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High-Temperature Steam Electrolysis Process Performance and Cost Estimates

March 2022

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EXECUTIVE SUMMARY

Technology readiness levels (TRLs) of water splitting electrolysis systems have dramatically increased in recent years as interest in clean hydrogen production and decarbonization of transportation, industry, and other sectors increases around the globe. High-temperature steam electrolysis (HTSE) systems can theoretically achieve relatively high overall system efficiencies. Over three decades of research and testing has been devoted to the discovery and development of oxygen ion ($O^{=}$)-conducting electrolysis cells. Several commercial vendors in the United States (U.S.) and Europe have now developed commercial solid oxide electrolysis cell (SOEC) stacks and have determined optimal operating current/voltage, temperatures, and pressures to maximize the stack thermodynamic efficiencies.

Recent testing of commercially manufactured SOEC stacks and thermally integrated systems with heat recuperation have confirmed the high theoretical efficiencies of HTSE can be achieved. With heat recuperation from the product hydrogen- and oxygen-bearing streams, any source of dry steam (i.e., steam above the saturation temperature) can be raised to the operating temperature of the electrolysis cell with a minimal percentage of electrical topping heat. Therefore, any source of heat that can produce steam, including electrical heating, can be used to convert de-ionized water into pure steam for electrolysis. This has incentivized the nuclear utilities and developers of advanced nuclear reactors to consider the use of the steam and electricity that is readily produced by nuclear power plants (NPPs) for hydrogen production.

Idaho National Laboratory (INL) has been very involved in collaborative materials research and modeling of HTSE components and systems for several years. Process modeling on a large variety of projects has led to foundational knowledge supporting technoeconomic assessments (TEAs) of hydrogen production plants. Several INL evaluations have been performed on behalf of commercial vendors under U.S. Department of Energy (DOE) Strategic Partnership Projects. In some cases, public versions of detailed, proprietary, investor-grade reports have been issued. INL has used the publicly accessible reports to create a HTSE baseline plant to help address gaps in overall HTSE systems. The INL reference design and cost analysis can be rationally used to determine steps to reducing levelized cost of hydrogen (LCOH) to disruptive levels that will meet the DOE cost target of producing hydrogen for less than $2/kilogram (kg) by 2025, and approaching the Earthshot goal of $1/kg by 2030.

This baseline design and cost analysis provides INL's best estimate of the operational parameters and costs for an n^{th}-of-a-kind (NOAK) HTSE plant in 2020 U.S. dollars (USD). Proprietary data are not disclosed in this report and INL acknowledges some commercial HTSE suppliers will have performance specifications, cost estimates, and test data that differ from the analysis presented in this document. Whereas this document is meant to be at best a conservative estimate of the technology and not an absolute reference, the intent is to periodically update the reference plant design, operating parameters, and capital and operating costs as these continue to be made available by vendors or by independent cost analysis efforts. For example, an ongoing effort by Strategic Analysis Inc. may corroborate or alter some of the assumptions and outcomes provided in this report.

A recent public release of a report for a nuclear utility lists many of the latest updates to the INL HTSE TEA methodology [1]. In that study, two different case studies were presented—a Base Case and an Advanced Case. Recent developments have resulted in slight modifications of the Base Case, which is presented in this summary. The following is a list of these updates, as well as their justifications:

- Stack costs were updated from $155/kilowatt, direct current (kW-dc) to $78/kW-dc. The updated value corresponds to the cost computed in a recent analysis using a Design for Manufacturing and Assembly (DFMA) bottom-up manufacturing cost analysis approach for an electrode-supported stack using standard manufacturing processes with a manufacturing rate of 1000 megawatt (MW)/year [2].

- Balance-of-plant (BoP) equipment component cost estimates provided using Aspen Process Economic Analyzer (APEA) software [3] were updated with the following material selection: SS304 is specified for components with operating temperature T<600°C, while SS347 is specified for components with an operating temperature T>600°C. Previously low alloy steel was specified for several low-temperature process components. Use of stainless steel throughout the process provides a more robust design.
- High-pressure compression of hydrogen and long-distance transport costs were excluded. This was done to separate the costs of the primary HTSE process components from the costs of the scenario-specific equipment, which enables a more direct comparison with other hydrogen production technologies.
- An updated set of financial input value specifications were selected to be representative of nuclear-integrated hydrogen production in an industrial application. The following parameters were specified with values that differ from those in the Current Central Solid Oxide Electrolysis H2A V3.2018 Case Study [4]: Reference Year updated to 2020 from 2016; Start-up Year updated to 2025 from 2015; Plant Life and Analysis period updated to 20 years from 40 years (20 years is representative of the duration of a light water reactor (LWR) NPP operating license extension); Debt Period updated to 20 years from Constant Debt; and After-Tax Real internal rate of return (IRR) updated to 10% from 8%.
- Grid-integrated analysis was excluded such that this current report considers only constant hydrogen production where the HTSE plant sources the entirety of the available NPP power for its operation (i.e., no dispatching of nuclear power between the HTSE plant and the grid is assumed).

Unless otherwise noted, all cost information for the analyses in this report was calculated in 2020 USD. It is recognized that inflationary pressures have recently caused a dramatic increase in costs of many industrial materials; however, it should be noted that this report does not attempt to quantify recent price inflation due to global factors as a result of the COVID pandemic. Any changes in equipment cost must be considered as part of future project feasibility studies. This report deals only with steady-state process design and the associated economic analysis; the cost estimate does not include hydrogen storage required to provide a steady output of hydrogen while operating the electrolysis plant in a transient operating mode (i.e., the electrolysis plant is not operated in a standby mode since the NPP is not assumed to divert electrical power to the grid during periods of peak electricity demand in this study).

For project study purposes, a gigawatt-scale LWR-HTSE process design model was developed and used to evaluate basic steady-state constant hydrogen production scenarios, where the full LWR capacity is dedicated to the HTSE plant. The integration with an LWR plant included basic cost estimates for nuclear process heat delivery equipment such as piping and heat exchangers (HXs) to transfer thermal energy to the HTSE process for use in vaporizing the process feedwater. No allowance was made to estimate costs of equipment modifications inside the nuclear plant boundary, specific nuclear permitting, nuclear code compliance, etc. The evaluation determined that an HTSE process utilizing all energy output from an LWR NPP would require approximately 5% of the LWR total steam flow to provide the process heat input needed to vaporize the HTSE process feedwater, while the balance of the LWR steam flow would continue to be used to drive the steam turbines/generator that produce the electrical power used to meet the HTSE plant (including BoP) process electrical power demands. The analysis specified the use of Therminol-66 as the heat transfer fluid (HTF) to transfer nuclear process heat an assumed distance of 1 km to the HTSE plant, which was determined to have **specific electricity and thermal energy requirements of 36.8 kWh-e/kg-H_2 (kilowatt hour electric per kg-H_2) and 6.4 kWh-t/kg-H_2 (kilowatt hour thermal per kg H_2)**, respectively. The HTSE plant system **efficiency was calculated as 90.2%** on a higher heating value (HHV) basis. These efficiency metrics are based on the energy requirements to produce a 20 bar hydrogen product (energy requirements for high-pressure compression are excluded).

In contrast to the 2021 INL HTSE process performance and cost analysis mentioned above, in which multiple SOEC technology cases were considered, the current study focuses on a single SOEC technology case and utilizes sensitivity analysis to determine the impact of selected parameters on the LCOH. To achieve NOAK plant status, manufacturing capacity sufficient to have deployed 2,500 megawatt electrical (MWe) of previous HTSE capacity (N = 100 count of 25 MWe modular HTSE blocks) is assumed to be available to support plant start-up in the year 2025. The analysis maintains the 4-year stack service life specification from the DOE Hydrogen and Fuel Cell Technologies Office (HFTO) SOEC Hydrogen Production Record Current Technology Case [5]. 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 design capacity of 702 tonne H_2/day** (1000 MW-dc stack power input; 1076 megawatt alternating current [MW-ac] total electrical power demand) has a direct capital cost (DCC) of $544/kW-dc (includes HTSE plant equipment BoP, alternating current [AC] to direct current [DC] conversion inefficiencies, and nuclear plant heat and power delivery equipment) and **a total capital investment (TCI) of $703/kW-dc** (includes project indirect costs in addition to the direct cost listed above). Assuming energy from the LWR for the HTSE system 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 can **produce hydrogen at a LCOH of $1.86/kg**, which does not include product high-pressure compression, storage, or transportation costs, as mentioned. A summary of the assumptions and results for the base HTSE analysis is shown below in Table ES-1.

Table ES-1. Summary of the Base HTSE Model Design Case.

Parameter	**Value**	**Notes**
Plant Design Capacity	702 metric tonnes hydrogen per day	
Power Requirement	1000 MW-dc 1076 MW-ac	-DC power corresponds to stack power input; -AC power corresponds to total power requirement (BoP, AC power to rectifier, pumps, compressors, topping heaters, etc.)
Thermal Requirement	188.2 MW-t	Provided directly by nuclear process heat
Efficiency (HHV)	90.2%	Includes both thermal and electrical energy consumption
Stack Pressure	5 bar	Based on maximizing system efficiency by trending operating pressure and steam utilization versus system efficiency
Steam Utilization (conversion of reactant steam)	80%	Based on maximizing system efficiency by trending operating pressure and steam utilization versus system efficiency
H_2 Product Pressure	20 bar	
H_2 Product Purity	99.9 mol% H_2	Water condensation from cooling and compression only; no PSA / TSA steps included
Electricity Required	36.8 kWh-e/kg-H_2	Process model result
Thermal Energy Required	6.4 kWh-t/kg-H_2	Process model result
Technology Horizon	NOAK, 95% learning rate	95% corresponds to a 5% cost reduction with every doubling of the number of units produced
Stack Cost	$78/kW-dc	Value reported from DFMA analysis of an electrode-supported cell stack with 1,000 MW/year manufacturing rate [2]
Service Life	4 years	Assumes annual stack replacements to restore the HTSE plant design capacity rating at the start of each operating year; consistent with the Current Technology Case in [5]

Parameter	Value	Notes
Direct Capital Cost	\$544/kW-dc	Includes the capital cost of the nuclear process heat delivery system; for HTSE applications not including nuclear process heat input, the CAPEX would be reduced accordingly
Total Capital Investment	\$703/kW-dc	Includes indirect costs (site preparation, engineering & design, contingency, land, etc.)
Levelized Cost of H_2 (LCOH for HTSE)	\$1.86/kg	At \$30/megawatt hour (MWh) electricity cost; Excludes storage and transport costs

Figure ES-1 below shows the capital cost contributions of the various components of the HTSE system evaluated, including direct and indirect costs. Figure ES-2 shows the relative contributions of individual cost components to the levelized cost of hydrogen (LCOH). As shown in Figure ES-2, the cost of electricity to power the electrolyzer is the most significant cost of production. The balance of plant system is the second most significant contribution. Design parameters including operating pressure, steam utilization, module size/capacity, product output purity/pressure, etc. impact process efficiency and capital costs. INL is currently collaborating in ongoing efforts to evaluate these and other system design parameters to identify the economic optimum process design specifications.

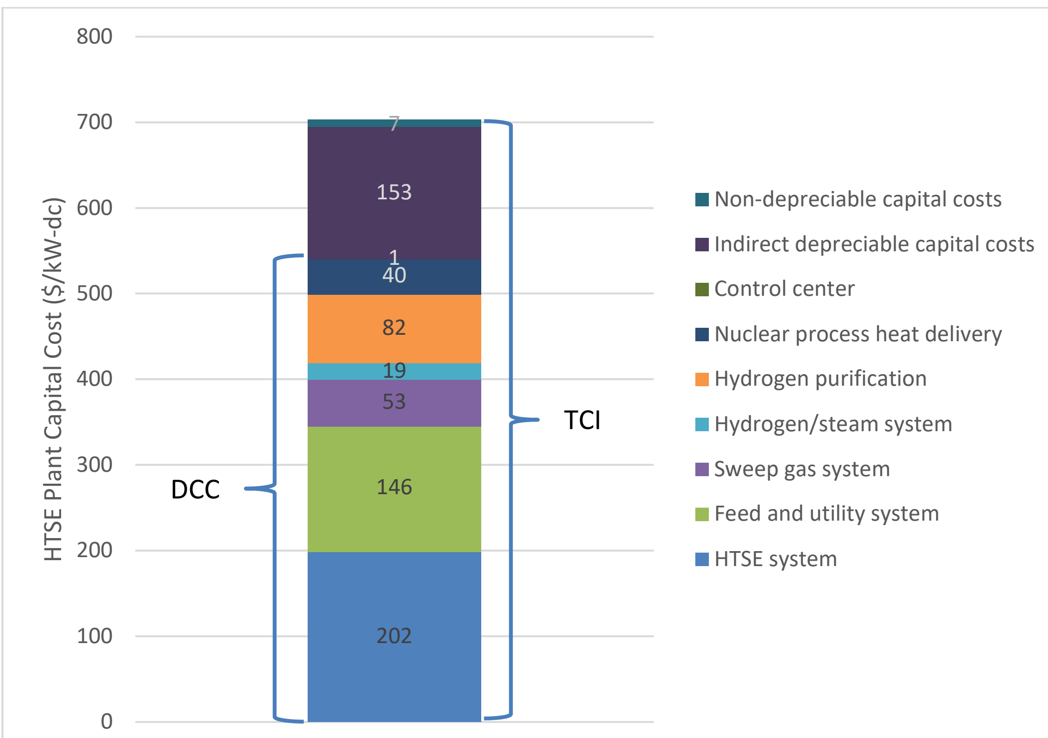

Figure ES-1. Estimated CAPEX distribution for a GW-scale, NOAK nuclear-integrated HTSE plant. Direct capital costs (DCC) and Total Capital Investment (TCI) contributions are identified in the annotation.

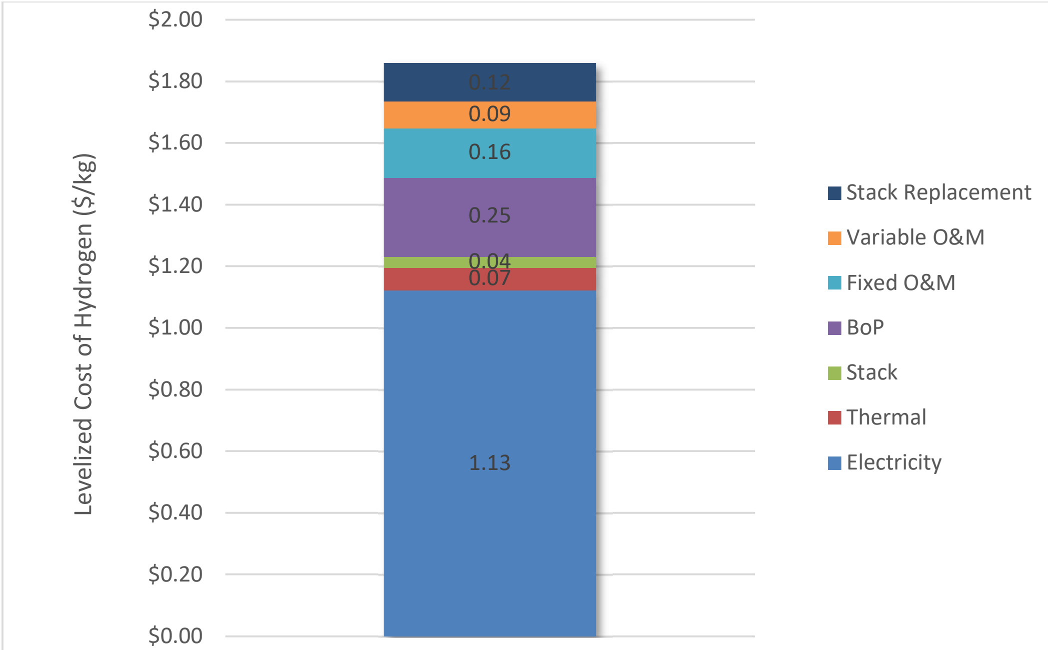

Figure ES-2. LCOH cost components for an NOAK constant hydrogen production LWR-HTSE system configuration with 611 tonnes per day actual hydrogen production rate (702 tonnes/day design production capacity), a direct CAPEX of $544/kW-dc, and an electricity price of $30/MWh-e.

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 encompass the possible range of expected technology advancements and/or variations in market conditions. Results of the sensitivity analysis are shown in Figure ES-3.

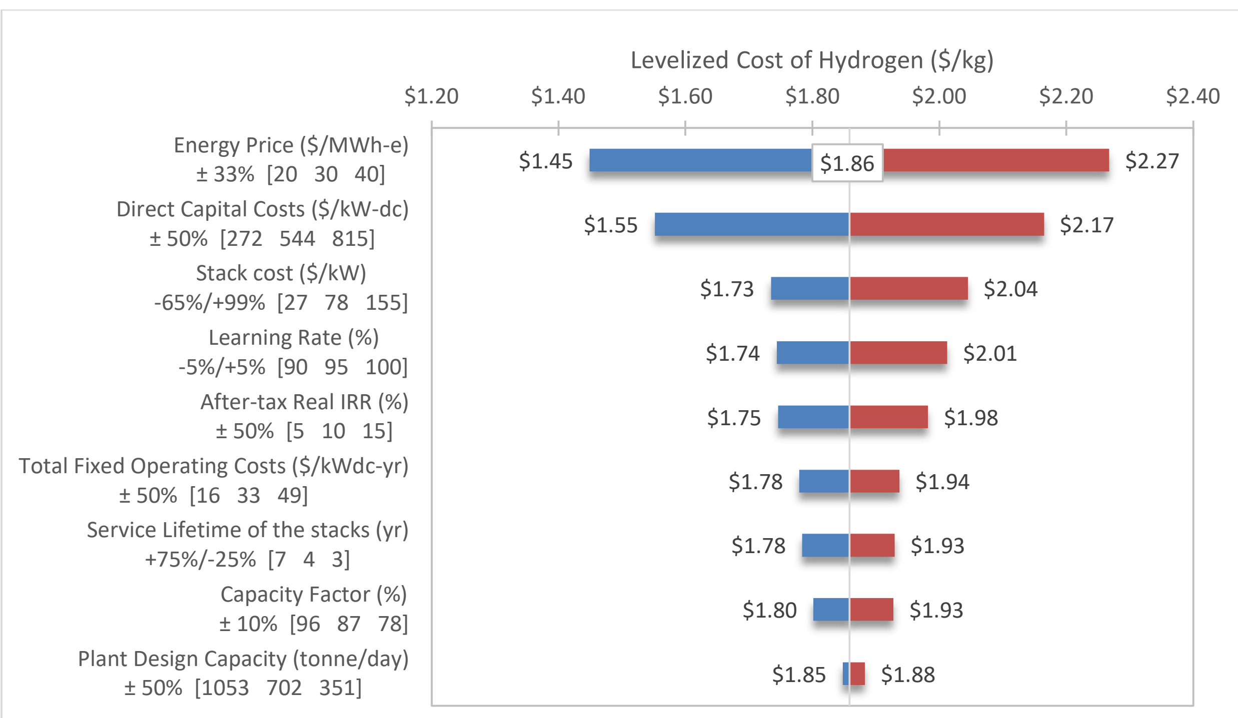

Figure ES-3. Sensitivity of LCOH to selected constant hydrogen production case input parameters.

Based on the selected range over which the sensitivity variables were perturbed, the parameters that have the greatest impact on LCOH are electricity energy price and the direct capital cost. A decrease in the electricity price from \$30/MWh-e to \$20/MWh-e results in an LCOH decrease of \$0.41/kg-H_2. A 50% decrease in the direct capital costs results in a decrease of \$0.31/kg in the LCOH. A second set of variables including the stack cost, after-tax real IRR, and learning rate (for decreases in modular equipment costs as a function of the number of units produced by the equipment manufacturer) have a medium impact on the LCOH. Additional results and observations from the sensitivity analysis are listed below:

- Stack costs are a significant driver of the HTSE LCOH. The stack costs contribute to the initial plant construction costs as well as the HTSE plant operations and maintenance (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 [5] versus the stack cost that specific SOEC vendors have reported would be possible using current technology with a manufacturing capacity of several hundred megawatts per year. Therefore, a prospective HTSE plant construction project developer could significantly reduce uncertainties in hydrogen production costs by obtaining competitive project-specific stack and system pricing information from SOEC vendors.
- The learning rate affects the HTSE plant modular equipment capital costs, and therefore, manipulation of this parameter impacts the HTSE plant direct capital costs (DCC is a sensitivity parameter that was perturbed separately from the learning rate; however, perturbation of the DCC affects the cost of both modular and conventional/scalable plant components). Variation in the learning rate of ±5% corresponds to a DCC ranging from \$418/kW-dc to \$713/kW-dc. Note that when the learning rate is equal to 100% there is no cost reduction attributed to mass production of the HTSE modules, which is representative of a first-of-a-kind (FOAK) plant type. 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 (e.g., several hundred megawatts), conventional/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 GW-scale HTSE.

A comparison of LWR-HTSE and natural gas (NG) steam methane reforming (SMR) (the conventional standard for large-scale hydrogen production) LCOH was performed to identify cases where HTSE could produce hydrogen at a cost that is competitive with SMR. The SMR LCOH is highly dependent on natural gas pricing. Figure ES-4 provides a comparison of the LWR-HTSE LCOH as a function of electrical power price versus the LCOH for a SMR plant with a production rate comparable to that of the LWR-HTSE plant at selected natural gas prices ranging from \$4/million British thermal unit (MMBtu) to \$10/MMBtu. Note that as of early February 2022, natural gas spot prices at the Henry Hub were reported in the range of \$6/MMBtu [6]. At \$6/MMBtu for natural gas pricing, the LWR-HTSE plant would need to purchase electrical power at a price of around \$21.5/MWh-e to produce hydrogen at an LCOH comparable to the SMR LCOH. As will be described below, this electricity price is within the range at which power could theoretically be purchased from an NPP. Additionally, the LWR-HTSE plant has the advantage of more stable pricing than does the SMR using natural gas, since natural gas prices can fluctuate and vary widely depending on local, national, and worldwide geopolitical factors.

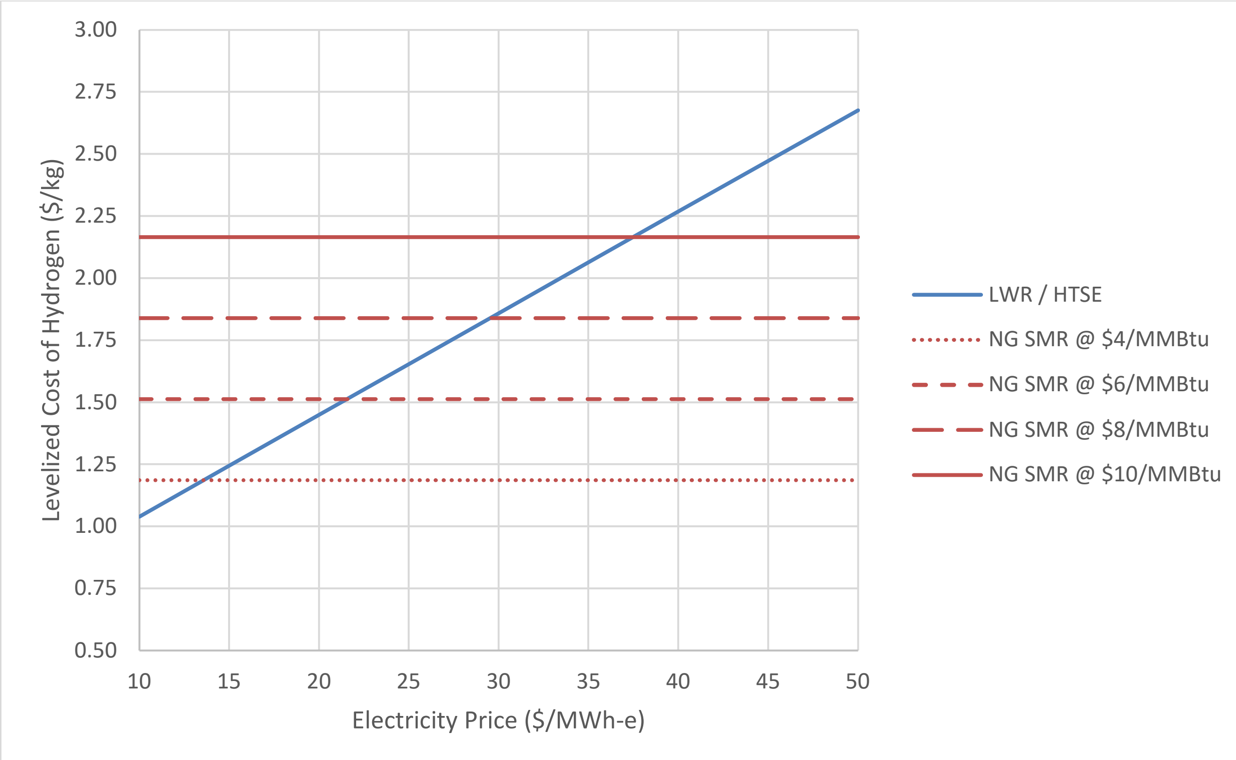

Figure ES-4. LWR-integrated HTSE plant LCOH as a function of electricity price. Also shown is the NG-SMR LCOH corresponding to selected natural gas pricing levels.

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 carbon price could increase the effective cost of SMR-derived hydrogen. The natural gas SMR LCOH is increased by approximately $0.01/kg for every $1/metric tonne (MT)-$CO_2$ tax that is applied. Specifically, the calculations performed indicate that a carbon tax of $25/tonne-$CO_2$ would increase the natural gas SMR LCOH by $0.22/kg [1]. In addition to the electricity price and HTSE plant capital costs, 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.

In addition, it should be noted that hydrogen produced by electrolysis is relatively pure. Hydrogen produced by SMR requires CO_2 separation and CO methanation to achieve hydrogen qualities that are required for power turbine cooling, ammonia production, metal alloys refining, and electronics manufacturing. **INL deduced the cost of SMR purification to meet the purity of electrolysis grade hydrogen is around $0.25-$0.30/kg-H_2.**

To show a possible path to reach a hydrogen production cost of $1/kg-$H_2$ a scenario informed by the sensitivity analysis was constructed and added to the 'waterfall' chart in Figure ES-5. While the tornado chart presented above identifies the LCOH changes that could result from changes to individual parameters, the waterfall chart in Figure ES-5 illustrates the cumulative LCOH decrease that could be achieved by combining multiple price-decreasing parameter changes. Note that the value of the capital cost specified in the waterfall chart differs from the lower bound specified in the tornado chart. While the proposed pathway to achieve $1/kg hydrogen from HTSE is viewed as aggressive, the parameters required to achieve this metric are not unfeasible. In addition to the reductions in energy price, operating parameters, and capital and operating costs, it may also be possible to obtain an additional source of revenue from oxygen byproduct sales or clean hydrogen production credits that could maintain the prospect of $1/kg hydrogen from HTSE in the event challenges are encountered in achieving the parameter specifications detailed in Figure ES-5.

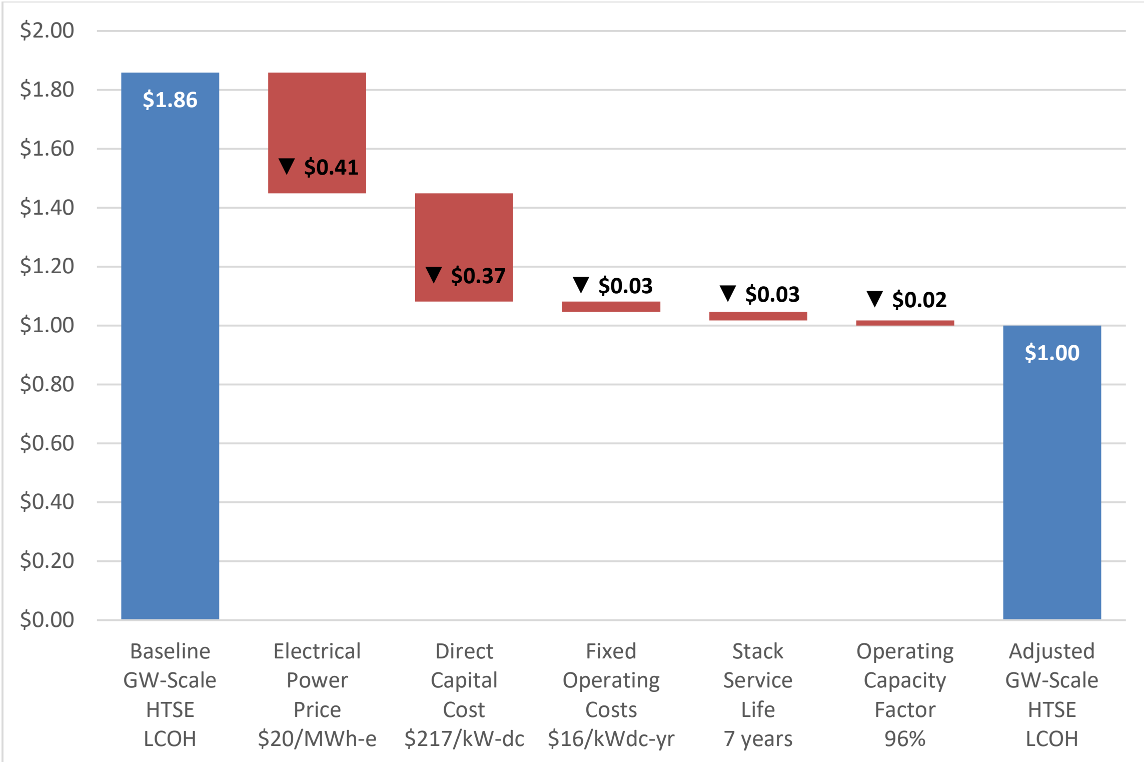

Figure ES-5. Waterfall chart illustrating a potential pathway to achieve an LCOH of $1/kg

As noted, and as shown in the sensitivity analysis in Figure ES-3, the levelized cost of electricity/energy (LCOE) is the most significant factor in the calculation of LCOH and therefore the most significant factor in the profitability of any system that aspires to produce hydrogen by electrolysis. Although historical LCOE values for nuclear power plants have been around $30/MWh, which is why this value was used as the baseline in the NPP-HTSE analysis described herein, there is significant area for improvement in these costs. This is due to many factors, including regulation uncertainty, that have led to a low degree of cost reduction initiatives in the nuclear power industry compared to other industries. Also, until the past decade, the nuclear industry has not been under the extreme price and competitive pressure that it is now.

Because of the current price pressure on the nuclear industry that has caused some NPPs to prematurely close, many studies have been done to outline roadmaps for decreasing operating costs. One such study [7] uses an 'Integration Options for Nuclear' (ION) approach to outline various possible improvements to nuclear power to reduce operating costs in the areas of technology, process, human performance, and governance. The ION Generation I analysis considered technologies and options that would be viable within the 3–5-year time frame. Table ES-2 below is reproduced from the mentioned reference with permission.

Table ES-2. Preliminary LCOE analysis showing identified pathways to reducing NPP operating costs

	Scenario 2: ION-Gen1 LCOE with Sustaining and Innovation Capital	Scenario 3: Significantly Reduced Capital	Scenario 4: Reduced Capital, Aggressive Reduction of Fixed O&M	Scenario 5: Reduced Capital, Improved Cost of Capital	Scenario 6: Nuclear Production Tax Credit
Generation Source	Nuclear	Nuclear	Nuclear	Nuclear	Nuclear
Plant Size (MW)	2200	2200	2200	2200	2200
Capacity Factor (%)	93	93	93	93	93
Fuel Cost ($/MMBtu)	0.65	0.65	0.65	0.65	0.65
Heat Rate (Btu/kWh)	10,300	10,300	10,300	10,300	10,300
Fixed O&M ($/kW-year)	71.36	71.36	64.55*	71.36	71.36
Variable O&M ($/MWh)	3.00	3.00	3.00	3.00	3.00
Overnight Costs ($/kW)	455 ($1B investment)	186 ($410M investment)	239 ($525M investment)	273 ($500M investment)	455 ($1B investment)
Interest Rate (%)	9.6	9.6	9.6	7.6	9.6
Production Tax Credit (PTC) ($/MWh)	0	0	0	0	$2.88
Levelized Cost of Energy ($/MWh)	**25.87**	**21.49**	**21.51**	**21.68**	**21.50**

The LCOE values in Table ES-2 represent industry averages. Single NPP operators have plans to have or already have LCOE values around $20/MWh. Full details can be found in the referenced document, but suffice to say, there is credible evidence to say that the LCOE average for the nuclear power industry will soon be on the lower end of the $20/MWh to 30/MWh range, which will, in turn, make hydrogen production via high-temperature electrolysis using nuclear electricity very competitive.

These analyses represent 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.

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CONTENTS

FIGURES

TABLES

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ACRONYMS

AC	alternating current
AEO	Annual Energy Outlook
APEA	Aspen Process Economic Analyzer
BoP	balance-of-plant
CAPEX	capital expenses
CCS	carbon capture and sequestration
CEPCI	Chemical Engineering Plant Cost Index
DC	direct current
DCC	direct capital costs
DFMA	Design for Manufacturing and Assembly
DOE	U.S. Department of Energy
EDR	Exchanger Design and Rating
EIA	Energy Information Agency
FEED	front end engineering design
FOAK	first-of-a-kind
GW	gigawatt
GWe	gigawatt electrical
HDSAM	Hydrogen Delivery Scenario Analysis Model
HHV	higher heating value
HTE	high-temperature electrolysis
HTF	heat transfer fluid
HTFO	Hydrogen and Fuel Cell Technologies Office
HTSE	high-temperature steam electrolysis
HX	heat exchanger
INL	Idaho National Laboratory
ION	Integration Options for Nuclear
IRR	internal rate of return
kg	kilogram
km	kilometer

kV	kilovolt
kW-dc	kilowatt direct current
LCOE	levelized cost of energy
LCOH	levelized cost of hydrogen
LTE	low-temperature electrolysis
LWR	light water reactor
MMBTU	million British thermal unit
MT	metric tonne
MW	megawatt
MW-ac	megawatt alternating current
MW-e	megawatt electric
MW-dc	megawatt direct current
MWh	megawatt hour
MWh-e	megawatt hour electric
NG	natural gas
NOAK	nth-of-a-kind
NPH	nuclear process heat
NPP	nuclear power plant
O&M	operations and maintenance
OEM	original equipment manufacturers
OPEX	operating expenses
P&ID	piping and instrumentation diagram
PEM	polymer exchange membrane
PFD	process flow diagram
PTC	production tax credits
PWR	Pressurized Water Reactor
SEL	steam extraction loop
SMR	steam methane reforming
SOEC	solid oxide electrolysis cell
TCI	total capital investment

TDL	thermal delivery loop
TEA	technoeconomic analysis
TPE	thermal power extraction
TRL	technology readiness level
U.S.	United States
USD	U.S. dollars

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High-Temperature Steam Electrolysis Process Performance and Cost Estimates

1. INTRODUCTION

This document presents a detailed engineering plant design model and analysis for the integration of hydrogen production via solid oxide electrolysis cell/high-temperature steam electrolysis (SOEC/HTSE) with a light water reactor (LWR) nuclear power plant (NPP). This analysis represents original modeling and analysis of current state-of-the-art HTSE technology integrated with nuclear power as well as forecasted performance improvements of HTSE technology. Plant design and analysis of input parameters, costs, and benefits of HTSE integrated with nuclear power is completed. Also described are various sensitivity studies on the cost to produce hydrogen (i.e., the Levelized cost of hydrogen [LCOH]) as well as a competitive comparison of hydrogen production via LWR-HTSE with the conventional steam methane reforming (SMR) process. Thus, not only current and forecasted technologies are modeled, but the sensitivity studies give a sense of what is possible with the improvement of the process input parameters.

Integration with an NPP using a small portion of heat from an LWR diverted to provide heat to the HTSE process, can significantly increase the HTSE system efficiency. A detailed process and control model of both the thermal delivery loop (TDL) and the nuclear reactor dynamics for thermal power extraction (TPE) from nuclear power have been separately performed outside of this current work [8, 9]. For the current analysis, a simplified model of TPE was used. Figure 1 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 heat after the high-pressure turbine, using steam instead of hot oil to transfer heat from the NPP, decreasing the distance between the steam extraction and the HTSE, and returning condensate to the first NPP feedwater heater versus to the condenser.

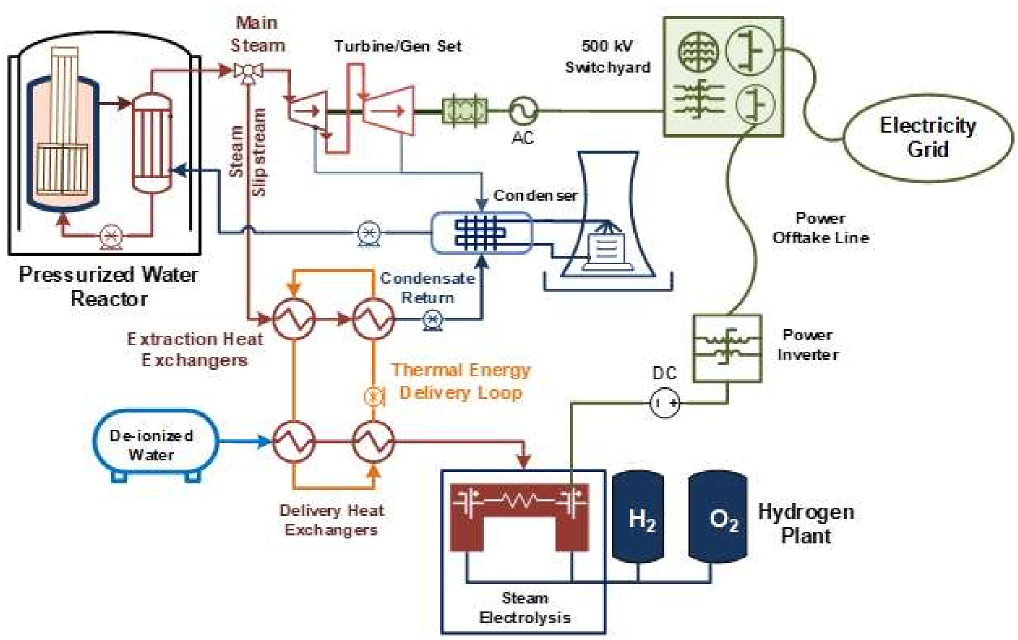

Figure 1. Overview of HTSE integrated with an NPP. Equipment added to the NPP includes the steam slip stream from the turbine inlet, TDL, HTSE hydrogen plant, and associated water and electricity supply tie-ins.

A process model of the NPP-HTSE was developed using AspenTech HYSYS to model the energy requirements and production rates (heat and material balances), as well as to derive the capital and operating expenses. 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 (H2A) to determine the overall costs of hydrogen production (the LCOH).

HTSE is a rapidly developing technology that has certain advantages over low-temperature electrolysis (LTE). The greatest advantage of HTSE includes higher efficiency of hydrogen production and, therefore, reduction in the cost to produce hydrogen, especially when integrated with heat from large sources of thermal power such as 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 very effectively 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. With the higher temperature operation, HTSE requires materials of construction that are more expensive than those used for LTE but the HTSE may optimize cost and performance by using stainless steels such as SS304 for components with operating temperature $T<600°C$ and SS347 for components with an operating temperature $T>600°C$. Previously, low alloy steel was specified for several low-temperature process components. The use of stainless steel throughout the process provides a more robust design.

The TDL modeled as a part of this study includes only assumptions of major 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 conclusions.

2. PROCESS MODELING DESIGN BASIS

An LWR-integrated HTSE process model was developed using AspenTech HYSYS simulation software [10] for: (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 2, which 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 2.1. Process operating conditions and equipment performance specifications are detailed in Section 2.2.

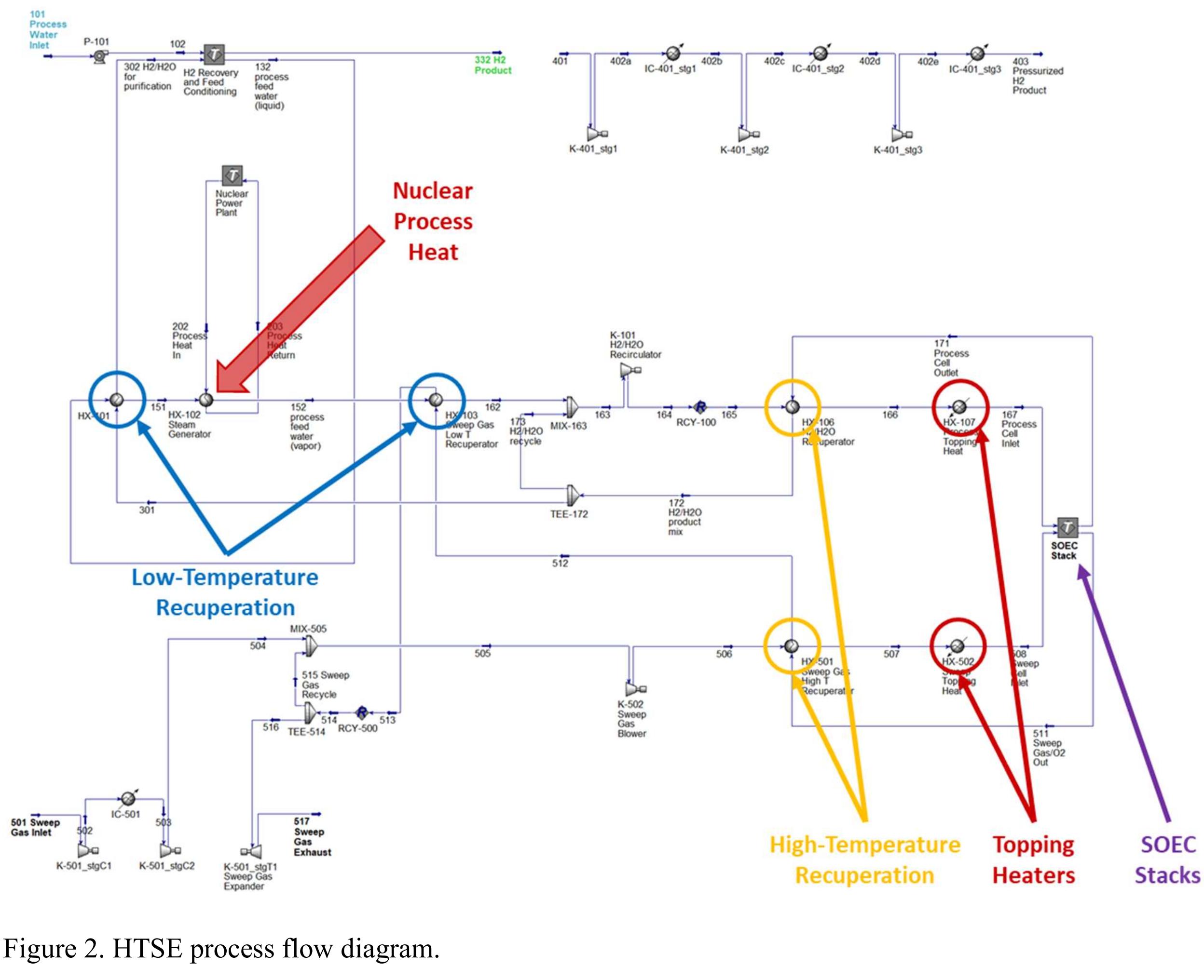

Figure 2. HTSE process flow diagram.

2.1 PROCESS OVERVIEW

The HTSE system evaluated includes several major process systems. These systems include: (1) the HTSE system; (2) the feed and utility system; (3) the air sweep gas system; (4) the hydrogen/steam system; (5) the hydrogen purification system; (6) the 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 identified process systems are included in Appendix A.

2.1.1 HTSE System

The HTSE system includes the SOEC stacks, the high-temperature recuperator heat exchangers (HXs), electric trim heaters, and insulated pressure containment vessel that houses the electrolysis stack array. 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 outlet streams to the lower-temperature stack inlet streams; the 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). An effective use of an external heat source, such as an NPP, is to vaporize the feedwater, given the large amount of energy that is required to change water from liquid to gas. In this study, an NPP was used as the external heat source, but another source such as a commercial natural gas (NG) boiler could be used as well if the electrolysis plant is not located near an NPP. 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 electrolysis stack operating temperature.

2.1.2 Feedwater and Utility System

The feedwater and utility system includes the process components necessary to prepare and stage a clean, demineralized feedwater stream (separate from the process and steam cycle water of the NPP) as a reactant 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 to condense and separate out water 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, rectifying the alternating current (AC) power from the transmission system to direct current (DC) power for use in the SOEC stacks and the bus bars for distributing the high amperage current from the rectifier to the stacks.

2.1.3 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) and an oxidizing environment can cause premature degradation of the SOEC materials of construction, it is important to reduce the oxygen concentration. An air sweep gas stream is used to dilute and evacuate high-concentration oxygen from the anode side of 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 produced within the stack), the net power requirements of the sweep gas compressor/expander are negligible in comparison with other HTSE system power demands.

2.1.4 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. It is desirable to maintain a reducing environment on the cathode side to prevent premature material degradation, so some amount of recycled hydrogen is always required. The hydrogen/steam system is comprised of low-temperature recuperators, the feedwater steam generator, high-temperature gas blowers, and piping/manifolds necessary to recycle a portion of the stack product gas. The low-temperature recuperators are used to preheat the liquid phase feedwater while simultaneously cooling the H_2/H_2O mixture enroute to the hydrogen purification system.

2.1.5 Hydrogen Purification System

The hydrogen/steam process gas mixture in the stack outlet stream flows through high- and then 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 at a pressure of 20 bar with a hydrogen purity of 99.9%.

2.1.6 Nuclear Process Heat Delivery System

The HTSE process efficiency can be increased by using external thermal heat to vaporize feedwater in preparation for splitting in the electrolyzer. This is beneficial due to the large amounts of energy required to vaporize water. In this analysis, the external heat for the HTSE process is assumed to come from an NPP via a TDL. Other heat sources, such as a commercial natural gas boiler could be substituted if desired. In the current case where a gigawatt-scale HTSE plant is coupled with an NPP such that the HTSE plant consumes all the energy output of the NPP (both thermal and electrical), approximately 5% of the nuclear plant steam flow is required to provide the heat duty required for vaporization of the HTSE process feedwater.

In this analysis, NPP steam is assumed to be 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 sub-cool the nuclear plant steam to transfer heat to a heat transfer fluid (HTF) in the TDL, which 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 synthetic heat transfer oil, such as Therminol-66 or DowTherm, in the TDL. Other analyses could consider the use of steam in place of the hot oil as the HTF.

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 km 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.

2.1.7 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 Aspen Process Economic Analyzer'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. 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.

2.2 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 a reducing environment at the cathode. A detailed listing of HTSE-process operating condition specifications is provided in Table 1.

Table 1. HTSE and related subsystem process operating condition specifications.

Parameter	Value	Reference or Note
Stack operating temperature	800°C	O'Brien et al. 2020 [11]
Stack operating pressure	5 bars	See Section 2.2.1
Operating mode	Constant V	
Cell voltage	1.29 V/cell	Thermoneutral stack operating point
Current density	1.5 A/cm²	James and Murphy 2021 [2]
Stack inlet H_2O composition	90 mol%	O'Brien et al. 2020 [11]
Steam utilization	80%	See Section 2.2.1
HTSE modular block capacity	25 MW-dc	1000x capacity increase [11]
Sweep gas	Air	O'Brien et al. 2020 [11]
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 [5]
Stack degradation rate	0.856%/1000 hr	HFTO Hydrogen Production Record 20006 [5]
Stack replacement schedule	Annual stack replacements completed to restore design production capacity	Based on H2A model stack replacement cost calculations

BoP equipment specifications are listed in Table 2. As detailed in Table 2 the system design basis includes purification of the hydrogen product to 99.9 mol% hydrogen and compression to a pressure of 20 bar. The system design basis specifically does not include high-pressure compression beyond 20 bar or hydrogen storage or transportation infrastructure capacity since the requirements for these operations are likely to vary for specific installations with potentially different product end-use applications.

Table 2. BoP equipment specifications.

Parameter	**Value**	**Reference or Note**
Heat Exchangers		
Heat exchanger ΔP: TDL, feedwater heating, low-temperature 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	20 bar (290 psia)	Corresponds to operating pressure of final product purification stage; additional compression may be needed for transport, storage, or specific hydrogen use applications
Thermal Delivery Loop		
TDL HTF	Therminol-66	O'Brien et al 2017 [12]; Frick et al 2019 [13]
TDL transport distance	1.0 km	Vedros et al 2020 [14]
Maximum HTF velocity	3.0 m/s	Basis for pipe diameter calculations

2.2.1 Stack Operating Conditions Selection

Because the majority of hydrogen production costs are generally associated with energy input, the HTSE system's normal operating mode must minimize energy use and, also, equipment capital costs. Steam utilization (the percentage of steam that is converted to H_2 and O_2 in the electrolysis stacks, the remainder is unreacted steam) 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 using 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 utilization has a direct impact on the HTSE system cooling and thermal energy input requirements. References [11, 15-18] indicate that HTSE steam utilization typically ranges anywhere from 40 to 90%. The lower the steam utilization, 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 utilization results in increased system cooling and thermal energy input requirements. Although the use of a very high steam utilization 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 utilization 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 utilization 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 increase to nearly 90% in the near future.

The effect of the stack operating pressure and steam utilization on the system's electrical energy consumption are shown in Figure 3. 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 utilization 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. The analysis shown in Figure 3 includes hydrogen product high-pressure compression and illustrates the reduction in energy usage from using liquid pumps to increase the pressure in the feed system in exchange for less hydrogen product compression. All other analyses in this report exclude hydrogen product compression. This analysis does not take into account the cost savings that could be achieved by using lower cost HTSE process materials at lower operating pressures.

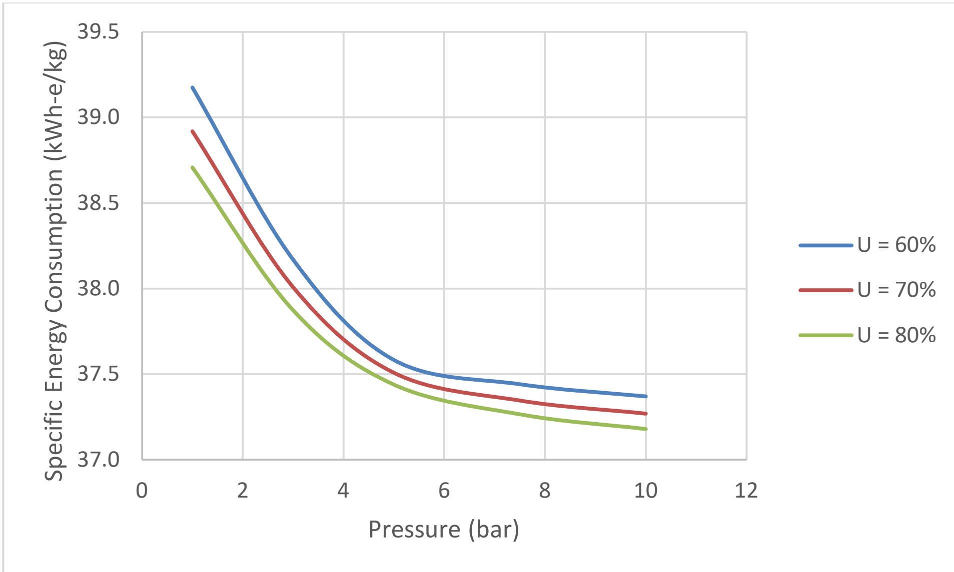

Figure 3. Electrical energy consumption as function of stack operating pressure with steam utilization as a parameter. Unlike the other analyses in this report, this analysis includes high pressure hydrogen compression.

The system's thermal energy consumption is most strongly affected by steam utilization, as shown in Figure 4. Increases in steam utilization result in a nearly linear decrease in thermal energy consumption over the range of values evaluated.

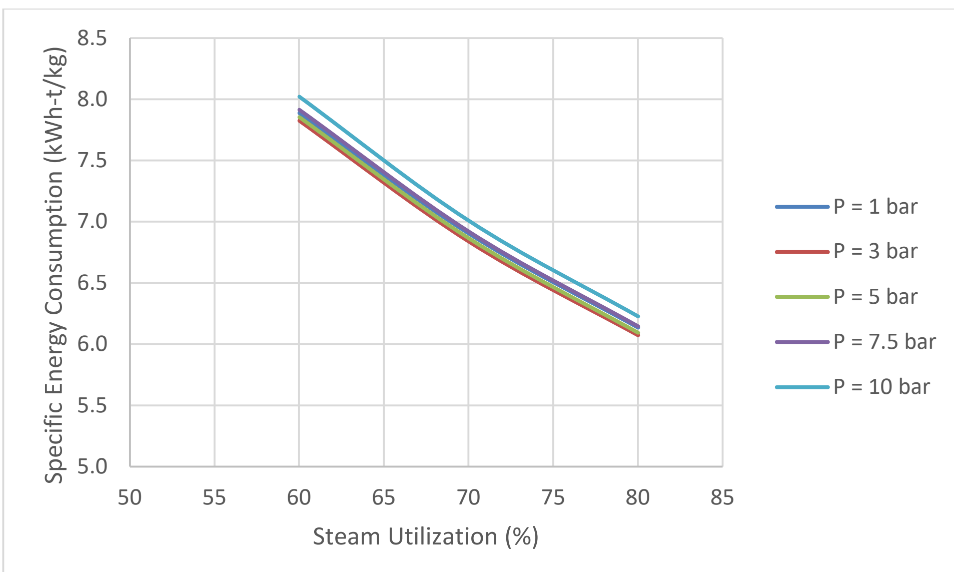

Figure 4. Thermal energy consumption as function of steam usage with stack operating pressure as a parameter.

The HTSE system efficiency is a metric that includes both thermal and electrical energy consumption. The impact of the stack operating pressure and steam utilization on the HTSE system efficiency is plotted in Figure 5. Increases in steam utilization 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.

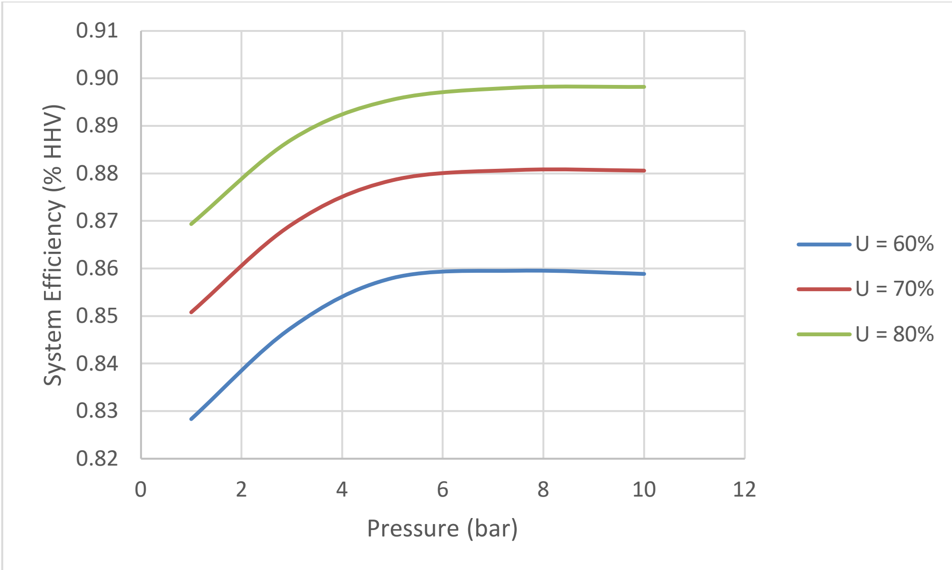

Figure 5. HTSE system efficiency as a function of stack operating pressure with steam usage as a parameter.

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. The pressure of 5 bar was selected currently, but the analysis of whether a pressure lower than 5 bar would be more cost effective, taking into account a detailed material selection is ongoing. 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 [11, 15-18].

2.2.2 Normal Operation

HTSE-system normal operation is characterized by the conditions specified in Table 1. 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. A grid-integrated HTSE plant may operate in the normal operating mode most 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. This current analysis focuses only on the constant hydrogen production mode and not grid-integrated optimizations so hot standby mode is not applicable in this report.

2.2.3 SOEC Performance Degradation

Actual annual hydrogen production may vary from the design production capacity for several reasons. 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 and a degradation rate of 0.856%/1000 hr [5]; these values were also used as the basis for stack life and degradation in the current analysis. Based on the specified degradation rate, the production capacity would be reduced to 93.45% 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 96.7% 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.

2.3 HTSE Process Model Performance Estimates

The LWR-HTSE process material balances, process energy requirements, and process efficiency are summarized in Table 3. LWR-HTSE process summaries are provided for a gigawatt-scale NPP using its full capacity to power an HTSE plant. Electrical and thermal power requirements by equipment type are shown for the GW-scale LWR/HTSE system in Figure 6.

Table 3. LWR-HTSE process summary.

Parameter	Value	Notes
Plant design capacity	702 tonnes/day	99.9 mol% hydrogen at 20 bar
Design point power consumption	1000 MW-dc; 1076 MW-ac	DC power corresponds to stack power input; AC power corresponds to total power requirement (AC power to rectifier, pumps, compressors, topping heaters, etc.)
Availability factor	90%	HTSE plant operating time; corresponds to nuclear plant availability [time]/[time]
Cell degradation factor	96.7%	Adjustment to production rate due to cell degradation
Operating capacity factor	87.1%	The ratio of actual production rate to design production rate. Calculated as product of availability and cell degradation factors.
Actual hydrogen production rate	611 tonnes/day	
Process power requirement: Electrical Thermal	 1076 MW-ac 188.2 MW-t	Design condition
Specific energy consumption: Electrical Thermal	 36.8 kWh-e/kg H_2 6.4 kWh-t/kg H_2	Includes compression of hydrogen product to 20 bar; additional energy required for high-pressure storge or transport compression
System H_2 production efficiency	90.2% higher heating value (HHV) basis	Energy content of product H_2 divided by electrical energy equivalent input
Utilities: Process water feed rate Cooling water circulation rate	 72.6 kg/s [1.2k gpm] 1170 kg/s [18.6k gpm]	

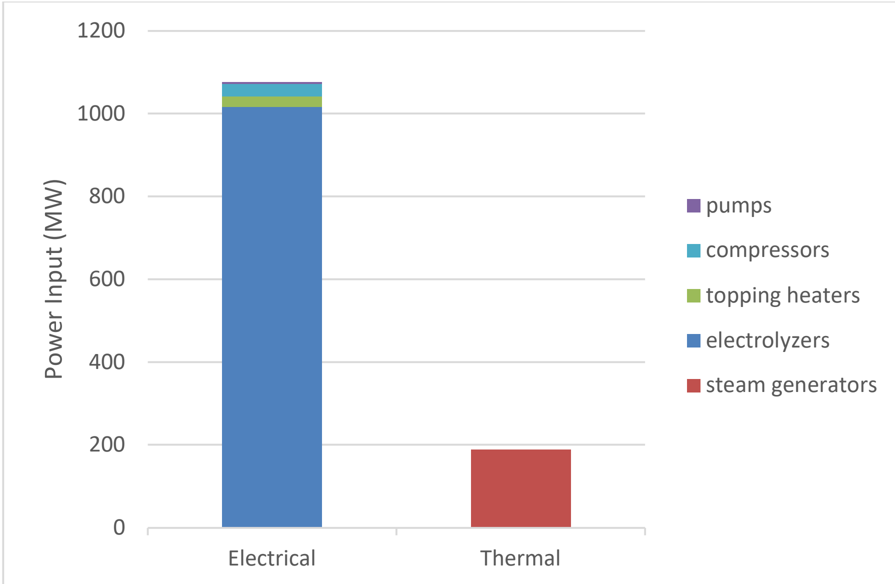

Figure 6. GW-scale LWR-HTSE electrical and thermal power requirements (design point).

2.4 HTSE Process Design Considerations

2.4.1 LWR/HTSE integration

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

Appendix B includes 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 4. Table 4 also includes equipment with functionality that may exist separately in both the LWR and the HTSE plant. Although this equipment is included in the Idaho National Laboratory (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 the purchase and installation of separate HTSE-specific equipment items may not be required. Other analyses could substitute a commercial natural gas boiler or another source of heat as required, depending on the application.

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 that 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 2.4.2.

Table 4. 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

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 7 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).

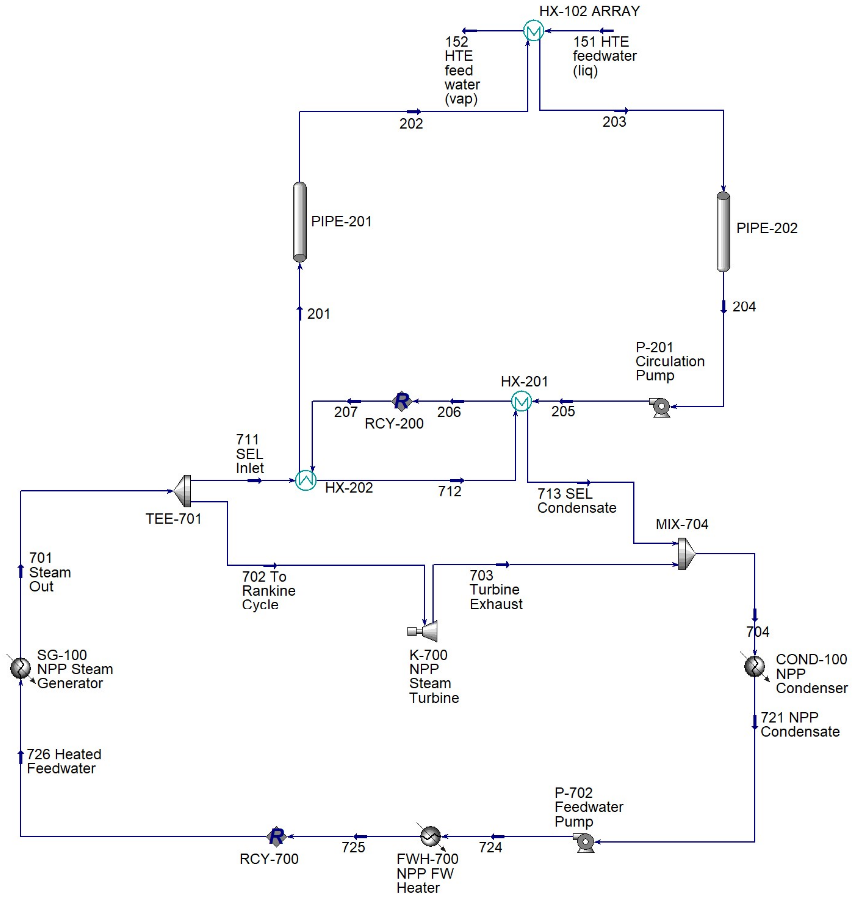

Figure 7. TDL integration with NPP (simplified NPP model).

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.

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 kilovolt (kV) to the rectifier supply voltage (assumed ~4 kV for equipment costing purposes). If equipment for distribution of 20 kV power is not present, then this equipment will need to be retrofitted, 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 rectifiers for converting AC power to DC power and DC bus bars for distributing the power from the inverters to each of the HTSE modules.

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 if 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 current analysis.

Water Purification

INL's HTSE-system design basis includes feedwater pretreatment and purified feedwater storage capacity. If the NPP includes water pretreatment equipment and storage capacity sufficient to supply the HTSE plant, these equipment items can be removed from the 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.

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 water 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 NPP 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.

2.4.1.2 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 the distribution of steam between the power cycle and nuclear process heat applications (e.g., HTSE)
- Leak monitoring and detection equipment (e.g., 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 (e.g., 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 shut-down, de-rating, or interruption of service or operations required to implement process modifications.

2.4.2 Thermal Delivery Loop Design Parameters Requiring Further Investigation

2.4.2.1 Options for Steam Extraction Loop Condensate Return

A detailed diagram of the TDL integration is shown in Figure 9, 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 (MIX-184), or a location in the high-pressure boiler feedwater heating train (MIX-188). The INL system design basis specifies the return of the SEL condensate to the point upstream of the condenser because NPP condensers are built with excess design capacity, that is suitable for handling excess steam input associated with plant start-up and shut-down, 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 a 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.

2.4.2.2 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 the 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 as observed in Figure 8:

- 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 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 the detection of radioactive contaminants that may have escaped from the NPP primary or secondary loops. Equivalent protocols for the 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 the 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.

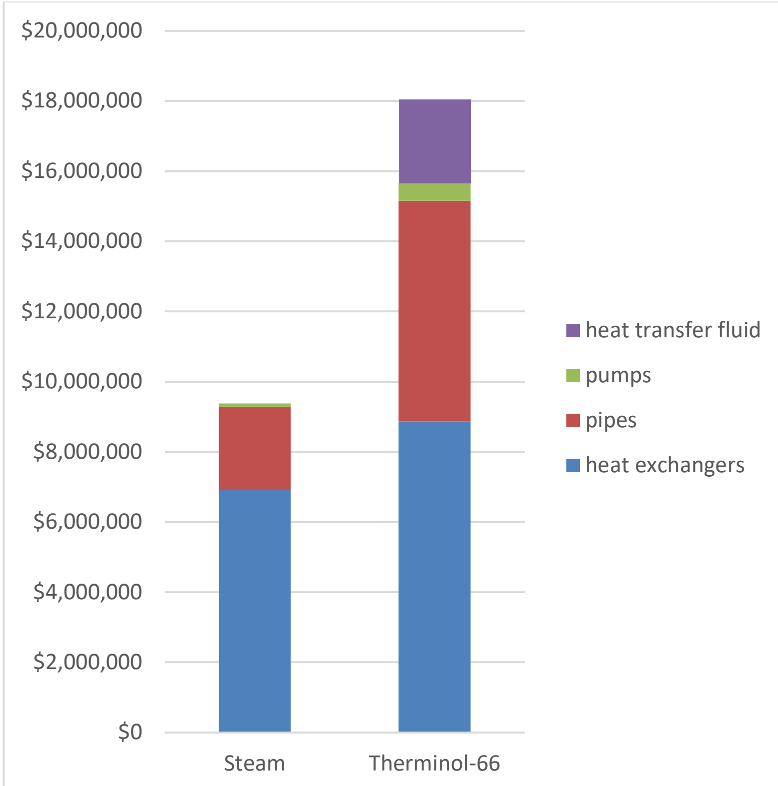

Figure 8. Steam versus synthetic heat transfer oil TDL capital costs for a gigawatt-scale LWR/HTSE plant installation.

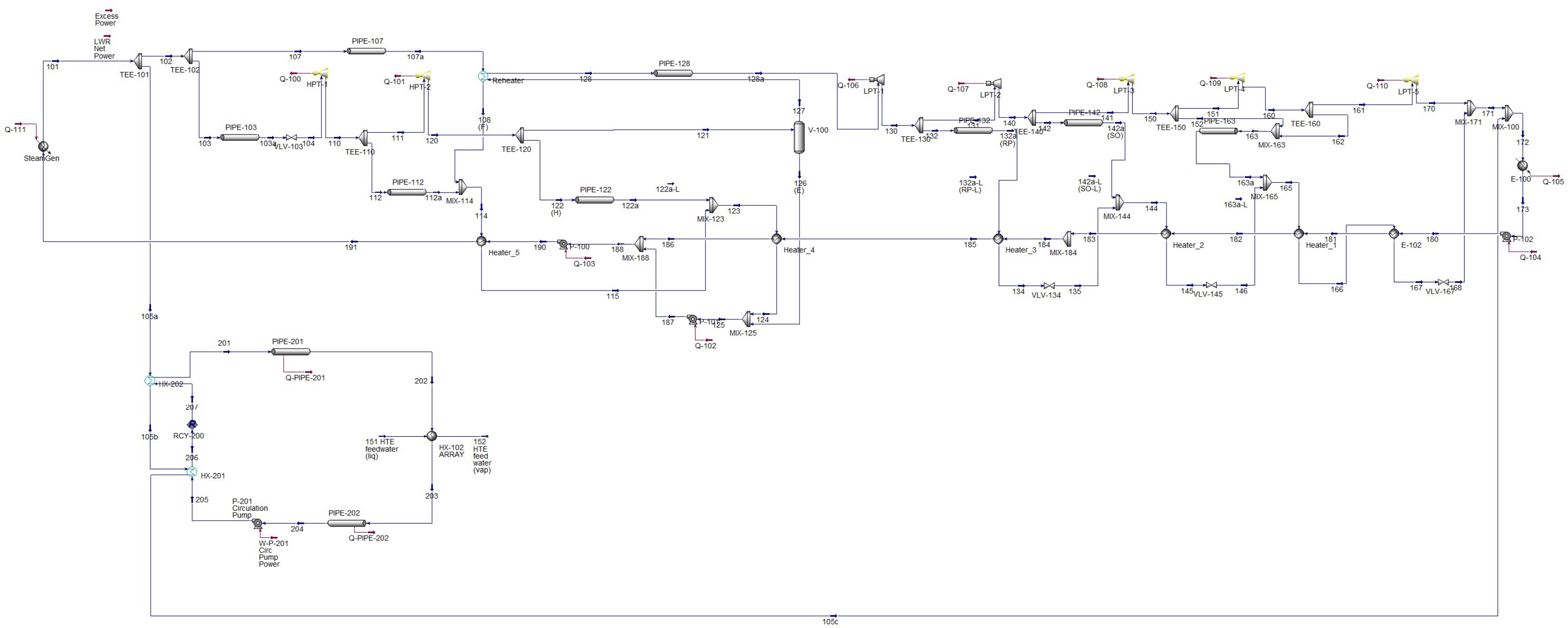

Figure 9. TDL integration with NPP (detailed NPP model).

3. HTSE PROCESS CAPITAL AND OPERATING COSTS

3.1 HTSE Process Capital Costs

3.1.1 Capital Cost Estimation Methodology

3.1.1.1 Modular Equipment

This analysis assumes that the HTSE plant is constructed using a modular concept, which involves the 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 stack input 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 the feedwater pumps, feedwater preheating equipment, steam generators, recuperators, topping heaters, product purification equipment (compressors, gas coolers, knock-out drums), and sweep gas system.

Modular construction is a logical approach that is used for the design of HTSE processes. It is envisioned that the HTSE modules produced by a given SOEC manufacturer would adhere to a standardized design, and that the modules would be mass produced at an industrial manufacturing facility.

The system components included in each of the HTSE modular blocks are identified as modular equipment. 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 AC and DC power requirements for each HTSE block exceed 25 megawatt electric (MW-e) total power input (the total power requirements for an HTSE plant configuration with 40 modular units will exceed 1 gigawatt electric [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 using a learning curve relationship to account for economies of mass production. A learning rate of 95% was specified in the development of the cost versus capacity correlation developed in this analysis. The learning rate of 95% corresponds to a 5% cost reduction with every doubling of the number of units produced, which is within the range of values reported in the literature for the energy sector [19-22]. The learning rate affects the HTSE plant modular equipment capital costs. Note that when the learning rate is equal to 100%, there is no cost reduction attributed to mass production of the HTSE modules, which is representative of a first-of-a-kind (FOAK) plant type and that the 95% assumed learning rate for the nth-of-a-kind (NOAK) plant described is considered conservative. Planned expansions in vendor-specific manufacturing capacity could affect the learning rate that is realized as the establishment of large-scale SOEC manufacturing capacity continues in the coming years. 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.

FOAK and NOAK Plant Construction

A gigawatt-scale HTSE plant has not yet been constructed; therefore, capital cost reductions are expected when advancing from the FOAK plant installation to an 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 versus 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 [3] and/or scaled based on data reported in previous HTSE process evaluations [5, 23-26]. 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.

FOAK 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 block 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 $8^{\log_2(0.95)} = 0.95^3 = 85.7\%$ 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 $\sum_{N=1}^{8} N^{log_2(0.95)} = 7.26$ times the cost of the first unit.

NOAK 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 (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."

3.1.1.2 Scalable Equipment

The feed and utility, nuclear process heat (NPH) delivery, multistage product compression (when applicable), 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 a single TDL unit, 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 from 25 to 1150 MW as mentioned previously. Item-specific scaling exponents for each of the scalable equipment components were then determined from the capacity versus 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. This data set was then used to derive a power law correlation to predict total scalable equipment costs as a function of plant capacity.

3.1.1.3 Indirect Costs

An indirect cost multiplier of 1.294 is applied to the installed capital costs predicted by the equation (see Table 5). The indirect costs include site preparation, engineering and design, project contingency, contractor and legal fees, and land. The engineering and design and process continency values assumed were reduced from the values in [5] by applying an 80% learning curve 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 5. Indirect cost multipliers.

Indirect Cost Category	HFTO Hydrogen Production Record	INL HTSE Process Analysis
Site Preparation	2%	2%
Engineering and Design	10%	2.3% $^{\alpha}$
Process Contingency	15% total	1.6% $^{\alpha}$
Project Contingency		7.2%
Contractor Fee	15% total	10%
Legal Fee		5%
Land	<1%	1%
Cumulative Multiplier	1.421	1.294
$^{\alpha}$ NOAK plant specifications were obtained by applying an 80% learning curve to the corresponding parameter values from HFTO Record 20006 [5].		

3.1.1.4 Total Capital Investment

A total capital investment (TCI) 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 TCI was derived by fitting the resulting cost versus capacity data set using a power law relation.

3.1.2 Gigawatt-Scale LWR-HTSE Estimated Process Capital Costs

As described in Section 2 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 Appendix B.

Capital costs reported correspond to an HTSE plant with 1000 MW-dc of electrolysis input capacity. This equates to 40 × 25 MW-dc HTSE units with a total plant power consumption of 1076 MW-ac (accounting for the power consumption of both the electrolyzer and the associated with the BoP equipment) and a design point hydrogen production rate of 702 tonnes/H_2-day.

Capital cost summary tables for the FOAK and NOAK GW-scale LWR-HTSE plant installations are shown in Table 6 and Table 7, respectively. The LWR-HTSE CAPEX estimates for FOAK and NOAK plant types are presented graphically in Figure 10. Capital costs for each of the equipment components within the LWR/HTSE plant boundary limits are obtained from references [3, 5, 23-26]. All capital costs were indexed to 2020 dollars using the Chemical Engineering Plant Cost Index (CEPCI) [27].

Table 6. CAPEX summary for a generic 1,000 MW-dc LWR-HTSE plant (FOAK plant type)

Cost Category	**Description**	**2020 Dollar Basis**	**% of DCC**	**% of TCI**
Direct capital costs (DCC)	HTSE system [α]	$215,770,000	36.6%	28.3%
	Feed and utility system	$160,810,000	27.3%	21.1%
	Sweep gas system	$58,560,000	9.9%	7.7%
	Hydrogen/steam system	$21,320,000	3.6%	2.8%
	Hydrogen purification	$90,650,000	15.4%	11.9%
	Nuclear steam delivery	$41,100,000	7.0%	5.4%
	H_2 compression and storage	$0	0.0%	0.0%
	Control center	$830,000	0.1%	0.1%
	Total	$589,040,000 ($590/kW-dc)	100.0%	77.3%
Indirect depreciable capital costs	Site preparation	$11,780,000		1.5%
	Engineering and design	$13,370,000		1.8%
	Contingencies and contractor fee	$110,750,000		14.5%
	Legal fee	$29,450,000		3.9%
	Total	$165,350,000		21.7%
Total depreciable capital costs		$754,390,000		99.0%
Non-depreciable capital costs	Land	$7,540,000		1.0%
Total capital investment (TCI)		$761,930,000 ($763/kW-dc)		100.0%

[α] Based on HTSE stack capital cost specification of $78/kW-dc [2]

Table 7. CAPEX summary for a generic 1,000 MW-dc LWR-HTSE plant (NOAK plant type)

Cost Category	**Description**	**2020 Dollar Basis**	**% of DCC**	**% of TCI**
Direct capital costs (DCC)	HTSE system $^{\alpha}$	$202,070,000	37.2%	28.8%
	Feed and utility system	$146,320,000	26.9%	20.8%
	Sweep gas system	$52,740,000	9.7%	7.5%
	Hydrogen/steam system	$19,200,000	3.5%	2.7%
	Hydrogen purification	$81,640,000	15.0%	11.6%
	Nuclear steam delivery	$40,340,000	7.4%	5.7%
	H_2 compression and storage	$0	0.0%	0.0%
	Control center	$830,000	0.2%	0.1%
	Total	$543,140,000 ($544/kW-dc)	100.0%	77.3%
Indirect depreciable capital costs	Site preparation	$10,860,000		1.5%
	Engineering and design	$12,330,000		1.8%
	Contingencies and contractor fee	$102,130,000		14.5%
	Legal fee	$27,160,000		3.9%
	Total	$152,480,000		21.7%
Total depreciable capital costs		$695,620,000		99.0%
Non-depreciable capital costs	Land	$6,960,000		1.0%
Total capital investment (TCI)		$702,580,000 ($703/kW-dc)		100.0%

$^{\alpha}$ Based on HTSE stack capital cost specification of $78/kW-dc [2]

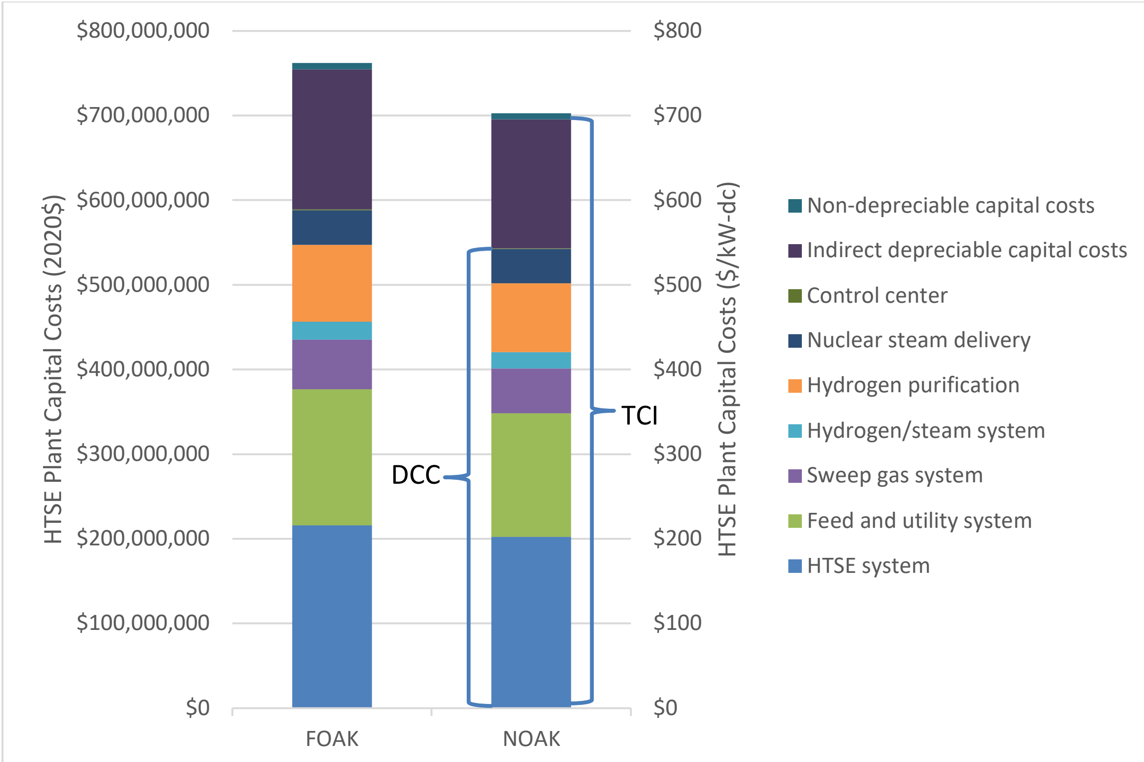

Figure 10. FOAK and NOAK plant CAPEX estimates for a generic 1,000 MW-dc LWR-HTSE plant (HTSE stack capital cost specification of $78/kW-dc) [2]

3.1.3 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 any follow-on grid-integrated LWR-HTSE plant optimization analyses. Figure 11 and Figure 12 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. Section 3.1.1 previously provided more information on these equipment categorizations. The capital cost correlation estimates the capital costs of the HTSE process areas described in Section 2.1. Note that retrofit costs required for the LWR to interface with the TDL system are not included in the cost estimates (see Section 2.4.1.2 for additional information).

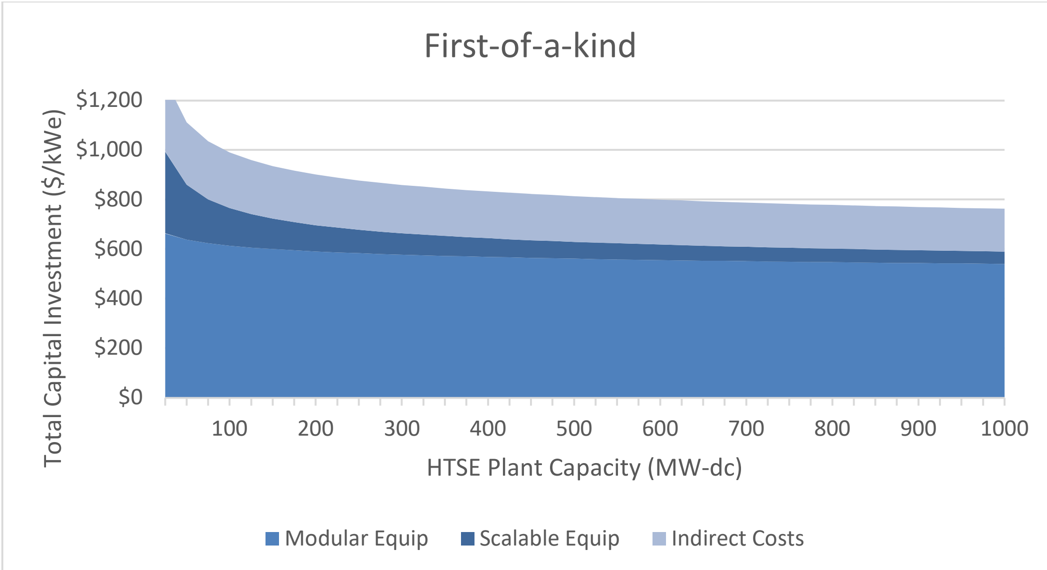

Figure 11. Total capital investment as a function of plant capacity for a FOAK HTSE plant (HTSE stack capital cost specification of $78/kW-dc) [2].

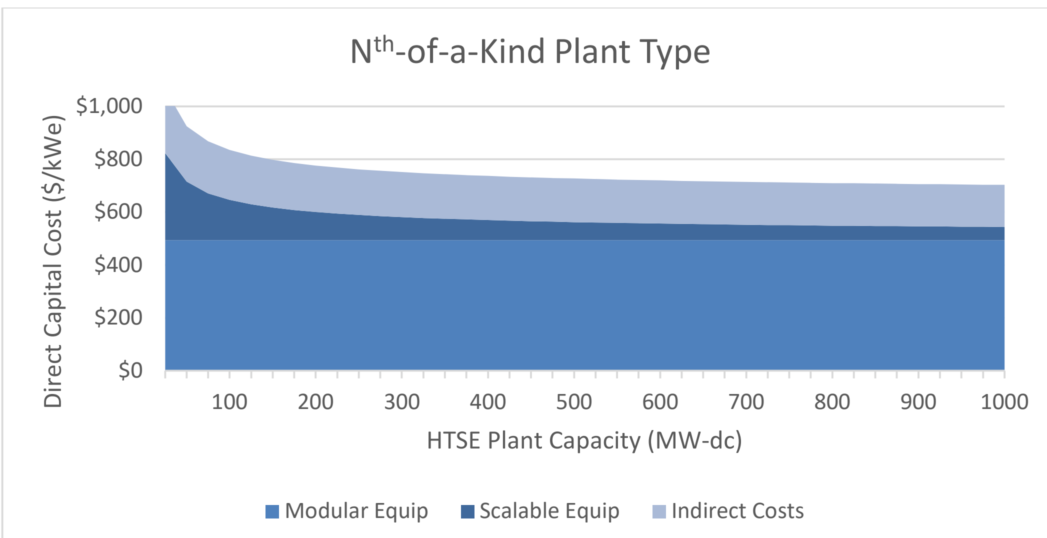

Figure 12. Total capital investment as a function of plant capacity for an NOAK HTSE plant (HTSE stack capital cost specification of $78/kW-dc) [2].

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 weighted average installation factor for all equipment items is applied to obtain installed equipment costs. 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 8.

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

where

TCI= Total Capital Investment ($/kW-dc)
P = HTSE system power (MW-dc)
m = indirect cost multiplier = 1.294
f = installation factor
$a_{scalable}$ = scalable equipment cost coefficient
$n_{scalable}$ = scalable equipment scaling exponent
$a_{modular}$ = modular equipment cost coefficient
$n_{modular}$ = modular equipment scaling exponent,

Table 8. LWR-HTSE capital cost correlation parameters (HTSE stack capital cost specification of $78/kW-dc; results reported in 2020 USD).

	FOAK	**NOAK**
f	1.387	1.382
$a_{scalable}$	862.9	862.9
$n_{scalable}$	-0.501	-0.501
$a_{modular}$	570.7	365.6
$n_{modular}$	-0.052	0

3.1.3.1 SOEC Technology Readiness Level Represented by Capital Cost Analysis

Capital cost analysis was performed using publicly available stack cost estimates. The stack cost of $78/kW-dc used in this analysis was obtained from a recent Design for Manufacturing and Assembly (DFMA) estimate for an electrode-supported stack with a manufacturing rate of 1,000 MW/yr [2]. The analysis uses the Current Hydrogen Production Case from the U.S. Department of Energy (DOE) HFTO Hydrogen Production Record #20006 [5] as the basis for stack service life (4 years). Therefore, the specified stack manufacturing capacity and stack service lifetimes would need to be achieved by the year 2024 to support the start-up of a gigawatt-scale HTSE plant in the year 2025 as specified in the hydrogen production cost analysis presented below (Section 4).

The BoP components are, in general, commercial technology, and the pricing information specified for these components corresponds to the current time. However, for the learning curve cost reductions specified for modular components to be realized, manufacturing capacity sufficient to produce 1,000 MW/yr or greater of BoP equipment components would be required (to achieve NOAK status by the specified plant start-up time).

3.2 HTSE Process Operations and Maintenance Costs

HTSE process operations and maintenance (O&M) costs were calculated according to the input specifications listed in Table 9. 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 depreciable capital 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 9. HTSE process O&M cost estimate basis.

Category	Value	Reference or Note
Fixed O&M Costs		
Total Plant Staff	15 (corresponds to 702 tonne/day design hydrogen production capacity)	8 person plant staff for a 50 tonne/day plant assumed [5]; 0.25 scaling exponent for varying plant capacity [28]
Burdened labor cost	$60/hr	
G&A rate/costs	20% of labor	
Licensing, Permits, and Fees	N/A	
Property Tax and Insurance	2% of TCI per year	
Rent	N/A	
Production Maintenance and Repairs	3% of DCC per year	
Variable O&M Costs		
Electricity Cost	$30/MWh-e	Baseline value assumed for electricity obtained from NPP
Replacement Costs	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	Cooling water cost is for make-up and chemical treatment [29]

O&M cost estimates for the generic 1,000 MW-dc LWR-HTSE plant are provided in Table 10. The O&M cost estimates correspond to the HTSE plant capacity specifications in Table 3 and the O&M cost estimate basis detailed in Table 9.

Table 10. Generic 1,000 MW-dc LWR-HTSE plant annual O&M cost estimate in 2020 USD.

Fixed O&M Costs					
Burdened labor cost, including overhead	$60	$/hour	15	FTEs	$1,933,000
G&A rate	20%	% of labor cost			$387,000
Property Tax and Insurance	2%	% of total capital investment			$14,029,000
Production Maintenance and Repairs	3%	% of installed direct capital costs			$16,268,000
Total Fixed O&M					$32,617,000 ($32.64/kWdc-yr)
Variable O&M Costs					
Replacement Costs					
Annual Stack Replacement Percentage [α]	27.3%	% of design capacity			$21,294,000
Total Unplanned Replacement	0.5%	% of total direct depreciable costs/year			$3,473,000
Electricity	30	$/MWh-e	1,076	MW-e	$246,157,000
Nuclear process heat [β]	10.2	$/MWh-t	188	MW-t	$14,639,000
Process Water	2	$/k-gal	1,660	k-gal/day	$1,055,000
Cooling Water (make-up and chemical treatment)	0.02	$/k-gal	26,800	k-gal/day	$170,000
Total Variable O&M (including energy costs)					$286,788,000 ($37.64/MWh-dc)
Total Variable O&M (excluding energy costs)					$25,992,000 ($3.41/MWh-dc)

[α] Based on HTSE stack capital cost specification of $78/kW-dc [2]
[β] Based on a thermal-to-electrical conversion efficiency of 34%

4. LWR-HTSE HYDROGEN PRODUCTION COST

4.1 LCOH Analysis

The current analysis is focused on constant hydrogen production. This analysis does not account for grid impacts or interactions and 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 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 user of this data should understand electricity pricing effects in the regional area of interest.

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 before HTSE plant start-up (i.e., 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.

The DOE H2A model [28] was configured with the LWR-HTSE process performance parameters described in Section 2.3, the capital costs described in Section 3.1.2, the O&M costs described in Section 3.2, and the project financial input parameters listed in Table 11 to calculate the non-grid-integrated LWR-HTSE plant LCOH.

Table 11. LWR-HTSE constant hydrogen production LCOH analysis input parameters.

Parameter	Value	Note or Reference
Financial Parameters		
Start-up year	2025	
Length of construction period	1 year	H2A Current SOEC Case Study [4]
Start-up time	1 year	H2A Current SOEC Case Study [4]
Plant life	20 years	
Depreciation schedule	20 year MACRS	H2A Current SOEC Case Study [4]
% Equity financing	40%	H2A Current SOEC Case Study [4]
Interest rate on debt	3.7%	H2A Current SOEC Case Study [4]
Debt period	20 years	
% of fixed operating costs during start-up	100%	H2A Current SOEC Case Study [4]
% of revenues during start-up	50%	H2A Current SOEC Case Study [4]
% of variable operating costs during start-up	75%	H2A Current SOEC Case Study [4]
Decommissioning costs (% of TDC)	10%	H2A Current SOEC Case Study [4]
Salvage value (% of TCI)	10%	H2A Current SOEC Case Study [4]
Inflation rate	1.9%	H2A Current SOEC Case Study [4]
After-tax real internal rate of return (IRR)	10%	
State taxes	6%	H2A Current SOEC Case Study [4]
Federal taxes	21%	H2A Current SOEC Case Study [4]
Indirect costs		
Site preparation (% of DCC)	2%	H2A Current SOEC Case Study [4]
Engineering and design (% of DCC)	2.3%	80% learning curve applied to value specified in [4]
Process contingency (% of DCC)	1.6%	80% learning curve applied to value specified in [4]
Project contingency (% of DCC)	7.2%	H2A Current SOEC Case Study [4]
Contractor's fee (% of DCC)	10%	H2A Current SOEC Case Study [4]
Legal fee (% of DCC)	5%	H2A Current SOEC Case Study [4]
Land (% of TDC)	1%	H2A Current SOEC Case Study [4]
Technology Case		
Plant type (NOAK)	NOAK	Nth-of-a-Kind specified as N=100 count of previous 25 MW-dc HTSE modular block installations
Learning rate for modular equipment cost reduction	95%	Specified learning rate is within the values reported for the energy sector [19-22]
NOAK plant stack cost	$78/kW-dc	Estimate for electrode supported cells with 1,000 MW/yr manufacturing capacity [2]

The LCOH analysis results for a baseline case with an HTSE plant providing actual hydrogen production capacity of 611 tonnes hydrogen per day (with a design capacity of 702 tonnes/day) and an energy price of \$30/MWh-e are presented in Figure 13. An HTSE plant of this capacity would use 1076 MW-ac of total power input (999.2 MW-dc stack power input). The LCOH for this baseline case is \$1.86/kg (in 2020 USD). It is apparent from Figure 13 that the largest contributor to the LCOH is the electricity cost.

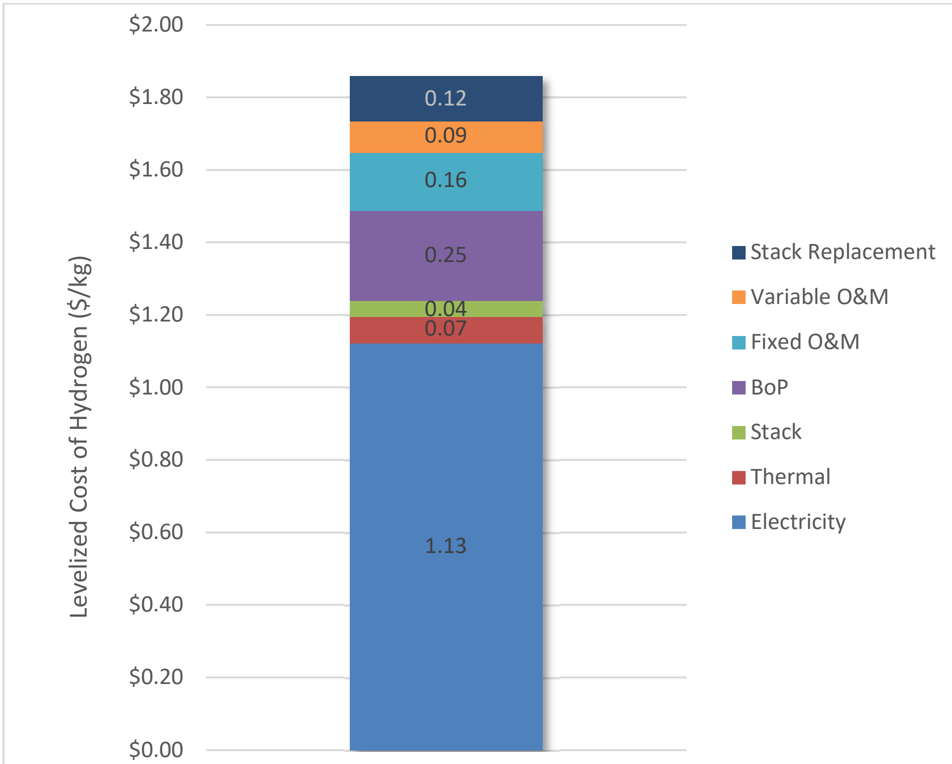

Figure 13. LCOH cost components for an NOAK constant hydrogen production LWR-HTSE system configuration with 611 tonnes per day actual hydrogen production capacity (702 tonnes/day design capacity), stack cost of \$78/kW-dc, and an energy price of \$30/MWh-e.

4.2 Sensitivity Analyses

Sensitivity analysis was performed to evaluate the impact of energy price and other key variables on the LCOH production. Figure 14 shows a 'tornado' chart that illustrates the LCOH sensitivity to the selected input parameters. Each of the sensitivity variables shown in the tornado chart is manipulated individually while all other variables are kept constant at the base values. 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 14 are sorted such that the variables that result in the largest net change in LCOH are positioned at the top of the chart giving the characteristic 'tornado' appearance.

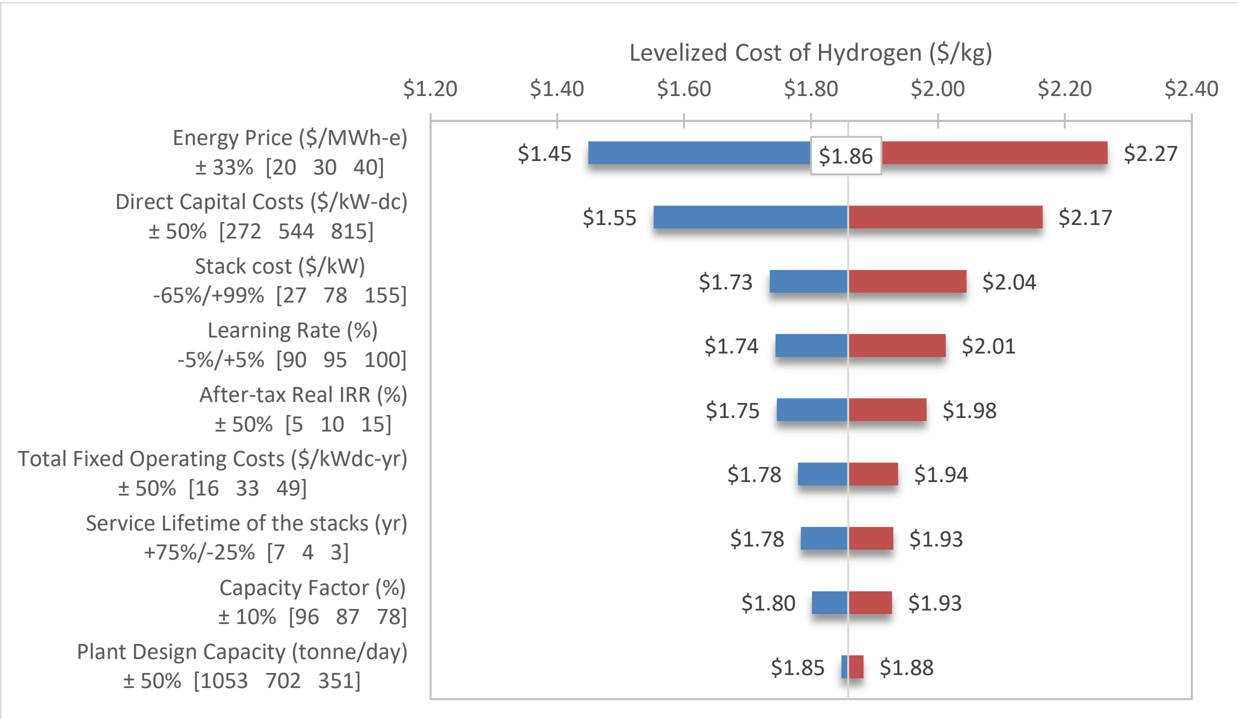

Figure 14. Sensitivity of LCOH to selected input parameters.

It can be observed from Figure 14 that, as expected, the specified changes in electricity price have the largest impact on LCOH. The range of electricity prices evaluated represents expected trends in future electricity market pricing (as well as typical LWR O&M costs). It can be observed 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 strong dependence of LCOH on energy price indicates that energy price is a key variable in determining the economic viability of an LWR-HTSE hydrogen production plant.

In addition to the energy costs, Figure 14 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 (stack replacement costs, maintenance costs, property tax and insurance costs, etc., are a function of the capital costs) as described in Section 3.2. The wide range of capital costs evaluated in the sensitivity analysis encompass capital cost estimates reported by other research organizations, stack manufacturers, etc. [5, 25, 30, 31], and are intended to provide perspective on the plausible range of LCOH changes that could result from capital cost estimates that differ from those estimated in the current analysis. The stack costs represent a subset of the capital costs, and therefore have a significant, but less impactful, effect on the LCOH than the overall system costs. The high stack cost value corresponds to the HFTO Hydrogen Production Record [5] Current Technology case, while the low stack cost value was calculated based on data reported by Tang et al. [25] for an SOEC stack module designed for manufacture in a mass production facility. Similarly, the learning rate is a parameter that affects the cost of the subset of equipment designated as "modular." Note that the stack cost is held constant in this analysis and changes to the learning rate do not affect the stack cost such that over the range of input parameters considered the LCOH is less sensitive to learning rate than the other capital cost related parameters (direct capital cost and stack cost).

5. NATURAL GAS STEAM METHANE REFORMING COMPARISON

5.1 NG-SMR LCOH Analysis

The incumbent competitor in commercial hydrogen production is the natural gas SMR process. As a result, the 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. For comparison, SMR LCOH costs were calculated using the H2A model [28] with input parameters defined in Table 12. Baseline SMR plant installed capital costs of $196,940,000 (in 2009 dollars) for a 316 tonne hydrogen/day production plant with hydrogen product purity of 99.9 mol% [32, 33] were adjusted for different production capacities using a scaling exponent of 0.6 and indexed to 2020 dollars using the Chemical Engineering Plant Cost Index [27].

Table 12. H2A model input parameters for SMR LCOH analysis.

Input Parameter	Value
Natural gas price	Varies
Plant capacity	680,000 kg/day
Start-up year	2025
Construction period	3 years
Start-up period	1 year
Plant life	40 years
Depreciation schedule	20-year MACRS
Equity financing	40%
Interest rate on debt	3.7%
Fixed operating costs during start-up	75%
Variable operating costs during start-up	75%
Revenues during start-up period	50%
Decommissioning costs	10%
Salvage value	10%
Inflation rate	1.9%
After-tax real IRR	10%
State tax rate	6%
Federal tax rate	21%
Total tax rate	25.74%

SMR plant capital costs were extrapolated (using a 0.6 scaling exponent) outside the H2A recommended range for plant capacity values below 235 tonne/day. For plant capacities above, the H2A 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.

5.2 Impact of Natural Gas Price

While fuel costs are low for an NPP, they are the main contributor for a natural gas plant. Natural gas prices historically have seen much variability. As a result, three conditions are considered in this subsection: (1) a medium gas price (which corresponds to the U.S. Energy Information Agency (EIA) 2021 Annual Energy Outlook (AEO) [34] Reference Case), (2) a low price corresponding to the EIA 2021 AEO High Oil and Gas Supply Case, and (3) a high natural gas price corresponding to the 2021

AEO Low Oil and Gas Supply Case. A plot of each of these natural gas price projections versus time is shown in Figure 15.

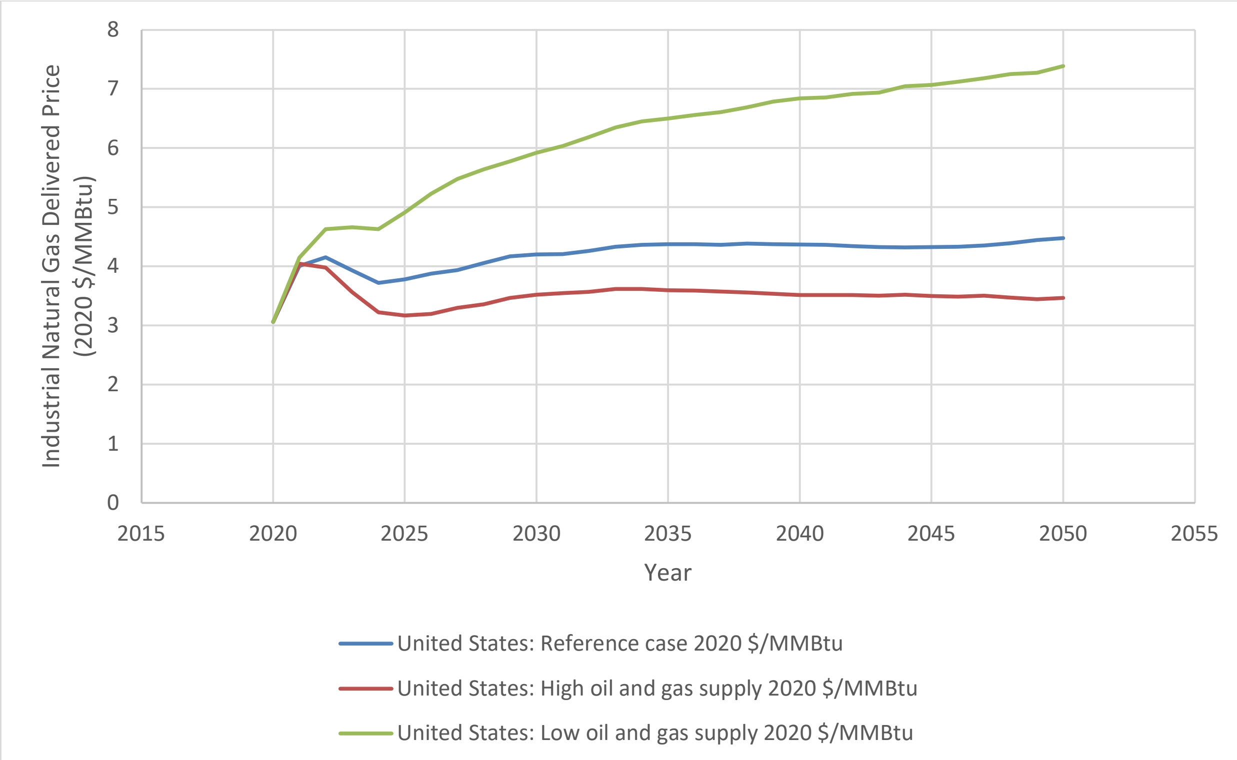

Figure 15. Projected industrial natural gas pricing as reported in selected EIA 2021 AEO Analysis Cases [34].

Figure 16 includes plots of SMR LCOH for each of the EIA 2020 AEO natural gas price projection cases as a function of SMR plant capacity. The plant capacity scaling range recommended by the H2A model falls between the vertical dotted lines. As previously described, the H2A model was modified to account for use of multiple process trains for SMR plant capacities above the suggested plant scaling capacity. As a result of the H2A 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. These results show that for a 600 to 700 MT/day hydrogen plant considered the LCOH of an SMR plant could be approximately \$1.15 to \$1.55/kg-H_2 depending on the price of natural gas.

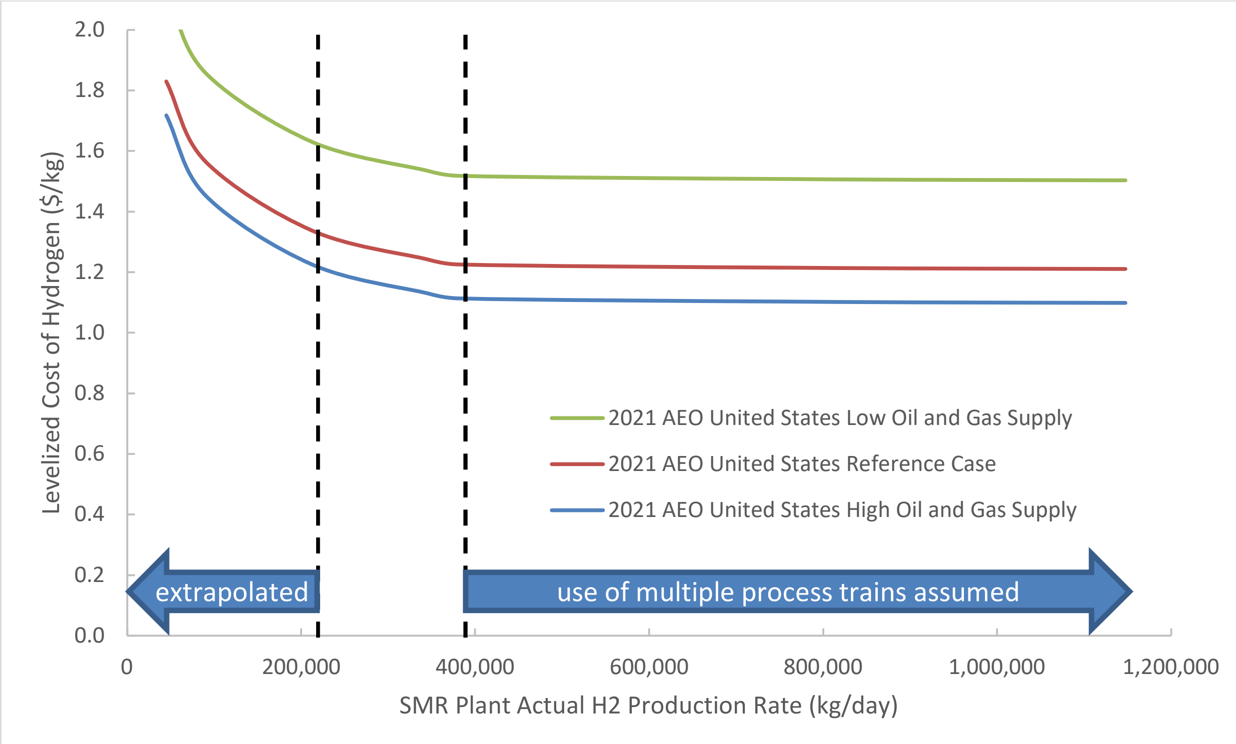

Figure 16. LCOH of SMR-based hydrogen production as a function of plant capacity and industrial natural gas pricing.

5.3 Cost of Carbon

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 [35]. For a hypothetical carbon price of $25/tonne-$CO_2$ [36] this corresponds to an added $0.22/kg-$H_2$ to hydrogen produced via SMR, while for a carbon price of $100/tonne-$CO_2$ (as may be required for deep decarbonization [37]) the NG-SMR LCOH would increase by $0.89/kg-$H_2$. In general, every $10/tonne-$CO_2$ unit increase in the carbon tax increases the NG-SMR LCOH by approximately $0.10/kg-$H_2$.

If SMR plants were to implement carbon capture and sequestration (CCS) the resulting LCOH of SMR + CCS could increase by a value of $0.34/kg-$H_2$ (600 tonne-H_2/day production) at the low end to $0.66/kg-$H_2$ (200 tonne-H_2/day) at the high end for a CO_2 transport distance of 100 miles. CCS becomes more costly with increased transport distance, and less costly for increased SMR plant capacities due to the economies of scale of the capture and transport equipment [28].

6. COMPARISON OF LWR-HTSE AND NG-SMR LCOH

The LCOH for the generic GW-scale LWR-HTSE plant is plotted as a function of the electricity cost (the sensitivity variable with the greatest impact on HTSE LCOH) in Figure 17. Figure 17 also includes the LCOH for a comparably sized NG-SMR plant with selected natural gas feedstock prices, while Figure 18 provides comparable data for a NG-SMR plant with CCS. The product outlet pressure is approximately 20 bar for both the LWR-HTSE and NG-SMR cases, and no high-pressure compression is included for either plant type. Additionally, no product transportation costs are included for either plant type.

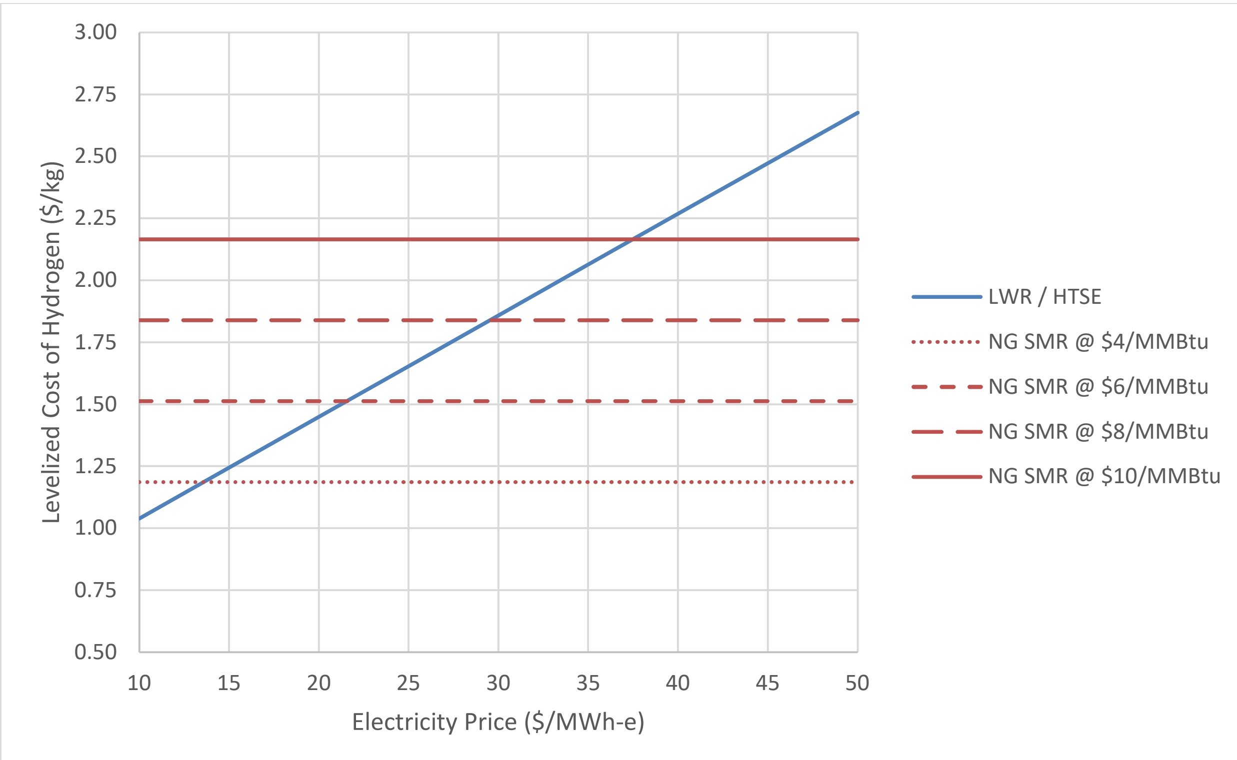

Figure 17. LCOH of 611 tonne/day production LWR-HTSE versus electricity price. LCOH of 612 tonne/day production SMR plant at selected natural gas prices included for comparison purposes.

It is apparent from Figure 17 that with electricity pricing of $30/MWh-e, an average natural gas price of approximately $8/MMBtu would be required for the LWR-HTSE plant to produce hydrogen on a cost competitive basis. If the LWR were able to provide electrical power to the HTSE plant at a price in the low $20/MWh-e range, which is close to the projected range of O&M costs attainable for LWRs [7], then the LWR-HTSE plant could complete with NG-SMR plants with NG feedstock prices of approximately $6/MMBtu. While the average NG price projected by the 2021 AEO High Oil & Gas Availability and Reference Cases is below these NG feedstock price points, the average NG price of $6.13/MMBtu projected by the 2021 AEO Low Oil & Gas Availability Case (between years 2020 to 2050) is in the approximate range of the NG price at which an HTSE plant obtaining low cost power from and LWR NPP could be cost competitive.

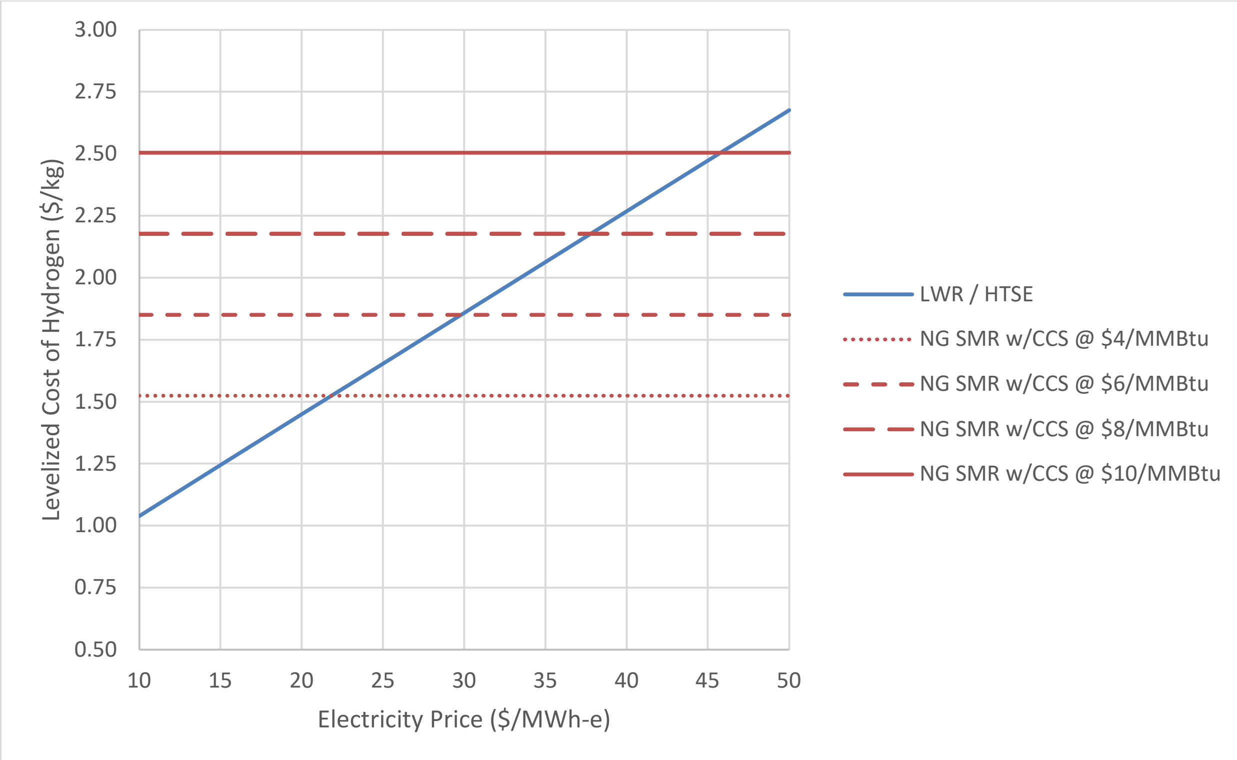

Figure 18. LCOH of 611 tonne/day production LWR-HTSE versus electricity price. LCOH of 612 tonne/day production SMR plant with CCS at selected NG prices included for comparison purposes.

When carbon capture and sequestration is included as part of the NG-SMR process, the LCOH increases accordingly. Figure 18 indicates that an NG-SMR plant with CCS would require NG feedstock pricing in the range of approximately $4/MMBtu to $6/MMBtu to be cost competitive with an HTSE plant purchasing electricity from an LWR NPP at a price of approximately $22/MWh-e to $30/MWh-e. These LWR NPP electricity prices are well within the range of O&M costs that future cost reductions could achieve [7]. Additionally, this range of natural gas pricing encompasses the average value of $4.19/MMBtu projected by the 2021 AEO Reference Case between the years 2020 and 2050.

Table 13 provides the average natural gas pricing at which the LCOH for LWR-HTSE plants would achieve LCOH parity with comparably sized (~610 tonne per day [tpd] H_2 production) NG-SMR plants both with and without CCS. The data in this table is provided as a function of the price at which the HTSE plant could obtain electricity from the LWR NPP.

Table 13. Calculation of NG prices at which 611 tpd LWR-HTSE and 612 tpd NG-SMR plants (with and without CCS) would achieve similar LCOH.

LWR-HTSE Power Price ($/MWh-e)	LCOH ($/kg)	SMR NG Breakeven Price ($/MMBtu)	SMR w/CCS NG Breakeven Price ($/MMBtu)
20	$1.45	$5.61	$3.54
25	$1.65	$6.86	$4.79
30	$1.86	$8.11	$6.04
35	$2.06	$9.37	$7.30
40	$2.27	$10.62	$8.55

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 the 2021 AE NG price projection, regulatory drivers to require CCS for NG-SMR based hydrogen production would likely be necessary for a GW-scale LWR-HTSE plant to produce hydrogen on a cost competitive basis. However, due to global events that have occurred in the time since the 2021 AEO was published, including COVID-related supply chain issues and current geopolitical events such as the 2022 Russian invasion of Ukraine, energy prices have generally increased. Continued pricing increases and/or sustained pricing of current natural gas prices will generally drive up the cost of NG-SMR and provide increased opportunities for LW-HTSE to produce clean hydrogen at a competitive cost.

Additionally, a government-provided clean hydrogen production credit or subsidy would further allow LWR-HTSE to produce hydrogen at a price competitive with NG-SMR. The existence of such a credit could allow LWR-HTSE to produce hydrogen at costs competitive with NG-SMR even without a carbon tax or other regulation that would otherwise drive NG-SMR plants to implement CCS technology. As previously noted, each kg of hydrogen produced by NG-SMR releases approximately 8.9 kg of CO_2 per kg of H_2 produced. This value is representative of a modern, high-tech NG-SMR plant; older, less efficient NG-SMR plants likely release closer to 10 kg of CO_2 per kg of H_2 produced. A government-offered subsidy or production credit of approximately $75/tonne-$CO_2$ of avoided carbon emissions (equal to about $0.67/kg-$H_2$) could allow a GW-scale NOAK LWR-HTSE plant with a power cost of $30/MWh-e to produce hydrogen at a cost competitive with an NG-SMR plant having an average NG feedstock cost of $4/MMBtu (unabated carbon emissions).

A final strategy that could be employed by LWR-HTSE plants to increase economic viability would be to operate the HTSE unit as a dispatchable load that could be placed in standby mode during periods of peak electrical demand. During the peak periods, the LWR could dispatch electrical power to the grid at a premium price to increase overall revenues. This operating strategy may require the addition of hydrogen storage to prevent interruption in the delivery of the hydrogen product to customers. Due to the cost implications associated with high-cost hydrogen storage and decreasing the operating capacity factor of the HTSE plant, this operating scheme requires investigation of electricity and hydrogen market pricing to optimize the hydrogen storage capacity and the HTSE plant operating schedule. These topics have been investigated in previous studies [1, 38, 39], and the analysis provided by this report may be used to support future studies on this topic.

7. LWR-HTSE $1/kg LCOH Target

To show a potential path to reach a hydrogen production cost of $1/kg-$H_2$ a possible scenario informed by the sensitivity analysis was constructed and added to the 'waterfall' chart in Figure 19. While the tornado chart presented in Section 4.2 identifies the LCOH changes that could result from changes to individual parameters, the waterfall chart in Figure 19 illustrates the cumulative LCOH decrease that could be achieved by combining multiple price-decreasing parameter changes. Note that the value of the capital cost specified in the waterfall chart differs from the lower bound specified in the tornado chart. While the pathway to achieve $1/kg hydrogen from HTSE is viewed as aggressive, the parameters required to achieve this metric are not unfeasible. In addition to the reductions in energy price, operating parameters, and capital and operating costs, it may also be possible to obtain an additional source of revenue from oxygen byproduct sales or clean hydrogen production credits that could maintain the prospect of $1/kg hydrogen from HTSE in the event challenges are encountered in achieving the parameter specifications detailed in Figure 19.

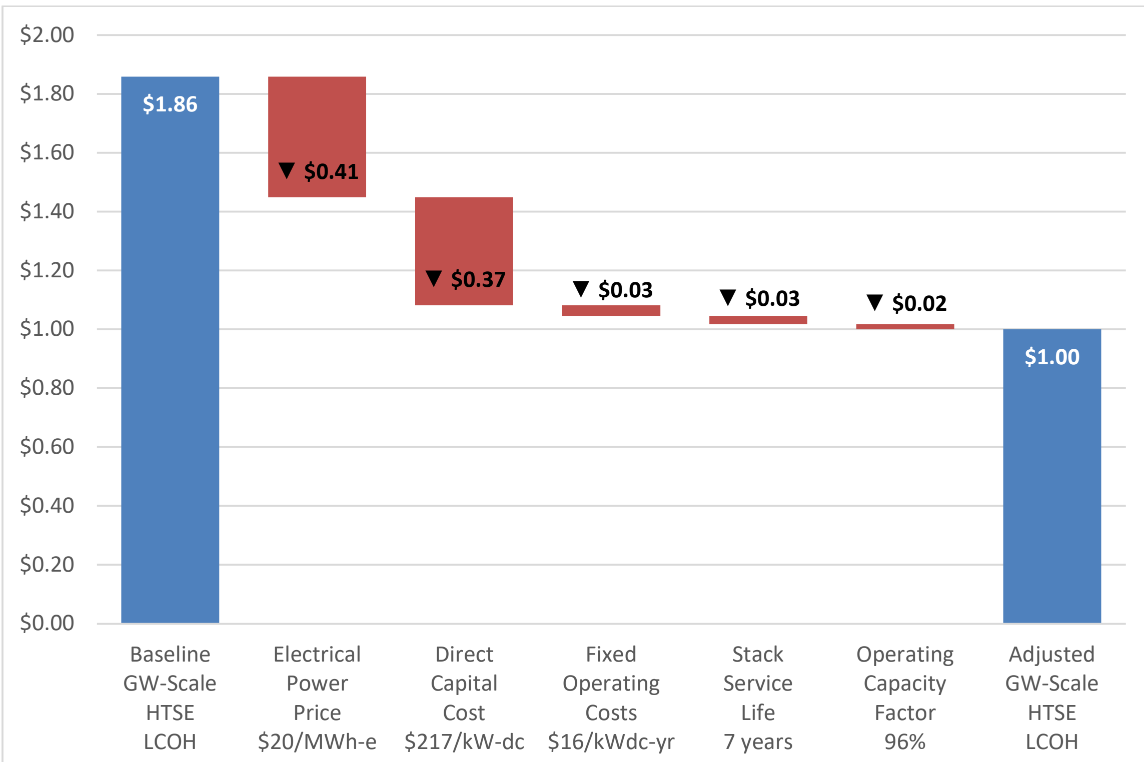

Figure 19. Waterfall chart illustrating a potential pathway to achieve an LCOH of $1/kg

As noted, and as shown in the sensitivity analysis in Figure 14, the levelized cost of electricity/energy (LCOE) is the most significant factor in the calculation of LCOH and therefore the most significant factor in the profitability of any system that aspires to produce hydrogen by electrolysis. Although historical LCOE values for nuclear power plants have been around $30/MWh there is significant area for improvement in these costs. This is because many factors, including regulation uncertainty, have led to a low degree of cost reduction initiatives in the nuclear power industry compared to other industries. Also, until the past decade, the nuclear industry has not been under the extreme price and competitive pressure that it is under now.

Because of current price pressure on the nuclear industry that has caused some NPPs to prematurely close, many studies have been done to outline roadmaps for decreasing the operating costs of nuclear power production. One such study [7] uses an 'Integration Options for Nuclear' (ION) approach to outline various possible improvements to nuclear power to reduce operating costs in the areas of technology, process, human performance, and governance. The ION Generation I analysis considered technologies and options that would be viable within the 3–5-year time frame. Table 14 is reproduced from the mentioned reference with permission.

Table 14. Preliminary LCOE analysis showing identified pathways to reducing NPP operating costs

	Scenario 2: **ION-Gen1 LCOE with Sustaining and Innovation Capital**	Scenario 3: **Significantly Reduced Capital**	Scenario 4: **Reduced Capital, Aggressive Reduction of Fixed O&M**	Scenario 5: **Reduced Capital, Improved Cost of Capital**	Scenario 6: **Nuclear Production Tax Credit**
Generation Source	Nuclear	Nuclear	Nuclear	Nuclear	Nuclear
Plant Size (MW)	2200	2200	2200	2200	2200
Capacity Factor (%)	93	93	93	93	93
Fuel Cost ($/MMBtu)	0.65	0.65	0.65	0.65	0.65
Heat Rate (Btu/kWh)	10,300	10,300	10,300	10,300	10,300
Fixed O&M ($/kW-year)	71.36	71.36	64.55*	71.36	71.36
Variable O&M ($/MWh)	3.00	3.00	3.00	3.00	3.00
Overnight Costs ($/kW)	455 ($1B investment)	186 ($410M investment)	239 ($525M investment)	273 ($500M investment)	455 ($1B investment)
Interest Rate (%)	9.6	9.6	9.6	7.6	9.6
Production Tax Credit (PTC) ($/MWh)	0	0	0	0	$2.88
Levelized Cost of Energy ($/MWh)	**25.87**	**21.49**	**21.51**	**21.68**	**21.50**

The LCOE values in Table 14 represent industry averages. Single nuclear power plant operators have plans to have or already have LCOE values around $20/MWh. Full details can be found in the referenced document, but suffice to say, there is credible evidence to say that the LCOE average for the nuclear power industry will soon be on the lower end of the $20/MWh to 30/MWh range, which will, in turn, make hydrogen production via high-temperature electrolysis using nuclear electricity very competitive.

8. SUMMARY

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 plant with 1000 MW-dc of electrolysis capacity would require a total electrical power input of 1076 MW-dc and a thermal power input of 188.2 MW-t. Steam flow from the NPP is used to drive the turbines to produce electrical power as well as to provide the thermal input to the HTSE plant. Of the portion of NPP steam flow ultimately used to provide thermal or electrical power input to the HTSE plant, approximately 5% is used for HTSE process heat input needed to vaporize the HTSE process feedwater while the remainder is used to drive the steam turbines to provide electrical power input. The analysis specified the 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 36.8 kWh-e/kg-H_2 and 6.4 kWh-t/kg-H_2, respectively, which excludes any additional power required for application-specific high-pressure product transport/storage compression. The HTSE plant efficiency was calculated as 90.2% on an HHV basis. An economic analysis of the HTSE process was performed based on the assumption of a steady-state (non-grid integrated) operating condition.

The SOEC stack cost is based on the value of $78/kW-dc reported for an electrode-supported cell construction with a 1,000 MW/yr manufacturing rate [2] and a stack service life specification of 4 years consistent with HFTO Hydrogen Production Record #20006 [5]. This analysis includes annual stack replacements to restore the HTSE plant design capacity rating at the start of each operating year. The GW-scale NOAK HTSE plant with a hydrogen production design capacity of 702 tonne H_2/day (1000 MW-dc; 1076 MW-ac) has DCC of $544/kW-dc (includes assumptions on HTSE plant equipment and nuclear plant heat and power delivery equipment) and a total capital investment of $703/kW-dc (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 the corresponding thermal energy price), the HTSE plant could produce hydrogen at an LCOH of $1.86/kg, which does not include high-pressure product hydrogen compression beyond 20 bar, product storage or transportation costs. A summary of the assumptions and results for the GW-scale HTSE process and economic analysis is shown below in Table 15.

Table 15. Summary of Base HTSE Model Design Case.

Description	Value	Note
Plant Design Capacity	702 tonnes/day	99.9 mol% hydrogen at 20 bar
Power Requirements	1000 MW-dc 1076 MW-ac	DC power corresponds to stack power input; AC power corresponds to total power requirement
Operating Capacity Factor	87.1%	Accounts for plant shut-downs as well as cell degradation
Actual Hydrogen Production Rate	611 tonnes/day	
Efficiency (HHV)	90.2%	Includes both thermal and electrical energy consumption
Stack Operating Pressure	5 bar	Based on maximizing system efficiency by trending operating pressure and steam utilization versus system efficiency
Steam Utilization (conversion of reactant steam)	80%	
Electricity Required	36.8 kWh-e/kg-H_2	
Thermal Energy Required	6.4 kWh-t/kg-H_2	
Technology Horizon	NOAK, 95% learning rate	Nth-of-a-Kind defined as 2.5 GW-e of previous HTSE plant installations
Stack Cost	$78/kW-dc	Electrode-supported with 1,000 MW/yr manufacturing rate
Stack Service Life	4 years	Assumes annual stack replacements to restore the HTSE plant design capacity rating at the start of each operating year
Direct Capital Cost	$544/kW-dc	GW-scale NOAK Plant
Total Capital Investment	$703/kW-dc	GW-scale NOAK Plant
Levelized Cost of H_2 (HTSE)	$1.86/kg	At $30/MWh electricity cost; excludes application- and/or site-specific product storage and transport costs. Does not include high-pressure product hydrogen compression beyond 20 bar.

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 electricity price and HTSE plant direct capital costs. A second set of variables including the stack cost, learning rate (for decreases in modular equipment costs as a function of the number of units produced by the equipment manufacturer), IRR, total fixed operating costs, stack service life, and capacity factor 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-dc 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.
- Direct capital costs are also a major driver of the HTSE 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 (stack replacement costs, maintenance costs, property tax and insurance costs, etc., are estimated as a function of the capital costs).
- The stack costs represent a subset of the capital costs, and therefore have a significant, but less impactful, effect on the LCOH than the overall system costs. The stack pricing of \$78/kW-dc considered in this analysis is sufficiently low that the balance-of-plant capital costs represent a larger opportunity for cost reductions. However, if the stack cost were approximately doubled from the baseline value to a stack cost of \$155/kW-dc as estimated in the HFTO Record #20006 Current Technology Case [5], the LCOH would increase by nearly \$0.18/kg, or nearly 10%, due to the increased initial capital cost as well as the recurring stack replacement costs. Therefore, a prospective HTSE plant developer could significantly reduce uncertainties in hydrogen production costs 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 hundreds to thousands of metric tonnes of hydrogen production per day.

A comparison of LWR-HTSE and natural gas SMR LCOH was performed to identify cases where HTSE could produce hydrogen at a cost that is competitive with SMR. The SMR LCOH is highly dependent on natural gas pricing. With electricity pricing of \$30/MWh-e, an average natural gas price of approximately \$8/MMBtu would be required for the LWR-HTSE plant to produce hydrogen on a cost competitive basis. If the LWR were able to provide electrical power to the HTSE plant at a price in the low \$20/MWh-e range, which is close to the projected range of O&M costs attainable for LWRs, then the LWR-HTSE plant could compete with NG-SMR plants with NG feedstock prices of approximately \$6/MMBtu. While the average NG price projected by the 2021 AEO High Oil & Gas Availability and Reference Cases are below these NG feedstock price points, the average NG price projected by the 2021 AEO Low Oil & Gas Availability Case is in the approximate range of the NG price at which an HTSE plant obtaining low-cost power from and LWR NPP could be cost competitive.

Because hydrogen produced via SMR is associated with significant carbon emissions, a carbon price 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 indicate that a carbon tax of \$25/tonne-CO_2 would result in a \$0.22/kg increase in the natural gas SMR LCOH. In addition to the electricity price and HTSE plant capital costs, 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. If an NG-SMR plant were to utilize CCS to reduce carbon emissions, the additional capital and operating costs would increase the NG-SMR LCOH by $0.34/kg for a plant with a production capacity approximately equal to that of the LWR-HTSE plant considered (the NG-SMR LCOH increase due to CCS is greater for smaller capacity NG-SMR plants due to the economies of scale associated with the CO_2 capture and transport equipment). Alternatively, a clean hydrogen production subsidy or credit could also reduce the LWR-HTSE LCOH in order to provide additional opportunities for clean nuclear-based hydrogen production to be cost competitive with NG-SMR hydrogen. Finally, some customers may be willing to pay a price premium for clean, carbon-free hydrogen.

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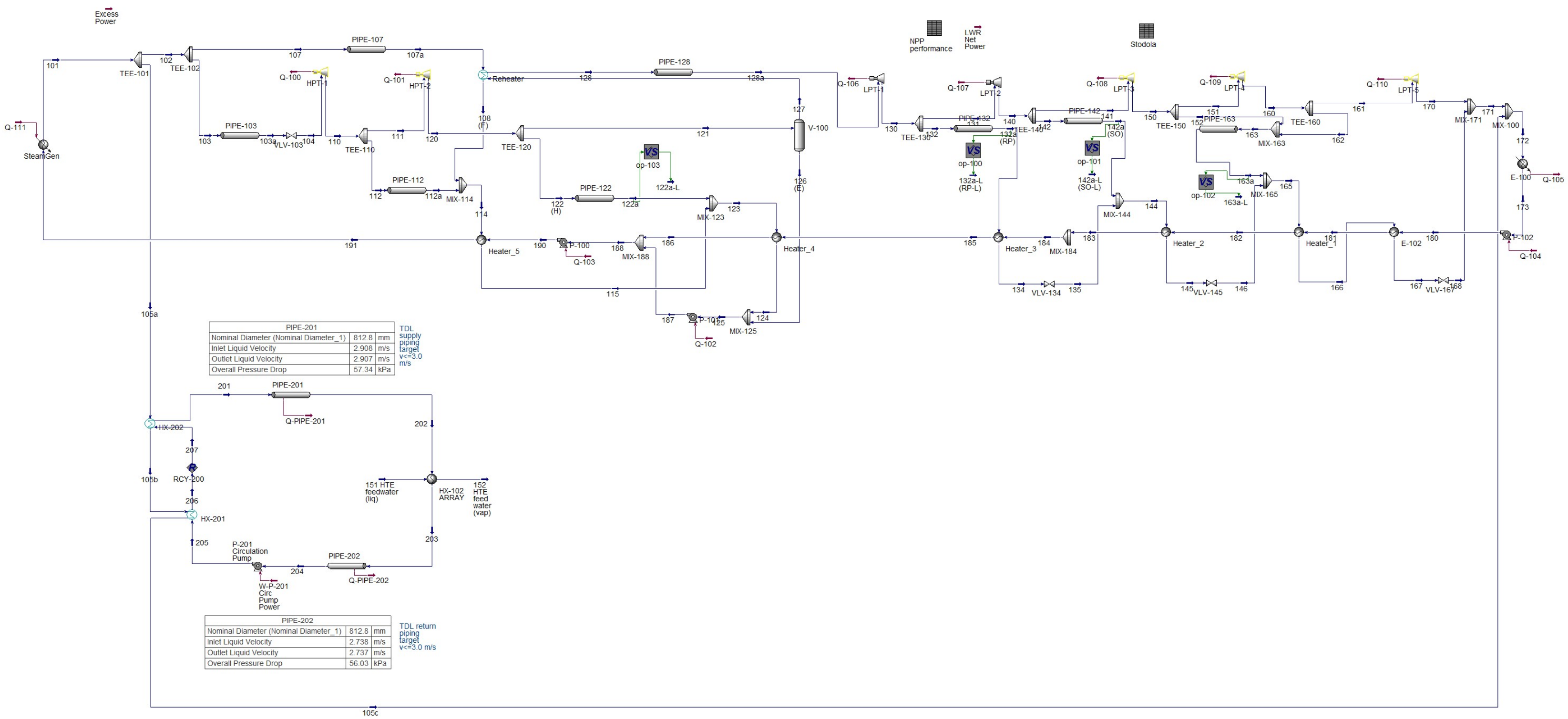
Appendix A
Process Flow Diagrams

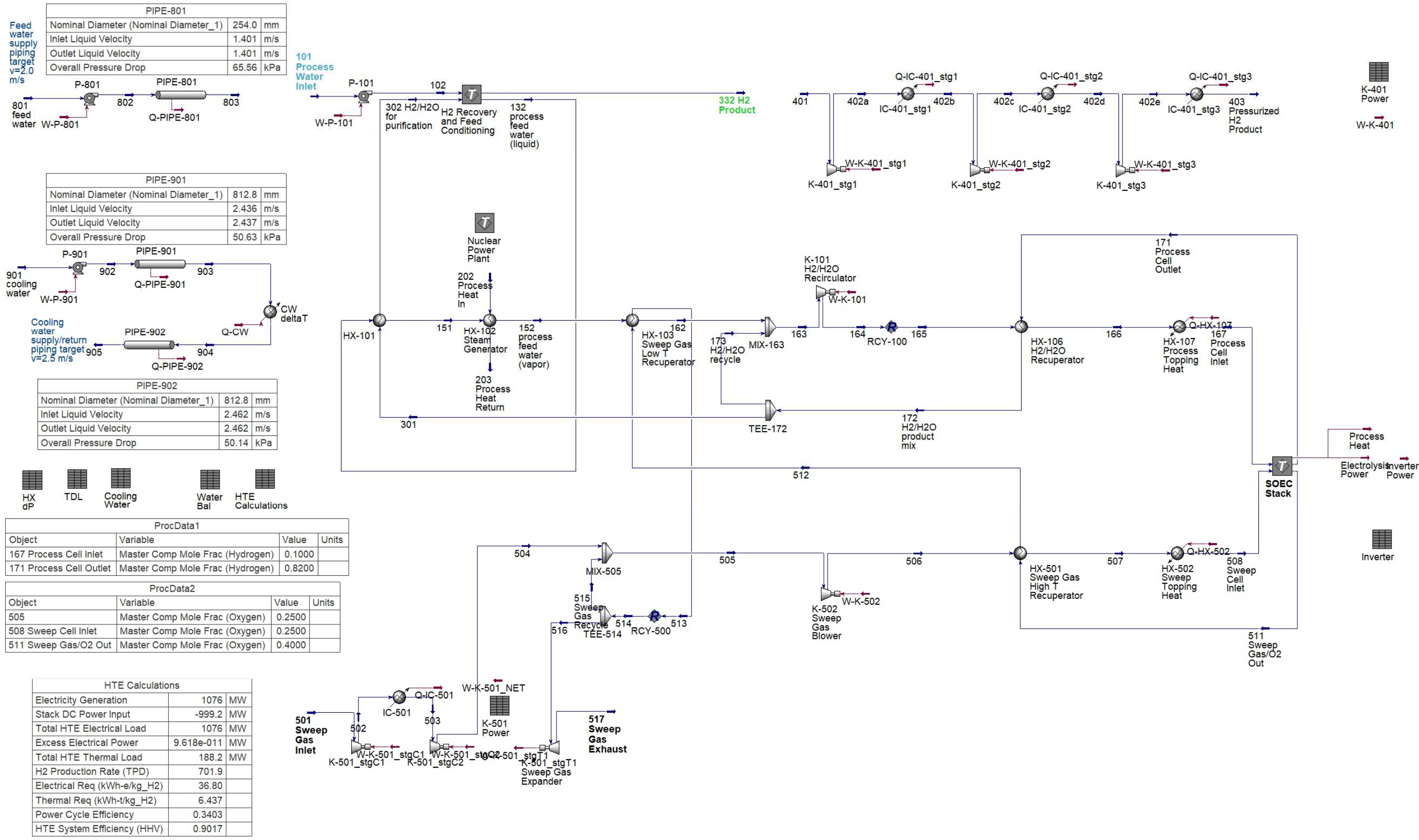
Nuclear Power Plant and Thermal Delivery Loop



PIPE-201		
Nominal Diameter (Nominal Diameter_1)	812.8	mm
Inlet Liquid Velocity	2.908	m/s
Outlet Liquid Velocity	2.907	m/s
Overall Pressure Drop	57.34	kPa

PIPE-202		
Nominal Diameter (Nominal Diameter_1)	812.8	mm
Inlet Liquid Velocity	2.738	m/s
Outlet Liquid Velocity	2.737	m/s
Overall Pressure Drop	56.03	kPa

High-Temperature Steam Electrolysis



PIPE-801		
Nominal Diameter (Nominal Diameter_1)	254.0	mm
Inlet Liquid Velocity	1.401	m/s
Outlet Liquid Velocity	1.401	m/s
Overall Pressure Drop	65.56	kPa

PIPE-901		
Nominal Diameter (Nominal Diameter_1)	812.8	mm
Inlet Liquid Velocity	2.436	m/s
Outlet Liquid Velocity	2.437	m/s
Overall Pressure Drop	50.63	kPa

PIPE-902		
Nominal Diameter (Nominal Diameter_1)	812.8	mm
Inlet Liquid Velocity	2.462	m/s
Outlet Liquid Velocity	2.462	m/s
Overall Pressure Drop	50.14	kPa

ProcData1

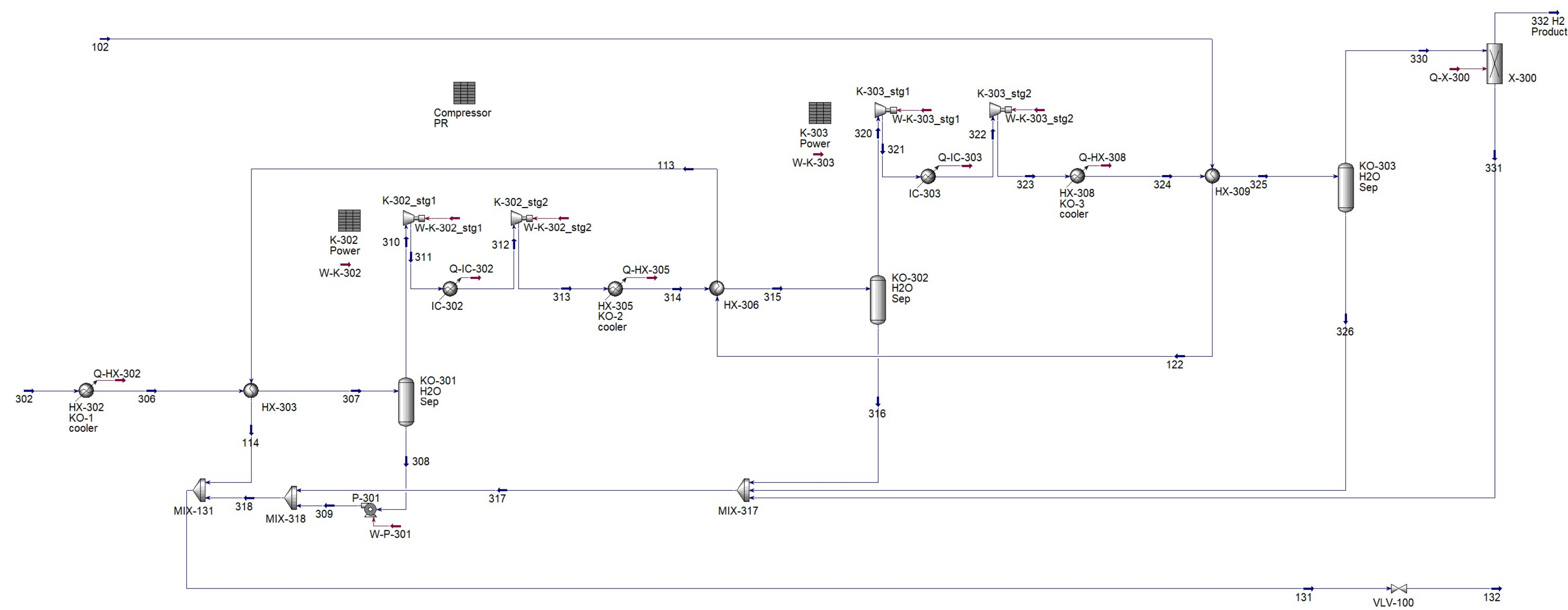
Object	Variable	Value	Units
167 Process Cell Inlet	Master Comp Mole Frac (Hydrogen)	0.1000	
171 Process Cell Outlet	Master Comp Mole Frac (Hydrogen)	0.8200	

ProcData2

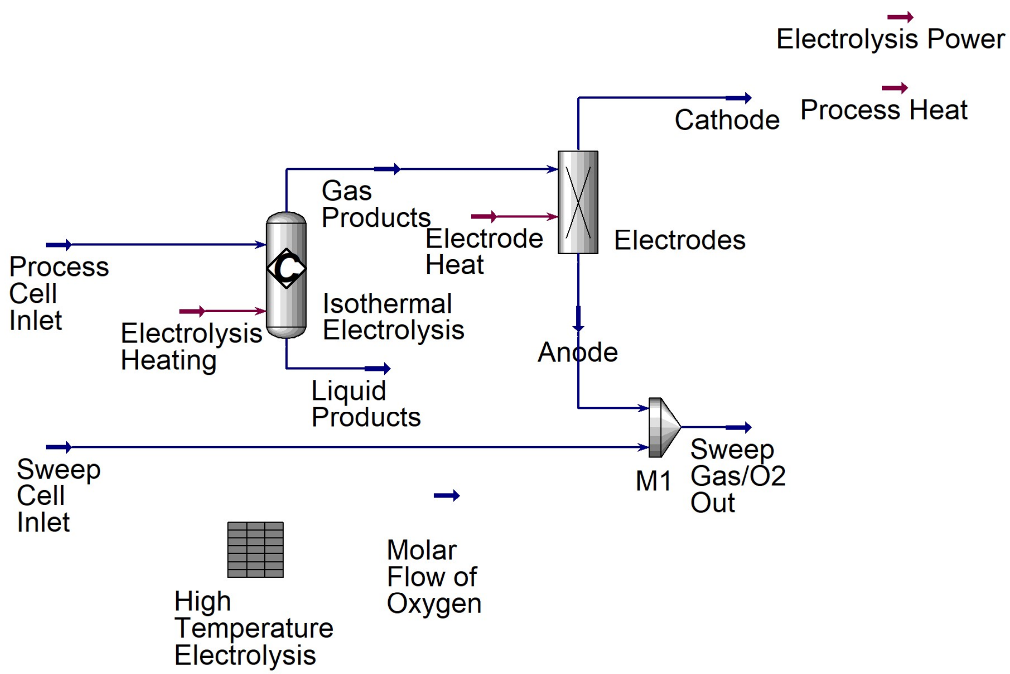
Object	Variable	Value	Units
505	Master Comp Mole Frac (Oxygen)	0.2500	
508 Sweep Cell Inlet	Master Comp Mole Frac (Oxygen)	0.2500	
511 Sweep Gas/O2 Out	Master Comp Mole Frac (Oxygen)	0.4000	

HTE Calculations		
Electricity Generation	1076	MW
Stack DC Power Input	-999.2	MW
Total HTE Electrical Load	1076	MW
Excess Electrical Power	9.618e-011	MW
Total HTE Thermal Load	188.2	MW
H2 Production Rate (TPD)	701.9	
Electrical Req (kWh-e/kg_H2)	36.80	
Thermal Req (kWh-t/kg_H2)	6.437	
Power Cycle Efficiency	0.3403	
HTE System Efficiency (HHV)	0.9017	

Product Recovery



Solid Oxide Electrolyzer Cell



Appendix B
Equipment Costs

Equipment	Conv or Mod	Process System	APEA model	Installed Cost (NOAK plant, 2020$)	Reference
HTSE Vessel Shell	Mod	HTSE System	HT HORIZ DRUM	$60,927,166	[3]
HTSE Vessel Isolation Valves	Mod	HTSE System		$15,366,734	[23]
SOE Cells	Mod	HTSE System		$77,937,600	[2]
SOEC Module Assembly	Mod	HTSE System		$9,441,348	[23]
SOEC Electrical Connector Assemblies	Mod	HTSE System		$3,056,511	[23]
Sleeved Process Connections	Mod	HTSE System		$11,231,338	[23]
HX-501 Sweep Gas High-Temperature Recuperator	Mod	HTSE System		$5,380,082	[23]
HX-106 H_2/H_2O Recuperator	Mod	HTSE System		$6,211,766	[23]
HX-502 Sweep Gas Topping Heater	Mod	HTSE System		$2,630,104	[23]
HX-107 H_2/H_2O Topping Heater	Mod	HTSE System		$8,448,673	[23]
Control Module	Mod	HTSE System		$131,334	[25]
Thermocouples	Mod	HTSE System		$65,667	[25]
Pressure Sensors	Mod	HTSE System		$985,003	[25]
HTSE Block Container (shipping container)	Mod	HTSE System		$256,036	[40]
Rectifier/Power Supply	Mod	Feed & Utility System		$106,920,755	[25]
Disconnect Switch	Mod	Feed & Utility System	BELSDISCNCT SW	$117,094	[3]
Transformer	Mod	Feed & Utility System	BELSTRANSFORM	$12,949,331	[3]
Switch Board	Mod	Feed & Utility System	BELSSWITCH BRD	$2,015,921	[3]
DC Bus Power Distribution	Mod	Feed & Utility System	BELSBUS DUCT	$3,893,495	[3]
Power Pole Lines	Conv	Feed & Utility System	BELSPOLE LINE	$928,884	[3]
Purified Water Storage Tank	Mod	Feed & Utility System	VT STORAGE	$5,438,570	[3]
Water Pretreatment Filter/Softener System	Conv	Feed & Utility System		$2,322,323	[23]
Water Treatment RO/EDI System	Conv	Feed & Utility System		$10,237,384	[23]
CT-901 Cooling Tower	Conv	Feed & Utility System	CTWCOOLING	$1,491,673	[3]
Air Filter	Mod	Air Sweep Gas System		$65,658	[25]
K-501C Sweep Gas Compressor	Mod	Air Sweep Gas System	AC CENTRIF M	$25,404,593	[3]
K-501T Sweep Gas Exhaust Turbine	Mod	Air Sweep Gas System	TURTURBOEXP	$26,323,151	[3]
K-502 Sweep Gas Blower	Mod	Air Sweep Gas System	FN PROPELLER	$947,623	[3]
P-101 Water Pump	Mod	Hydrogen/Steam System	CP CENTRIF	$1,248,870	[3]
F-101 Water Filter	Mod	Hydrogen/Steam System	F CARTRIDGE	$541,539	[3]
DI Polisher	Mod	Hydrogen/Steam System		$197,001	[25]
Water Flow Meter	Mod	Hydrogen/Steam System		$328,334	[25]
HX-101 Condenser & Water Preheater	Mod	Hydrogen/Steam System	HE TEMA EXCH	$3,338,965	[3]

Equipment	Conv or Mod	Process System	APEA model	Installed Cost (NOAK plant, 2020$)	Reference
HX-102 Feed Water Vaporizer	Mod	Hydrogen/Steam System	HE TEMA EXCH	$6,161,049	[3]
HX-103 Sweep Gas Low Temp Recuperator	Mod	Hydrogen/Steam System	HE TEMA EXCH	$6,503,370	[3]
K-101 Hydrogen Recycle Blower	Mod	Hydrogen/Steam System	FN PROPELLER	$880,034	[3]
HX-303 Feedwater Heater #1	Mod	Hydrogen Purification System	HE TEMA EXCH	$2,532,805	[3]
HX-306 Feedwater Heater #2	Mod	Hydrogen Purification System	HE TEMA EXCH	$2,447,934	[3]
HX-309 Feedwater Heater #3	Mod	Hydrogen Purification System	HE TEMA EXCH	$3,082,643	[3]
HX-302 Separation Vessel #1 Precooler	Mod	Hydrogen Purification System	HE TEMA EXCH	$2,640,596	[3]
HX-305 Separation Vessel #2 Precooler	Mod	Hydrogen Purification System	HE TEMA EXCH	$2,549,418	[3]
HX-308 Separation Vessel #3 Precooler	Mod	Hydrogen Purification System	HE TEMA EXCH	$2,606,864	[3]
P-301 KO-1 Outlet Pump	Mod	Hydrogen Purification System	CP CENTRIF	$1,002,486	[3]
K-302 H_2 Purification Multistage Compressor #2	Mod	Hydrogen Purification System	GC RECIP MOTR	$28,209,794	[3]
K-303 H_2 Purification Multistage Compressor #3	Mod	Hydrogen Purification System	GC RECIP MOTR	$23,647,539	[3]
KO-301 H_2 Separation Vessel #1	Mod	Hydrogen Purification System	VT CYLINDER	$4,365,200	[3]
KO-302 H_2 Separation Vessel #2	Mod	Hydrogen Purification System	VT CYLINDER	$4,228,882	[3]
KO-303 H_2 Separation Vessel #3	Mod	Hydrogen Purification System	VT CYLINDER	$4,328,849	[3]
Backup Electric Boiler	Mod	NPH Delivery System		$6,871,164	[41]
PIPE-201 Nuclear Process Heat Piping (supply)	Conv	NPH Delivery System	BPIPIPE	$11,754,884	[3]
PIPE-202 Nuclear Process Heat Piping (return)	Conv	NPH Delivery System	BPIPIPE	$11,754,884	[3]
P-201 Nuclear Process Heat Circulation Pump	Conv	NPH Delivery System	CP CENTRIF	$1,441,999	[3]
HX-201 Nuclear Process Heat TDL HX	Conv	NPH Delivery System	HE TEMA EXCH	$869,499	[3]
HX-202 Nuclear Process Heat TDL HX	Conv	NPH Delivery System	HE TEMA EXCH	$4,154,673	[3]
Therminol-66 HTF	Conv	NPH Delivery System		$3,492,267	[42]
CB-101 Control Building	Conv	Control System	BCIVBUILDING	$498,879	[3]
OC-101 Operator Center	Conv	Control System	BINSOPER CENT	$329,225	[3]
Total				$543,135,463	

Appendix C
Process Model Stream Tables

BATTELLE ENERGY ALLIANCE Bedford, MA USA	Case Name:	Generic HTSE PFD_v3.00_Therm66_5bar_U80 40x25 MWe unit (INL-RP
	Unit Set:	HTSE PFD
	Date/Time:	Thu Mar 31 17:00:00 2022

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)	362.7	362.7	442.5	442.5	442.5
Mass Flow	(kg/h)	6534	6534	7969	7969	7969
Actual Volume Flow	(m3/h)	6.414	6.413	8.903	16.51	2821
Mass Density	(kg/m3)	1019	1019	895.1	482.6	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	328.8	328.8 *	707.8
Pressure	(bar)	4.900	4.900	5.206	5.206 *	5.102
Molar Flow	(kgmole/h)	442.5	503.7	503.7	503.7	503.7
Mass Flow	(kg/h)	7969	8269	8269	8269 *	8269
Actual Volume Flow	(m3/h)	4371	5009	4798	4798	8043
Mass Density	(kg/m3)	1.823	1.651	1.723	1.723	1.028

Name		167 Process Cell Inlet	171 Process Cell Outle	172 H2/H2O product n	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	343.8	343.8	247.1
Pressure	(bar)	5.000	5.000	4.900	4.900	3.427
Molar Flow	(kgmole/h)	503.7	503.7	503.7	61.21	437.9
Mass Flow	(kg/h)	8269	2466	2466	299.7	1.104e+005
Actual Volume Flow	(m3/h)	8983	8995	5276	641.0	129.5
Mass Density	(kg/m3)	0.9205	0.2742	0.4675	0.4675	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)	172.6	343.8	94.23	15.00	15.00
Pressure	(bar)	3.167	4.900	4.640	19.95	19.95
Molar Flow	(kgmole/h)	437.9	442.5	442.5	362.7	1.451e+004
Mass Flow	(kg/h)	1.104e+005	2166	2166	731.2	2.925e+004
Actual Volume Flow	(m3/h)	121.9	4635	2909	438.7	1.755e+004
Mass Density	(kg/m3)	905.1	0.4675	0.7448	1.667	1.667

Name		402a	402b	402c	402d	402e
Vapour Fraction		1.0000	1.0000	1.0000	1.0000	1.0000
Temperature	(C)	15.00	15.00 *	15.00	15.00 *	15.00
Pressure	(bar)	19.95	19.95	19.95	19.95	19.95
Molar Flow	(kgmole/h)	1.451e+004	1.451e+004	1.451e+004	1.451e+004	1.451e+004
Mass Flow	(kg/h)	2.925e+004	2.925e+004	2.925e+004	2.925e+004	2.925e+004
Actual Volume Flow	(m3/h)	1.755e+004	1.755e+004	1.755e+004	1.755e+004	1.755e+004
Mass Density	(kg/m3)	1.667	1.667	1.667	1.667	1.667

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)	15.00 *	20.00 *	109.4	50.00 *	147.7
Pressure	(bar)	19.95	1.013 *	2.190	2.146	4.640
Molar Flow	(kgmole/h)	1.451e+004	572.6	572.6	572.6	572.6
Mass Flow	(kg/h)	2.925e+004	1.652e+004 *	1.652e+004	1.652e+004	1.652e+004
Actual Volume Flow	(m3/h)	1.755e+004	1.376e+004	8314	7162	4321
Mass Density	(kg/m3)	1.667	1.200	1.987	2.306	3.823

* Specified by user.

Workbook: Case (Main) (continued)

Material Streams (continued)

Fluid Pkg: All

Name		505	506	507	508 Sweep Cell Inlet	511 Sweep Gas/O2 Ou
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)	725.3	725.3	725.3	725.3	906.6
Mass Flow	(kg/h)	2.104e+004	2.104e+004	2.104e+004	2.104e+004	2.684e+004
Actual Volume Flow	(m3/h)	5727	5311	1.252e+004	1.296e+004	1.620e+004
Mass Density	(kg/m3)	3.674	3.962	1.680	1.624	1.657

Name		512	513	514	515 Sweep Gas Recyc	516
Vapour Fraction		1.0000	1.0000	1.0000	1.0000	1.0000
Temperature	(C)	330.7	238.8	238.8 *	238.8	238.8
Pressure	(bar)	4.900	4.640	4.640 *	4.640	4.640
Molar Flow	(kgmole/h)	906.6	906.6	907.1 *	152.8	754.3
Mass Flow	(kg/h)	2.684e+004	2.684e+004	2.686e+004	4523	2.233e+004
Actual Volume Flow	(m3/h)	9301	8325	8329	1403	6927
Mass Density	(kg/m3)	2.886	3.224	3.224	3.224	3.224

Name		517 Sweep Gas Exhau	801 feed water	802	803	901 cooling water
Vapour Fraction		1.0000	0.0000	0.0000	0.0000	0.0000
Temperature	(C)	98.31	10.00 *	10.01	10.02	20.00 *
Pressure	(bar)	1.013 *	1.034 *	1.691 *	1.035	1.034 *
Molar Flow	(kgmole/h)	754.3	1.451e+004	1.451e+004	1.451e+004	2.339e+005
Mass Flow	(kg/h)	2.233e+004	2.613e+005	2.613e+005	2.613e+005	4.214e+006
Actual Volume Flow	(m3/h)	2.299e+004	256.6	256.6	256.6	4168
Mass Density	(kg/m3)	0.9715	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	20.02	34.00 *	34.01	800.0
Pressure	(bar)	2.073 *	1.567	1.536	1.034	5.000
Molar Flow	(kgmole/h)	2.339e+005	2.339e+005	2.339e+005	2.339e+005	181.3
Mass Flow	(kg/h)	4.214e+006	4.214e+006	4.214e+006	4.214e+006	5803
Actual Volume Flow	(m3/h)	4168	4168	4212	4212	3239
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 @Ce
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)	503.7	685.1	0.0000	181.3	503.7
Mass Flow	(kg/h)	2466	8269	0.0000	5803	8269
Actual Volume Flow	(m3/h)	8995	1.223e+004	0.0000	3239	8983
Mass Density	(kg/m3)	0.2742	0.6759	0.6759	1.791	0.9205

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)	725.3	906.6	0.1917	362.7	362.7
Mass Flow	(kg/h)	2.104e+004	2.684e+004	1.000 *	6534	6534
Actual Volume Flow	(m3/h)	1.296e+004	1.620e+004	6.265	6.413	6.559
Mass Density	(kg/m3)	1.624	1.657	0.1596	1019	996.2

* Specified by user.

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	94.23
Pressure	(bar)	9.876	10.88	9.189	6.400	4.640
Molar Flow	(kgmole/h)	362.7	362.7	442.5	442.5	442.5
Mass Flow	(kg/h)	6534	6534	7969	7969	2166
Actual Volume Flow	(m3/h)	6.664	6.460	8.650	8.903	2909
Mass Density	(kg/m3)	980.4	1011	921.3	895.1	0.7448
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		0.8824	0.8605	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)	442.5	442.5	61.75	61.75	380.8
Mass Flow	(kg/h)	2166	2166	1112	1112	1054
Actual Volume Flow	(m3/h)	2455	2463	1.135	1.134	2462
Mass Density	(kg/m3)	0.8826	0.8796	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)	380.8	380.8	380.8	380.8	380.8
Mass Flow	(kg/h)	1054	1054	1054	1054	1054
Actual Volume Flow	(m3/h)	1885	1702	1303	1108	1040
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)	14.88	18.08	79.84	365.9	365.9
Mass Flow	(kg/h)	268.1	322.9	1435	785.9	785.9
Actual Volume Flow	(m3/h)	0.2691	0.8347	2.003	1040	798.0
Mass Density	(kg/m3)	996.2	386.8	716.4	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)	365.9	365.9	365.9	365.9	2.684
Mass Flow (kg/h)	785.9	785.9	785.9	785.9	48.36
Actual Volume Flow (m3/h)	717.6	551.1	468.1	439.3	4.763e-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)	363.2	0.5172	362.7	3.660e+005	3.469e+005
Mass Flow (kg/h)	737.6	6.414	731.2	6.594e+006 *	6.250e+006
Actual Volume Flow (m3/h)	439.2	0.2256	438.7	2.467e+005	2.339e+005
Mass Density (kg/m3)	1.679	28.44	1.667	26.72	26.72
Name	**103 @NPP**	**103a @NPP**	**104 @NPP**	**105a @NPP**	**105b @NPP**
Vapour Fraction	1.0000	0.9991	0.9978	1.0000	0.0000
Temperature (C)	267.1	265.2	261.8	267.1	264.3
Pressure (bar)	52.54	50.96	48.31	52.54	51.91
Molar Flow (kgmole/h)	3.217e+005	3.217e+005	3.217e+005	1.906e+004	1.906e+004
Mass Flow (kg/h)	5.796e+006	5.796e+006	5.796e+006	3.434e+005	3.434e+005
Actual Volume Flow (m3/h)	2.169e+005	2.238e+005	2.365e+005	1.285e+004	441.9
Mass Density (kg/m3)	26.72	25.89	24.50	26.72	777.2
Name	**105c @NPP**	**107 @NPP**	**107a @NPP**	**108 (F) @NPP**	**110 @NPP**
Vapour Fraction	0.0000	1.0000	0.9989	0.0220	0.9387
Temperature (C)	192.7	267.1	264.5	263.9	223.5
Pressure (bar)	51.29	52.54	50.44	49.93	24.75
Molar Flow (kgmole/h)	1.906e+004	2.523e+004	2.523e+004	2.523e+004	3.217e+005
Mass Flow (kg/h)	3.434e+005	4.545e+005	4.545e+005	4.545e+005	5.796e+006
Actual Volume Flow (m3/h)	392.1	1.701e+004	1.774e+004	965.8	4.397e+005
Mass Density (kg/m3)	875.9	26.72	25.62	470.6	13.18
Name	**111 @NPP**	**112 @NPP**	**112a @NPP**	**114 @NPP**	**115 @NPP**
Vapour Fraction	0.9387	0.9387	0.9392	0.5525	0.0000
Temperature (C)	223.5	223.5	221.8	221.8	186.6
Pressure (bar)	24.75	24.75	24.01	24.01	24.01
Molar Flow (kgmole/h)	2.940e+005	2.767e+004	2.767e+004	5.290e+004	5.290e+004
Mass Flow (kg/h)	5.297e+006	4.984e+005	4.984e+005	9.529e+005	9.529e+005
Actual Volume Flow (m3/h)	4.019e+005	3.782e+004	3.899e+004	4.432e+004	1082
Mass Density (kg/m3)	13.18	13.18	12.78	21.50	880.8
Name	**120 @NPP**	**121 @NPP**	**122 (H) @NPP**	**122a @NPP**	**122a-L @NPP**
Vapour Fraction	0.8948	0.8948	0.8948	0.8962	0.0000 *
Temperature (C)	183.3	183.3	183.3	181.0	181.0
Pressure (bar)	10.80	10.80	10.80	10.26	10.26
Molar Flow (kgmole/h)	2.940e+005	2.664e+005	2.759e+004	2.759e+004	2.759e+004
Mass Flow (kg/h)	5.297e+006	4.800e+006	4.971e+005	4.971e+005	4.971e+005
Actual Volume Flow (m3/h)	8.565e+005	7.761e+005	8.037e+004	8.453e+004	561.1
Mass Density (kg/m3)	6.185	6.185	6.185	5.880	885.9

* Specified by user.

Workbook: Case (Main) (continued)

Material Streams (continued)

Fluid Pkg: All

Name		123 @NPP	124 @NPP	125 @NPP	126 (E) @NPP	127 @NPP
Vapour Fraction		0.3155	0.0000	0.0000 *	0.0000	1.0000
Temperature	(C)	181.0	180.9	181.0	181.5	181.5
Pressure	(bar)	10.26	10.26	10.26	10.37	10.37
Molar Flow	(kgmole/h)	8.049e+004	8.049e+004	1.082e+005	2.773e+004	2.387e+005
Mass Flow	(kg/h)	1.450e+006	1.450e+006	1.950e+006	4.995e+005	4.301e+006
Actual Volume Flow	(m3/h)	8.786e+004	1636	2201	564.2	8.074e+005
Mass Density	(kg/m3)	16.50	886.1	885.9	885.4	5.327

Name		128 @NPP	128a @NPP	130 @NPP	131 @NPP	132 @NPP
Vapour Fraction		1.0000	1.0000	1.0000	1.0000	1.0000
Temperature	(C)	252.9	252.5	147.2	147.2	147.2
Pressure	(bar)	9.956	9.757	3.477	3.477	3.477
Molar Flow	(kgmole/h)	2.387e+005	2.387e+005	2.387e+005	2.212e+005	1.751e+004
Mass Flow	(kg/h)	4.301e+006	4.301e+006	4.301e+006	3.985e+006	3.155e+005
Actual Volume Flow	(m3/h)	1.011e+006	1.032e+006	2.323e+006	2.153e+006	1.704e+005
Mass Density	(kg/m3)	4.252	4.167	1.851	1.851	1.851

Name		132a (RP) @NPP	132a-L (RP-L) @NPP	134 @NPP	135 @NPP	140 @NPP
Vapour Fraction		1.0000	0.0000 *	0.0000	0.0053	0.9595
Temperature	(C)	146.5	136.9	102.4	99.63	101.1
Pressure	(bar)	3.303	3.303	3.303	0.9998	1.052
Molar Flow	(kgmole/h)	1.751e+004	1.751e+004	1.751e+004	1.751e+004	2.212e+005
Mass Flow	(kg/h)	3.155e+005	3.155e+005	3.155e+005	3.155e+005	3.985e+006
Actual Volume Flow	(m3/h)	1.794e+005	339.6	329.8	3142	6.177e+006
Mass Density	(kg/m3)	1.759	928.9	956.8	100.4	0.6452

Name		141 @NPP	142 @NPP	142a (SO) @NPP	142a-L (SO-L) @NPP	144 @NPP
Vapour Fraction		0.9595	0.9595	0.9606	0.0000 *	0.4075
Temperature	(C)	101.1	101.1	99.63	99.63	99.63
Pressure	(bar)	1.052	1.052	0.9998	0.9998	0.9998
Molar Flow	(kgmole/h)	2.085e+005	1.274e+004	1.274e+004	1.274e+004	3.025e+004
Mass Flow	(kg/h)	3.756e+006	2.295e+005	2.295e+005	2.295e+005	5.450e+005
Actual Volume Flow	(m3/h)	5.821e+006	3.557e+005	3.736e+005	239.4	3.767e+005
Mass Density	(kg/m3)	0.6452	0.6452	0.6143	958.7	1.447

Name		145 @NPP	146 @NPP	150 @NPP	151 @NPP	151 HTE feedwater (lic
Vapour Fraction		0.0000	0.0050	0.9418	0.9418	0.0030
Temperature	(C)	73.57	70.80	87.38	87.38	152.5
Pressure	(bar)	0.9998	0.3227	0.6342	0.6342	5.900
Molar Flow	(kgmole/h)	3.025e+004	3.025e+004	2.085e+005	2.072e+005	1.770e+004
Mass Flow	(kg/h)	5.450e+005	5.450e+005	3.756e+006	3.733e+006	3.188e+005
Actual Volume Flow	(m3/h)	558.6	1.390e+004	9.176e+006	9.120e+006	660.6
Mass Density	(kg/m3)	975.7	39.20	0.4093	0.4093	482.6

Name		152 @NPP	152 HTE feed water (v	160 @NPP	161 @NPP	162 @NPP
Vapour Fraction		0.9418	1.0000	0.9217	0.9217	0.9217
Temperature	(C)	87.38	155.3	71.99	71.99	71.99
Pressure	(bar)	0.6342	5.400	0.3396	0.3396	0.3396
Molar Flow	(kgmole/h)	1268	1.770e+004	2.072e+005	1.946e+005	1.260e+004
Mass Flow	(kg/h)	2.284e+004	3.188e+005	3.733e+006	3.506e+006	2.270e+005
Actual Volume Flow	(m3/h)	5.580e+004	1.130e+005	1.602e+007	1.504e+007	9.741e+005
Mass Density	(kg/m3)	0.4093	2.821	0.2330	0.2330	0.2330

* Specified by user.

Workbook: Case (Main) (continued)

Material Streams (continued)

Fluid Pkg: All

Name		163 @NPP	163a @NPP	163a-L @NPP	165 @NPP	166 @NPP
Vapour Fraction		0.9247	0.9256	0.0000 *	0.2944	0.0000 *
Temperature	(C)	71.99	70.80	70.80	70.80	70.80
Pressure	(bar)	0.3396	0.3227	0.3227	0.3227	0.3227
Molar Flow	(kgmole/h)	1.387e+004	1.387e+004	1.387e+004	4.412e+004	4.412e+004
Mass Flow	(kg/h)	2.498e+005	2.498e+005	2.498e+005	7.948e+005	7.948e+005
Actual Volume Flow	(m3/h)	1.076e+006	1.130e+006	255.7	1.144e+006	813.3
Mass Density	(kg/m3)	0.2323	0.2212	977.3	0.6950	977.3
Name		167 @NPP	168 @NPP	170 @NPP	171 @NPP	172 @NPP
Vapour Fraction		0.0000	0.0102	0.8717	0.7125	0.6805
Temperature	(C)	40.20	34.32	34.32	34.32	34.32
Pressure	(bar)	0.3227	5.419e-002	5.419e-002 *	5.419e-002	5.419e-002
Molar Flow	(kgmole/h)	4.412e+004	4.412e+004	1.946e+005	2.387e+005	2.578e+005
Mass Flow	(kg/h)	7.948e+005	7.948e+005	3.506e+006	4.301e+006	4.644e+006
Actual Volume Flow	(m3/h)	801.2	2.121e+005	7.985e+007	8.006e+007	8.258e+007
Mass Density	(kg/m3)	992.1	3.747	4.390e-002	5.371e-002	5.624e-002
Name		173 @NPP	180 @NPP	181 @NPP	182 @NPP	183 @NPP
Vapour Fraction		0.0000 *	0.0000	0.0000	0.0000	0.0000
Temperature	(C)	34.32	34.64	39.89	68.02	96.85
Pressure	(bar)	5.419e-002	30.66 *	30.66	30.66	30.66
Molar Flow	(kgmole/h)	2.578e+005	2.578e+005	2.578e+005	2.578e+005	2.578e+005
Mass Flow	(kg/h)	4.644e+006	4.644e+006	4.644e+006	4.644e+006	4.644e+006
Actual Volume Flow	(m3/h)	4671	4665	4674	4738	4827
Mass Density	(kg/m3)	994.2	995.5	993.5	980.2	962.0
Name		184 @NPP	185 @NPP	186 @NPP	187 @NPP	188 @NPP
Vapour Fraction		0.0000	0.0000	0.0000	0.0000	0.0000
Temperature	(C)	96.85	134.1	179.9	181.5	180.4
Pressure	(bar)	30.66	30.66	30.66	30.66 *	30.66
Molar Flow	(kgmole/h)	2.578e+005	2.578e+005	2.578e+005	1.082e+005	3.660e+005
Mass Flow	(kg/h)	4.644e+006	4.644e+006	4.644e+006	1.950e+006	6.594e+006
Actual Volume Flow	(m3/h)	4827	4978	5227	2198	7425
Mass Density	(kg/m3)	962.0	932.8	888.5	886.8	888.0
Name		190 @NPP	191 @NPP	201 @NPP	202 @NPP	203 @NPP
Vapour Fraction		0.0000	0.0000	0.0000	0.0000	0.0000
Temperature	(C)	181.1	219.3	247.1	247.1	172.6
Pressure	(bar)	63.43 *	63.43	4.000	3.427	3.167
Molar Flow	(kgmole/h)	3.660e+005	3.660e+005	1.752e+004	1.752e+004	1.752e+004
Mass Flow	(kg/h)	6.594e+006	6.594e+006	4.414e+006	4.414e+006	4.414e+006
Actual Volume Flow	(m3/h)	7414	7806	5180	5180	4877
Mass Density	(kg/m3)	889.4	844.7	852.2	852.2	905.1
Name		204 @NPP	205 @NPP	206 @NPP	207 @NPP	
Vapour Fraction		0.0000	0.0000	0.0000	0.0000	
Temperature	(C)	172.6	172.7	186.0	186.0 *	
Pressure	(bar)	2.606	5.000 *	4.500	4.500 *	
Molar Flow	(kgmole/h)	1.752e+004	1.752e+004	1.752e+004	1.752e+004	
Mass Flow	(kg/h)	4.414e+006	4.414e+006	4.414e+006	4.414e+006 *	
Actual Volume Flow	(m3/h)	4877	4877	4927	4927	
Mass Density	(kg/m3)	905.1	905.1	895.9	896.0	

Workbook: Case (Main) (continued)

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	0.9000	0.9000	0.9000 *	0.9000
Comp Mole Frac (Hydrogen)	0.0004	0.1000	0.1000	0.1000 *	0.1000
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 n	173 H2/H2O recycle	202 Process Heat In
Comp Mole Frac (H2O)	0.9000	0.1800	0.1800	0.1800	***
Comp Mole Frac (Hydrogen)	0.1000	0.8200	0.8200	0.8200	***
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	203 Process Heat Ret	301	302 H2/H2O for purific	332 H2 Product	401
Comp Mole Frac (H2O)	***	0.1800	0.1800	0.0000	0.0000
Comp Mole Frac (Hydrogen)	***	0.8200	0.8200	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)	***	***	***	***	***

* Specified by user.

Workbook: Case (Main) (continued)

Compositions (continued)

Fluid Pkg: All

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 *	0.4000
Comp Mole Frac (Nitrogen)	0.7500	0.7500	0.7500	0.7500 *	0.6000
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)	0.4000	0.4000	0.3999 *	0.3999	0.3999
Comp Mole Frac (Nitrogen)	0.6000	0.6000	0.6001 *	0.6001	0.6001
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	**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)	0.3999	0.0000 *	0.0000	0.0000	0.0000 *
Comp Mole Frac (Nitrogen)	0.6001	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)	***	***	***	***	***

* Specified by user.

Workbook: Case (Main) (continued)

Compositions (continued)

Fluid Pkg: All

Name	Cathode @Cell	Gas Products @Cell	Liquid Products @Cell	Molar Flow of Oxygen	Process Cell Inlet @Ce
Comp Mole Frac (H2O)	0.1800	0.1324	0.1324	0.0000 *	0.9000
Comp Mole Frac (Hydrogen)	0.8200	0.6030	0.6030	0.0000 *	0.1000
Comp Mole Frac (Oxygen)	0.0000	0.2647	0.2647	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	0.4000	0.0000 *	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.7500	0.6000	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	0.1800
Comp Mole Frac (Hydrogen)	0.0000	0.0000	0.0004	0.0004	0.8200
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	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)	***	***	***	***	***

* Specified by user.

Workbook: Case (Main) (continued)

Compositions (continued)

Fluid Pkg: All

Name	306 @H2rec	307 @H2rec	308 @H2rec	309 @H2rec	310 @H2rec
Comp Mole Frac (H2O)	0.1800	0.1800	1.0000	1.0000	0.0470
Comp Mole Frac (Hydrogen)	0.8200	0.8200	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)	***	***	***	***	***
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)	***	***	***	***	***

* Specified by user.

Workbook: Case (Main) (continued)

Compositions (continued)

Fluid Pkg: All

Name	103 @NPP	103a @NPP	104 @NPP	105a @NPP	105b @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	105c @NPP	107 @NPP	107a @NPP	108 (F) @NPP	110 @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	111 @NPP	112 @NPP	112a @NPP	114 @NPP	115 @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	120 @NPP	121 @NPP	122 (H) @NPP	122a @NPP	122a-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	123 @NPP	124 @NPP	125 @NPP	126 (E) @NPP	127 @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	128 @NPP	128a @NPP	130 @NPP	131 @NPP	132 @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	132a (RP) @NPP	132a-L (RP-L) @NPP	134 @NPP	135 @NPP	140 @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	141 @NPP	142 @NPP	142a (SO) @NPP	142a-L (SO-L) @NPP	144 @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	145 @NPP	146 @NPP	150 @NPP	151 @NPP	151 HTE feedwater (lic
Comp Mole Frac (H2O)	1.0000	1.0000	1.0000	1.0000	0.9997
Comp Mole Frac (Hydrogen)	***	***	***	***	0.0003
Comp Mole Frac (Oxygen)	***	***	***	***	0.0000
Comp Mole Frac (Nitrogen)	***	***	***	***	0.0000
Comp Mole Frac (CO2)	***	***	***	***	0.0000
Comp Mole Frac (CO)	***	***	***	***	0.0000
Comp Mole Frac (DTRM-A)	***	***	***	***	***
Comp Mole Frac (Therminol-66)	***	***	***	***	***
Name	152 @NPP	152 HTE feed water (v	160 @NPP	161 @NPP	162 @NPP
Comp Mole Frac (H2O)	1.0000	0.9997	1.0000	1.0000	1.0000
Comp Mole Frac (Hydrogen)	***	0.0003	***	***	***
Comp Mole Frac (Oxygen)	***	0.0000	***	***	***
Comp Mole Frac (Nitrogen)	***	0.0000	***	***	***
Comp Mole Frac (CO2)	***	0.0000	***	***	***
Comp Mole Frac (CO)	***	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	163 @NPP	163a @NPP	163a-L @NPP	165 @NPP	166 @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	167 @NPP	168 @NPP	170 @NPP	171 @NPP	172 @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	173 @NPP	180 @NPP	181 @NPP	182 @NPP	183 @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	184 @NPP	185 @NPP	186 @NPP	187 @NPP	188 @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	190 @NPP	191 @NPP	201 @NPP	202 @NPP	203 @NPP
Comp Mole Frac (H2O)	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	1.0000	1.0000

* Specified by user.

Workbook: Case (Main) (continued)

Compositions (continued)

Fluid Pkg: All

Name	204 @NPP	205 @NPP	206 @NPP	207 @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 *	

Energy Streams

Fluid Pkg: All

Name		Electrolysis Power	Inverter Power	Process Heat	Q-CW	Q-HX-107
Heat Flow	(MW)	-24.98	-25.49	2.776e-002	70.59	0.5233
Mass Flow	(kg/h)	---	---	---	---	---
Name		Q-HX-502	Q-IC-501	Q-IC-401_stg1	Q-IC-401_stg2	Q-IC-401_stg3
Heat Flow	(MW)	0.1011	0.2791	1.088e-004	1.088e-004	1.088e-004
Mass Flow	(kg/h)	---	1.716e+004	6.690	6.688	6.688
Name		Q-PIPE-801	Q-PIPE-901	Q-PIPE-902	W-K-101	W-K-401
Heat Flow	(MW)	1.831e-006	6.902e-003	7.585e-003	5.202e-002	3.262e-004
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)	0.1080	1.087e-004	1.087e-004	1.087e-004	-9.698e-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)	0.4175	0.4600	0.8871	1.474e-003	6.239e-003
Mass Flow	(kg/h)	---	---	---	---	---
Name		W-P-901	Electrode Heat @Cell	Electrolysis Heating @	Electrolysis Power @C	Process Heat @Cell
Heat Flow	(MW)	0.1604	-1.639e-005	25.01	-24.98	2.776e-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	0.6975	0.2768	0.1989	0.1358
Mass Flow	(kg/h)	5.866	4.288e+004	1.702e+004	1.223e+004	8348
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)	0.1256	6.886e-007	6.846e-007	6.882e-007	6.864e-007
Mass Flow	(kg/h)	7719	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)	-4.124e-003	9.210e-008	0.3278	0.2968	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	0.1618
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 Power @NPP
Heat Flow	(MW)	0.1660	0.1466	0.1502	2.060e-004	7.958e-012
Mass Flow	(kg/h)	---	---	---	---	---
Name		LWR Net Power @NP	Q-100 @NPP	Q-101 @NPP	Q-102 @NPP	Q-103 @NPP
Heat Flow	(MW)	1076	164.7	175.4	1.662	9.012
Mass Flow	(kg/h)	---	---	---	---	---
Name		Q-104 @NPP	Q-105 @NPP	Q-106 @NPP	Q-107 @NPP	Q-108 @NPP
Heat Flow	(MW)	5.295	2124	236.0	182.3	66.95
Mass Flow	(kg/h)	---	---	---	---	---

* Specified by user.

Workbook: Case (Main) (continued)

Energy Streams (continued)

Fluid Pkg: All

Name		Q-109 @NPP	Q-110 @NPP	Q-111 @NPP	Q-PIPE-201 @NPP	Q-PIPE-202 @NPP
Heat Flow	(MW)	77.50	189.5	3388	7.492e-002	5.137e-002
Mass Flow	(kg/h)	---	---	---	---	---
Name		W-P-201 Circ Pump P				
Heat Flow	(MW)	0.4324				
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	No	500.0 *
			301		
TEE-514	Tee	514	515 Sweep Gas Recycle	No	500.0 *
			516		
HX-106 H2/H2O Recuperator	Heat Exchanger	171 Process Cell Outlet	172 H2/H2O product mix	No	500.0 *
		165	166		
HX-103 Sweep Gas Low T Re	Heat Exchanger	512	513	No	500.0 *
		152 process feed water (vapo	162		
HX-501 Sweep Gas High T Re	Heat Exchanger	511 Sweep Gas/O2 Out	512	No	500.0 *
		506	507		
HX-102 Steam Generator	Heat Exchanger	202 Process Heat In	203 Process Heat Return	No	500.0 *
		151	152 process feed water (vapo		
HX-101	Heat Exchanger	301	302 H2/H2O for purification	No	500.0 *
		132 process feed water (liquid	151		
MIX-163	Mixer	173 H2/H2O recycle	163	No	500.0 *
		162			
MIX-505	Mixer	504	505	No	500.0 *
		515 Sweep Gas Recycle			
K-502 Sweep Gas Blower	Compressor	505	506	No	500.0 *
		W-K-502			
K-101 H2/H2O Recirculator	Compressor	163	164	No	500.0 *
		W-K-101			
K-401_stg2	Compressor	402b	402c	No	500.0 *
		W-K-401_stg2			
K-401_stg3	Compressor	402d	402e	No	500.0 *
		W-K-401_stg3			
K-401_stg1	Compressor	401	402a	No	500.0 *
		W-K-401_stg1			
K-501_stgC1	Compressor	501 Sweep Gas Inlet	502	No	500.0 *
		W-K-501_stgC1			
K-501_stgC2	Compressor	503	504	No	500.0 *
		W-K-501_stgC2			
HX-107 Process Topping Hea	Heater	166	167 Process Cell Inlet	No	500.0 *
		Q-HX-107			
HX-502 Sweep Topping Heat	Heater	507	508 Sweep Cell Inlet	No	500.0 *
		Q-HX-502			
CW deltaT	Heater	903	904	No	500.0 *
		Q-CW			
SOEC Stack	Standard Sub-Flowsheet	167 Process Cell Inlet	171 Process Cell Outlet	No	2500 *
		508 Sweep Cell Inlet	511 Sweep Gas/O2 Out		
			Electrolysis Power		
			Process Heat		
H2 Recovery and Feed Condi	Standard Sub-Flowsheet	302 H2/H2O for purification	132 process feed water (liquid	No	2500 *

* Specified by user.

Workbook: Case (Main) (continued)

Unit Ops (continued)

Operation Name	Operation Type	Feeds	Products	Ignored	Calc Level
H2 Recovery and Feed Condi	Standard Sub-Flowsheet	102	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 *
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-901	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 *

* Specified by user.

Workbook: Case (Main) (continued)

Unit Ops (continued)

Operation Name	Operation Type	Feeds	Products	Ignored	Calc Level
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 *
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 *

* Specified by user.