Exergy and economic comparison between kW-scale hybrid and stand-alone solid oxide fuel cell systems

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Abstract

Although hybrid solid oxide fuel cell (SOFC) microturbine systems generate power more efficiently than stand-alone SOFC systems, hybrid systems remain in the demonstration phase. The present study compares a hybrid system's exergetic and economic performance with that of a stand-alone system. Both systems meet a university building's kW-scale power demand. The hybrid system operates at 66% exergetic efficiency, and the stand-alone system operates at 59% exergetic efficiency. Increasing the fuel cell's operating voltage increases the systems' exergetic efficiencies, and varying the fuel cell's temperature, pressure, and fuel utilization influences the systems' exergetic performances, though to a lesser extent. The present study calculates the systems' life cycle costs. We find that the systems' life cycle costs depend significantly on the systems' operation. During baseline operation,

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the hybrid system costs less than the stand-alone system. After optimizing the systems during cogeneration operation, the hybrid system costs slightly more than the stand-alone system. Overall, our findings support hybrid systems continued research and development; it is recommended that future work simulate hybrid and stand-alone systems under a range of thermal-to-electric ratios to reflect different building types and operation.

Keywords: solid oxide fuel cell; microturbine; hybrid; economic; energy; exergy

NOMENCLATURE

A	cross-sectional area (m^2)
A_{system}	system's availability
C_i	cost of component i ($\$ \text{kW}^{-1}$)
$C_{i,0}$	reference cost of component i ($\$ \text{kW}^{-1}$)
$\bar{c}_p$	specific heat capacity (molar basis) ($\text{J} (\text{mol K})^{-1}$)
C_{sys}	system's total capital cost ($\$ \text{kW}^{-1}$)
CEPCI	Chemical Engineering Plant Cost Index
CEPCI ₀	Chemical Engineering Plant Cost Index (reference value)
CF _e	electric capacity factor
CHP	combined heat and power (cogeneration)
ΔT_{lm}	log mean temperature difference (K)
E	system's initial capital cost (2014 USD)
$\dot{E}_d$	exergy destruction rate (kW)
EO	electric-only
$\bar{e}_f$	flow exergy (molar basis) (J mol^{-1})
$\bar{e}^{\text{ch}}$	flow exergy's chemical contribution (molar basis) (J mol^{-1})
F	first year operating cost (2014 USD)
$F_{\text{c,electric}}$	electric utility cost ($\$ \text{kWh}^{-1}$)
$F_{\text{c,ngas}}$	natural gas cost ($\$ \text{kWh}^{-1}$)
$\bar{g}$	specific Gibbs function (molar basis) (J mol^{-1})

HHV	higher heating value (J mol^{-1})
h	specific enthalpy (mass basis) (J kg^{-1})
$\bar{h}$	specific enthalpy (molar basis) (J mol^{-1})
LCC	life cycle cost
LCS	life cycle savings
LHV	lower heating value (J mol^{-1})
$\dot{m}$	mass flow rate (kg s^{-1})
n	capacity scaling exponent
$\dot{n}$	molar flow rate (mol s^{-1})
P_1	ratio of life cycle operating cost to first year operating cost
P_2	ratio of life cycle capital cost to initial capital cost
p	pressure (Pa)
p_0	reference pressure (Pa)
$\dot{Q}_{\text{demand,average}}$	annual average thermal energy demand (kW_{th})
$\dot{Q}_{\text{recovery}}$	thermal energy recovery (kW)
$\bar{R}$	universal gas constant ($8.314 \text{ J (mol K)}^{-1}$)
S_i	capacity of component i
$S_{i,0}$	reference capacity of component i
$\bar{s}$	entropy (molar basis) (J (mol K)^{-1})
T	temperature (K)
T_0	reference temperature (K)
T_b	boundary temperature (K)
T_{PEN}	average PEN temperature (K)
U	overall heat transfer coefficient ($\text{kW (m}^2 \text{ K)}^{-1}$)
$U_{\text{f,sys}}$	fuel utilization
V_{cell}	operating fuel cell voltage (V)
v	velocity (m s^{-1})
$\dot{W}_{\text{demand,average}}$	annual average power demand (kW)
$\dot{W}_{\text{net}}$	net power (kW)
$\dot{W}_{\text{SOFC,DC}}$	fuel cell stack's DC power (kW)
y_i^e	mole fraction of species i in the environment

Greek symbols

ε_{sys}	system's exergetic efficiency
$\eta_{\text{SOFC,e}}$	fuel cell's electric efficiency
$\eta_{\text{sys,CHP}}$	system's cogeneration efficiency
$\eta_{\text{sys,e}}$	system's electric efficiency
λ_{air}	air ratio (-)

Subscripts

0	environmental property (dead state)
blower	air blower
comp	compressor
m	mixed flow
p	primary flow
PEN	positive electrode-electrolyte-negative electrode
s	secondary flow
sys	system
turb	turbine

1. Introduction

Integrating a solid oxide fuel cell (SOFC) stack with a microturbine (MT) enables an SOFC system to generate power relatively efficiently. Hybrid SOFC-MT systems have the potential to achieve electric efficiencies approaching 70% (LHV) [1–6]. On their own, microturbines generate power at approximately 30% (LHV) electric efficiency [7, 8], and stand-alone SOFC systems operate between 52%–65% (LHV) electric efficiency [9]. Hybrid systems integrate a fuel cell stack with a microturbine, enabling these systems to generate greater power per unit input energy than either system alone. In addition, unlike most commercial SOFC systems, which operate the SOFC stack near

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9 atmospheric pressure, hybrid systems may pressurize the SOFC stack, placing
10 the stack immediately after the air compressor. Pressurization enhances the
11 fuel cell's performance by increasing the fuel cell's reversible potential. Hy-
12 brid systems also have the potential to reduce CO₂ emissions relative to more
13 conventional systems due to their relatively high electric efficiency. Other
14 emissions, such as NO_x and SO_x, may fall below those of more conventional
15 systems. SOFCs emit negligible SO_x, in particular, as the natural gas is
16 desulfurized prior to entering the fuel cell stack [10].

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18 While hybrid systems operate at high electric and overall efficiencies,
19 these systems remain in the demonstration phase. The National Fuel Cell
20 Research Center at the University of California, Irvine tested a 220 kW hy-
21 brid system in the early 2000s [11–13], and FuelCell Energy, Inc. developed
22 a hybrid system consisting of a molten carbonate fuel cell stack and gas tur-
23 bine [14]. Neither system, however, is commercially developed. Stand-alone
24 systems, on the other hand, are available. Bloom Energy [9], Atrex Energy
25 [15], and Protonex [16] have developed stand-alone systems for a variety of
26 kW and sub-kW-scale applications, including communications and sensors,
27 data centers, retail stores, and remote power applications. In order for hy-
28 brid systems to compete in the distributed generation market, they must
29 demonstrate favorable performance on the kW-scale.

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31 The present study compares a hybrid system's exergetic and economic
32 performance with that of a stand-alone system. Using exergy, we identify
33 sources of inefficiency in both systems. Moran, et al. define exergy as follows:
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37 Exergy is the maximum theoretical work obtainable from an *over-*
38 *all* system consisting of a system and the environment as the sys-

tem comes into equilibrium with the environment (passes to the dead state). [17, p. 362, emphasis in the original]

We also investigate the influence of the fuel cell’s operating parameters on the systems’ exergetic efficiencies, finding that increasing the fuel cell’s operating voltage significantly improves the systems’ exergetic efficiencies. In the second part of this study, we calculate each system’s life cycle cost from the owner’s perspective. The life cycle cost formula incorporates the systems’ equipment, operation, and supplemental energy costs. The present study parametrically varies the systems’ operating parameters in order to identify local minima in the systems’ life cycle costs.

Throughout this study, the term “hybrid” refers to the integration of the SOFC stack with a microturbine. The fuel cell stack operates under pressure, situated between the air compressor and turbine. Alternatively, the fuel cell stack may be placed after the turbine [18–21]. Such a configuration, however, typically reduces the system’s efficiency [18, 20]. SOFCs may be integrated with other devices, such as a polymer electrolyte membrane fuel cell stack [21, 22], absorption chiller [21, 23], photovoltaic array [24], and wind turbine [21, 24]. We have chosen to integrate the fuel cell stack with a microturbine, as microturbines are established combined heat and power (CHP) technologies whose operating temperature and pressure coincide with those of SOFCs. The working fluid passes from the fuel cell stack to the turbine, obviating the need for a heat exchanger. Zhang, et al. [21] refer to this configuration as a “direct thermal coupling” scheme.

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9 Previous exergy analyses have focused largely on either hybrid or stand-
10 alone systems, typically on the MW-scale. Few (if any) studies, however,
11 have compared hybrid and stand-alone systems in the low to mid-kW range
12 (50 kW–250 kW). Calise, et al. [25] performed one of the earliest exergy
13 studies on a 1.5 MW hybrid system. The authors found that the largest
14 exergy destruction occurred in the SOFC stack, followed by the afterburner.
15 Calise, et al. [26] performed an exergy analysis of a 1.5 MW hybrid system
16 during part-load operation, finding that the system’s part-load performance
17 depended largely on the system’s control strategy. Gandiglio, et al. [27] per-
18 formed an exergy and thermoeconomic comparison between 280 MW hybrid
19 and stand-alone systems. Similar to Calise, et al. [25], these authors found
20 that the hybrid system’s SOFC stack and afterburner destroyed the most
21 exergy, whereas in the stand-alone system, the air preheater and afterburner
22 destroyed the most exergy. Chan, et al. [28] found that the afterburner and
23 vaporizer destroyed the most exergy in a methane-fueled, MW-scale, stand-
24 alone system. Douvartzides, et al. [29] also performed an exergy analysis of
25 a methane-fueled, stand-alone system, finding that decreasing the reforming
26 factor increased the system’s exergetic efficiency.

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43 Previous economic studies have considered small (residential-scale, 1–
44 2 kW) and large (MW) systems. To the authors’ knowledge, however, no
45 previous economic studies have compared hybrid and stand-alone systems
46 in the 50–250 kW range. Gandiglio, et al. [27] performed a thermoecono-
47 mic comparison between multi-MW pressurized and atmospheric power
48 plants. The authors found that the pressurized and atmospheric systems
49 exhibited similar levelized costs of electricity, whereas the hybrid system ex-
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hibited a lower thermoeconomic levelized cost of electricity. Santin, et al. [30] performed a techno-economic analysis of a 500 kW hybrid system with different fuels, finding that methanol-fueled systems yielded higher internal rates of return on investment than kerosene-fueled systems. Fontell, et al. found a 250 kW stand-alone system to be economically competitive with a conventional gas engine [31]. Numerous other studies have considered large (baseload) SOFC power plants [32–44]. On the residential-scale, Braun [45–47] performed a techno-economic analysis of several different SOFC system configurations with heat recovery. Braun found that the methane-fueled system with anode gas recycle and 100% internal reforming saved the most money among the feasible options. Hawkes and Leach [48–50] and Hawkes, et al. [51–54] also considered residential-scale, CHP SOFC systems.

Section 2 describes the atmospheric and pressurized systems, as well as the previously developed fuel cell stack model. Section 2 also describes the systems’ baseline operating parameters and building loads. Section 3 presents the systems’ exergetic performances during baseline operation and the systems’ exergy accountings. The sensitivity analysis varies the fuel cell’s operating parameters and microturbine pressure ratio. Section 4 describes the cost model. The cost model incorporates the systems’ first costs, operating costs, and supplemental energy costs. The parametric analysis varies the fuel cell’s and turbomachinery’s operating parameters. Section 5 concludes the present study.

2. System descriptions

2.1. Stand-alone system

The present study integrates a previously developed fuel cell stack model [55–58] with two overall systems derived from previous studies [1–6, 21, 30, 40, 41, 44, 46, 59–66]. The system models were developed in the iterative equation solver software, Engineering Equation Solver (EES) [67]. Figure 1 presents the stand-alone system. This configuration is based on the methane-fueled, 100% internal reforming design with anode gas recycle presented by Braun [46]. Braun found this configuration to save more money than other configurations without requiring the complex controllability and equipment costs associated with cathode gas recycle. The chosen system also reflects stand-alone systems developed by U.S. companies Atrex Energy and FuelCell Energy, Inc. Atrex Energy [59] develops an internal reforming tubular SOFC. FuelCell Energy, Inc. [60] develops molten carbonate fuel cell systems that employ 100% internal reforming. Similar to the systems modeled in the present study, both of these commercial systems operate at sufficiently high temperatures to cogenerate steam or hot water.

Table 1 presents the systems’ baseline operating parameters. In the stand-alone system, air enters the air blower at state 1, where it is compressed to overcome the system’s air-side pressure drop. The air is then pre-heated to a temperature suitable for SOFC operation between states 2 and 3. The temperature of air entering the air channel is iteratively determined by the temperature rise across the air channel, which is specified to be 150°C in both systems based on the maximum allowable thermal strain in a fuel cell [6, 62, 68]. The average positive electrode-electrolyte-negative electrode (PEN)

Figure 1 The stand-alone system's process flow diagram (adapted from Refs. [46, 59, 60]). Thermodynamic properties are provided during baseline operation (Table 1).

temperature, specified to be 750°C, iteratively determines the inlet air flow rate in both systems. Intermediate-temperature SOFCs typically operate between 550°C and 800°C [69]. Methane enters the fuel compressor at state 4, where it is compressed to overcome the system's fuel-side pressure drop. The fuel is desulfurized (states 5 and 6) and pre-heated (states 8 and 9). In both systems, the temperature of the fuel entering the stack is specified to be 10°C less than the average PEN temperature. The fuel cell stack generates DC power, which is inverted to AC power at 95% efficiency. Part of the anode exhaust circulates back to the fuel pre-heater at state 7. Anode gas recycle maintains a steam-to-carbon (S/C) ratio of 2.5 at the fuel cell stack's inlet,

Table 1 Baseline operating parameters.

Parameter	Value	
	Stand-alone	Hybrid
Net power	84.2 kW	84.2 kW
Thermal energy demand	170.8 kW _{th}	170.8 kW _{th}
Fuel cell voltage	0.7 V	0.7 V
System fuel utilization	0.85	0.85
Pressure ratio	-	4
Turbine inlet temperature	-	1,125°C
Average PEN temperature	750°C	750°C
Fuel channel inlet temperature	740°C	740°C
Air channel temperature rise	150°C	150°C
S/C ratio	2.5	2.5
Inverter efficiency	95%	95%
Air blower efficiency	70%	-
Air compressor efficiency	-	80%
Fuel compressor efficiency	70%	70%
Turbine efficiency	-	80%
Stack thermal energy loss (% inlet HHV)	3%	3%

thereby reducing the risk of solid carbon formation inside the fuel cell stack [70]. Unreacted fuel exits the fuel cell stack at state 10 and reacts in the auxiliary combustor. The exhaust pre-heats the incoming fuel (states 13 and 14) and air (states 14 and 15). The heat recovery heat exchanger recovers thermal energy for cogeneration between states 15 and 16. In both systems, the exhaust exits the system 25°C above the dew point temperature.

2.2. Hybrid system

Figure 2 presents the hybrid system. The present model resembles previous configurations, which have considered fuel and air bypasses and external reforming [1–6, 30, 40, 41, 44, 61–66]. The configuration in Fig. 2 most closely

Figure 2 The hybrid system's process flow diagram (adapted from Refs. [1–6, 21, 30, 40, 41, 44, 61–66]). Thermodynamic properties are provided during baseline operation (Table 1).

resembles Chan, et al.'s configuration [1, 3]. These authors bypassed the fuel directly to the combustor in order to increase the turbine inlet temperature. The 220 kW experimental system developed by Siemens Westinghouse and tested at the National Fuel Cell Research Center [11–13] similarly employed a fuel bypass to control the mass flow through the system.

In the hybrid system, air enters the air compressor at state 1, where it

is compressed to a pressure ratio of four [1, 3]. The air is then pre-heated between states 2 and 3. Air enters the hybrid system’s fuel cell stack at a temperature higher than it enters the fuel cell stack in the stand-alone system. Air flows at a lower mass flow rate in the hybrid system, giving rise to a higher stack exit air temperature and correspondingly higher inlet air temperature to maintain a fixed cathode temperature rise. Methane enters the fuel compressor at state 4, where it is compressed to a pressure ratio slightly greater than four due to the fuel-side component pressure drops. The fuel is desulfurized between states 7 and 8 and pre-heated between states 10 and 11. Similar to the stand-alone system, part of the anode exhaust circulates back to the fuel cell stack’s inlet at state 9. Unlike the stand-alone system, however, part of the hybrid system’s fuel may pass directly to the auxiliary combustor at state 6 in order to maintain the turbine inlet temperature at 1,125°C. The system’s exhaust pre-heats the incoming fuel (states 16 and 17) and air (states 17 and 18). Finally, the heat recovery heat exchanger recovers thermal energy for cogeneration between states 18 and 19.

2.3. Fuel cell stack

The hybrid and stand-alone systems incorporate a previously developed 1-D, co-flow, numerical SOFC stack model [55–58]. The fuel cell model is a 10 cm × 10 cm, 18 channel, planar design whose material and geometric properties are presented in the International Energy Agency’s Annex II reports [71, 72 (citing [73])] and repeated in Braun [45]. The fuel cell model discretizes the gas channels, PEN structure, fuel-side and air-side interconnects along the flow direction, accounting for axial changes in current

density, temperature, composition, and pressure. The activation (Butler-Volmer equation), concentration (Fick's law of diffusion), and ohmic (electrical resistivity) polarizations reduce the fuel cell's operating voltage below the Nernst potential. The water-gas shift reaction occurs at equilibrium, and the methane steam reforming reaction proceeds according to Achenbach and Riensche's kinetic expression [74]. The fuel cell's energy balance accounts for convection in the gas channels, and conduction and cross-channel radiation in the solid components. Heat transfer also occurs between adjacent fuel cells' fuel and air channels. The momentum balance accounts for momentum change and wall shear stress, giving rise to the gas channel pressure drops. Identical equations and operating parameters govern each fuel cell in the stack. The charges, species mass, energy, and momentum balances may be simulated dynamically, although the present study considers only steady-state behavior.

The present study modifies the fuel cell stack model since its presentation in our previous study in two respects [55]. Previously, the fuel cell stack model accounted for radiation between the stack and stainless steel container, treating radiation as a boundary condition. We found, however, that this boundary condition gave rise to an unreasonably small air temperature rise in the air channel. Thus, we chose to remove the radiation boundary condition and instead assume that the fuel cell stack's container loses 3% of the system's inlet fuel energy (HHV) to the environment [33, 39, 47]. Second, we modified the fuel cell's design to reflect intermediate-temperature operation. Our previous study [55] considered an electrolyte-supported design. The present study considers an anode-supported design. Anode-supported fuel

cells operate at approximately 750°C, thereby permitting the use of metallic interconnects. Metallic interconnects reduce the fuel cell stack’s cost. Thijsen [75] provides typical dimensions for anode-supported fuel cells, and the present work adopts these values. The anode is 325 μm thick, the electrolyte is 8 μm thick, and the cathode is 50 μm thick. The interconnect’s thermal conductivity equals 25 W (m K)^{−1} [69].

2.4. Balance of plant (BOP) components

The BOP components comprise the fuel and air compressors, fuel and air pre-heaters, desulfurizer, ejector, auxiliary combustor, heat recovery heat exchanger, and inverter. Each BOP model calculates the exit mass flow rate, temperature, and pressure based on the corresponding inlet quantities and 0-D mass and energy balances. The energy balance applied to the air blower, air compressor, and fuel compressor yields the exit temperature [17]:

$$T_{\text{exit}} = T_{\text{inlet}} + \frac{T_{\text{inlet}}}{\eta_{\text{blower(comp)}}} \left(\left(\frac{p_{\text{exit}}}{p_{\text{inlet}}} \right)^{\frac{\bar{R}}{\bar{c}_p}} - 1 \right) \quad (1)$$

where T_{exit} is the exit temperature, T_{inlet} is the inlet temperature, $\eta_{\text{blower(comp)}}$ is the blower’s (or compressor’s) isentropic efficiency, p_{exit} is the exit pressure, p_{inlet} is the inlet pressure, $\bar{R}$ is the universal gas constant, and $\bar{c}_p$ is the ideal gas mixture’s molar specific heat capacity. The turbine’s exit temperature is calculated similarly [17]:

$$T_{\text{exit}} = T_{\text{inlet}} + \eta_{\text{turb}} T_{\text{inlet}} \left(\left(\frac{p_{\text{exit}}}{p_{\text{inlet}}} \right)^{\frac{\bar{R}}{\bar{c}_p}} - 1 \right) \quad (2)$$

where η_{turb} is the turbine's isentropic efficiency. The power consumed or generated by each rotating device equals the enthalpy change across the device.

The fuel and air pre-heaters operate in counter-flow. The heat exchangers calculate the UA product, which is the product of a heat exchanger's overall heat transfer coefficient (U) and heat exchange area (A). The UA product indicates a heat exchanger's size, which is used in the cost model to estimate the heat exchanger's capital cost (Section 4). The log mean temperature difference relates the UA product to the heat exchanger's heat rate as follows:

$$q = UA\Delta T_{\text{lm}} \quad (3)$$

where ΔT_{lm} is the log mean temperature difference [76].

The ejector overcomes the anode recycle loop's pressure drop by entraining the recycled gas. The anode recycle loop's pressure drop comprises the fuel pre-heater and fuel-side stack pressure drops. The mixed flow exits the ejector at a pressure greater than the secondary flow but less than the primary flow. Bird, et al. [77] present the ejector model, which Liso [78] and Braun, et al. [47] implemented in SOFC system models. The primary fluid's diameter equals 3.31 mm, and the mixed fluid's diameter equals 20.4 mm [79, 80]. Mass, momentum, and energy balances yield the mixed flow's mass flow rate, primary flow's pressure, and mixed flow's temperature:

$$\dot{m}_{\text{m}} = \dot{m}_{\text{p}} + \dot{m}_{\text{s}} \quad (4)$$

$$v_{\text{m}}\dot{m}_{\text{m}} + p_{\text{m}}A_{\text{m}} = v_{\text{p}}\dot{m}_{\text{p}} + p_{\text{p}}A_{\text{p}} + v_{\text{s}}\dot{m}_{\text{s}} + p_{\text{s}}A_{\text{s}} \quad (5)$$

$$\dot{m}_{\text{m}} \left(h_{\text{m}} + \frac{1}{2}v_{\text{m}}^2 \right) = \dot{m}_{\text{p}} \left(h_{\text{p}} + \frac{1}{2}v_{\text{p}}^2 \right) + \dot{m}_{\text{s}} \left(h_{\text{s}} + \frac{1}{2}v_{\text{s}}^2 \right) \quad (6)$$

where $\dot{m}$ denotes the mass flow rate, v denotes the velocity, and A denotes the cross-sectional area. The subscripts m , p , and s denote the mixed, primary, and secondary flows, respectively.

Lastly, pressure drops occur across each system component. Braun [47] estimated the pressure drops across fuel cell system components based on several previous studies [81–83], and the present study adopts Braun’s estimates. The pressure drops across the fuel and air-preheaters equal 50 mbar and 100 mbar, respectively, and the heat recovery heat exchanger’s pressure drop equals 15 mbar. The desulfurizer’s pressure drop equals 100 mbar, and the auxiliary combustor’s pressure drop equals 20 mbar. The fuel-side and air-side fuel cell stack pressure drops both equal 30 mbar. These values remain constant, regardless of the system’s operating point. For more detail, the reader is referred to Ref. [57].

2.5. Building loads

The hybrid and stand-alone systems meet the annual average power demand of a recent addition to the University of Pittsburgh’s engineering building, the Mascaro Center for Sustainable Innovation (MCSI). We collected data during the winter semester, winter break, summer semester, and summer break. Based on data collected during these time periods, MCSI consumes 84.2 kW on average annually. The electric loads include plug loads, lights, fans for the air-handling units, and miscellaneous equipment. MCSI consumes 170.8 kW_{th} of hot water on average annually. The hot water demand includes the hot water used by the radiators and air-handling units. The hot water demand excludes cooling energy, which is produced externally in the central chilled water system. Both systems produce exhaust streams

at approximately 440°C prior to heat recovery, which is suitable for heating water.

3. Energetic and exegetic analyses

3.1. Exergy balance

A system reaches equilibrium with its environment when it assumes the environment's temperature, pressure, and composition. In order to model the systems' exergetic performances, the present work adopts the equations and methodology presented in Moran, et al. [17] and Klein and Nellis [84]. The steady-state exergy balance applied to each system component is given by the following equation:

$$0 = \sum_{i=1}^{\# \text{ transfers}} \left(1 - \frac{T_0}{T_i}\right) \dot{Q}_i - \dot{W} + \sum_{i=1}^{\# \text{ inlets}} \dot{n}_{\text{in},i} \bar{e}_{\text{f},\text{in},i} - \sum_{i=1}^{\# \text{ exits}} \dot{n}_{\text{out},i} \bar{e}_{\text{f},\text{out},i} - \dot{E}_{\text{d}} \quad (7)$$

where T_0 is the environment's temperature, T_i is the boundary temperature at which heat transfer $\dot{Q}_i$ occurs, $\dot{W}$ is the power generation, $\dot{n}_{\text{in},i}$ is the molar flow rate entering inlet i , $\bar{e}_{\text{f},\text{in},i}$ is the molar flow exergy entering inlet i , $\dot{n}_{\text{out},i}$ is the molar flow rate leaving exit i , $\bar{e}_{\text{f},\text{out},i}$ is the molar flow exergy leaving exit i , and $\dot{E}_{\text{d}}$ is the exergy destruction. The specific flow exergy is given by the following formula:

$$\bar{e}_{\text{f}} = \bar{h} - \bar{h}_0 - T_0 (\bar{s} - \bar{s}_0) + \bar{e}^{\text{ch}} \quad (8)$$

where $\bar{h}$ is the fluid's molar enthalpy evaluated at the inlet or exit conditions, $\bar{h}_0$ is the fluid's molar enthalpy evaluated at the environment's conditions,

T_0 is the environment's temperature, $\bar{s}$ is the molar entropy evaluated at the inlet or exit conditions, $\bar{s}_0$ is the fluid's molar entropy evaluated at the environment's conditions, and $\bar{e}^{\text{ch}}$ is the chemical exergy. The present work assumes $T_0 = 25^\circ\text{C}$ and $p_0 = 1 \text{ atm}$.

3.2. Performance metrics

The system's exergetic efficiency is defined as the ratio of the desired products' exergy to the inlet exergy:

$$\varepsilon_{\text{sys}} = \frac{\dot{W}_{\text{net}} + \left(1 - \frac{T_0}{T_b}\right) \dot{Q}_{\text{recovery}}}{(\dot{n}_{\text{fuel}} \cdot \bar{e}_{\text{f}})_{\text{system,inlet}}} \quad (9)$$

where $\dot{W}_{\text{net}}$ is the net power, T_0 is the environment's temperature (298.15 K), T_b is the system's boundary temperature, $\dot{Q}_{\text{recovery}}$ is the thermal energy recovered, $\dot{n}_{\text{fuel}}$ is the fuel molar flow rate, and $\bar{e}_{\text{f}}$ is the flow exergy [84]. The numerator in Eqn. (9) accounts for exergy associated with both power generation and heat recovery. The factor $(1 - T_0/T_b)$ may be interpreted as the Carnot efficiency, which represents the maximum amount of work that could be obtained from $\dot{Q}_{\text{recovery}}$. The system's overall (cogeneration) efficiency is defined as the ratio of the system's output energy to the inlet fuel energy:

$$\eta_{\text{sys,CHP}} = \frac{\dot{W}_{\text{net}} + \dot{Q}_{\text{recovery}}}{(\dot{n}_{\text{fuel}} \cdot \text{HHV}_{\text{fuel}})_{\text{system,inlet}}} \quad (10)$$

where HHV_{fuel} is the fuel's higher heating value [47]. The system's electric efficiency is defined as the ratio of the net power to the fuel's energy [47]:

$$\eta_{\text{sys,e}} = \frac{\dot{W}_{\text{net}}}{(\dot{n}_{\text{fuel}} \cdot \text{HHV}_{\text{fuel}})_{\text{system,inlet}}} \quad (11)$$

and the fuel cell's stack's electric efficiency is defined similarly:

$$\eta_{\text{stack,e}} = \frac{\dot{W}_{\text{SOFC,DC}}}{(\dot{n}_{\text{fuel}} \cdot \text{HHV}_{\text{fuel}})_{\text{system,inlet}}} \quad (12)$$

where $\dot{W}_{\text{SOFC,DC}}$ is the DC power generated by the fuel cell [45].

The air ratio and fuel utilization indicate the system's reactant consumption. The air ratio is defined as the oxygen molar flow rate divided by twice the methane molar flow rate [47]:

$$\lambda_{\text{air}} = \frac{\dot{n}_{\text{O}_2,\text{system,inlet}}}{2\dot{n}_{\text{CH}_4,\text{system,inlet}}} \quad (13)$$

The system's fuel utilization represents the portion of the system's fuel that is oxidized inside the fuel cell stack. The fuel utilization is defined as follows:

$$U_{\text{f,sys}} = \frac{\dot{n}_{\text{H}_2,\text{consumed}}}{4\dot{n}_{\text{CH}_4,\text{sys}}} \quad (14)$$

where $U_{\text{f,sys}}$ is the system's fuel utilization, $\dot{n}_{\text{H}_2,\text{consumed}}$ is the molar flow rate of H_2 consumed by the fuel cell stack, and $\dot{n}_{\text{CH}_4,\text{sys}}$ is the molar flow rate of methane entering the system [47]. The denominator in Eqn. (14) represents the amount of H_2 yielded by the reforming and water-gas shift reactions, assuming that all of the fuel species are converted to H_2 , CO_2 , and H_2O .

3.3. Results

3.3.1. Baseline operation

Table 2 presents the stand-alone and hybrid systems' energetic and exergetic results during baseline operation. The hybrid system's exergetic efficiency exceeds that of the stand-alone system by about six percentage points.

Table 2 The stand-alone and hybrid systems' results during baseline operation.

	Stand-alone	Hybrid
Exergetic efficiency	59.2%	65.5%
Cogeneration efficiency	79.7%	80.2%
System's electric efficiency	45.3%	52.9%
Stack's electric efficiency	51.5%	44.3%
Thermal energy recovered	64.2 kW _{th}	43.5 kW _{th}
Thermal energy loss	5.6 kW _{th}	4.8 kW _{th}
Air ratio	3.0	2.6

The stand-alone system recovers approximately 50% more thermal energy than the hybrid system recovers, but the hybrid system operates at a higher electric efficiency than the stand-alone system. Table 3 presents the stand-alone system's exergy accounting. Exergy accounting is the evaluation of each component's exergy destruction, exergy loss, power, and cogeneration, presented in a form that allows for easy comparison between components [17]. The left column in Table 3 represents the exergy output associated with power generation and thermal energy recovery, and the right column represents the exergy that is destroyed or lost. Based on the results in Table 3, the air pre-heater destroys the most exergy. Air experiences a large temperature rise across the air pre-heater (nearly 600°C, as shown in Fig. 1). Air also forms the system's bulk flow, as indicated by the air ratio presented in Table 2 ($\lambda_{\text{air}} = 3.0$). Thus, significant heat transfer and exergy destruction occur in the air pre-heater. The fuel cell stack and auxiliary combustor also destroy and lose significant exergy. The chemical and electrochemical reactions inside these components give rise to irreversibilities [25]. The remaining components, including the heat recovery heat exchanger, inverter,

Table 3 The stand-alone system's exergy accounting. All values are expressed as a percentage of the system's inlet exergy.

	Power or cogeneration (%)	Exergy destroyed or lost (%)
Fuel cell stack	52.4	9.9
Fuel compressor	-0.7	0.1
Air blower	-3.3	0.9
Desulfurizer	-	0.0
Ejector	-	1.6
Fuel pre-heater	-	0.6
Air pre-heater	-	13.7
Heat recovery	10.8	3.6
Auxiliary combustor	-	5.9
Inverter	-	2.8
Exhaust	-	1.7
Sum	59.2	40.8

ejector, air blower, and fuel compressor, destroy smaller amounts of exergy, and a small fraction of the inlet exergy exits with the system exhaust.

Table 4 presents the hybrid system's exergy accounting. The fuel cell stack and auxiliary combustor destroy and lose significant amounts of exergy. The remaining components destroy smaller amounts of exergy. Unlike the stand-alone system, the hybrid system's air pre-heater destroys about 5% of the system's inlet exergy. In the hybrid system, the air experiences a lower temperature rise across the air pre-heater compared to the stand-alone system due to the compressor's pre-heating. The hybrid system's air flow rate is also less than the stand-alone system's air flow rate. Consequentially, the hybrid system's air pre-heater transfers less heat, and thus, destroys less exergy than the stand-alone system's air pre-heater. The hybrid system's

Table 4 The hybrid system’s exergy accounting. All values are expressed as a percentage of the system’s inlet exergy.

	Power or cogeneration (%)	Exergy destroyed or lost (%)
Fuel cell stack	45.1	8.4
Fuel compressor	−1.0	0.2
Air compressor	−12.3	1.7
Turbine	24.8	1.8
Desulfurizer	-	0.0
Ejector	-	1.3
Air pre-heater	-	4.9
Fuel pre-heater	-	0.3
Heat recovery	8.9	3.0
Power conditioning (stack)	-	2.4
Power conditioning (turbomachinery)	-	0.6
Auxiliary combustor	-	8.1
Exhaust	-	1.9
Sum	65.5	34.6

exergetic performance furthermore depends on the turbomachinery’s operation. The air compressor consumes 12% of the system’s inlet exergy, and the turbine converts nearly 25% of the system’s inlet exergy to mechanical power.

3.3.2. Sensitivity analysis

Figure 3 presents the stand-alone system’s energetic and exergetic results vs. the fuel cell’s operating voltage (Fig. 3a), fuel utilization (Fig. 3b), and average PEN temperature (Fig. 3c). During each sensitivity analysis, all operating parameters otherwise assume their baseline values (Table 1). As shown in Fig. 3a, increasing the fuel cell’s operating voltage from slightly over 0.5 V to 0.8 V increases the system’s exergetic efficiency from 45% to 67%. Higher voltages correspond to higher fuel cell electric efficiencies, re-

sulting in higher exergetic efficiencies. Figure 3a also indicates that the heat recovery declines at higher voltages, which is likely due to the declining waste heat production inside the fuel cell stack at higher voltages. The waste heat production equals the enthalpy change of reaction minus the power generated by the fuel cell stack. As shown in Fig. 3b, increasing the system's fuel utilization from 0.77 to 0.90 decreases the system's exergetic efficiency by about two percentage points, despite the system's higher electric efficiency at higher fuel utilizations. Increasing the system's fuel utilization decreases the amount of fuel combusted in the auxiliary combustor, resulting in lower exhaust temperatures. Lower exhaust temperatures, in turn, reduce both the quantity and quality (temperature) of the recovered thermal energy, resulting in lower exergetic efficiencies. Increasing the average PEN temperature from 650°C to 850°C increases the system's exergetic efficiency by about two percentage points (Fig. 3c). Increasing the PEN temperature increases the fuel cell stack's power density, thereby decreasing the required fuel cell stack area.

(a)

(b)

(c)

Figure 3 Sensitivity of the stand-alone system's energetic and exergetic results to changes in the fuel cell's operating parameters: (a) Varying the SOFC's voltage, (b) Varying the system's fuel utilization, (c) Varying the average PEN temperature. The dashed lines indicate the minimum value of the independent variable at which a solution was found.

Figure 4 presents the hybrid system's energetic and exergetic results as a function of the fuel cell's operating voltage (Fig. 4a), fuel utilization (Fig. 4b), average PEN temperature (Fig. 4c), and the microturbine's pressure ratio (Fig. 4d). Increasing the fuel cell's operating voltage increases the system's exergetic efficiency by 12 percentage points. In Fig. 4b, increasing the system's fuel utilization decreases the system's exergetic efficiency. Increasing the system's fuel utilization reduces the amount of fuel combusted in the auxiliary combustion, resulting in lower turbine inlet temperatures. In order to maintain the turbine inlet temperature at 1,125°C, the system requires additional methane (stream 6 in Fig. 2). The additional methane reduces the system's electric efficiency. As shown in Fig. 4c, increasing the average PEN temperature increases the system's exergetic efficiency. The heat recovery, on the other hand, declines at higher temperatures. The reduction in required stack cooling with increasing PEN temperature likely explains the reduction in air ratio and heat recovery. Finally, Fig. 4d indicates that increasing the pressure ratio reduces the amount of thermal energy recovered, but not sufficiently to reduce the system's exergetic efficiency. The rise in net turbine power, which equals the power generated by the turbine minus the power consumed by the air compressor, explains the exergetic efficiency's rise with increasing pressure ratio.

Figure 4 Sensitivity of the hybrid system's energetic and exergetic results to changes in the fuel cell's operating parameters: (a) Varying the SOFC's voltage, (b) Varying the system's fuel utilization, (c) Varying the average PEN temperature, (d) Varying the pressure ratio. The dashed lines indicate the minimum value of the independent variable at which a solution was found.

4. Economic analysis

While energy and exergy provide useful metrics for evaluating a system's thermodynamic performance, investment decisions depend on a system's life cycle cost. A system's life cycle cost includes fuel costs, capital costs, maintenance costs, and salvage value. A system's life cycle cost also includes utility costs if utilities provide supplemental thermal energy and electricity to a building. The present study adopts Duffie and Beckman's P_1 - P_2 method [85] to calculate the hybrid and stand-alone systems' life cycle costs. Subsection 4.1 describes the cost model. Subsection 4.2.1 presents the systems' baseline costs, and Subsections 4.2.2 and 4.2.3 vary the systems' operating parameters.

4.1. Cost model

Duffie and Beckman's P_1 - P_2 method yields a system's life cycle cost:

$$LCC = P_1 F + P_2 E \quad (15)$$

where P_1 is the ratio of the system's life cycle operating cost to the first year's operating cost, F is the first year operating cost in dollars, P_2 is the ratio of the system's life cycle capital cost to the initial capital cost, and E is the initial capital cost in dollars [85]. The constant P_1 is a present worth factor that accounts for inflation. The constant P_2 accounts for numerous other factors, including a down payment, loan interest, income tax, property tax, insurance, maintenance, depreciation, and the system's salvage value. The P_1 - P_2 method, as applied to grid-connected fuel cell systems, also accounts for utility costs. In the present study, the power grid provides supplemental

power when the fuel cell stack is unable to meet demand, and an auxiliary boiler provides supplemental thermal energy. The P_1 - P_2 method applied to grid-connected fuel cell systems takes the following form [46]:

$$\begin{aligned} \text{LCC}_{\text{system}} = & P_{1,\text{utility,ngas}}(F_{\text{system,ngas}} + F_{\text{utility,ngas,supp}}) \\ & + P_{1,\text{utility,electric}}F_{\text{utility,electric,supp}} + P_{2,\text{system}}E_{\text{system}} \end{aligned} \quad (16)$$

where the first term on the right-hand side of Eqn. (16) represents the cost of natural gas consumed by the system and auxiliary boiler, the second term represents the cost of supplemental grid power, and the third term represents the system's capital cost.

The present work adopts Braun's methodology [46] to calculate the first year operating costs. The system's first year natural gas cost is given by:

$$F_{\text{system,ngas}} = \frac{\dot{W}_{\text{net}} \times 8,760 \text{ h} \times \text{CF}_e A_{\text{system}} F_{\text{C,ngas}}}{\eta_{\text{sys,e}}} \quad (17)$$

where CF_e is the electric capacity factor, A_{system} is the system's availability, and $F_{\text{C,ngas}}$ is the natural gas price. In Eqn. (17), notice that the net power is multiplied by 8,760 hours to yield the power generated per year. The cost of natural gas consumed by the auxiliary boiler depends on the supplemental thermal energy required by the building, which is calculated according to the following formula:

$$F_{\text{utility,ngas,supp}} = \frac{(\dot{Q}_{\text{demand,average}} - \dot{Q}_{\text{recovery}} \text{CF}_e A_{\text{system}}) \times 8,760 \text{ h} \times F_{\text{C,ngas}}}{\eta_{\text{boiler}}} \quad (18)$$

where $\dot{Q}_{\text{demand,average}}$ is the average thermal energy demand, and η_{boiler} is the auxiliary boiler's thermal efficiency. Similarly, the cost associated with

drawing supplemental power is given by:

$$F_{\text{utility,electric,supp}} = \dot{W}_{\text{net}} \times 8,760 \text{ h} \times F_{\text{C,electric}} \times (1 - \text{CF}_e A_{\text{system}}) \quad (19)$$

where $F_{\text{C,electric}}$ is the electric utility price. The power grid meets demand when the fuel cell stack is unavailable or operates at less than capacity.

The present study also calculates the systems' life cycle savings. A system's life cycle savings reflects a system's economic performance relative to that of a conventional system. The conventional system consists of the power grid and an auxiliary boiler. The present work adopts Braun's methodology [46] to calculate the life cycle savings. The life cycle savings is given by:

$$\text{LCS} = \text{LCC}_{\text{utility}} - \text{LCC}_{\text{system}} \quad (20)$$

where $\text{LCC}_{\text{utility}}$ is the utility's life cycle cost. The utility's life cycle cost is given by:

$$\text{LCC}_{\text{utility}} = P_{1,\text{utility,ngas}} F_{\text{utility,ngas,only}} + P_{1,\text{utility,electric}} F_{\text{utility,electric,only}} \quad (21)$$

In Eqn. (21), $F_{\text{utility,ngas,only}}$ is the auxiliary boiler's operating cost during utility-only operation, and $F_{\text{utility,electric,only}}$ is the electric utility's cost during utility-only operation. The auxiliary boiler's operating cost is given by:

$$F_{\text{utility,ngas,only}} = \frac{\dot{Q}_{\text{demand,average}} \times 8,760 \text{ h} \times F_{\text{C,ngas}}}{\eta_{\text{boiler}}} \quad (22)$$

and the electric utility's cost is given by:

$$F_{\text{utility,electric,only}} = \dot{W}_{\text{demand,average}} \times 8,760 \text{ h} \times F_{\text{C,electric}} \quad (23)$$

where $\dot{W}_{\text{demand,average}}$ is the average annual power demand.

The present study adopts a bottom-up approach to estimating the system's capital cost, approximating each component cost based on previous estimates in the literature and scaling these estimates to the desired size. The component cost is given by:

$$C_i = C_{i,0} \left(\frac{S_i}{S_{i,0}} \right)^n \frac{\text{CEPCI}}{\text{CEPCI}_0} \quad (24)$$

where C_i is the cost of component i in $\$ \text{ kW}^{-1}$, $C_{i,0}$ is the reference cost of component i in $\$ \text{ kW}^{-1}$, S_i is the capacity of component i , $S_{i,0}$ is the reference capacity of component i , n is the capacity scaling exponent, CEPCI is the Chemical Engineering Plant Cost Index, and CEPCI_0 is the reference Chemical Engineering Plant Cost Index [86].¹ Table 5 presents each component's reference capacity, reference cost, and scaling exponent. Numerous studies provide these parameters [37, 45, 46, 87–91]. The National Energy Technology Laboratory [37] estimates the fuel cell cost to be $\$540 \text{ m}^{-2}$ (2007 USD), based on the U.S. Department of Energy's 2010 cost goal. The Pacific Northwest National Laboratory [87] estimates the inverter cost to be $\$39,052$ (2013 USD) for a 270 kW hybrid system. The technology development company

¹If the reference cost, $C_{i,0}$, is not already expressed in $\$ \text{ kW}^{-1}$ (see Table 5), then the right-hand side of Eqn. (24) is divided by the net power generated to yield units of $\$ \text{ kW}^{-1}$.

TIAX [88] estimates the air blower and fuel pre-heater costs to be on par with one another, each costing approximately \$11,500 (2002 USD) in a 250 kW stand-alone system. TIAX estimates the air pre-heater cost to be \$144,948 (2002 USD). R&D Dynamics Corporation [92] estimated the fuel compressor cost to be \$500 per unit, assuming a production quantity of 50,000 units yr⁻¹. Based on Battelle’s analysis of 1 kW and 5 kW SOFC auxiliary power units [89], we estimate that assembly and testing costs to be slightly over 4% of the system’s equipment cost. Boehm [90] and Turton, et al. [91] provide the generator and turbine cost parameters, respectively. The scaling exponents in Table 5 draw primarily from Braun’s techno-economic analysis of residential-scale SOFC systems [45, 46]. The system’s capital cost is given by:

$$C_{\text{system}} = \sum_i C_i \quad (25)$$

where C_{system} is the system’s capital cost in \$ kW⁻¹. The uncertainty associated with the capital cost is $\pm 30\%$. Such an estimate reflects a “study estimate,” or “factored estimate” [93].

Finally, the system’s initial cost (2014 USD), appearing in the LCC formula (Eqn. (16)), is given by:

$$E_{\text{system}} = C_{\text{system}} \dot{W}_{\text{net}} \quad (26)$$

The study’s economic and technical parameters appear in Table 6. The equipment inflation rate is based on changes in the Chemical Engineering Plant Cost Index over the previous 10 years [94]. Similarly, the electric utility’s inflation rate is estimated based on historical annual retail electricity prices in

Table 5 Cost parameters for estimating the system components' costs.^a

System component	Reference capacity ($S_{i,0}$)	Reference cost ($C_{i,0}$)	Scaling exponent (n)
Fuel cell stack	$A_{\text{stack}} = 116.29 \text{ m}^2$	\$68,856	[37] 1.00
Stack enclosure	$p_{\text{stack}} = 1.21 \text{ atm}$	\$2,496	[37] 0.33 [35]
Air blower	$\dot{W}_{\text{blower}} = 5.7 \text{ kW}$	\$6,643	[88] 0.807 [45, 46]
Air filter	$\dot{W}_{\text{blower}} = 5.7 \text{ kW}$	\$306	[88] 0.807 [45, 46]
Fuel compressor	$\dot{W}_{\text{comp,fuel}} = 1.1 \text{ kW}$	\$500	[92] 0.67 [45]
Desulfurizer	$\dot{m}_{\text{desulfizer}} = 2.47 \text{ g s}^{-1}$	\$1,176	[87] 0.67 [45, 46]
Ejector	$\dot{m}_{\text{ejector}} = 36.27 \text{ g s}^{-1}$	\$116	[81] 0.67 [45, 46]
Air pre-heater	$UA_{\text{air}} = 343.74 \text{ W K}^{-1}$	\$22,444	[88] 0.67 [45]
Fuel pre-heater	$UA_{\text{fuel}} = 18.08 \text{ W K}^{-1}$	\$13,601	[88] 0.67 [45]
Heat recovery	$UA_{\text{cogen}} = 387.69 \text{ W K}^{-1}$	\$26,951	[10] 0.67 [45]
Inverter	$\dot{W}_{\text{net}} = 84.2 \text{ kW}$	\$190 kW^{-1}	[87] -0.22 [45, 46]
Auxiliary combustor	$\dot{m}_{\text{comb}} = 175.19 \text{ g s}^{-1}$	\$4,211	[89] 0.82 [46]
Generator	$\dot{W}_{\text{net}} = 84.2 \text{ kW}$	\$61 kW^{-1}	[90] -0.34 [90]
Turbine ^c	-	-	- -
Additional BOP ^b	$\dot{W}_{\text{net}} = 84.2 \text{ kW}$	\$58 kW^{-1}	[89] -0.829 [46]
Assembly/testing	-	4.31% $\times$ equip. cost	[89] - -
Installation	-	25% $\times$ equip. cost	[93] - -

^a Costs are expressed in 2014 USD during baseline operation (84.2 kW net power).

^b Includes electronics and controls, instrumentation, and assembly components.

^c The turbine's capital cost is calculated according to Turton, et al.'s formula [91]: $\log_{10} C_{\text{turbine}} = 2.2476 + 1.4965 \log_{10} (\dot{W}_{\text{turbine}}) - 0.1618 [\log_{10} (\dot{W}_{\text{turbine}})]^2$ (2001 USD), where $\dot{W}_{\text{turbine}}$ is expressed in kW.

Table 6 The stand-alone and hybrid systems' economic and technical parameters.

Parameter	Value
System lifetime	20 years
Equipment inflation rate	2.78%
Electric utility inflation rate	2.83%
Discount rate	10%
Income tax rate	35%
Interest rate	8%
Down payment (% capital cost)	10%
Insurance and maintenance (% capital cost)	7%
Property tax	0.75%
Salvage value	None
Electric capacity factor, CF_e	95%
System availability, A_{system}	100%
Boiler's efficiency, η_{boiler}	80% (HHV)
Natural gas price	2.94 ¢/kWh
Electricity price	10.74 ¢/kWh

the U.S. over the previous 10 years [95]. The annual maintenance cost equals 6% of the system's capital cost based on estimates provided by Peters, et al. [93]. The maintenance cost accounts for the SOFC stack's and desulfurizer sorbent's replacement costs, assuming a fuel cell stack lifetime of five years [34] and desulfurizer sorbent lifetime of one year [46]. The property tax rate is the same as that used in the Pacific Northwest National Laboratory's cost analysis of a 270 kW hybrid system [87]. The auxiliary boiler's efficiency is based on ASHRAE's estimates of conventional steam generation technologies [96]. The average retail electricity price for commercial customers is 10.74 ¢ kWh⁻¹ (2014 USD) [95], and the fuel cost is the average commercial natural gas price in 2014 of 2.94 ¢ kWh⁻¹ [97].

4.2. Results

4.2.1. Baseline performance

Figure 5a presents the systems' life cycle costs during electric-only (EO) and cogeneration (CHP) operation. EO operation assumes that the system generates power only, meaning $\dot{Q}_{\text{recovery}} = 0$ kW in Eqn. (18), and the auxiliary boiler meets the thermal energy demand entirely. The present study normalizes all life cycle costs by the stand-alone system's life cycle cost during EO operation. As shown in Fig. 5a, the hybrid system costs less than the stand-alone system during EO and CHP operation. The fuel cell stack's cost drives the stand-alone system's capital cost upward. The stand-alone system's fuel cell stack costs \$818/kW, constituting 30% of the system's capital cost. The hybrid system's fuel cell stack costs \$282/kW, constituting 12% of the system's capital cost. As shown in Fig. 5b, the amount of money saved by the hybrid and stand-alone systems depends on the systems' operation. During EO operation, both systems lose money relative to the conventional system, but the hybrid system comes closer to breaking even. During CHP operation, the hybrid system saves nearly three times more money than the stand-alone system.

(a)

(b)

Figure 5 The stand-alone and hybrid systems' economic results during baseline operation: (a) Life cycle cost, normalized by the reference case, during baseline operation, (b) Life cycle savings.

4.2.2. Multivariable parametric analysis

In order to identify local minima in the systems' life cycle costs, the present study varies the fuel cell's operating parameters. The stand-alone system's operating voltage (V_{cell}) varies between 0.6 V and 0.8 V, the fuel utilization ($U_{\text{f,sys}}$) varies between 0.7 and 0.9, and the average PEN temperature ($T_{\text{PEN,average}}$) varies between 700°C and 850°C. Braun [46] considered these ranges when minimizing a stand-alone system's cost. Varying the system's operating parameters over these ranges, especially the PEN temperature, significantly influences the component costs (Eqn. (24)). Increasing the temperature above 850°C requires ceramic interconnect [46], the cost of which is not included in the present study. Figure 6 presents the stand-alone system's life cycle cost vs. PEN temperature and operating voltage during CHP operation at 85% fuel utilization (Fig. 6a) and 90% fuel utilization (Fig. 6b). In both subfigures, the life cycle cost exhibits parabolic curvature with respect to voltage. Such behavior reflects the trade-off between fuel and capital costs. That is, increasing the fuel cell's operating voltage increases the fuel cell's electric efficiency, thereby reducing the system's fuel costs. Increasing the fuel cell's operating voltage also reduces the fuel cell's power density, however, thereby increasing the fuel cell stack's capital cost. In both subfigures, increasing the PEN temperature reduces the system's life cycle cost. Increasing the fuel cell's temperature increases the fuel cell's power density, thereby reducing the system's capital cost. Lastly, increasing the fuel utilization from 85% (Fig. 6a) to 90% (Fig. 6b) elevates the life cycle cost across all values of the temperature and voltage considered. The stand-alone system yields a minimum normalized life cycle cost of 0.74 at $V_{\text{cell}} = 0.725$ V, $U_{\text{f,sys}} = 0.775$, and $T_{\text{PEN,average}} = 850^\circ\text{C}$.

Similar to the stand-alone system, the hybrid system's life cycle cost depends on numerous operating parameters. The present study varies the hybrid system's operating voltage between 0.6 V and 0.8 V, fuel utilization between 0.7 and 0.9, and average PEN temperature between 650°C and 850°C. The pressure ratio and turbine inlet temperature remain constant during these analyses. Figure 6 presents the hybrid system's life cycle cost vs. PEN temperature and operating voltage during CHP operation at 85% fuel utilization (Fig. 6c) and 90% fuel utilization (Fig. 6d). The surface in Fig. 6c spans $V_{\text{cell}} = 0.625$ V and 0.725 V. Outside this range, the model fails to yield a solution at certain temperatures. Similar to the stand-alone system, the hybrid system's life cycle cost exhibits parabolic curvature with respect to voltage, particularly at lower temperatures and 90% fuel utilization (Fig. 6d), again reflecting the trade-off between the system's fuel and capital costs. The hybrid system's life cycle cost also exhibits parabolic curvature with respect to temperature. As the temperature increases, the additional methane and stack area decrease, whereas the heat recovery decreases and air pre-heater UA product increases. After performing the parametric analysis, the hybrid system exhibits a minimum normalized life cycle cost of 0.79 at the following operating parameters: $V_{\text{cell}} = 0.725$ V, $U_{\text{f,sys}} = 0.80$, and $T_{\text{PEN,average}} = 725^\circ\text{C}$.

(a)

(b)

(c)

(d)

Figure 6 The stand-alone and hybrid systems' optimization results, varying the SOFC's operating voltage and average PEN temperature: (a) Stand-alone, $U_{f,system} = 0.85$, (b) Stand-alone, $U_{f,system} = 0.90$, (c) Hybrid, $U_{f,system} = 0.85$, (d) Hybrid, $U_{f,system} = 0.90$.

Figure 7 Varying the turbine inlet temperature and pressure ratio at $V_{\text{cell}} = 0.725$ V, $U_{\text{f,sys}} = 0.80$, and $T_{\text{PEN,average}} = 725^\circ\text{C}$.

In addition to varying the hybrid system's fuel cell operating parameters, the present study separately varied the hybrid system's pressure ratio (PR) between three and four, and the turbine inlet temperature (TIT) between $1,125^\circ\text{C}$ and $1,250^\circ\text{C}$. Martinez, et al. recommended maintaining the hybrid system's turbine inlet temperature below $1,250^\circ\text{C}$ [6]. Figure 7 presents the hybrid system's life cycle cost vs. the turbine inlet temperature and pressure ratio. Increasing the turbine inlet temperature steadily reduces the system's life cycle cost. Higher turbine inlet temperatures generally correspond to higher microturbine efficiencies, thus reducing the system's operating cost. Increasing the system's pressure ratio from three to four reduces the system's life cycle cost. The minimum life cycle cost of 0.78 occurs at $\text{PR} = 4$ and $\text{TIT} = 1,250^\circ\text{C}$, and the optimal fuel cell operating parameters stated in the previous paragraph. Following the turbomachinery's optimization, we again varied the fuel cell's operating parameters, assuming $\text{PR} = 4$ and $\text{TIT} = 1,250^\circ\text{C}$. The minimum life cycle cost of 0.76 occurs at $V_{\text{cell}} = 0.75$ V, $U_{\text{f,sys}} = 0.80$, and $T_{\text{PEN,average}} = 775^\circ\text{C}$.

(a)

(b)

Figure 8 The stand-alone and hybrid systems' economic results. The systems have been optimized during CHP operation: (a) Normalized life cycle cost, (b) Life cycle savings.

4.2.3. Optimized performance

Figure 8 presents the systems' economic results after optimizing the systems. Figure 8a compares the systems' life cycle costs. During EO and CHP operation, the system costs differ slightly. Figure 8b presents the systems' life

cycle savings. Unlike the baseline results, both systems save money relative to the conventional system during EO operation. The hybrid system saves over twice as much money as the stand-alone system during EO operation. During CHP operation, the hybrid system saves 16% less money than the stand-alone system. Thus, the stand-alone system outperforms the hybrid system after optimization and during cogeneration.

5. Conclusions

The present study compared a hybrid system's exergetic and economic performance with that of a stand-alone system. The hybrid system's exergetic efficiency exceeded that of the stand-alone system by six percentage points during baseline operation. Increasing the fuel cell's operating voltage from slightly over 0.5 V to 0.8 V increased the hybrid system's exergetic efficiency by 12 percentage points, and the stand-alone system's exergetic efficiency by 22 percentage points. Varying other operating parameters, including the fuel utilization and pressure ratio, also influenced the systems' exergetic performance, though to a lesser extent. The hybrid system cost less than the stand-alone system during baseline operation. The stand-alone system outperformed the hybrid system after optimizing the system and when cogenerating thermal energy. Under these conditions, the stand-alone system saved approximately \$153,000 relative to a conventional system, whereas the hybrid system saved approximately \$129,000.

The present study's findings support hybrid systems' continued development. Hybrid systems perform particularly well during electric-only operation, which may include remote power applications. Stand-alone systems

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9 outperform hybrid systems following optimization and during cogeneration
10 operation. Stand-alone systems may thus be better suited for cogeneration
11 applications, including manufacturing processes, commercial buildings, and
12 similar applications that demand coincident power and thermal energy. It is
13 recommended that future work simulate hybrid and stand-alone systems un-
14 der a range of thermal-to-electric ratios to reflect different building types and
15 operation. We also recommend that future work optimize the systems dur-
16 ing electric-only operation. The systems' part-load performance, dynamic
17 behavior, and environmental impacts serve as additional topics for future
18 work.
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List of Tables

Table 1	Baseline operating parameters.
Table 2	The stand-alone and hybrid systems' results during base-line operation.
Table 3	The stand-alone system's exergy accounting. All values are expressed as a percentage of the system's inlet exergy.
Table 4	The hybrid system's exergy accounting. All values are expressed as a percentage of the system's inlet exergy.
Table 5	Cost parameters for estimating the system components' costs.
Table 6	The stand-alone and hybrid systems' economic and technical parameters.

List of Figures

- Figure 1 The stand-alone system's process flow diagram (adapted from Refs. [46, 59, 60]). Thermodynamic properties are provided during baseline operation (Table 1).
- Figure 2 The hybrid system's process flow diagram (adapted from Refs. [1–6, 21, 30, 40, 41, 44, 61–66]). Thermodynamic properties are provided during baseline operation (Table 1).
- Figure 3 Sensitivity of the stand-alone system's energetic and exergetic results to changes in the fuel cell's operating parameters: (a) Varying the SOFC's voltage, (b) Varying the system's fuel utilization, (c) Varying the average PEN temperature. The dashed lines indicate the minimum value of the independent variable at which a solution was found.
- Figure 4 Sensitivity of the hybrid system's energetic and exergetic results to changes in the fuel cell's operating parameters: (a) Varying the SOFC's voltage, (b) Varying the system's fuel utilization, (c) Varying the average PEN temperature, (d) Varying the pressure ratio. The dashed lines indicate the minimum value of the independent variable at which a solution was found.
- Figure 5 The stand-alone and hybrid systems' economic results during baseline operation: (a) Life cycle cost, normalized by the reference case, during baseline operation, (b) Life cycle savings.
- Figure 6 The stand-alone and hybrid systems' optimization results, varying the SOFC's operating voltage and average PEN temperature: (a) Stand-alone, $U_{f,system} = 0.85$, (b) Stand-alone, $U_{f,system} = 0.90$, (c) Hybrid, $U_{f,system} = 0.85$, (d) Hybrid, $U_{f,system} = 0.90$.

Figure 7 Varying the turbine inlet temperature and pressure ratio, assuming $V_{\text{cell}} = 0.725 \text{ V}$, $U_{\text{f,sys}} = 0.80$, and $T_{\text{PEN,average}} = 725^\circ\text{C}$.

Figure 8 The stand-alone and hybrid systems' economic results. The systems have been optimized during CHPoperation: (a) Normalized life cycle cost, (b) Life cycle savings.