



# Exergy & economic analysis of biogas fueled solid oxide fuel cell systems



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## HIGHLIGHTS

- Exergy analysis of a pressurized SOFC integrated with an anaerobic digester.
- Capital cost estimation of all major equipment in AD-SOFC power plant.
- IRR economic analyses of AD-SOFC system with sensitivity analysis.
- Compared values of IRR with other available technologies for biogas utilization.
- Of options studied, AD-SOFC achieved both highest efficiency and highest IRR.

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## ABSTRACT

We present an exergy and an economic analysis of a power plant that uses biogas produced from a thermophilic anaerobic digester (AD) to fuel a solid oxide fuel cell (SOFC). We performed a 4-variable parametric analysis of the AD-SOFC system in order to determine the optimal design operation conditions, depending on the objective function of interest. We present results on the exergy efficiency (%), power normalized capital cost (\$ kW<sup>-1</sup>), and the internal rate of return on investment, IRR, (% yr<sup>-1</sup>) as a function of the current density, the stack pressure, the fuel utilization, and the total air stoichiometric ratio. To the authors' knowledge, this is the first AD-SOFC paper to include the cost of the AD when conducting economic optimization of the AD-SOFC plant. Our calculations show that adding a new AD-SOFC system to an existing waste water treatment (WWT) plant could yield positives values of IRR at today's average electricity prices and could significantly out-compete other options for using biogas to generate electricity. AD-SOFC systems could likely convert WWT plants into net generators of electricity rather than net consumers of electricity while generating economically viable rates of return on investment if the costs of SOFC systems are within a factor of two of the DOE/SECA cost targets.

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## 1. Introduction

Biomass in municipal waste water is one of many promising fuels for cost-effectively generating electricity without increasing the cost of food or increasing the concentration of carbon dioxide in the atmosphere. In 2004, the United States generated over 120 billion liters per day of municipal waste water [1], which amounts to approximately 400 L of waste water per day per person. Typical values of waste water chemical oxygen demand (COD) in Europe and the US vary between 0.2 and 0.6 g per liter [2]. Assuming an average COD of 0.5 g L<sup>-1</sup>, the waste water in the US has a potential chemical enthalpy of about 14 GW<sub>th</sub>. Assuming that the chemical enthalpy in

the waste water can be converted into electricity at an efficiency of 50%, the waste water produced in the US could potentially generate on average roughly 7 GW of electricity. While this value is only a little more than 1% of the roughly 500 GW of time-averaged electricity generation in the US in 2010 [3], it should not be overlooked. In addition, the conversion of the chemical exergy in the waste water into electricity represents a viable market for fuel cell systems in the near future because the scale of typical waste water treatment plants corresponds closely with the size of today's large-scale fuel cell systems (100 kW–1 MW) and because the high efficiency of the SOFC system is important for reducing the power normalized upfront capital cost (\$ kW<sup>-1</sup>) of the anaerobic digester (AD).

Currently, most waste water treatment plants in the US use aerobic bacteria as opposed to anaerobic bacteria to remove the COD in waste water [4]. For example, of the roughly 130 waste water treatment plants in Massachusetts as of 2011, only 6 of them

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were using or were in the process of installing anaerobic digesters at the treatment facility [5]. Aerobic digester facilities consume approximately 2% of the electricity generated in the US [6]. The pumps and air blowers required to operate aerobic digesters are the largest source of electricity demand at waste water treatment plants. In an aerobic digester, there is near complete destruction of the original exergy [7,8] in the waste water because the chemical reactions are occurring near the temperature, pressure, and chemical composition of the environment; hence cannot be used to generate useful work.

However, there are now a growing number of waste water treatment plants globally that use anaerobic digesters [6]. In addition, there are now a few cases in which the biogas from the waste water anaerobic digesters (AD) is sent directly to fuel cell systems. One example is the King County Carbonate Fuel Cell Demonstration Project [9], which in 2004–2005 integrated an anaerobic digester with a 1-MW molten carbonate fuel cell produced by Fuel Cell Energy (AD-MCFC.) A couple more examples are the recent start-up of a 1.4 MW molten carbonate fuel cell at a waste water treatment facility in California [10] and the planned 0.2 MW molten carbonate fuel cell at a waste water treatment facility in Wyoming [11]. There have been a number of previous numerical simulations and economic studies on integrating MCFCs with anaerobic digesters [12–15], as well as simulations on integrating SOFCs with anaerobic digesters [16–20]. However, to the best of the author's knowledge, there has not been an economic analysis of an AD-MCFC or AD-SOFC system that includes the upfront capital cost of the AD and that uses fuel cell capital costs in  $\$ m^{-2}$  to graph the effect of current density and pressure on the economic figures of merit.

While MCFC systems (a) can use biogas as fuel, (b) can generate electricity at a relatively high system efficiency, and (c) currently have a larger installed capacity than SOFC systems, we decided here to model a SOFC rather than a MCFC because: (a) the all-ceramic SOFC materials can achieve higher current densities at a given voltage [21], and therefore may potentially achieve lower capital costs per power generated than MCFC technology [22]; (b) the Department of Energy (DOE) via the solid state energy conversion alliance (SECA) publishes cost estimates and cost targets for SOFC stack technologies; and (c) LG-Rolls Royce Fuel Cell Systems (Canton, OH) has published the  $V-i$  curves for their pressurized SOFC systems [23]. The most studied means of achieving the high system efficiencies is to pressurize the fuel cell stacks and generate net electricity from the combined compressor–expander [24–30].

SOFC's can operate on biogas provided that the  $CH_4$  is reformed by  $CO_2$  or  $H_2O$  either internally or externally. The advantage of internal reforming is that the endothermic reforming reaction inside of the SOFC offset some of the thermal energy generated by Ohmic and electrode resistance. Since an internally reformed SOFC has less net thermal energy generation, here we have only modeled an AD-SOFC system that has internal-reforming of biogas.

A number of research groups have already obtained successful experimental results using internal reforming of biogas to power a SOFC [17,20,31–41]. For example, Shiratori et al. [33] operated a SOFC for 50 h with direct biogas using a Ni-ScSZ cermet as the anode material without external reforming of the biogas. The gas composition sent to the SOFC was approximately 62%  $CH_4$ , 36%  $CO_2$ , 2%  $H_2O$  from a mesophilic anaerobic digester. After 50 h of operation, they saw no carbon formation on the anode. They did measure a roughly 100 mV drop in operating voltage when 1 ppm of  $H_2S$  was added into the biogas stream. The cell voltage recovered the 100 mV drop in voltage after the 1 ppm level of  $H_2S$  was removed. In a subsequent paper, Shiratori et al. [32] operated the SOFC at 800 °C, and they measured continuous degradation in operating voltage with a 790 ppm concentration of  $H_2S$ , which caused fatal

shutdown of the SOFC after 45 h. Shiratori et al. [32] also saw significant deposit of carbon when they ran at 800 °C for over 700 h of operation. Interestingly, there was no carbon deposit when they operated using simulated biogas. This suggests that carbon deposition is more complicated than simply knowing the C:H:O ratio in the anode. On the other hand, Staniforth and Kendall [39] determined that there would be major issues with carbon deposition unless small amounts of air were added to the fuel stream before entering the anode inlet of the SOFC.

The conclusions we draw from this prior research are the following: (a) the concentration of  $H_2S$  must be less than 1 ppm if pure Ni-YSZ is the anode material, and (b) there must either be a significant amount of anode gas recycle or enough carbon dioxide and water vapor in the biogas in order to reform methane and to minimize carbon deposition. Even though future SOFC anode materials may be sulfur tolerant [31,34,42], in this study, we do not assume that the Ni-YSZ anode is doped with materials to improve the sulfur tolerance. Therefore, we discuss in later sections cost effective means of reducing the  $H_2S$  concentration to less than 1 ppm.

The goal of this paper is to present a detailed exergy and economic analysis of the combined power plant. We plot the efficiency (%), power normalized capital cost ( $\$ kW^{-1}$ ), and the internal rate of return on investment ( $\% yr^{-1}$ ) as a function of the current density, the stack pressure, the fuel utilization, and the total air stoichiometric ratio. All four independent variables were simultaneously varied in order to find the optimal, steady-state, design conditions. To the authors' knowledge, this is the first AD-SOFC paper to include the cost of the AD when conducting economic optimization of the AD-SOFC plant.

## 2. Review & modeling of equipment in AD-SOFC plant

Fig. 1 shows a process flow diagram of the AD-SOFC power plant system. The main components of this system are: (a) the anaerobic digester, (b) the SOFC, (c) the gas turbine (i.e. compressor, combustor, & expander), (d) the heat exchangers, and (e) the balance of plant equipment, such as filters, pumps, and fuel compressor. The AD was modeled using HSC Chemistry 6.0 (Outotec, Espoo, Finland), and the model was exported to Excel (Microsoft, Seattle, WA) in order to perform a 4-variable parametric analysis using a Visual Basic macro. Before going into detailed discussions of each of the main components, our goal here is to discuss the overall process flow diagram. Additional pieces of equipment that would need to be added to the WWT plant are outside of the dashed box, and assumed existing pieces of equipment are shown inside of the dashed box. The proposed system uses a hydrocyclone to generate both a high COD waste stream ( $12 g L^{-1}$ ) to send to the new AD and a low COD waste water stream ( $<0.5 g L^{-1}$ ) to send to the existing aerobic digesters at the WWT plant. The un-reacted solid and liquid biomass from the AD is sent to the existing aerobic digester. The biogas from the AD is compressed, is heated in a heat exchanger using the exhaust gases, and then is sent to the  $H_2S$  polishing and siloxane removal reactor. Bulk  $H_2S$  capture is accomplished in the AD through the use of sacrificial iron oxides [43]. The  $H_2S$  polishing step is done using ZnO and a separate reactor that regenerates the ZnO (not shown in Fig. 1.) This high temperature reactor catalyzes the breakdown of siloxanes into silicates, effectively removing siloxanes before the SOFC. The gases exiting the anode and cathode of the SOFC are combusted and sent to a gas turbine before the exhaust gas is used to heat exchange with the incoming fuel and air and transfer thermal energy to the AD to maintain its temperature at 55 °C.

The majority of the equipment presented in Fig. 1 is commercially available and mature technology, with the exception of the

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