar Flow	(kgmole/h)	3777	3777	3777	3777	3777
Mass Flow	(kg/h)	1.058e+005	1.058e+005	1.058e+005	1.058e+005	1.058e+005
Actual Volume Flow	(m3/h)	4.093e+004	2.600e+004	1.997e+004	1.271e+004	9749
Mass Density	(kg/m3)	2.586	4.070	5.299	8.325	10.86
Name		MSC-4A @MSC	MSC-4B @MSC	MSC-5A @MSC	N2 @MSC	N2-2- @MSC
Vapour Fraction		1.0000	1.0000	1.0000	1.0000	1.0000
Temperature	(C)	133.9	40.00 *	133.9	36.37	40.00 *
Pressure	(bar)	20.69	20.69	42.40	1.172	42.40
Molar Flow	(kgmole/h)	3777	3777	3777	3777	3777
Mass Flow	(kg/h)	1.058e+005	1.058e+005	1.058e+005	1.058e+005	1.058e+005
Actual Volume Flow	(m3/h)	6230	4762	3069	8.292e+004	2332
Mass Density	(kg/m3)	16.99	22.22	34.49	1.276	45.38
Name		AN-NH3 @NH3syn	AQ-NH3 @NH3syn	H2O-VAP @NH3syn	MKUP-H2O @NH3syn	NH3-1 @NH3syn
Vapour Fraction		0.0000	0.0000	1.0000	0.0000	0.0000
Temperature	(C)	-34.44	74.57	197.4	30.56 *	29.29
Pressure	(bar)	1.034	17.24	14.82	1.034 *	17.24 *
Molar Flow	(kgmole/h)	7460	8668	0.0000	0.4795	7258
Mass Flow	(kg/h)	1.270e+005	1.544e+005	0.0000	8.638 *	1.236e+005
Actual Volume Flow	(m3/h)	188.3	179.8	0.0000	8.611e-003	208.7
Mass Density	(kg/m3)	674.8	858.5	7.308	1003	592.2
Name		NH3-2 @NH3syn	NH3-LIQ1 @NH3syn	NH3-VAP @NH3syn	NH3-VAP1 @NH3syn	NH3-VAP2 @NH3syn
Vapour Fraction		0.0000	0.0000 *	1.0000	1.0000	0.2466
Temperature	(C)	-32.19	40.00 *	38.29	88.33 *	-28.49
Pressure	(bar)	1.034 *	15.64	14.82	15.77	1.241 *
Molar Flow	(kgmole/h)	5684	---	1778	---	---
Mass Flow	(kg/h)	9.680e+004	---	3.029e+004	---	---
Actual Volume Flow	(m3/h)	144.0	---	2757	---	---
Mass Density	(kg/m3)	672.1	576.4	10.99	9.643	4.267
Name		NH3-VAP3 @NH3syn	PROD-1 @NH3syn	PROD-2 @NH3syn	PROD-3 @NH3syn	PURGE @NH3syn
Vapour Fraction		1.0000 *	0.0000	0.0000	0.0002	1.0000
Temperature	(C)	-30.86	36.07	15.52	32.72	40.00
Pressure	(bar)	1.103	196.2	194.8	194.8	17.24
Molar Flow	(kgmole/h)	---	6324	1270	7594	107.6
Mass Flow	(kg/h)	---	1.070e+005	2.152e+004	1.285e+005	980.8
Actual Volume Flow	(m3/h)	---	183.9	35.12	219.1	164.2
Mass Density	(kg/m3)	0.9474	581.5	612.8	586.3	5.974

* Specified by user.

BATTELLE ENERGY ALLIANCE
Bedford, MA
USA

Case Name: Generic HTSE+NH3 PFD_v3.00_Therm66_5bar_U80 (3049 tpd NH3)_re
Unit Set: HTSE PFD
Date/Time: Fri Apr 30 14:25:18 2021

Workbook: Case (Main) (continued)

Material Streams (continued)

Fluid Pkg: All

Name		PURGE-3 @NH3syn	PURGE-4 @NH3syn	PURGE-5 @NH3syn	PURGE-7 @NH3syn	RECY-1 @NH3syn
Vapour Fraction		1.0000	1.0000	0.0246	0.2239	1.0000
Temperature	(C)	29.29	-32.19	40.00 *	32.21	15.52
Pressure	(bar)	17.24 *	1.034 *	17.56	17.24	194.8
Molar Flow	(kgmole/h)	335.9	1577	1577	1913	2.380e+004
Mass Flow	(kg/h)	4886	2.684e+004	2.684e+004	3.173e+004	1.772e+005
Actual Volume Flow	(m3/h)	458.9	3.010e+004	96.85	627.3	3283
Mass Density	(kg/m3)	10.65	0.8917	277.2	50.58	53.97
Name		RECY-2 @NH3syn	RECY-3 @NH3syn	RECY-4 @NH3syn	RX1-LIQ @NH3syn	RX1-OUT @NH3syn
Vapour Fraction		1.0000	1.0000	1.0000	0.0000	1.0000
Temperature	(C)	30.28	37.79	37.79 *	525.8	525.8
Pressure	(bar)	194.1	206.8 *	206.8 *	203.7	203.7
Molar Flow	(kgmole/h)	2.380e+004	2.380e+004	2.379e+004 *	0.0000	3.520e+004
Mass Flow	(kg/h)	1.772e+005	1.772e+005	1.772e+005	0.0000	3.057e+005
Actual Volume Flow	(m3/h)	3455	3343	3341	0.0000	1.206e+004
Mass Density	(kg/m3)	51.28	53.01	53.02	25.34	25.35
Name		RX2-LIQ @NH3syn	RX2-OUT @NH3syn	RX3-LIQ @NH3syn	RX3-OUT @NH3syn	SYN-1 @NH3syn
Vapour Fraction		0.0000	1.0000	0.0000	1.0000	1.0000
Temperature	(C)	464.7	464.7	431.1	431.1	39.17
Pressure	(bar)	201.3	201.3	199.6	199.6	42.40 *
Molar Flow	(kgmole/h)	0.0000	3.293e+004	0.0000	3.140e+004	1.509e+004
Mass Flow	(kg/h)	0.0000	3.057e+005	0.0000	3.057e+005	1.285e+005
Actual Volume Flow	(m3/h)	0.0000	1.055e+004	0.0000	9677	9484
Mass Density	(kg/m3)	28.92	28.96	31.58	31.59	13.55
Name		SYN-2 @NH3syn	SYN-3 @NH3syn	SYN-6 @NH3syn	SYN-7 @NH3syn	SYN-8 @NH3syn
Vapour Fraction		1.0000	1.0000	1.0000	1.0000	1.0000
Temperature	(C)	40.00 *	86.11 *	56.07	56.09	320.0 *
Pressure	(bar)	42.40	206.8	206.8	206.2	205.5
Molar Flow	(kgmole/h)	1.509e+004	1.509e+004	3.889e+004	3.889e+004	3.889e+004
Mass Flow	(kg/h)	1.285e+005	1.285e+005	3.057e+005	3.057e+005	3.057e+005
Actual Volume Flow	(m3/h)	9508	2442	5782	5800	9999
Mass Density	(kg/m3)	13.52	52.62	52.86	52.70	30.57
Name		SYN-9 @NH3syn	SYN-10 @NH3syn	SYN-11 @NH3syn	SYN-12 @NH3syn	SYN-13 @NH3syn
Vapour Fraction		1.0000	1.0000	1.0000	1.0000	1.0000
Temperature	(C)	357.2 *	488.5	357.2 *	357.2 *	357.2 *
Pressure	(bar)	204.8	203.1	202.4	200.6	198.9
Molar Flow	(kgmole/h)	3.889e+004	3.520e+004	3.520e+004	3.293e+004	3.140e+004
Mass Flow	(kg/h)	3.057e+005	3.057e+005	3.057e+005	3.057e+005	3.057e+005
Actual Volume Flow	(m3/h)	1.062e+004	1.156e+004	9689	9101	8716
Mass Density	(kg/m3)	28.79	26.45	31.55	33.59	35.07
Name		SYN-14 @NH3syn	SYN-15 @NH3syn	SYN-16 @NH3syn	SYN-17 @NH3syn	SYN-18 @NH3syn
Vapour Fraction		0.9964	0.8088	0.7984	1.0000	0.9493
Temperature	(C)	75.30	40.00 *	36.11 *	36.07	15.56 *
Pressure	(bar)	198.2	197.5	196.8	196.2	195.5
Molar Flow	(kgmole/h)	3.140e+004	3.140e+004	3.140e+004	2.507e+004	2.507e+004
Mass Flow	(kg/h)	3.057e+005	3.057e+005	3.057e+005	1.987e+005	1.987e+005
Actual Volume Flow	(m3/h)	4783	3866	3804	3632	3308
Mass Density	(kg/m3)	63.91	79.06	80.35	54.72	60.07

* Specified by user.

Workbook: Case (Main) (continued)

Material Streams (continued)

Fluid Pkg: All

Name	SYN-IN @NH3syn	VENT-1 @NH3syn	VENT-2 @NH3syn	WATER-1 @NH3syn	WATER-2 @NH3syn
Vapour Fraction	1.0000	1.0000	1.0000	0.0000	0.0000
Temperature (C)	39.17	-34.44 *	-34.44 *	30.65	197.4
Pressure (bar)	42.40	1.034	1.034 *	14.82 *	14.82
Molar Flow (kgmole/h)	1.509e+004	3.061	3.062 *	0.4795	6890
Mass Flow (kg/h)	1.285e+005	52.55	52.56	8.638	1.241e+005
Actual Volume Flow (m3/h)	9484	58.02	58.02	8.608e-003	145.3
Mass Density (kg/m3)	13.55	0.9058	0.9058	1004	854.0
Name	WATER-3 @NH3syn	WATER-4 @NH3syn	WATER-5 @NH3syn	WATER-6A @NH3syn	WATER-6B @NH3syn
Vapour Fraction	0.0000	0.0000	0.0000	0.0000	0.0000
Temperature (C)	197.4	197.4	131.1 *	40.00 *	40.00 *
Pressure (bar)	14.82	17.24 *	17.24	17.24 *	17.24
Molar Flow (kgmole/h)	6890	6890	6890	6863	6890
Mass Flow (kg/h)	1.241e+005	1.241e+005	1.241e+005	1.236e+005 *	1.241e+005
Actual Volume Flow (m3/h)	145.3	145.3	135.0	124.1	124.8
Mass Density (kg/m3)	854.0	854.0	919.3	996.5	994.3
Name	151 HTE feedwater (li	152 HTE feed water (v	201 @NPP	202 @NPP	202 Process Heat Sup
Vapour Fraction	0.4403	1.0000	0.0000	0.0000	0.0000
Temperature (C)	158.2	155.3	247.1	247.0	247.0
Pressure (bar)	5.900	5.400	4.000	3.870	3.870
Molar Flow (kgmole/h)	1.381e+004	1.381e+004	8148	8148	259.2
Mass Flow (kg/h)	2.487e+005	2.488e+005	2.053e+006	2.053e+006	6.531e+004
Actual Volume Flow (m3/h)	3.584e+004	8.817e+004	2410	2409	76.63
Mass Density (kg/m3)	6.940	2.821	852.2	852.2	852.2
Name	203 @NPP	203 Process Heat Ret	204 @NPP	205 @NPP	206 @NPP
Vapour Fraction	0.0000	0.0000	0.0000	0.0000	0.0000
Temperature (C)	178.2	178.2	178.1	178.2	189.7
Pressure (bar)	3.610	3.610	3.481	5.000 *	4.500
Molar Flow (kgmole/h)	8148	259.2	8148	8148	8148
Mass Flow (kg/h)	2.053e+006	6.531e+004	2.053e+006	2.053e+006	2.053e+006
Actual Volume Flow (m3/h)	2278	72.46	2278	2278	2298
Mass Density (kg/m3)	901.3	901.3	901.3	901.3	893.4
Name	207 @NPP	701 Steam Out @NPP	702 To Rankine Cycle	703 Turbine Exhaust @	711 SEL Inlet @NPP
Vapour Fraction	0.0000	1.0000 *	1.0000	0.7931	1.0000
Temperature (C)	189.7 *	267.1	267.1	79.34	267.1
Pressure (bar)	4.500 *	52.54 *	52.54	0.4611	52.54
Molar Flow (kgmole/h)	8148	3.664e+005	3.580e+005	3.580e+005	8348
Mass Flow (kg/h)	2.053e+006 *	6.601e+006	6.450e+006	6.450e+006	1.504e+005
Actual Volume Flow (m3/h)	2298	2.470e+005	2.414e+005	1.788e+007	5628
Mass Density (kg/m3)	893.4	26.72	26.72	0.3607	26.72
Name	712 @NPP	713 SEL Condensate (	721 NPP Condensate	722 @NPP	723 @NPP
Vapour Fraction	0.0000	0.0000	0.0000 *	0.0000	0.0000
Temperature (C)	264.3	198.2	78.84	79.25	82.04
Pressure (bar)	51.91	51.29	0.4519	30.66 *	30.66
Molar Flow (kgmole/h)	8348	8348	3.580e+005	3.580e+005	3.664e+005
Mass Flow (kg/h)	1.504e+005	1.504e+005	6.450e+006	6.450e+006	6.601e+006
Actual Volume Flow (m3/h)	193.5	172.9	6633	6625	6792
Mass Density (kg/m3)	777.2	869.6	972.5	973.6	971.9

* Specified by user.

Workbook: Case (Main) (continued)

Material Streams (continued)

Fluid Pkg: All

Name		724 @NPP	725 @NPP	726 Heated Feedwate	101 @NPP-2	102 @NPP-2
Vapour Fraction		0.0000	0.0000	0.0000	1.0000 *	1.0000
Temperature	(C)	82.51	222.2 *	222.2 *	267.1	267.1
Pressure	(bar)	64.73	63.43 *	63.43 *	52.54 *	52.54
Molar Flow	(kgmole/h)	3.664e+005	3.664e+005	3.664e+005	3.663e+005	3.592e+005
Mass Flow	(kg/h)	6.601e+006	6.601e+006	6.601e+006 *	6.599e+006 *	6.471e+006
Actual Volume Flow	(m3/h)	6783	7848	7848	2.469e+005	2.421e+005
Mass Density	(kg/m3)	973.1	841.1	841.1	26.72	26.72
Name		103 @NPP-2	103a @NPP-2	104 @NPP-2	107 @NPP-2	107a @NPP-2
Vapour Fraction		1.0000	0.9991	0.9986	1.0000	0.9989
Temperature	(C)	267.1	265.2	263.9	267.1	264.5
Pressure	(bar)	52.54	50.96	49.97	52.54	50.44
Molar Flow	(kgmole/h)	3.340e+005	3.340e+005	3.340e+005	2.522e+004	2.522e+004
Mass Flow	(kg/h)	6.016e+006	6.016e+006	6.016e+006	4.543e+005	4.543e+005
Actual Volume Flow	(m3/h)	2.251e+005	2.323e+005	2.371e+005	1.700e+004	1.773e+004
Mass Density	(kg/m3)	26.72	25.89	25.37	26.72	25.62
Name		108 (F) @NPP-2	110 @NPP-2	111 @NPP-2	112 @NPP-2	112a @NPP-2
Vapour Fraction		0.0000	0.9383	0.9383	0.9383	0.9387
Temperature	(C)	262.8	225.3	225.3	225.3	223.6
Pressure	(bar)	49.93	25.60	25.60	25.60	24.83
Molar Flow	(kgmole/h)	2.522e+004	3.340e+005	3.052e+005	2.872e+004	2.872e+004
Mass Flow	(kg/h)	4.543e+005	6.016e+006	5.499e+006	5.174e+005	5.174e+005
Actual Volume Flow	(m3/h)	582.8	4.412e+005	4.033e+005	3.795e+004	3.913e+004
Mass Density	(kg/m3)	779.5	13.64	13.64	13.64	13.22
Name		114 @NPP-2	115 @NPP-2	120 @NPP-2	121 @NPP-2	122 (H) @NPP-2
Vapour Fraction		0.5476	0.0000	0.8940	0.8940	0.8940
Temperature	(C)	223.6	188.2	184.8	184.8	184.8
Pressure	(bar)	24.83	24.83	11.18	11.18	11.18
Molar Flow	(kgmole/h)	5.394e+004	5.394e+004	3.052e+005	2.776e+005	2.762e+004
Mass Flow	(kg/h)	9.717e+005	9.717e+005	5.499e+006	5.001e+006	4.977e+005
Actual Volume Flow	(m3/h)	4.334e+004	1105	8.600e+005	7.821e+005	7.783e+004
Mass Density	(kg/m3)	22.42	879.0	6.394	6.394	6.394
Name		122a @NPP-2	122a-L @NPP-2	123 @NPP-2	124 @NPP-2	125 @NPP-2
Vapour Fraction		0.8953	0.0000 *	0.3118	0.0000	0.0000 *
Temperature	(C)	182.5	182.5	182.5	182.4	182.5
Pressure	(bar)	10.62	10.62	10.62	10.62	10.62
Molar Flow	(kgmole/h)	2.762e+004	2.762e+004	8.156e+004	8.156e+004	1.107e+005
Mass Flow	(kg/h)	4.977e+005	4.977e+005	1.469e+006	1.469e+006	1.994e+006
Actual Volume Flow	(m3/h)	8.186e+004	562.8	8.525e+004	1661	2255
Mass Density	(kg/m3)	6.079	884.3	17.24	884.5	884.3
Name		126 (E) @NPP-2	127 @NPP-2	128 @NPP-2	128a @NPP-2	130 @NPP-2
Vapour Fraction		0.0000	1.0000	1.0000	1.0000	1.0000
Temperature	(C)	183.0	183.0	252.9	252.5	147.2
Pressure	(bar)	10.73	10.73	10.30	10.09	3.597
Molar Flow	(kgmole/h)	2.914e+004	2.485e+005	2.485e+005	2.485e+005	2.485e+005
Mass Flow	(kg/h)	5.249e+005	4.476e+006	4.476e+006	4.476e+006	4.476e+006
Actual Volume Flow	(m3/h)	593.9	8.136e+005	1.016e+006	1.037e+006	2.335e+006
Mass Density	(kg/m3)	883.8	5.502	4.404	4.316	1.917

* Specified by user.

Workbook: Case (Main) (continued)

Material Streams (continued)

Fluid Pkg: All

Name		131 @NPP-2	132 @NPP-2	132a (RP) @NPP-2	132a-L (RP-L) @NPP-	134 @NPP-2
Vapour Fraction		1.0000	1.0000	1.0000	0.0000 *	0.0000
Temperature	(C)	147.2	147.2	146.5	138.0	103.3
Pressure	(bar)	3.597	3.597	3.417	3.417	3.417
Molar Flow	(kgmole/h)	2.310e+005	1.752e+004	1.752e+004	1.752e+004	1.752e+004
Mass Flow	(kg/h)	4.161e+006	3.156e+005	3.156e+005	3.156e+005	3.156e+005
Actual Volume Flow	(m3/h)	2.170e+006	1.646e+005	1.733e+005	340.1	330.1
Mass Density	(kg/m3)	1.917	1.917	1.821	927.9	956.1
Name		135 @NPP-2	140 @NPP-2	141 @NPP-2	142 @NPP-2	142a (SO) @NPP-2
Vapour Fraction		0.0053	0.9583	0.9583	0.9583	0.9594
Temperature	(C)	100.6	102.0	102.0	102.0	100.6
Pressure	(bar)	1.034	1.089	1.089	1.089	1.034
Molar Flow	(kgmole/h)	1.752e+004	2.310e+005	2.182e+005	1.273e+004	1.273e+004
Mass Flow	(kg/h)	3.156e+005	4.161e+006	3.931e+006	2.293e+005	2.293e+005
Actual Volume Flow	(m3/h)	3060	6.241e+006	5.897e+006	3.439e+005	3.612e+005
Mass Density	(kg/m3)	103.1	0.6667	0.6667	0.6667	0.6348
Name		142a-L (SO-L) @NPP	144 @NPP-2	145 @NPP-2	146 @NPP-2	150 @NPP-2
Vapour Fraction		0.0000 *	0.4068	0.0000	0.0051	0.9406
Temperature	(C)	100.6	100.6	74.38	71.57	88.25
Pressure	(bar)	1.034	1.034	1.034	0.3335	0.6559
Molar Flow	(kgmole/h)	1.273e+004	3.025e+004	3.025e+004	3.025e+004	2.182e+005
Mass Flow	(kg/h)	2.293e+005	5.449e+005	5.449e+005	5.449e+005	3.931e+006
Actual Volume Flow	(m3/h)	239.4	3.643e+005	558.8	1.368e+004	9.296e+006
Mass Density	(kg/m3)	958.0	1.496	975.2	39.84	0.4229
Name		151 @NPP-2	152 @NPP-2	160 @NPP-2	161 @NPP-2	162 @NPP-2
Vapour Fraction		0.9406	0.9406	0.9205	0.9205	0.9205
Temperature	(C)	88.25	88.25	72.77	72.77	72.77
Pressure	(bar)	0.6559	0.6559	0.3510	0.3510	0.3510
Molar Flow	(kgmole/h)	2.169e+005	1284	2.169e+005	2.042e+005	1.277e+004
Mass Flow	(kg/h)	3.908e+006	2.314e+004	3.908e+006	3.678e+006	2.300e+005
Actual Volume Flow	(m3/h)	9.242e+006	5.471e+004	1.624e+007	1.528e+007	9.557e+005
Mass Density	(kg/m3)	0.4229	0.4229	0.2407	0.2407	0.2407
Name		163 @NPP-2	163a @NPP-2	163a-L @NPP-2	165 @NPP-2	166 @NPP-2
Vapour Fraction		0.9234	0.9244	0.0000 *	0.2967	0.0000 *
Temperature	(C)	72.77	71.57	71.57	71.57	71.57
Pressure	(bar)	0.3510	0.3335	0.3335	0.3335	0.3335
Molar Flow	(kgmole/h)	1.405e+004	1.405e+004	1.405e+004	4.430e+004	4.430e+004
Mass Flow	(kg/h)	2.532e+005	2.532e+005	2.532e+005	7.981e+005	7.981e+005
Actual Volume Flow	(m3/h)	1.055e+006	1.108e+006	259.2	1.122e+006	817.0
Mass Density	(kg/m3)	0.2399	0.2284	976.8	0.7113	976.8
Name		167 @NPP-2	168 @NPP-2	170 @NPP-2	171 @NPP-2	172 @NPP-2
Vapour Fraction		0.0000	0.0100	0.8698	0.7165	0.7008
Temperature	(C)	40.12	34.32	34.32	34.32	34.32
Pressure	(bar)	0.3335	5.419e-002	5.419e-002 *	5.419e-002	5.419e-002
Molar Flow	(kgmole/h)	4.430e+004	4.430e+004	2.042e+005	2.485e+005	2.556e+005
Mass Flow	(kg/h)	7.981e+005	7.981e+005	3.678e+006	4.476e+006	4.605e+006
Actual Volume Flow	(m3/h)	804.4	2.101e+005	8.360e+007	8.381e+007	8.432e+007
Mass Density	(kg/m3)	992.1	3.799	4.400e-002	5.341e-002	5.461e-002

Workbook: Case (Main) (continued)

Material Streams (continued)

Fluid Pkg: All

Name		173 @NPP-2	180 @NPP-2	181 @NPP-2	182 @NPP-2	183 @NPP-2
Vapour Fraction		0.0000 *	0.0000	0.0000	0.0000	0.0000
Temperature	(C)	34.32	34.64	40.10	68.79	97.79
Pressure	(bar)	5.419e-002	30.66 *	30.66	30.66	30.66
Molar Flow	(kgmole/h)	2.556e+005	2.556e+005	2.556e+005	2.556e+005	2.556e+005
Mass Flow	(kg/h)	4.605e+006	4.605e+006	4.605e+006	4.605e+006	4.605e+006
Actual Volume Flow	(m3/h)	4632	4626	4635	4700	4790
Mass Density	(kg/m3)	994.2	995.5	993.5	979.8	961.3

Name		184 @NPP-2	185 @NPP-2	186 @NPP-2	187 @NPP-2	188 @NPP-2
Vapour Fraction		0.0000	0.0000	0.0000	0.0000	0.0000
Temperature	(C)	97.79	135.3	181.4	183.0	181.9
Pressure	(bar)	30.66	30.66	30.66	30.66 *	30.66
Molar Flow	(kgmole/h)	2.556e+005	2.556e+005	2.556e+005	1.107e+005	3.663e+005
Mass Flow	(kg/h)	4.605e+006	4.605e+006	4.605e+006	1.994e+006	6.599e+006
Actual Volume Flow	(m3/h)	4790	4942	5192	2253	7445
Mass Density	(kg/m3)	961.3	931.8	886.9	885.2	886.4

Name		190 @NPP-2	191 @NPP-2	711 @NPP-2	713 @NPP-2	
Vapour Fraction		0.0000	0.0000	1.0000	0.0000	
Temperature	(C)	182.6	221.1	267.1	121.1 *	
Pressure	(bar)	63.43 *	63.43	52.54	49.99 *	
Molar Flow	(kgmole/h)	3.663e+005	3.663e+005	7125	7125	
Mass Flow	(kg/h)	6.599e+006	6.599e+006	1.284e+005	1.284e+005	
Actual Volume Flow	(m3/h)	7433	7833	4803	135.9	
Mass Density	(kg/m3)	887.8	842.5	26.72	944.6	

Compositions

Fluid Pkg: All

Name	101 Process Water Inl	132 process feed wate	202 Process Heat Sup	203 Process Heat Ret	302 H2/H2O for purific
Comp Mole Frac (H2O)	1.0000 *	0.9996	***	***	0.1801
Comp Mole Frac (Hydrogen)	0.0000 *	0.0004	***	***	0.8199
Comp Mole Frac (Oxygen)	0.0000 *	0.0000	***	***	0.0000
Comp Mole Frac (Nitrogen)	0.0000 *	0.0000	***	***	0.0000
Comp Mole Frac (CO2)	0.0000 *	0.0000	***	***	0.0000
Comp Mole Frac (CO)	0.0000 *	0.0000	***	***	0.0000
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	1.0000	1.0000	***
Comp Mole Frac (Ammonia)	***	***	***	***	***
Comp Mole Frac (Argon)	***	***	***	***	***

Name	332 H2 Product	401	403 Pressurized H2 P	801 feed water	802
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000	1.0000 *	1.0000
Comp Mole Frac (Hydrogen)	1.0000	1.0000	1.0000	0.0000 *	0.0000
Comp Mole Frac (Oxygen)	0.0000	0.0000	0.0000	0.0000 *	0.0000
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000 *	0.0000
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000 *	0.0000
Comp Mole Frac (CO)	0.0000	0.0000	0.0000	0.0000 *	0.0000
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	***	***	***	***	***
Comp Mole Frac (Argon)	***	***	***	***	***

* Specified by user.

Workbook: Case (Main) (continued)

Compositions (continued)

Fluid Pkg: All

Name	803	901 cooling water	902	903	904
Comp Mole Frac (H2O)	1.0000	1.0000 *	1.0000	1.0000	1.0000
Comp Mole Frac (Hydrogen)	0.0000	0.0000 *	0.0000	0.0000	0.0000
Comp Mole Frac (Oxygen)	0.0000	0.0000 *	0.0000	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.0000	0.0000 *	0.0000	0.0000	0.0000
Comp Mole Frac (CO2)	0.0000	0.0000 *	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	0.0000	0.0000 *	0.0000	0.0000	0.0000
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	***	***	***	***	***
Comp Mole Frac (Argon)	***	***	***	***	***

Name	905	AIR	AN-NH3	BURN AIR-1	BURN H2
Comp Mole Frac (H2O)	1.0000	0.0100 *	0.0000	0.0000 *	0.0000 *
Comp Mole Frac (Hydrogen)	0.0000	0.0000 *	0.0000	0.0000 *	1.0000 *
Comp Mole Frac (Oxygen)	0.0000	0.2080 *	0.0000	0.2100 *	0.0000 *
Comp Mole Frac (Nitrogen)	0.0000	0.7730 *	0.0000	0.7900 *	0.0000 *
Comp Mole Frac (CO2)	0.0000	0.0000 *	0.0000	0.0000 *	0.0000 *
Comp Mole Frac (CO)	0.0000	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	***	0.0000 *	1.0000	0.0000 *	0.0000 *
Comp Mole Frac (Argon)	***	0.0090 *	0.0000	0.0000 *	0.0000 *

Name	BURN N2-5	N2	N2-2+	N2-2-	O2
Comp Mole Frac (H2O)	0.0025	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Hydrogen)	0.0142	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Oxygen)	0.0000	0.0000	0.0000	0.0000	0.9642
Comp Mole Frac (Nitrogen)	0.9834	0.9990	1.0000	0.9990	0.0000
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	***	***	0.0000	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	0.0000	0.0000	***	0.0000	0.0000
Comp Mole Frac (Argon)	0.0000	0.0010	***	0.0010	0.0358

Name	SYN-IN	AIR @ASU	AIR-1 @ASU	AIR-2 @ASU	AIR-3 @ASU
Comp Mole Frac (H2O)	0.0000	0.0100	0.0100	0.0100	0.0000
Comp Mole Frac (Hydrogen)	0.7500	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Oxygen)	0.0000	0.2080	0.2080	0.2080	0.2101
Comp Mole Frac (Nitrogen)	0.2500	0.7730	0.7730	0.7730	0.7808
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	0.0000	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	***	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Argon)	***	0.0090	0.0090	0.0090	0.0091

Workbook: Case (Main) (continued)

Compositions (continued)

Fluid Pkg: All

Name	AIR-4 @ASU	AIR-5 @ASU	AIR-6 @ASU	AIR-3A @ASU	AIR-3B @ASU
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Hydrogen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Oxygen)	0.2101	0.2101	0.2101	0.2101	0.2101
Comp Mole Frac (Nitrogen)	0.7808	0.7808	0.7808	0.7808	0.7808
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Argon)	0.0091	0.0091	0.0091	0.0091	0.0091

Name	AIR_A @ASU	AIR_B @ASU	AIR_C @ASU	AIR_D @ASU	AIR_E @ASU
Comp Mole Frac (H2O)	0.0100	0.0100	0.0100	0.0100	0.0100
Comp Mole Frac (Hydrogen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Oxygen)	0.2080	0.2080	0.2080	0.2080	0.2080
Comp Mole Frac (Nitrogen)	0.7730	0.7730	0.7730	0.7730	0.7730
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Argon)	0.0090	0.0090	0.0090	0.0090	0.0090

Name	C2-WASTE @ASU	HP-BOT-1 @ASU	HP-BOT-2 @ASU	HP-TOP-2 @ASU	HP-TOP-3 @ASU
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Hydrogen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Oxygen)	0.1134	0.3889	0.3889	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.7929	0.5951	0.5951	0.9990	0.9990
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Argon)	0.0937	0.0160	0.0160	0.0010	0.0010

Name	HPCOLBOT @ASU	HPCOLTOP @ASU	KO-LIQ @ASU	LPCOLBOT @ASU	LPCOLTOP @ASU
Comp Mole Frac (H2O)	0.0000	0.0000	1.0000	0.0000	0.0000
Comp Mole Frac (Hydrogen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Oxygen)	0.3889	0.0000	0.0000	0.9642	0.0000
Comp Mole Frac (Nitrogen)	0.5951	0.9990	0.0000	0.0000	0.9990
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Argon)	0.0160	0.0010	0.0000	0.0358	0.0010

Workbook: Case (Main) (continued)

Compositions (continued)

Fluid Pkg: All

Name	MS-LIQ @ASU	N2 @ASU	N2-1 @ASU	O2 @ASU	O2-1 @ASU
Comp Mole Frac (H2O)	1.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Hydrogen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Oxygen)	0.0000	0.0000	0.0000	0.9642	0.9642
Comp Mole Frac (Nitrogen)	0.0000	0.9990	0.9990	0.0000	0.0000
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Argon)	0.0000	0.0010	0.0010	0.0358	0.0358

Name	TO-HPCOL @ASU	TO-LPCOL @ASU	VENT-GAS @ASU	WASTE @ASU	Anode @Cell
Comp Mole Frac (H2O)	0.0000	0.0000	1.0000	0.0000	0.0000
Comp Mole Frac (Hydrogen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Oxygen)	0.2101	0.2101	0.0000	0.1134	1.0000
Comp Mole Frac (Nitrogen)	0.7808	0.7808	0.0000	0.7929	0.0000
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	***	***	***	***	0.0000
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	0.0000	0.0000	0.0000	0.0000	***
Comp Mole Frac (Argon)	0.0091	0.0091	0.0000	0.0937	***

Name	Cathode @Cell	Gas Products @Cell	Liquid Products @Cell	Molar Flow of Oxygen	Process Cell Inlet @C
Comp Mole Frac (H2O)	0.1801	0.1324	0.1324	0.0000 *	0.9007
Comp Mole Frac (Hydrogen)	0.8199	0.6027	0.6027	0.0000 *	0.0993
Comp Mole Frac (Oxygen)	0.0000	0.2648	0.2648	1.0000 *	0.0000
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000 *	0.0000
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000 *	0.0000
Comp Mole Frac (CO)	0.0000	0.0000	0.0000	0.0000 *	0.0000
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	***	***	***	***	***
Comp Mole Frac (Argon)	***	***	***	***	***

Name	Sweep Cell Inlet @Ce	Sweep Gas/O2 Out @	AQ-NH3 @COL4	PURGE @COL4	PURGE-7 @COL4
Comp Mole Frac (H2O)	0.0000	0.0000	0.7917	0.0045	0.0000
Comp Mole Frac (Hydrogen)	0.0000	0.0000	0.0000	0.7253	0.0409
Comp Mole Frac (Oxygen)	0.2500	0.4001	0.0000	0.0010	0.0001
Comp Mole Frac (Nitrogen)	0.7500	0.5999	0.0000	0.2692	0.0153
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	0.0000	0.0000	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	***	***	0.2083	0.0000	0.9438
Comp Mole Frac (Argon)	***	***	0.0000	0.0000	0.0000

* Specified by user.

Workbook: Case (Main) (continued)

Compositions (continued)

Fluid Pkg: All

Name	WATER-6A @COL4	HPCOLBOT @COL5	HPCOLTOP @COL5	Reflux @COL5	To Condenser @COL
Comp Mole Frac (H2O)	1.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Hydrogen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Oxygen)	0.0000	0.3889	0.0000	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.0000	0.5951	0.9990	0.9981	0.9985
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Argon)	0.0000	0.0160	0.0010	0.0019	0.0015

Name	TO-HPCOL @COL5	Boilup @COL6	C2-WASTE @COL6	FlshV @COL6	HP-BOT-2 @COL6
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Hydrogen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Oxygen)	0.2101	0.9612	0.1134	0.9612	0.3889
Comp Mole Frac (Nitrogen)	0.7808	0.0000	0.7929	0.0000	0.5951
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Argon)	0.0091	0.0388	0.0937	0.0388	0.0160

Name	HP-TOP-3 @COL6	LPCOLBOT @COL6	LPCOLTOP @COL6	SumpV @COL6	To Reboiler @COL6
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Hydrogen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Oxygen)	0.0000	0.9642	0.0000	0.9445	0.9622
Comp Mole Frac (Nitrogen)	0.9990	0.0000	0.9990	0.0000	0.0000
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Argon)	0.0010	0.0358	0.0010	0.0555	0.0378

Name	TO-LPCOL @COL6	ToFlsh @COL6	ToReb @COL6	ToSump @COL6	ToTee @COL6
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Hydrogen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Oxygen)	0.2101	0.9642	0.9642	0.9752	0.9642
Comp Mole Frac (Nitrogen)	0.7808	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Argon)	0.0091	0.0358	0.0358	0.0248	0.0358

Workbook: Case (Main) (continued)

Compositions (continued)

Fluid Pkg: All

Name	1 @H2burn	2 @H2burn	3 @H2burn	4 @H2burn	5 @H2burn
Comp Mole Frac (H2O)	0.9998	0.9998	0.9998	0.9998	0.9998 *
Comp Mole Frac (Hydrogen)	0.0000	0.0000	0.0000	0.0000	0.0000 *
Comp Mole Frac (Oxygen)	0.0000	0.0000	0.0000	0.0000	0.0000 *
Comp Mole Frac (Nitrogen)	0.0002	0.0002	0.0002	0.0002	0.0002 *
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000 *
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	0.0000	0.0000	0.0000	0.0000	0.0000 *
Comp Mole Frac (Argon)	0.0000	0.0000	0.0000	0.0000	0.0000 *

Name	6 @H2burn	BURN AIR-1 @H2burn	BURN AIR-2 @H2burn	BURN AIR-3 @H2burn	BURN H2 @H2burn
Comp Mole Frac (H2O)	0.7626	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Hydrogen)	0.0034	0.0000	0.0000	0.0000	1.0000
Comp Mole Frac (Oxygen)	0.0000	0.2100	0.2100	0.2100	0.0000
Comp Mole Frac (Nitrogen)	0.2340	0.7900	0.7900	0.7900	0.0000
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Argon)	0.0000	0.0000	0.0000	0.0000	0.0000

Name	BURN N2-1 @H2burn	BURN N2-2 @H2burn	BURN N2-3 @H2burn	BURN N2-4 @H2burn	BURN N2-5 @H2burn
Comp Mole Frac (H2O)	0.7626	0.7626	0.7626	0.7626	0.0025
Comp Mole Frac (Hydrogen)	0.0034	0.0034	0.0034	0.0034	0.0142
Comp Mole Frac (Oxygen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.2340	0.2340	0.2340	0.2340	0.9834
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Argon)	0.0000	0.0000	0.0000	0.0000	0.0000

Name	BURNER LIQ @H2bu	WATER @H2burn	1 @H2rec	101 Process Water Inl	102 @H2rec
Comp Mole Frac (H2O)	0.7626	0.9998	0.2000 *	1.0000	1.0000
Comp Mole Frac (Hydrogen)	0.0034	0.0000	0.8000 *	0.0000	0.0000
Comp Mole Frac (Oxygen)	0.0000	0.0000	0.0000 *	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.2340	0.0002	0.0000 *	0.0000	0.0000
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000 *	0.0000	0.0000
Comp Mole Frac (CO)	***	***	0.0000 *	0.0000	0.0000
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	0.0000	0.0000	***	***	***
Comp Mole Frac (Argon)	0.0000	0.0000	***	***	***

Workbook: Case (Main) (continued)

Compositions (continued)

Fluid Pkg: All

Name	113 @H2rec	114 @H2rec	122 @H2rec	131 @H2rec	132 process feed wate
Comp Mole Frac (H2O)	1.0000	1.0000	1.0000	0.9996	0.9996
Comp Mole Frac (Hydrogen)	0.0000	0.0000	0.0000	0.0004	0.0004
Comp Mole Frac (Oxygen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	***	***	***	***	***
Comp Mole Frac (Argon)	***	***	***	***	***

Name	302 H2/H2O for purific	303 @H2rec	304a @H2rec	304b @H2rec	304c @H2rec
Comp Mole Frac (H2O)	0.1801	0.2000	0.2000	0.2000	0.2000
Comp Mole Frac (Hydrogen)	0.8199	0.8000	0.8000	0.8000	0.8000
Comp Mole Frac (Oxygen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	***	***	***	***	***
Comp Mole Frac (Argon)	***	***	***	***	***

Name	304d @H2rec	304e @H2rec	304f @H2rec	304g @H2rec	304h @H2rec
Comp Mole Frac (H2O)	0.2000	0.2000	0.2000	0.2000	0.2000
Comp Mole Frac (Hydrogen)	0.8000	0.8000	0.8000	0.8000	0.8000
Comp Mole Frac (Oxygen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	***	***	***	***	***
Comp Mole Frac (Argon)	***	***	***	***	***

Name	305 @H2rec	306 @H2rec	307 @H2rec	308 @H2rec	309 @H2rec
Comp Mole Frac (H2O)	0.2000	0.1801	0.1801	1.0000	1.0000
Comp Mole Frac (Hydrogen)	0.8000	0.8199	0.8199	0.0000	0.0000
Comp Mole Frac (Oxygen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	***	***	***	***	***
Comp Mole Frac (Argon)	***	***	***	***	***

Workbook: Case (Main) (continued)

Compositions (continued)

Fluid Pkg: All

Name	310 @H2rec	311 @H2rec	312 @H2rec	313 @H2rec	314 @H2rec
Comp Mole Frac (H2O)	0.0470	0.0470	0.0470	0.0470	0.0470
Comp Mole Frac (Hydrogen)	0.9530	0.9530	0.9530	0.9530	0.9530
Comp Mole Frac (Oxygen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	***	***	***	***	***
Comp Mole Frac (Argon)	***	***	***	***	***

Name	315 @H2rec	316 @H2rec	317 @H2rec	318 @H2rec	320 @H2rec
Comp Mole Frac (H2O)	0.0470	1.0000	0.9900	0.9977	0.0083
Comp Mole Frac (Hydrogen)	0.9530	0.0000	0.0100	0.0023	0.9917
Comp Mole Frac (Oxygen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	***	***	***	***	***
Comp Mole Frac (Argon)	***	***	***	***	***

Name	321 @H2rec	322 @H2rec	323 @H2rec	324 @H2rec	325 @H2rec
Comp Mole Frac (H2O)	0.0083	0.0083	0.0083	0.0083	0.0083
Comp Mole Frac (Hydrogen)	0.9917	0.9917	0.9917	0.9917	0.9917
Comp Mole Frac (Oxygen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	***	***	***	***	***
Comp Mole Frac (Argon)	***	***	***	***	***

Name	326 @H2rec	330 @H2rec	331 @H2rec	332 H2 Product @H2r	132 process feed wate
Comp Mole Frac (H2O)	1.0000	0.0009	0.6492	0.0000	0.9996
Comp Mole Frac (Hydrogen)	0.0000	0.9991	0.3508	1.0000	0.0004
Comp Mole Frac (Oxygen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	***	***	***	***	***
Comp Mole Frac (Argon)	***	***	***	***	***

Workbook: Case (Main) (continued)

Compositions (continued)

Fluid Pkg: All

Name	132B @HTSE	151 @HTSE	151B @HTSE	152 process feed wate	162 @HTSE
Comp Mole Frac (H2O)	0.9996	0.9996	0.9996	0.9996	0.9996
Comp Mole Frac (Hydrogen)	0.0004	0.0004	0.0004	0.0004	0.0004
Comp Mole Frac (Oxygen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	***	***	***	***	***
Comp Mole Frac (Argon)	***	***	***	***	***

Name	163 @HTSE	164 @HTSE	165 @HTSE	166 @HTSE	167 Process Cell Inlet
Comp Mole Frac (H2O)	0.9007	0.9007	0.9007 *	0.9007	0.9007
Comp Mole Frac (Hydrogen)	0.0993	0.0993	0.0993 *	0.0993	0.0993
Comp Mole Frac (Oxygen)	0.0000	0.0000	0.0000 *	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000 *	0.0000	0.0000
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000 *	0.0000	0.0000
Comp Mole Frac (CO)	0.0000	0.0000	0.0000 *	0.0000	0.0000
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	***	***	***	***	***
Comp Mole Frac (Argon)	***	***	***	***	***

Name	171 Process Cell Outl	172 H2/H2O product n	173 H2/H2O recycle @	202 Process Heat Sup	203 Process Heat Ret
Comp Mole Frac (H2O)	0.1801	0.1801	0.1801	***	***
Comp Mole Frac (Hydrogen)	0.8199	0.8199	0.8199	***	***
Comp Mole Frac (Oxygen)	0.0000	0.0000	0.0000	***	***
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	***	***
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	***	***
Comp Mole Frac (CO)	0.0000	0.0000	0.0000	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	1.0000 *	1.0000
Comp Mole Frac (Ammonia)	***	***	***	***	***
Comp Mole Frac (Argon)	***	***	***	***	***

Name	301 @HTSE	302 H2/H2O for purific	501 Sweep Gas Inlet @	502 @HTSE	503 @HTSE
Comp Mole Frac (H2O)	0.1801	0.1801	0.0000 *	0.0000	0.0000
Comp Mole Frac (Hydrogen)	0.8199	0.8199	0.0000 *	0.0000	0.0000
Comp Mole Frac (Oxygen)	0.0000	0.0000	0.2100 *	0.2100	0.2100
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.7900 *	0.7900	0.7900
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000 *	0.0000	0.0000
Comp Mole Frac (CO)	0.0000	0.0000	0.0000 *	0.0000	0.0000
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	***	***	***	***	***
Comp Mole Frac (Argon)	***	***	***	***	***

* Specified by user.

Workbook: Case (Main) (continued)

Compositions (continued)

Fluid Pkg: All

Name	504 @HTSE	505 @HTSE	506 @HTSE	507 @HTSE	508 Sweep Cell Inlet @
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000	0.0000	0.0000 *
Comp Mole Frac (Hydrogen)	0.0000	0.0000	0.0000	0.0000	0.0000 *
Comp Mole Frac (Oxygen)	0.2100	0.2500	0.2500	0.2500	0.2500 *
Comp Mole Frac (Nitrogen)	0.7900	0.7500	0.7500	0.7500	0.7500 *
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000 *
Comp Mole Frac (CO)	0.0000	0.0000	0.0000	0.0000	0.0000 *
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	***	***	***	***	***
Comp Mole Frac (Argon)	***	***	***	***	***

Name	511 Sweep Gas/O2 O	512 @HTSE	513 @HTSE	514 @HTSE	515 Sweep Gas Recy
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000	0.0000 *	0.0000
Comp Mole Frac (Hydrogen)	0.0000	0.0000	0.0000	0.0000 *	0.0000
Comp Mole Frac (Oxygen)	0.4001	0.4001	0.4001	0.4001 *	0.4001
Comp Mole Frac (Nitrogen)	0.5999	0.5999	0.5999	0.5999 *	0.5999
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000 *	0.0000
Comp Mole Frac (CO)	0.0000	0.0000	0.0000	0.0000 *	0.0000
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	***	***	***	***	***
Comp Mole Frac (Argon)	***	***	***	***	***

Name	516 @HTSE	517 Sweep Gas Exhau	401 @KhpH2	402a @KhpH2	402b @KhpH2
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Hydrogen)	0.0000	0.0000	1.0000	1.0000	1.0000
Comp Mole Frac (Oxygen)	0.4001	0.4001	0.0000	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.5999	0.5999	0.0000	0.0000	0.0000
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	***	***	***	***	***
Comp Mole Frac (Argon)	***	***	***	***	***

Name	402c @KhpH2	402d @KhpH2	402e @KhpH2	403 Pressurized H2 Pr	PURGE-4 @Krecov
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Hydrogen)	1.0000	1.0000	1.0000	1.0000	0.0015
Comp Mole Frac (Oxygen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000	0.0011
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	0.0000	0.0000	0.0000	0.0000	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	***	***	***	***	0.9974
Comp Mole Frac (Argon)	***	***	***	***	0.0000

* Specified by user.

Workbook: Case (Main) (continued)

Compositions (continued)

Fluid Pkg: All

Name	PURGE-5 @Krecov	PURGE-4A @Krecov	PURGE-4B @Krecov	PURGE-4C @Krecov	PURGE-4D @Krecov
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Hydrogen)	0.0015	0.0015	0.0015	0.0015	0.0015
Comp Mole Frac (Oxygen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.0011	0.0011	0.0011	0.0011	0.0011
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	0.9974	0.9974	0.9974	0.9974	0.9974
Comp Mole Frac (Argon)	0.0000	0.0000	0.0000	0.0000	0.0000

Name	PURGE-4E @Krecov	PURGE-4F @Krecov	PURGE-4G @Krecov	NH3-VAP1 @KRU	NH3-VAP3 @KRU
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Hydrogen)	0.0015	0.0015	0.0015	0.0000	0.0000
Comp Mole Frac (Oxygen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.0011	0.0011	0.0011	0.0000	0.0000
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	0.9974	0.9974	0.9974	1.0000	1.0000
Comp Mole Frac (Argon)	0.0000	0.0000	0.0000	0.0000	0.0000

Name	NH3-VAP3A @KRU	NH3-VAP3B @KRU	NH3-VAP3C @KRU	NH3-VAP3D @KRU	NH3-VAP3E @KRU
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Hydrogen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Oxygen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	1.0000	1.0000	1.0000	1.0000	1.0000
Comp Mole Frac (Argon)	0.0000	0.0000	0.0000	0.0000	0.0000

Name	NH3-VAP3F @KRU	NH3-VAP3G @KRU	SYN-2 @Ksyn	SYN-3 @Ksyn	SYN-2A @Ksyn
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Hydrogen)	0.0000	0.0000	0.7500	0.7500	0.7500
Comp Mole Frac (Oxygen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.2500	0.2500	0.2500
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	1.0000	1.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Argon)	0.0000	0.0000	0.0000	0.0000	0.0000

Workbook: Case (Main) (continued)

Compositions (continued)

Fluid Pkg: All

Name	SYN-2B @Ksyn	SYN-2C @Ksyn	SYN-2D @Ksyn	SYN-2E @Ksyn	MSC-1A @MSC
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Hydrogen)	0.7500	0.7500	0.7500	0.7500	0.0000
Comp Mole Frac (Oxygen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.2500	0.2500	0.2500	0.2500	0.9990
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Argon)	0.0000	0.0000	0.0000	0.0000	0.0010

Name	MSC-1B @MSC	MSC-2A @MSC	MSC-2B @MSC	MSC-3A @MSC	MSC-3B @MSC
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Hydrogen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Oxygen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.9990	0.9990	0.9990	0.9990	0.9990
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Argon)	0.0010	0.0010	0.0010	0.0010	0.0010

Name	MSC-4A @MSC	MSC-4B @MSC	MSC-5A @MSC	N2 @MSC	N2-2- @MSC
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Hydrogen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Oxygen)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.9990	0.9990	0.9990	0.9990	0.9990
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Argon)	0.0010	0.0010	0.0010	0.0010	0.0010

Name	AN-NH3 @NH3syn	AQ-NH3 @NH3syn	H2O-VAP @NH3syn	MKUP-H2O @NH3syn	NH3-1 @NH3syn
Comp Mole Frac (H2O)	0.0000	0.7917	0.9754	1.0000 *	0.0000
Comp Mole Frac (Hydrogen)	0.0000	0.0000	0.0000	0.0000 *	0.0003
Comp Mole Frac (Oxygen)	0.0000	0.0000	0.0000	0.0000 *	0.0000
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000 *	0.0002
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000 *	0.0000
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	1.0000	0.2083	0.0246	0.0000 *	0.9995
Comp Mole Frac (Argon)	0.0000	0.0000	0.0000	0.0000 *	0.0000

* Specified by user.

Workbook: Case (Main) (continued)

Compositions (continued)

Fluid Pkg: All

Name	NH3-2 @NH3syn	NH3-LIQ1 @NH3syn	NH3-VAP @NH3syn	NH3-VAP1 @NH3syn	NH3-VAP2 @NH3syn
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000	0.0000 *	0.0000
Comp Mole Frac (Hydrogen)	0.0000	0.0000	0.0001	0.0000 *	0.0000
Comp Mole Frac (Oxygen)	0.0000	0.0000	0.0000	0.0000 *	0.0000
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0001	0.0000 *	0.0000
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000 *	0.0000
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	1.0000	1.0000	0.9998	1.0000 *	1.0000
Comp Mole Frac (Argon)	0.0000	0.0000	0.0000	0.0000 *	0.0000

Name	NH3-VAP3 @NH3syn	PROD-1 @NH3syn	PROD-2 @NH3syn	PROD-3 @NH3syn	PURGE @NH3syn
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000	0.0000	0.0045
Comp Mole Frac (Hydrogen)	0.0000	0.0109	0.0073	0.0103	0.7253
Comp Mole Frac (Oxygen)	0.0000	0.0000	0.0000	0.0000	0.0010
Comp Mole Frac (Nitrogen)	0.0000	0.0040	0.0027	0.0038	0.2692
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	1.0000	0.9851	0.9899	0.9859	0.0000
Comp Mole Frac (Argon)	0.0000	0.0000	0.0000	0.0000	0.0000

Name	PURGE-3 @NH3syn	PURGE-4 @NH3syn	PURGE-5 @NH3syn	PURGE-7 @NH3syn	RECY-1 @NH3syn
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Hydrogen)	0.2257	0.0015	0.0015	0.0409	0.7674
Comp Mole Frac (Oxygen)	0.0002	0.0000	0.0000	0.0001	0.0001
Comp Mole Frac (Nitrogen)	0.0819	0.0011	0.0011	0.0153	0.1760
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	0.6921	0.9974	0.9974	0.9438	0.0565
Comp Mole Frac (Argon)	0.0000	0.0000	0.0000	0.0000	0.0000

Name	RECY-2 @NH3syn	RECY-3 @NH3syn	RECY-4 @NH3syn	RX1-LIQ @NH3syn	RX1-OUT @NH3syn
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000 *	0.0000	0.0000
Comp Mole Frac (Hydrogen)	0.7674	0.7674	0.7673 *	0.6833	0.6832
Comp Mole Frac (Oxygen)	0.0001	0.0001	0.0001 *	0.0001	0.0001
Comp Mole Frac (Nitrogen)	0.1760	0.1760	0.1761 *	0.1739	0.1739
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000 *	0.0000	0.0000
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	0.0565	0.0565	0.0565 *	0.1427	0.1428
Comp Mole Frac (Argon)	0.0000	0.0000	0.0000 *	0.0000	0.0000

* Specified by user.

Workbook: Case (Main) (continued)

Compositions (continued)

Fluid Pkg: All

Name	RX2-LIQ @NH3syn	RX2-OUT @NH3syn	RX3-LIQ @NH3syn	RX3-OUT @NH3syn	SYN-1 @NH3syn
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Hydrogen)	0.6278	0.6268	0.5844	0.5843	0.7500
Comp Mole Frac (Oxygen)	0.0001	0.0001	0.0001	0.0001	0.0000
Comp Mole Frac (Nitrogen)	0.1517	0.1513	0.1344	0.1344	0.2500
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	0.2204	0.2218	0.2812	0.2813	0.0000
Comp Mole Frac (Argon)	0.0000	0.0000	0.0000	0.0000	0.0000

Name	SYN-2 @NH3syn	SYN-3 @NH3syn	SYN-6 @NH3syn	SYN-7 @NH3syn	SYN-8 @NH3syn
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Hydrogen)	0.7500	0.7500	0.7606	0.7606	0.7606
Comp Mole Frac (Oxygen)	0.0000	0.0000	0.0001	0.0001	0.0001
Comp Mole Frac (Nitrogen)	0.2500	0.2500	0.2048	0.2048	0.2048
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	0.0000	0.0000	0.0346	0.0346	0.0346
Comp Mole Frac (Argon)	0.0000	0.0000	0.0000	0.0000	0.0000

Name	SYN-9 @NH3syn	SYN-10 @NH3syn	SYN-11 @NH3syn	SYN-12 @NH3syn	SYN-13 @NH3syn
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Hydrogen)	0.7606	0.6832	0.6832	0.6268	0.5843
Comp Mole Frac (Oxygen)	0.0001	0.0001	0.0001	0.0001	0.0001
Comp Mole Frac (Nitrogen)	0.2048	0.1739	0.1739	0.1513	0.1344
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	0.0346	0.1428	0.1428	0.2218	0.2813
Comp Mole Frac (Argon)	0.0000	0.0000	0.0000	0.0000	0.0000

Name	SYN-14 @NH3syn	SYN-15 @NH3syn	SYN-16 @NH3syn	SYN-17 @NH3syn	SYN-18 @NH3syn
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Hydrogen)	0.5843	0.5843	0.5843	0.7289	0.7289
Comp Mole Frac (Oxygen)	0.0001	0.0001	0.0001	0.0001	0.0001
Comp Mole Frac (Nitrogen)	0.1344	0.1344	0.1344	0.1672	0.1672
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	0.2813	0.2813	0.2813	0.1038	0.1038
Comp Mole Frac (Argon)	0.0000	0.0000	0.0000	0.0000	0.0000

Workbook: Case (Main) (continued)

Compositions (continued)

Fluid Pkg: All

Name	SYN-IN @NH3syn	VENT-1 @NH3syn	VENT-2 @NH3syn	WATER-1 @NH3syn	WATER-2 @NH3syn
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000 *	1.0000	0.9960
Comp Mole Frac (Hydrogen)	0.7500	0.0413	0.0412 *	0.0000	0.0000
Comp Mole Frac (Oxygen)	0.0000	0.0002	0.0002 *	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.2500	0.0685	0.0685 *	0.0000	0.0000
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000 *	0.0000	0.0000
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	0.0000	0.8900	0.8900 *	0.0000	0.0040
Comp Mole Frac (Argon)	0.0000	0.0000	0.0000 *	0.0000	0.0000

Name	WATER-3 @NH3syn	WATER-4 @NH3syn	WATER-5 @NH3syn	WATER-6A @NH3syn	WATER-6B @NH3syn
Comp Mole Frac (H2O)	0.9960	0.9960	0.9960	1.0000 *	0.9960
Comp Mole Frac (Hydrogen)	0.0000	0.0000	0.0000	0.0000 *	0.0000
Comp Mole Frac (Oxygen)	0.0000	0.0000	0.0000	0.0000 *	0.0000
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000 *	0.0000
Comp Mole Frac (CO2)	0.0000	0.0000	0.0000	0.0000 *	0.0000
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	0.0040	0.0040	0.0040	0.0000 *	0.0040
Comp Mole Frac (Argon)	0.0000	0.0000	0.0000	0.0000 *	0.0000

Name	151 HTE feedwater (li	152 HTE feed water (v	201 @NPP	202 @NPP	202 Process Heat Sup
Comp Mole Frac (H2O)	0.9997	0.9997	***	***	***
Comp Mole Frac (Hydrogen)	0.0003	0.0003	***	***	***
Comp Mole Frac (Oxygen)	0.0000	0.0000	***	***	***
Comp Mole Frac (Nitrogen)	0.0000	0.0000	***	***	***
Comp Mole Frac (CO2)	0.0000	0.0000	***	***	***
Comp Mole Frac (CO)	0.0000	0.0000	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	1.0000	1.0000	1.0000
Comp Mole Frac (Ammonia)	***	***	***	***	***
Comp Mole Frac (Argon)	***	***	***	***	***

Name	203 @NPP	203 Process Heat Ret	204 @NPP	205 @NPP	206 @NPP
Comp Mole Frac (H2O)	***	***	***	***	***
Comp Mole Frac (Hydrogen)	***	***	***	***	***
Comp Mole Frac (Oxygen)	***	***	***	***	***
Comp Mole Frac (Nitrogen)	***	***	***	***	***
Comp Mole Frac (CO2)	***	***	***	***	***
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	1.0000	1.0000	1.0000	1.0000	1.0000
Comp Mole Frac (Ammonia)	***	***	***	***	***
Comp Mole Frac (Argon)	***	***	***	***	***

* Specified by user.

Workbook: Case (Main) (continued)

Compositions (continued)

Fluid Pkg: All

Name	207 @NPP	701 Steam Out @NPF	702 To Rankine Cycle	703 Turbine Exhaust @	711 SEL Inlet @NPP
Comp Mole Frac (H2O)	***	1.0000	1.0000	1.0000	1.0000
Comp Mole Frac (Hydrogen)	***	***	***	***	***
Comp Mole Frac (Oxygen)	***	***	***	***	***
Comp Mole Frac (Nitrogen)	***	***	***	***	***
Comp Mole Frac (CO2)	***	***	***	***	***
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	1.0000 *	***	***	***	***
Comp Mole Frac (Ammonia)	***	***	***	***	***
Comp Mole Frac (Argon)	***	***	***	***	***

Name	712 @NPP	713 SEL Condensate (	721 NPP Condensate	722 @NPP	723 @NPP
Comp Mole Frac (H2O)	1.0000	1.0000	1.0000	1.0000	1.0000
Comp Mole Frac (Hydrogen)	***	***	***	***	***
Comp Mole Frac (Oxygen)	***	***	***	***	***
Comp Mole Frac (Nitrogen)	***	***	***	***	***
Comp Mole Frac (CO2)	***	***	***	***	***
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	***	***	***	***	***
Comp Mole Frac (Argon)	***	***	***	***	***

Name	724 @NPP	725 @NPP	726 Heated Feedwate	101 @NPP-2	102 @NPP-2
Comp Mole Frac (H2O)	1.0000	1.0000	1.0000 *	1.0000 *	1.0000
Comp Mole Frac (Hydrogen)	***	***	***	***	***
Comp Mole Frac (Oxygen)	***	***	***	***	***
Comp Mole Frac (Nitrogen)	***	***	***	***	***
Comp Mole Frac (CO2)	***	***	***	***	***
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	***	***	***	***	***
Comp Mole Frac (Argon)	***	***	***	***	***

Name	103 @NPP-2	103a @NPP-2	104 @NPP-2	107 @NPP-2	107a @NPP-2
Comp Mole Frac (H2O)	1.0000	1.0000	1.0000	1.0000	1.0000
Comp Mole Frac (Hydrogen)	***	***	***	***	***
Comp Mole Frac (Oxygen)	***	***	***	***	***
Comp Mole Frac (Nitrogen)	***	***	***	***	***
Comp Mole Frac (CO2)	***	***	***	***	***
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	***	***	***	***	***
Comp Mole Frac (Argon)	***	***	***	***	***

Workbook: Case (Main) (continued)

Compositions (continued)

Fluid Pkg: All

Name	108 (F) @NPP-2	110 @NPP-2	111 @NPP-2	112 @NPP-2	112a @NPP-2
Comp Mole Frac (H2O)	1.0000	1.0000	1.0000	1.0000	1.0000
Comp Mole Frac (Hydrogen)	***	***	***	***	***
Comp Mole Frac (Oxygen)	***	***	***	***	***
Comp Mole Frac (Nitrogen)	***	***	***	***	***
Comp Mole Frac (CO2)	***	***	***	***	***
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	***	***	***	***	***
Comp Mole Frac (Argon)	***	***	***	***	***

Name	114 @NPP-2	115 @NPP-2	120 @NPP-2	121 @NPP-2	122 (H) @NPP-2
Comp Mole Frac (H2O)	1.0000	1.0000	1.0000	1.0000	1.0000
Comp Mole Frac (Hydrogen)	***	***	***	***	***
Comp Mole Frac (Oxygen)	***	***	***	***	***
Comp Mole Frac (Nitrogen)	***	***	***	***	***
Comp Mole Frac (CO2)	***	***	***	***	***
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	***	***	***	***	***
Comp Mole Frac (Argon)	***	***	***	***	***

Name	122a @NPP-2	122a-L @NPP-2	123 @NPP-2	124 @NPP-2	125 @NPP-2
Comp Mole Frac (H2O)	1.0000	1.0000	1.0000	1.0000	1.0000
Comp Mole Frac (Hydrogen)	***	***	***	***	***
Comp Mole Frac (Oxygen)	***	***	***	***	***
Comp Mole Frac (Nitrogen)	***	***	***	***	***
Comp Mole Frac (CO2)	***	***	***	***	***
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	***	***	***	***	***
Comp Mole Frac (Argon)	***	***	***	***	***

Name	126 (E) @NPP-2	127 @NPP-2	128 @NPP-2	128a @NPP-2	130 @NPP-2
Comp Mole Frac (H2O)	1.0000	1.0000	1.0000	1.0000	1.0000
Comp Mole Frac (Hydrogen)	***	***	***	***	***
Comp Mole Frac (Oxygen)	***	***	***	***	***
Comp Mole Frac (Nitrogen)	***	***	***	***	***
Comp Mole Frac (CO2)	***	***	***	***	***
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	***	***	***	***	***
Comp Mole Frac (Argon)	***	***	***	***	***

Workbook: Case (Main) (continued)

Compositions (continued)

Fluid Pkg: All

Name	131 @NPP-2	132 @NPP-2	132a (RP) @NPP-2	132a-L (RP-L) @NPP-	134 @NPP-2
Comp Mole Frac (H2O)	1.0000	1.0000	1.0000	1.0000	1.0000
Comp Mole Frac (Hydrogen)	***	***	***	***	***
Comp Mole Frac (Oxygen)	***	***	***	***	***
Comp Mole Frac (Nitrogen)	***	***	***	***	***
Comp Mole Frac (CO2)	***	***	***	***	***
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	***	***	***	***	***
Comp Mole Frac (Argon)	***	***	***	***	***

Name	135 @NPP-2	140 @NPP-2	141 @NPP-2	142 @NPP-2	142a (SO) @NPP-2
Comp Mole Frac (H2O)	1.0000	1.0000	1.0000	1.0000	1.0000
Comp Mole Frac (Hydrogen)	***	***	***	***	***
Comp Mole Frac (Oxygen)	***	***	***	***	***
Comp Mole Frac (Nitrogen)	***	***	***	***	***
Comp Mole Frac (CO2)	***	***	***	***	***
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	***	***	***	***	***
Comp Mole Frac (Argon)	***	***	***	***	***

Name	142a-L (SO-L) @NPP	144 @NPP-2	145 @NPP-2	146 @NPP-2	150 @NPP-2
Comp Mole Frac (H2O)	1.0000	1.0000	1.0000	1.0000	1.0000
Comp Mole Frac (Hydrogen)	***	***	***	***	***
Comp Mole Frac (Oxygen)	***	***	***	***	***
Comp Mole Frac (Nitrogen)	***	***	***	***	***
Comp Mole Frac (CO2)	***	***	***	***	***
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	***	***	***	***	***
Comp Mole Frac (Argon)	***	***	***	***	***

Name	151 @NPP-2	152 @NPP-2	160 @NPP-2	161 @NPP-2	162 @NPP-2
Comp Mole Frac (H2O)	1.0000	1.0000	1.0000	1.0000	1.0000
Comp Mole Frac (Hydrogen)	***	***	***	***	***
Comp Mole Frac (Oxygen)	***	***	***	***	***
Comp Mole Frac (Nitrogen)	***	***	***	***	***
Comp Mole Frac (CO2)	***	***	***	***	***
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	***	***	***	***	***
Comp Mole Frac (Argon)	***	***	***	***	***

Workbook: Case (Main) (continued)

Compositions (continued)

Fluid Pkg: All

Name	163 @NPP-2	163a @NPP-2	163a-L @NPP-2	165 @NPP-2	166 @NPP-2
Comp Mole Frac (H2O)	1.0000	1.0000	1.0000	1.0000	1.0000
Comp Mole Frac (Hydrogen)	***	***	***	***	***
Comp Mole Frac (Oxygen)	***	***	***	***	***
Comp Mole Frac (Nitrogen)	***	***	***	***	***
Comp Mole Frac (CO2)	***	***	***	***	***
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	***	***	***	***	***
Comp Mole Frac (Argon)	***	***	***	***	***

Name	167 @NPP-2	168 @NPP-2	170 @NPP-2	171 @NPP-2	172 @NPP-2
Comp Mole Frac (H2O)	1.0000	1.0000	1.0000	1.0000	1.0000
Comp Mole Frac (Hydrogen)	***	***	***	***	***
Comp Mole Frac (Oxygen)	***	***	***	***	***
Comp Mole Frac (Nitrogen)	***	***	***	***	***
Comp Mole Frac (CO2)	***	***	***	***	***
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	***	***	***	***	***
Comp Mole Frac (Argon)	***	***	***	***	***

Name	173 @NPP-2	180 @NPP-2	181 @NPP-2	182 @NPP-2	183 @NPP-2
Comp Mole Frac (H2O)	1.0000	1.0000	1.0000	1.0000	1.0000
Comp Mole Frac (Hydrogen)	***	***	***	***	***
Comp Mole Frac (Oxygen)	***	***	***	***	***
Comp Mole Frac (Nitrogen)	***	***	***	***	***
Comp Mole Frac (CO2)	***	***	***	***	***
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	***	***	***	***	***
Comp Mole Frac (Argon)	***	***	***	***	***

Name	184 @NPP-2	185 @NPP-2	186 @NPP-2	187 @NPP-2	188 @NPP-2
Comp Mole Frac (H2O)	1.0000	1.0000	1.0000	1.0000	1.0000
Comp Mole Frac (Hydrogen)	***	***	***	***	***
Comp Mole Frac (Oxygen)	***	***	***	***	***
Comp Mole Frac (Nitrogen)	***	***	***	***	***
Comp Mole Frac (CO2)	***	***	***	***	***
Comp Mole Frac (CO)	***	***	***	***	***
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Comp Mole Frac (Ammonia)	***	***	***	***	***
Comp Mole Frac (Argon)	***	***	***	***	***

* Specified by user.

Workbook: Case (Main) (continued)

Compositions (continued)

Fluid Pkg: All

Name	190 @NPP-2	191 @NPP-2	711 @NPP-2	713 @NPP-2	
Comp Mole Frac (H2O)	1.0000	1.0000	1.0000	1.0000	
Comp Mole Frac (Hydrogen)	***	***	***	***	
Comp Mole Frac (Oxygen)	***	***	***	***	
Comp Mole Frac (Nitrogen)	***	***	***	***	
Comp Mole Frac (CO2)	***	***	***	***	
Comp Mole Frac (CO)	***	***	***	***	
Comp Mole Frac (DTRM-A)	***	***	***	***	
Comp Mole Frac (Therminol-66)	***	***	***	***	
Comp Mole Frac (Ammonia)	***	***	***	***	
Comp Mole Frac (Argon)	***	***	***	***	

Energy Streams

Fluid Pkg: All

Name		Q-CW	Q-PIPE-801	Q-PIPE-901	Q-PIPE-902	W-P-801
Heat Flow	(MW)	62.80	1.150e-006	7.531e-003	8.384e-003	3.047e-003
Mass Flow	(kg/h)	---	---	---	---	---
Name		W-P-901	COL-DUTY @ASU	Q-CW-EXCH1 @ASU	Q-MOLSIEVE @ASU	Q_ACOMP-IC1 @ASU
Heat Flow	(MW)	6.445e-002	3.694	5.378e-002	-0.5917	2.247
Mass Flow	(kg/h)	---	---	3306	3.637e+004	1.381e+005
Name		Q_ACOMP-IC2 @ASU	Q_ACOMP-IC3 @ASU	W_A-EXPAND @ASU	W_ACOMP-2 @ASU	W_ACOMP-ST1 @AS
Heat Flow	(MW)	3.198	3.152	9.741e-002	5.317e-002	2.991
Mass Flow	(kg/h)	1.966e+005	1.938e+005	---	---	---
Name		W_ACOMP-ST2 @AS	W_ACOMP-ST3 @AS	Electrode Heat @Cell	Electrolysis Heating @	Electrolysis Power @C
Heat Flow	(MW)	3.183	3.126	-1.626e-005	24.82	-24.80
Mass Flow	(kg/h)	---	---	---	---	---
Name		Process Heat @Cell	COL-DUTY @COL5	COL-DUTY @COL6	Q_AIR-COMP @H2bu	Q_N2_COND @H2bu
Heat Flow	(MW)	2.755e-002	3.694	3.694	24.87	31.01
Mass Flow	(kg/h)	---	---	---	---	---
Name		Q_N2_LPS @H2burn	Q_N2_MPS @H2burn	W_AIR-COMP @H2bu	W_N2-PUMP @H2bur	Q-HX-301 @H2rec
Heat Flow	(MW)	34.74	98.03	30.21	8.284e-003	9.542e-005
Mass Flow	(kg/h)	---	---	---	---	5.866
Name		Q-HX-302 @H2rec	Q-HX-305 @H2rec	Q-HX-308 @H2rec	Q-IC-302 @H2rec	Q-IC-303 @H2rec
Heat Flow	(MW)	0.7115	0.2747	0.1975	0.1348	0.1246
Mass Flow	(kg/h)	4.374e+004	1.689e+004	1.214e+004	8286	7662
Name		Q-IC-301_stg1 @H2re	Q-IC-301_stg2 @H2re	Q-IC-301_stg3 @H2re	Q-IC-301_stg4 @H2re	Q-X-300 @H2rec
Heat Flow	(MW)	6.886e-007	6.846e-007	6.882e-007	6.864e-007	-4.093e-003
Mass Flow	(kg/h)	4.233e-002	4.208e-002	4.230e-002	4.219e-002	---
Name		W-K-301 @H2rec	W-K-302 @H2rec	W-K-303 @H2rec	W-K-301_stg1 @H2re	W-K-301_stg2 @H2re
Heat Flow	(MW)	9.210e-008	0.3254	0.2946	2.003e-008	1.136e-008
Mass Flow	(kg/h)	---	---	---	---	---
Name		W-K-301_stg3 @H2re	W-K-301_stg4 @H2re	W-K-301_stg5 @H2re	W-K-302_stg1 @H2re	W-K-302_stg2 @H2re
Heat Flow	(MW)	2.017e-008	2.024e-008	2.031e-008	0.1606	0.1648
Mass Flow	(kg/h)	---	---	---	---	---
Name		W-K-303_stg1 @H2re	W-K-303_stg2 @H2re	W-P-101 @H2rec	W-P-301 @H2rec	Electrolysis Power @H
Heat Flow	(MW)	0.1455	0.1491	1.463e-003	2.047e-004	-24.80
Mass Flow	(kg/h)	---	---	---	---	---
Name		Inverter Power @HTS	Process Heat @HTSE	Q-HX-107 @HTSE	Q-HX-502 @HTSE	Q-IC-501 @HTSE
Heat Flow	(MW)	-25.30	2.755e-002	0.5193	0.1003	0.2769
Mass Flow	(kg/h)	---	---	---	---	1.702e+004
Name		Q-NH3 HPS @HTSE	Q-NH3-LPS @HTSE	W-K-101 @HTSE	W-K-502 @HTSE	W-K-501_NET @HTS
Heat Flow	(MW)	1.761 *	0.3438 *	5.159e-002	0.1071	-9.231e-003
Mass Flow	(kg/h)	---	---	---	---	---

* Specified by user.

Workbook: Case (Main) (continued)

Energy Streams (continued)

Fluid Pkg: All

Name		W-K-501_stgC1 @HT	W-K-501_stgC2 @HT	W-K-501_stgT1 @HT	Q-IC-401_stg1 @Khpl	Q-IC-401_stg2 @Khpl
Heat Flow	(MW)	0.4142	0.4563	0.8798	0.6146	3.114
Mass Flow	(kg/h)	---	---	---	3.778e+004	1.914e+005
Name		Q-IC-401_stg3 @Khpl	W-K-401 @KhpH2	W-K-401_stg1 @KhpH	W-K-401_stg2 @KhpH	W-K-401_stg3 @KhpH
Heat Flow	(MW)	3.122	9.103	2.861	3.116	3.126
Mass Flow	(kg/h)	1.919e+005	---	---	---	---
Name		Q_REC-COMP-IC1 @	Q_REC-COMP-IC2 @	Q_REC-COMP-IC3 @	Q_REC-COMP-IC4 @	W_REC-COMP @Kre
Heat Flow	(MW)	0.0000 *	0.9395	1.241	9.671	4.217
Mass Flow	(kg/h)	0.0000	5.775e+004	7.629e+004	5.945e+005	---
Name		W_REC-COMP-ST1 @	W_REC-COMP-ST2 @	W_REC-COMP-ST3 @	W_REC-COMP-ST4 @	Q_RUC-INTC1 @KRU
Heat Flow	(MW)	0.8888	1.101	1.133	1.094	0.0000 *
Mass Flow	(kg/h)	---	---	---	---	0.0000
Name		Q_RUC-INTC2 @KRU	Q_RUC-INTC3 @KRU	Q_RUC-INTC4 @KRU	W_RU-Comp @KRU	W_RUC-STG1 @KRU
Heat Flow	(MW)	0.0000 *	---	---	---	---
Mass Flow	(kg/h)	0.0000	---	---	---	---
Name		W_RUC-STG2 @KRU	W_RUC-STG3 @KRU	W_RUC-STG4 @KRU	Q_SGC-IC1 @Ksyn	Q_SGC-IC2 @Ksyn
Heat Flow	(MW)	---	---	---	2.290	9.145
Mass Flow	(kg/h)	---	---	---	1.408e+005	5.622e+005
Name		Q_SGC-IC3 @Ksyn	W_SGC-STG1 @Ksyn	W_SGC-STG2 @Ksyn	W_SGC-STG3 @Ksyn	W_SYNGAS-COMP @
Heat Flow	(MW)	9.267	8.044	9.376	9.691	27.11
Mass Flow	(kg/h)	5.697e+005	---	---	---	---
Name		Q_MSC-INTC1 @MSC	Q_MSC-INTC2 @MSC	Q_MSC-INTC3 @MSC	Q_MSC-INTC4 @MSC	Q_MSC-INTC5 @MSC
Heat Flow	(MW)	2.762	2.916	2.934	2.970	3.040
Mass Flow	(kg/h)	1.698e+005	1.792e+005	1.803e+005	1.826e+005	---
Name		W_MSC-STG1 @MSC	W_MSC-STG2 @MSC	W_MSC-STG3 @MSC	W_MSC-STG4 @MSC	W_MSC-STG5 @MSC
Heat Flow	(MW)	2.866	2.901	2.904	2.911	2.926
Mass Flow	(kg/h)	---	---	---	---	---
Name		W_MSComp @MSC	Q_COOL-1 @NH3syn	Q_COOLER @NH3sy	Q_CW-EXCH @NH3s	Q_RU-COND @NH3s
Heat Flow	(MW)	14.51	-0.1009	14.08	37.19	---
Mass Flow	(kg/h)	---	---	8.658e+005	2.286e+006	---
Name		Q_STM-GEN @NH3s	Q_STM-GEN1 @NH3	Q_STM-GEN2 @NH3	Q_STM-GEN3 @NH3	Q_STRIP-COND @N
Heat Flow	(MW)	10.81	42.22	33.87	22.92	-10.62
Mass Flow	(kg/h)	1.772e+004	8.924e+004	7.160e+004	4.845e+004	---
Name		Q_STRIP-REB @NH3	Q_STRT-HTR @NH3s	R1-DUTY @NH3syn	R2-DUTY @NH3syn	RU-DUTY @NH3syn
Heat Flow	(MW)	43.65	0.0000 *	85.89	3.054	---
Mass Flow	(kg/h)	9.228e+004	0.0000	---	---	---
Name		RU-DUTY1 @NH3syn	RU-DUTY2A @NH3sy	W_BST-COMP @NH3	W_PUMP-1 @NH3syn	W_PUMP-2 @NH3syn
Heat Flow	(MW)	11.10	-13.11	1.598	4.398e-006	1.299e-002
Mass Flow	(kg/h)	2.979e+007	3.519e+007	---	---	---
Name		W_STG1 @NH3syn	Excess Electricity @N	Q-COND-100 @NPP	Q-FWH-700 @NPP	Q-PIPE-201 @NPP
Heat Flow	(MW)	-1.096e-005	---	3286	1108	7.490e-002
Mass Flow	(kg/h)	---	---	---	---	---
Name		Q-PIPE-202 @NPP	Q-SG-100 Thermal En	Rankine Energy @NP	W-K-700 Electricity Ge	W-P-201 Circ Pump P
Heat Flow	(MW)	5.313e-002	3368	3291	1124	0.1282
Mass Flow	(kg/h)	---	---	---	---	---
Name		W-P-701 Condensate	W-P-702 Feedwater P	Q-100 @NPP-2	Q-101 @NPP-2	Q-102 @NPP-2
Heat Flow	(MW)	7.421	8.569	170.9	182.1	1.674
Mass Flow	(kg/h)	---	---	---	---	---
Name		Q-103 @NPP-2	Q-104 @NPP-2	Q-105 @NPP-2	Q-106 @NPP-2	Q-107 @NPP-2
Heat Flow	(MW)	9.036	5.250	2169	245.3	190.4
Mass Flow	(kg/h)	---	---	---	---	---

* Specified by user.

Workbook: Case (Main) (continued)

Energy Streams (continued)

Fluid Pkg: All

Name		Q-108 @NPP-2	Q-109 @NPP-2	Q-110 @NPP-2	Q-111 @NPP-2	Q-TDL @NPP-2
Heat Flow	(MW)	70.16	81.29	202.2	3376	81.28
Mass Flow	(kg/h)	---	---	---	---	---

Unit Ops

Operation Name	Operation Type	Feeds	Products	Ignored	Calc Level
CW deltaT	Heater	903 Q-CW	904	No	500.0 *
Nuclear Power Plant (simple)	Standard Sub-Flowsheet	203 Process Heat Return	202 Process Heat Supply	No	3600 *
H2 burner	Standard Sub-Flowsheet	BURN H2 BURN AIR-1	BURN N2-5	No	2500 *
Air Separation Unit	Standard Sub-Flowsheet	AIR	N2 O2	No	2500 *
NH3 synthesis	Standard Sub-Flowsheet	SYN-IN	AN-NH3	No	2500 *
HP Product Compression	Standard Sub-Flowsheet	401	403 Pressurized H2 Product	No	2500 *
HTSE	Standard Sub-Flowsheet	202 Process Heat Supply 132 process feed water (liquid	203 Process Heat Return 302 H2/H2O for purification	No	2500 *
H2 Recovery and Feed Condi	Standard Sub-Flowsheet	101 Process Water Inlet 302 H2/H2O for purification	132 process feed water (liquid 332 H2 Product	No	2500 *
Nuclear Power Plant (detailed	Standard Sub-Flowsheet			No	2500 *
Multi-Stage Compressor	Standard Sub-Flowsheet	N2	N2-2-	No	2500 *
ADJ CW Pump dP	Adjust			No	3500 *
ADJ FW pump dP	Adjust			No	3500 *
ADJ ASU inlet flow rate	Adjust			No	3500 *
ADJ combustor inlet flow rate	Adjust			No	3500 *
HTE Calculations	Spreadsheet			No	500.0 *
Cooling Water	Spreadsheet			No	500.0 *
TDL	Spreadsheet			No	500.0 *
Water Bal	Spreadsheet			No	500.0 *
HX dP	Spreadsheet			No	500.0 *
NH3 heat use	Spreadsheet			No	500.0 *
Anhydrous NH3 production ra	Spreadsheet			No	500.0 *
PIPE-801	Pipe Segment	802	803 Q-PIPE-801	No	500.0 *
PIPE-901	Pipe Segment	902	903 Q-PIPE-901	No	500.0 *
PIPE-902	Pipe Segment	904	905 Q-PIPE-902	No	500.0 *
P-801	Pump	801 feed water W-P-801	802	No	500.0 *
P-901	Pump	901 cooling water W-P-901	902	No	500.0 *
SET number of HTSE blocks	Set			No	500.0 *
SET dP CW delta T	Set			No	500.0 *
K-400 T-P-c	Virtual Stream Extn v2.0.0	332 H2 Product	401	No	500.0 *
SG-MIX	Mixer	403 Pressurized H2 Product N2-2+	SYN-IN	No	500.0 *
N2-2-Cutter	Stream Cutter	N2-2-	N2-2+	No	500.0 *

* Specified by user.

Appendix F
Market Analyses

APPENDIX F
MARKET ANALYSES

F-1 Market and Demand Points in the Minnesota Region

Table F-1. Hydrogen demand within 100 miles of the Prairie Island Nuclear Power Plant.

Name	Demand Type	Potential H_2 Demand, kilotonnes/year Current (2017)	Future (2030)	Distance, miles
Red Wing: Northern States Power Co., Minnesota	Natural gas electricity generators	0.02	0.02	13
Goodhue County, Minnesota	FCEV	–	0.00	17
Cannon Falls Energy Center: Invenergy Services, LLC	Natural gas electricity generators	0.35	0.35	20
LSP-Cottage Grove Limited Partnership (LP): Cottage Grove Operating Services, LLC	Natural gas electricity generators	1.65	1.65	23
Inver Hills: Northern States Power Co, Minnesota	Natural gas electricity generators	0.09	0.09	26
Koch Industries, Inc., Flint Hills Resources Pine Bend Refinery	Syngas: Hydrogen, SMR*	–	158.12	27
Marathon Petroleum Corp. St. Paul Park Refinery	Refinery	28.90	36.61	27
Koch Industries Inc., Flint Hills Resources Pine Bend Refinery	Refinery	85.06	107.76	27
Marathon Petroleum Corp. St. Paul Park Refinery	Syngas: Hydrogen, SMR*	–	15.13	27
Gerdau Long Steel North America, St. Paul, Minnesota	DRI	0.40	1.42	30
Pierce County, Wisconsin	FCEV	–	0.00	32
Dakota County, Minnesota	FCEV	–	0.01	33
St. Paul Cogeneration: St. Paul Cogeneration, LLC	Natural gas electricity generators	0.38	0.38	38
High Bridge: Northern States Power Co, Minnesota	Natural gas electricity generators	10.19	10.19	39
Washington County, Minnesota	FCEV	–	0.00	40
Allen S King: Northern States Power Co, Minnesota	Natural gas electricity generators	0.04	0.04	40
Ramsey County, Minnesota	FCEV	–	0.01	40
Wabasha County, Minnesota	FCEV	–	0.00	40

Name	Demand Type	Potential H_2 Demand, kilotonnes/year Current (2017)	Future (2030)	Distance, miles
Rice County, Minnesota	FCEV	–	0.00	41
University of Minnesota combined heat and power (CHP) Plant: Veolia Energy	Natural gas electricity generators	0.83	0.83	45
Southeast Steam Plant: Veolia Energy	Natural gas electricity generators	0.38	0.38	46
St. Croix County, Wisconsin	FCEV	–	0.00	48
Scott County, Minnesota	FCEV	–	0.00	48
Faribault Energy Park: Minnesota Municipal Power Agency	Natural gas electricity generators	3.41	3.41	48
Covanta Hennepin Energy: Covanta Energy Co	Natural gas electricity generators	0.03	0.03	49
Saint Marys Hospital Power Plant: St Mary's Hospital	Natural gas electricity generators	0.68	0.68	50
New Prague: New Prague Utilities Commission	Natural gas electricity generators	0.01	0.01	51
Hennepin County, Minnesota	FCEV	–	0.02	52
Water Reclamation Plant: City of Rochester	Natural gas electricity generators	0.00	0.00	52
Cascade Creek: Rochester Public Utilities	Natural gas electricity generators	0.21	0.21	53
Dodge County, Minnesota	FCEV	–	0.00	53
Riverside (Minnesota): Northern States Power Co, Minnesota	Natural gas electricity generators	8.48	8.48	53
Olmsted County, Minnesota	FCEV	–	0.00	54
Franklin Heating Station: Franklin Heating Station	Natural gas electricity generators	0.86	0.86	55

Name	Demand Type	Potential H_2 Demand, kilotonnes/year Current (2017)	Potential H_2 Demand, kilotonnes/year Future (2030)	Distance, miles
Pepin County, Wisconsin	FCEV	–	0.00	55
Blue Lake: Northern States Power Co, Minnesota	Natural gas electricity generators	0.65	0.65	56
Olmsted Waste Energy: Olmsted County Public Works	Natural gas electricity generators	0.01	0.01	57
Shakopee Energy Park: Minnesota Municipal Power Agency	Natural gas electricity generators	0.06	0.06	58
Al-Corn Clean Fuel, Claremont	Syngas: ethanol	–	20.00	59
Anoka County, Minnesota	FCEV	–	0.00	59
Koda Biomass Plant: Koda Energy LLC	Natural gas electricity generators	0.16	0.16	63
Minnesota River: Minnesota Municipal Power Agency	Natural gas electricity generators	0.00	0.00	63
Buffalo County, Wisconsin	FCEV	–	0.00	64
Steele County, Minnesota	FCEV	–	0.00	67
Carver County, Minnesota	FCEV	–	0.00	68
Le Sueur County, Minnesota	FCEV	–	0.00	69
Owatonna: City of Owatonna, (Minnesota)	Natural gas electricity generators	0.01	0.01	69
Big River Resources Boyceville LLC, Boyceville	Syngas: ethanol	–	20.00	73
Elk River: Great River Energy	Natural gas electricity generators	0.24	0.24	74
Dunn County, Wisconsin	FCEV	–	0.00	74
Winona County, Minnesota	FCEV	–	0.00	75

Name	Demand Type	Potential H_2 Demand, kilotonnes/year		Distance, miles
		Current (2017)	Future (2030)	
Chisago County, Minnesota	FCEV	–	0.00	76
Polk County, Wisconsin	FCEV	–	0.00	77
Pleasant Valley (Minnesota): Great River Energy	Natural gas electricity generators	0.42	0.42	78
Waseca County, Minnesota	FCEV	–	0.00	78
Janesville: City of Janesville, (Minnesota)	Natural gas electricity generators	0.00	0.00	80
Spring Valley: Spring Valley Pub Utils Commission	Natural gas electricity generators	0.00	0.00	80
Guardian Energy LLC, Janesville	Syngas: Ethanol	–	60.00	81
Arcadia Electric: City of Arcadia, (Wisconsin)	Natural gas electricity generators	0.00	0.00	83
Eau Claire County, Wisconsin	FCEV	–	0.00	84
Sibley County, Minnesota	FCEV	–	0.00	85
Isanti County, Minnesota	FCEV	–	0.00	85
Wright County, Minnesota	FCEV	–	0.00	86
Cambridge CT: Great River Energy	Natural gas electricity generators	0.20	0.20	86
Elk Mound: Dairyland Power Coop	Natural gas electricity generators	0.14	0.14	87
Sherburne County, Minnesota	FCEV	–	0.00	87
Nicollet County, Minnesota	FCEV	–	0.00	88
Fillmore County, Minnesota	FCEV	–	0.00	89

Name	Demand Type	Potential H_2 Demand, kilotonnes/year Current (2017)	Future (2030)	Distance, miles
Wheaton: Northern States Power Co, Minnesota	Natural gas electricity generators	0.80	0.80	90
Preston (Minnesota): Preston Public Utilities Commission	Natural gas electricity generators	0.00	0.00	90
Heartland Corn Products, Winthrop	Syngas: ethanol	–	30.00	91
Mower County, Minnesota	FCEV	–	0.00	91
Glencoe: Glencoe Light & Power Commission	Natural gas electricity generators	0.00	0.00	91
Wilmarth: Northern States Power Co, Minnesota	Natural gas electricity generators	0.02	0.02	94
Mankato Energy Center: Southern Power Co	Natural gas electricity generators	3.80	3.80	94
Pro Corn LLC, Preston	Syngas: ethanol	–	20.00	94
Blue Earth County, Minnesota	FCEV	–	0.00	96
Trempealeau County, Wisconsin	FCEV	–	0.00	96
Cumberland (Wisconsin): City of Cumberland, (Wisconsin)	Natural gas electricity generators	0.00	0.00	98
Hutchinson Plant #2: Hutchinson Utilities Commission	Natural gas electricity generators	0.14	0.14	99
Hutchinson Plant #1: Hutchinson Utilities Commission	Natural gas electricity generators	0.06	0.06	100
Freeborn County, Minnesota	FCEV	–	0.00	100

*H_2 demand for synfuel production using CO_2 from SMR units in the refineries is exclusive of H_2 demand by these refineries

Table F-2. Hydrogen demand within 100 miles of the Monticello Nuclear Power Plant.

Name	Demand Type	Potential H_2 Demand, kilotonnes/year Current (2017)	Potential H_2 Demand, kilotonnes/year Future (2030)	Distance (miles)
Distance Sherburne County, Minnesota	FCEV	0.00	0.00	12
Wright County, Minnesota	FCEV	0.00	0.00	13
Elk River: Great River Energy	Natural gas electricity generators	0.24	0.24	17
Granite City: Northern States Power Co–Minnesota	Natural gas electricity generators	0.00	0.00	27
Stearns County, Minnesota	FCEV	0.00	0.00	32
Benton County, Minnesota	FCEV	0.00	0.00	32
Anoka County, Minnesota	FCEV	0.00	0.00	34
Hennepin County, Minnesota	FCEV	0.00	0.02	39
Riverside, Minnesota: Northern States Power Co., Minnesota	Natural gas electricity generators	8.48	8.48	39
Covanta Hennepin Energy: Covanta Energy Co	Natural gas electricity generators	0.03	0.03	42
Southeast Steam Plant: Veolia Energy	Natural gas electricity generators	0.38	0.38	43
Mille Lacs County, Minnesota	FCEV	0.00	0.00	45
Isanti County, Minnesota	FCEV	0.00	0.00	46
Univ Minnesota CHP plant, Veolia Energy	Natural gas electricity generators	0.83	0.83	46
Meeker County, Minnesota	FCEV	0.00	0.00	49
Ramsey County, Minnesota	FCEV	0.00	0.01	49
Litchfield: Litchfield Public Utilities	Natural gas electricity generators	0.00	0.00	50
Blue Lake: Northern States Power Co., Minnesota	Natural gas electricity generators	0.65	0.65	52
Cambridge CT: Great River Energy	Natural gas electricity generators	0.20	0.20	53
St. Paul Cogeneration: St. Paul Cogeneration, LLC	Natural gas electricity generators	0.38	0.38	53
Koda Biomass Plant: Koda Energy, LLC	Natural gas electricity generators	0.16	0.16	54
Shakopee Energy Park: Minnesota Municipal Power Agency	Natural gas electricity generators	0.06	0.06	54
Minnesota River: Minnesota Municipal Power Agency	Natural gas electricity generators	0.00	0.00	54
McLeod County, Minnesota	FCEV	0.00	0.00	55
Glencoe: Glencoe Light & Power Commission	Natural gas electricity generators	0.00	0.00	55

Name	Demand Type	Potential H_2 Demand, kilotonnes/year Current (2017)	Future (2030)	Distance (miles)
High Bridge: Northern States Power Co–Minnesota	Natural gas electricity generators	10.19	10.19	55
Hutchinson Plant #2: Hutchinson Utilities Commission	Natural gas electricity generators	0.14	0.14	57
Washington County, Minnesota	FCEV	0.00	0.00	57
Black Dog: Northern States Power Co–Minnesota	Natural gas electricity generators	3.68	3.68	57
Scott County, Minnesota	FCEV	0.00	0.00	58
Hutchinson Plant #1: Hutchinson Utilities Commission	Natural gas electricity generators	0.06	0.06	58
Morrison County, Minnesota	FCEV	0.00	0.00	59
Gerdau Long Steel North America–St. Paul	DRI	0.00	1.00	59
Carver County, Minnesota	FCEV	0.00	0.00	59
Allen S King: Northern States Power Co–Minnesota	Natural gas electricity generators	0.04	0.04	62
Marathon Petroleum Corp. St. Paul Park Refinery	Refinery	29.00	37.00	62
Mora: City of Mora–(Minnesota)	Natural gas electricity generators	0.00	0.00	62
Marathon Petroleum Corp. St. Paul Park Refinery	Syngas: Hydrogen, SMR*	–	15.00	62
Kanabec County, Minnesota	FCEV	0.00	0.00	63
Bushmills Ethanol Inc, Atwater	Syngas: Ethanol	–	30.00	64
Dakota County, Minnesota	FCEV	0.00	0.01	64
Inver Hills: Northern States Power Co–Minnesota	Natural gas electricity generators	0.09	0.09	65
Koch Industries Inc Flint Hills Resources Pine Bend Refinery	Syngas: Hydrogen, SMR*	–	158.00	65
Koch Industries Inc Flint Hills Resources Pine Bend Refinery	Refinery	85.00	108.00	66
LSP-Cottage Grove LP: Cottage Grove Operating Services LLC	Natural gas electricity generators	1.65	1.65	67
Sibley County, Minnesota	FCEV	0.00	0.00	68
Chisago County, Minnesota	FCEV	0.00	0.00	70
St. Croix County, Wisconsin	FCEV	0.00	0.00	74
New Prague: New Prague Utilities Commission	Natural gas electricity generators	0.01	0.01	77
Heartland Corn Products, Winthrop	Syngas: Ethanol	–	30.00	78
Kandiyohi County, Minnesota	FCEV	0.00	0.00	80

Name	Demand Type	Potential H_2 Demand, kilotonnes/year Current (2017)	Future (2030)	Distance (miles)
Willmar: Willmar Municipal Utilities	Natural gas electricity generators	0.06	0.06	82
Cannon Falls Energy Center: Invenergy Services LLC	Natural gas electricity generators	0.35	0.35	84
Pine County, Minnesota	FCEV	0.00	0.00	86
Rice County, Minnesota	FCEV	0.00	0.00	87
Pierce County, Wisconsin	FCEV	0.00	0.00	88
Polk County, Wisconsin	FCEV	0.00	0.00	88
Faribault Energy Park: Minnesota Municipal Power Agency	Natural gas electricity generators	3.41	3.41	89
Todd County, Minnesota	FCEV	0.00	0.00	90
Le Sueur County, Minnesota	FCEV	0.00	0.00	90
Pope County, Minnesota	FCEV	0.00	0.00	92
Douglas County, Minnesota	FCEV	0.00	0.00	95
Crow Wing County, Minnesota	FCEV	0.00	0.00	95
Renville County, Minnesota	FCEV	0.00	0.00	96
Red Wing: Northern States Power Co, Minnesota	Natural gas electricity generators	0.02	0.02	97

*H_2 demand for synfuel production using CO_2 from SMR units in the refineries is exclusive of H_2 demand by these refineries

Table F-3. Hydrogen demand locations covered for transportation near Xcel Energy’s Prairie Island Nuclear Power Plant

Labels/ demand points	Name	Demand Type	Potential Demand, kilotonnes	Future Potential Demand, kilotonnes
1	Koch Industries Inc Flint Hills Resources Pine Bend Refinery	Refinery	85.06	107.76
2	Marathon Petroleum Corp. St. Paul Park Refinery	Refinery	28.90	36.61
3	LSP-Cottage Grove LP: Cottage Grove Operating Services LLC	Natural gas electricity generators	1.65	1.65
4	Gerdau Long Steel North America–St. Paul	DRI	0.40	1.42
5	High Bridge: Northern States Power Co–Minnesota	Natural gas electricity generators	10.19	10.19
6	St. Paul Cogeneration: St. Paul Cogeneration LLC	Natural gas electricity generators	0.38	0.38
7	Univ Minnesota CHP Plant: Veolia Energy	Natural gas electricity generators	0.83	0.83
8	Southeast Steam Plant: Veolia Energy	Natural gas electricity generators	0.38	0.38
9	Covanta Hennepin Energy: Covanta Energy Co	Natural gas electricity generators	0.03	0.03
10	Riverside (Minnesota): Northern States Power Co–Minnesota	Natural gas electricity generators	8.48	8.48
11	Black Dog: Northern States Power Co–Minnesota	Natural gas electricity generators	3.68	3.68
12	Blue Lake: Northern States Power Co–Minnesota	Natural gas electricity generators	0.65	0.65
13	Shakopee Energy Park: Minnesota Municipal Power Agency	Natural gas electricity generators	0.06	0.06
14	Koda Biomass Plant: Koda Energy LLC	Natural gas electricity generators	0.16	0.16
15	Minnesota River: Minnesota Municipal Power Agency	Natural gas electricity generators	0.00	0.00

Table F-4. Overlapping hydrogen demand within 50 miles of Xcel Energy's Prairie Island and Monticello Nuclear Power Plants.
*H_2 demand for synfuel production using CO_2 from SMR units in the refineries is exclusive of H_2 demand by these refineries

		Potential H_2 Demand, kilotonnes		Prairie Island	Monticello
Name	Demand Type	Current (2017)	Future (2030)	Distance, miles	Distance, miles
Red Wing: Northern States Power Co–Minnesota	Natural gas electricity generators	0.02	0.02	13	97
Cannon Falls Energy Center: Invenergy Services LLC	Natural gas electricity generators	0.35	0.35	20	84
LSP-Cottage Grove LP: Cottage Grove Operating Services LLC	Natural gas electricity generators	1.65	1.65	23	67
Inver Hills: Northern States Power Co–Minnesota	Natural gas electricity generators	0.09	0.09	26	65
Koch Industries Inc Flint Hills Resources Pine Bend Refinery	Synfuels: Hydrogen, SMR*	–	158	27	65
Marathon Petroleum Corp. St. Paul Park Refinery	Refinery	29	37	27	62
Koch Industries Inc Flint Hills Resources Pine Bend Refinery	Refinery	85	108	27	66
Marathon Petroleum Corp. St. Paul Park Refinery	Synfuels Hydrogen, SMR*	–	15	27	62
Gerdau Long Steel North America–St. Paul	DRI	0	1	30	59
St. Paul Cogeneration: St. Paul Cogeneration LLC	Natural gas electricity generators	0.38	0.38	38	53
High Bridge: Northern States Power Co–Minnesota	Natural gas electricity generators	10.19	10.19	39	55
Allen S King: Northern States Power Co–Minnesota	Natural gas electricity generators	0.04	0.04	40	62
Univ Minnesota CHP Plant: Veolia Energy	Natural gas electricity generators	0.83	0.83	45	46
Southeast Steam Plant: Veolia Energy	Natural gas electricity generators	0.38	0.38	46	43
Faribault Energy Park: Minnesota Municipal Power Agency	Natural gas electricity generators	3.41	3.41	48	89
Covanta Hennepin Energy: Covanta Energy Co	Natural gas electricity generators	0.03	0.03	49	42
Saint Marys Hospital Power Plant: St Mary's Hospital	Natural gas electricity generators	0.68	0.68	50	

Table F-5. Hydrogen demand locations covered for transportation near Xcel Energy’s Monticello NPP.

Labels/ demand points	Name	Demand Type	Potential Demand, kilotonnes	Future Potential Demand, kilotonnes
2	Hutchinson Plant #1: Hutchinson Utilities Commission	Natural gas electricity generators	0.06	0.06
3	Hutchinson Plant #2: Hutchinson Utilities Commission	Natural gas electricity generators	0.14	0.14
4	Bushmills Ethanol Inc, Atwater	Syngas: Ethanol	–	30
5	Heartland Corn Products, Winthrop	Syngas: Ethanol	–	30
6	Willmar: Willmar Municipal Utilities	Natural gas electricity generators	0.06	0.06