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Thermo-economic analysis of PEM fuel cell fuelled with biomethane obtained from human waste by computer simulation

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ABSTRACT

Fuel cells, especially Proton Exchange Membrane (PEMFC) is considered as perfect alternative energy source that can either replace or complement existing energy source which is fossil fuel. However, the technology is still at developmental stage due to the lack of availability of fuel sources that is safe and economical. This study, therefore focused on thermo-economic analysis of PEMFC fuelled with biogas from human waste. To achieve the purpose of this study, thermo-economic analysis was used to analyse PEMFC designed to generate 1.45 MW of electricity from domestic human waste. The stages involved are biogas generation, hydrogen production and fuel cell application. The processes were modelled using Aspen HYSYS V8.8, a software developed by Aspen Tech®. The Hydrogen was produced at a rate of 358 kgmole/h, temperature of 330 K, pressure of 4.8 bar and 99% purity from the Preferential oxidation (PrOx) exit stream. Data generated from the simulation model were subsequently used for the thermodynamic and economic analysis and a fuel processor efficiency of 83.5% was obtained. Energy analysis of the process showed that the principle of energy conservation was satisfied, requiring and producing energy simultaneously and a net electrical efficiency of 42.32% was realized, while the result of exergy analysis showed that the unit associated with high irreversibilities and maximum exergy destruction is the steam generator. From the economic analysis, a rate of return of 20% was realized which is an indication that the investment is safe and profitable. The general overview of the processes based on economic and thermodynamics performance shows that the investment is worthwhile and indeed waste can be turned to wealth.

Introduction

Due to the fact that the universe is moving towards a more effective, efficient and zero-pollution energy source, the need to utilize fuel cell as an alternative for generating electricity becomes imperative. A fuel cell can be defined as an electrical cell, which can be continuously supplied with a fuel to keep electrical power output sustained indefinitely [6]. It is an electrochemical device that can generate electricity without combustion through the reaction of hydrogen and oxygen

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to produce water and heat [19]. Out of all the numerous types of fuel cell, Proton exchange membrane fuel cell (PEMFC) is considered to be the best, owing to its high efficiency, high current density, simple stack structure, no loss of electrolyte during operation and low operating temperature [1]. It also has the advantages of quick start-up and response, long endurance and flexibility of fuel usage. But the most common explanation limiting its development, use and commercialization is high cost, which may be connected to the way the fuel (hydrogen) is being sourced and processed [6]. The use of pure hydrogen as the energy carrier requires an expensive hydrogen-fuelling network which leads to high costs in the fuel processing system [8]. Therefore, biomethane sourced from human waste can be a perfect alternative for hydrogen production in fuel cell application. Using biomethane from human waste as an energy source for generation of electric power in fuel cell overcome the problem of rapid depletion of fossil fuels coupled with global environmental challenges.

There are several thermo-chemical reforming processes that are used to produce hydrogen from biogas fuel or any other hydrocarbon fuel, the most common include steam reforming (SR), partial oxidation reforming (POR) and auto-thermal reforming (ATR) [7]. Dodds et al. [6] and Suleiman et al. [19] studied the steam reforming of methane as the traditional and common method for producing hydrogen based on industrial scale. Though this method can produce a high concentration of hydrogen of about 70% on dry basis, it is highly endothermic and therefore requires a considerable supply of external heat. On the other hand, partial oxidation as reported by Silberova et al., (2005) does not have the disadvantage of being endothermic, but it generates a significant amount of carbon (ii) oxide which is undesirable for proton exchange membrane fuel cells.

Auto-thermal reforming as studied by Cozzolino [5], Ersoz et al. [7] and Bae et al. (2005) combines the thermal effects of the oxidation and steam reforming reactions by supplying the fuel, water and air as input into the reactor. The ATR was selected for this study because the thermal energy generated from oxidation is absorbed by SR thereby lowering the overall temperature of the process. This favours the water-gas shift reaction which consumes carbon (ll) oxide and produces more hydrogen. Hence, the auto-thermal reactor is more compact, practical and suitable for use.

The selection and justification of the processes adopted for generation and purification of hydrogen for fuel cell application cannot only be based on the theoretical efficiencies alone, but it must also be examined from the energy consumption point of view, its level of reversibility, energy loses and thermodynamic efficiencies which can be best evaluated through thermodynamic analysis.

Thermodynamics has been used years back to model energy systems, which includes advanced process plants (Mousafarash and Ameri, 2003). The first law of thermodynamics has been use d generally on a large scale to model system or analyse the energy utilization, but it cannot account for the quality aspect of energy or the source of irreversibilities in the system that is being considered (Rosen and Dincer, 2012). Since energy balance is unable to answer the question of real thermodynamic in efficiencies and the process that cause them, the cost and environment al impact associated with equipment, then the lost work caused by irreversibility can be identified with the aid of an exergetic analysis [11]. For this reason, the thermodynamic concept in which the second law of thermodynamics is considered to be used, without limitation for conducting an effective analysis of all processes, overcomes the limitation and uncertainty that are created due to first law analyses and criteria. However, Exergy calculation is incomplete without economic analysis. The combination of both is termed thermo-economic analysis [14].

Kamaruddin et al. [13] conducted a research on the simulation of catalytic autothermal reforming (ATR) of methane (CH_4) for hydrogen (H_2) production using HYSYS 2004.1 as a simulation tool to investigate the conversion behaviour of the reformer. Air to fuel ratio of 2.5 and water to fuel ratio of 1.5 were conditions used for high CH_4 conversion and high H_2 yield. Results of their findings revealed an H_2 yield of 44% on wet basis and the system efficiency was 87.7%. In the same light, Henry [12] investigated the autothermal reforming of propane process and the optimization of the operational conditions using Aspen HYSYS 2004.1 for PEMFC application. In the research, the exit streams from the auto-thermal reactor was subjected to heat integration process. From the result, 100 kgmole/hr of propane with a specified ratio of air and water 1: 7: 4.3, was able to produce 41.62% of hydrogen with CO concentration lower than 10 ppm, and fuel processor efficiency of 83.14%.

Suleiman et al. [19] conducted thermo-economic analysis of PEMFC, fuelled with methanol and methane using Thermolib simulation tool box. The result of their findings favoured methane system as the most preferred choice for PEMFC based on trade-off between economic, energy and exergy performance of methane and methanol system. Ay et al. [3] conducted a study on exergetic performance analysis for a PEM fuel cell and investigated the effects of operating temperature and pressure on the system efficiency and irreversibilities. The conclusion of the study revealed that the exergetic efficiency of PEM fuel cell is lowered with increase in membrane thickness and current density, and the efficiency increases with the rise of cell operating pressure and decrease of current density for the same membrane thickness. Kazim, [14] also conducted a brief study on exergo-economic analysis of a PEM fuel cell at varying operating pressure, temperature and air stoichiometries. The study extended its investigation to the air stoichiometry ratio ranging from 2 to 4 from the exergy perspective. It was indicated that the exergy cost of the fuel cell can witness greater improvement by adopting any or combination of higher operating pressure, inlet air stoichiometry or cell voltage which demonstrates a goodt improvement in the exergy cost.

Sophie et al. [18] investigated the systematic analysis of fuels derived from biomass for fuel cells application. The study presented an analysis of the recent techniques for biomass conversion and the pathways required for converting biomass feedstock into fuel cell. Fuel processing methods adopted were anaerobic digestion, metabolic processing, fermentation, gasification, and supercritical water gasification, which were compared to natural gas and fossil hydrogen reference cases. Solid

Table 1
Composition of Biomass feed and their properties.

Component	% Composition	Chemical formula	Density (kg/m^3)	Boiling point (°C)
Dextrose	0.55	$C_6H_{12}O_6$	1181	343.9
Sucrose	0.25	$C_{12}H_{22}O_{11}$	1514	461.9
Lysine	0.1998	$C_6H_{14}N_2O_2$	956.3	341.9
H_2S	0.0002	H_2S	788.4	−59.65
Total	1.0	–	–	–

oxide fuel cell was adopted for the bio-derived fuel gases. The selection was based on its wide range of fuel options, high scalability from single kW to multi 100 kW, and good efficiency. The conclusion suggested SOFCs to be a more favourable technology than PEMFCs in areas of wide range of fuel choices but higher thermal efficiency favoured PEMFC.

Attuluri et al. [2] conducted a parametric study of PEMFC for investigation of enhanced performance used in Fuel Cell Vehicle (FCV). The study involved producing the membrane electrode assembly at a catalyst Pt/C loading of 40% and conducting experiment at different parameters, viz, cell temperatures, oxygen and hydrogen flow rates and cathode and anode humidification temperatures. The results revealed that cell temperature has a major effect on the performance of the PEMFC, meanwhile other parameters produce variation in the activation and concentration polarization region alone.

George et al. [9] presented a study on thermodynamic analysis of biogas fed SOFC power plant. Through the consideration of energy and exergy balances for the system, a comprehensive thermodynamic model (THERMAS) was designed and implemented. A case study was selected using a specific SOFC-based system, equipped with three heat exchangers: a reformer, a SOFC-stack system and an afterburner. The simulation tool THERMAS was utilised to conduct a sensitivity analysis on parameters that affect system's efficiency based on exergy. The process of optimization relies on the difference between the efficiencies of energy and exergy by considering an innovative Optimization Factor (OPF) for all system simulation, which is dynamically affected by operational parameters, such as fuel composition, extension of chemical reactions and temperatures. It was found that the use of a pure fuel seems to be meaningless without optimization.

The main contributions of this study are to develop a complete detailed model that fully simulate a PEMFC fuelled with biomethane obtained from Human waste. The proposed model is composed of three individual models namely: biogas plant, hydrogen production with CO_2 clean-up (Fuel processor) and a PEMFC. Each model can be implemented individually or collectively to simulate the power plant, identify potential design issues and obtain a preliminary estimate of the expected system efficiency, and also perform energy, exergy and economic analysis on the system. Significant operating conditions could then be identified, and their effect on the overall system performance or efficiency could be evaluated.

Methodology

The data used for the simulation of biogas feed was obtained from a pilot plant which belongs to Lumus Labs in Lugbe, Abuja, Nigeria. The composition of the biomass is shown in Table 1. The simulation was done with Aspen HYSYS V8.8, a software developed by Aspen Tech®. The selection of this software was based on its flexibility in design, ease in handling, user friendly and its use in optimization of chemical, physical and operational processes.

Simulation of process configuration

In the simulation of the process configuration, four different stages were developed into a sub flowsheet. They include: biogas generation, hydrogen production (fuel processing), CO clean up and the fuel cell.

In the simulation, several building blocks (unit equipment) were used to simulate the process, which consist of the following equipment:

- Ø Biogas Section: Bio-digesters, amine contactor, amine regenerator and flash separator
- Ø Fuel processing and clean-up section: Autothermal reformer, high and low temperature shift reactor (HTS and LTS) and preferential oxidation reactor (PrOx)
- Ø PEM fuel cell section: Anode, cathode and combustor
- Ø Auxiliary units: Pumps, compressor, expander, heat exchanger, heater, cooler and mixers.

The first step in the simulation was to select the appropriate fluid package, the components, conditions and feed streams entering the system can be specified. Extended Non-Random-Two-Liquids (NRTL) and Acid gas equations of state (EOS) were selected as the thermodynamic fluid package on the basis of their compatibility and suitability, and they are worldwide accepted for the required simulation.

Simulation of biogas

The biogas was modelled as a combination of proteins (lysine), carbohydrates (sucrose and dextrose) and sulphide (H_2S). Two digesters were used in series to increase yield, although autothermal digesting was utilized since other methods require more reactors and have a longer reactor resident time. The Predicted rates of reaction are on the basis of small-scale

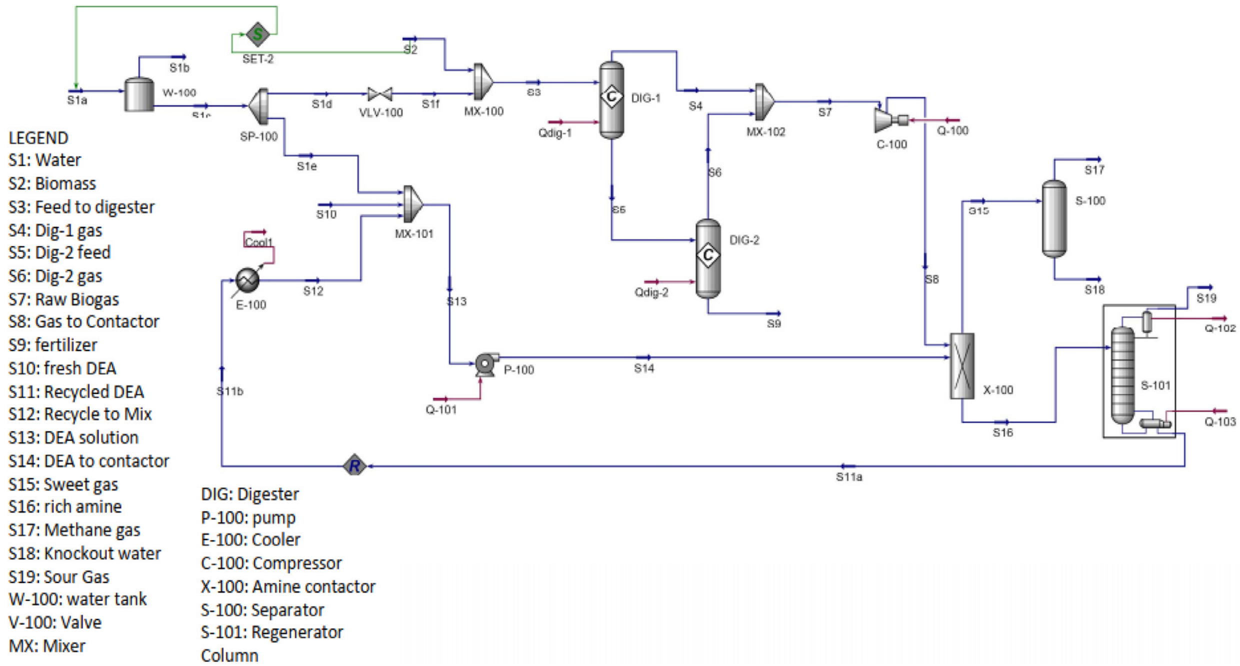

Fig. 1. Developed Aspen Hysys model for biogas production.

reactions that have been carried out and thus the bacteria employed in the reaction were not included in this model even though the temperature before they begin to die out was considered. The CO_2 removal section was modelled using an absorber regenerator combination which is the conventional method. The plant capacity has been made adjustable, so it can accommodate several inputs from different sources. The reactions involved in the bio-digesters for biomass conversion include:

$$C_6H_{12}O_6\,(\text{Dextrose}) \rightarrow 3CH_4 + 3CO_2 \tag{1}$$

$$C_{12}H_{22}O_{11}\,(\text{Sucrose}) + 4H_2O \rightarrow 2H_2 + 3CH_4 + 3CO_2 + 2NH_3 \tag{2}$$

$$C_6H_{14}N_2O_2\,(\text{Lysine}) + H_2O \rightarrow 6CO_2 + 6CH_4 \tag{3}$$

Water and Biomass at 30 °C and 40 °C respectively, were channeled into the digester operating at 60 °C and biogas was generated as depicted in Fig. 1, the effluent from the first digester serve as the feed for the second digester operating at 70 °C for more yield. The two streams of biogas were then directed through a tower containing an amine solution where all the impurities and acid gases were expelled. Amine absorbs sulfur compounds from biogas generated and it was recycled repeatedly. Di-Ethanol Amine (DEA) was used as treating solution because of its outstanding efficiency [16]

Simulation of hydrogen production (fuel processor)

In a way to generate a rich stream of hydrogen from biogas, the fuel is first preheated and then converted in a reforming unit that involves autothermal reforming by feeding the pre-heated fuel, steam, and air at the inlet. The autothermal is operated in such a way that the biogas and air are first fed into the reformer for combustion so that the catalyst of the reformer can be heated up. When the catalyst temperature reaches about 300 °C [5], a temperature at which the autothermal reaction can be self-activated (known as light-off), the predetermined mixture of biogas, air, and water is fed to the reformer. The purpose was to convert as much biogas into hydrogen gas at acceptable yields in an efficient manner while decreasing CO formation. A considerable range of air to water to fuel (A/W/F) ratio (1:2.5:4) was selected to observe its effect on hydrogen yield and CO formation as reported by Ersoz et al. (2003) and the conversion percentage reached was 80%. In brief, this model takes the following reactions into account (Eq. (4)).

$$4CH_4 + O_2 + 2H_2O = 4CO + 10H_2\,\Delta H_{298} = 8.5e4\,\text{kJ/kgmole}\,\text{Conversion}(\%) = 80 \tag{4}$$

Since the stoichiometry of the reaction and the conversion of the base component are known, the reformer was set up as a conversion reactor as shown in Fig. 2. By using conversion reactor, HYSYS was used to calculate the composition at the exit stream.

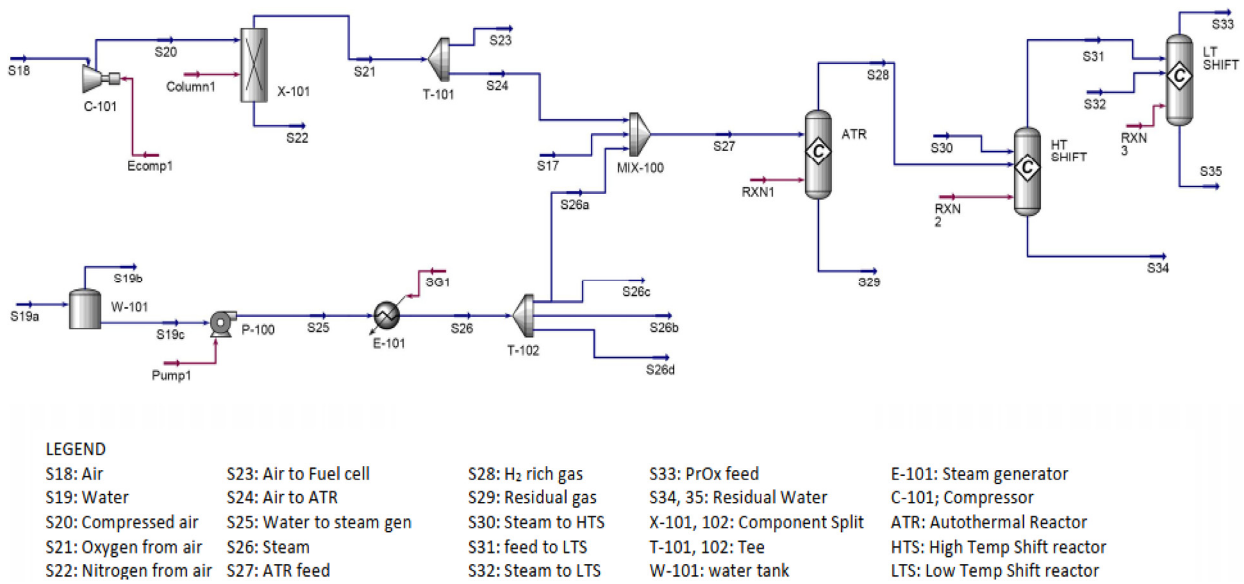

Fig. 2. Developed Aspen Hysys model for hydrogen production.

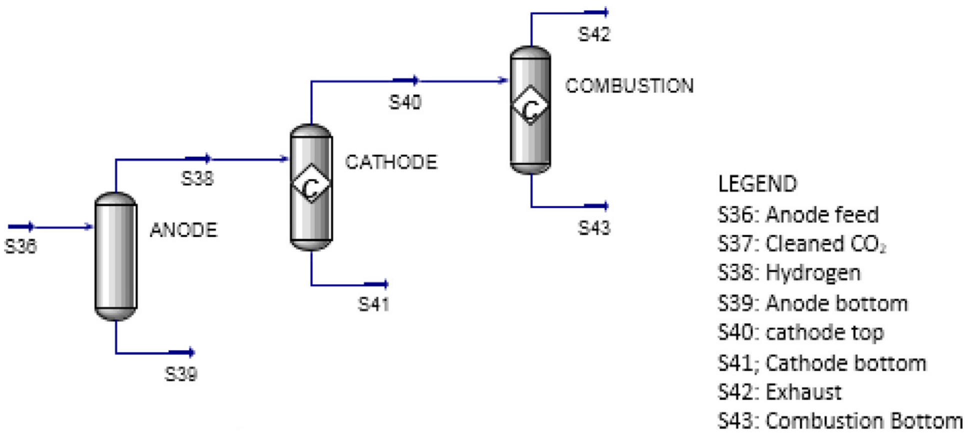

Fig. 3. Developed Aspen Hysys model for fuel cell.

Simulation of fuel cell

The fuel cell was simulated using an Hysys based model as shown in Fig. 3. Separator was used for the fuel cell anode and conversion reactor was used for the fuel cell cathode where hydrogen and oxygen react to form water. Gibbs reactor was used for the combustor where the heat is being recovered.

Exergy calculation

The following equations were used for the calculation of exergy:

For a system, the physical exergy is given by the relationship presented in Eq. (5) [20]:

$$b_{phy} = c_p\left[(T - T_o) - T_o ln\left(\frac{T}{T_o}\right)\right] + RT_o ln\left(\frac{P}{P_o}\right) \tag{5}$$

The basic exergy expression for a steady flow stream is presented in Eq. (6) [20]:

$$\varepsilon(T, p, x) = [H(T, p) - H^\circ(T^\circ, p^\circ)] - T^\circ[S(T, p) - S^\circ(T^\circ, p^\circ)] \tag{6}$$

The environmental parameters are assumed as:

p°=101.325 kPa and T° =298.15 K. The exergy of a stream (specified temperature T, pressure p, and composition x) [20]:

For a mixture, the chemical exergy is represented by Eq. (7);

$$b_{ch} = \sum_{i=1}^{n} x_i b_{ch,\ i} \tag{7}$$

While the exergy efficiency (ψ) calculation is represented by Eq. (8):

$$\psi = \frac{Exergy\ in\ product}{Total\ exergy\ input} \tag{8}$$

Thermodynamic efficiency calculation

The thermal (energy) efficiencies of the biogas plant, ATR, HTS, LTS, and PrOx reactors are represented as η_{Biogas}, η_{ATR}, η_{HTS}, η_{LTS}, and η_{PrOx}, respectively. They are expressed as the ratio of the heating values of the outlet streams to the heating value of inlet streams. The heating value of the streams were calculated by the multiplication of their lower heating value (LHV) with their corresponding molar flow rate [15].

Formulae of the efficiency calculations for various components and subsystems are presented in Eqs. (9)–(14).

$$\eta_{Biogas} = \frac{m_{S17} \times LHV_{S17}}{m_{S2} \times LHV_{S2}} \tag{9}$$

$$\eta_{ATR} = \frac{m_{S28} \times LHV_{S28}}{m_{S27} \times LHV_{S27}} \tag{10}$$

$$\eta_{HTS} = \frac{m_{S31} \times LHV_{S31}}{m_{S28} \times LHV_{S28}} \tag{11}$$

$$\eta_{LTS} = \frac{m_{S33} \times LHV_{S33}}{m_{S31} \times LHV_{S31}} \tag{12}$$

$$\eta_{PrOx} = \frac{m_{S36} \times LHV_{S36}}{m_{S33} \times LHV_{S33}} \tag{13}$$

$$\eta_{FC} = \frac{m_{S42} \times LHV_{S42}}{m_{S36} \times LHV_{S36}} \tag{14}$$

Also, the cumulative thermal efficiencies are defined as the fraction of the reactor exit stream heating value to the heating value of the total fuel feed to the system. η_6 represents the fuel cell electrical efficiency.

Efficiency of the fuel processor is defined by the relationship presented in Eq. (15) [13]

$$\text{Efficiency }\% = 100\ \times \frac{LHV\ of\ _{H2}\ produced}{LHV\ of\ _{Fuel\ \ used}} \tag{15}$$

The lower heating value (LHV) of the product hydrogen is depicted by Eq. (16)

$$\text{LHV of } H_2 \text{ produced} = H_2 \text{ yield} \times \text{heat of combustion of } H_2 \tag{16}$$

Result and discussion

The temperature profile of the fuel processor is presented in Fig. 4. It shows the temperature range of unit equipment from ATR to PrOx reactors. The focus here is to maximize hydrogen production and reduce the concentration of CO to the lowest through parameter variation. Hence, it is paramount to monitor hydrogen yield and CO concentrations at every stage. Since immediately after the ATR, there is a CO clean-up system where the rich hydrogen syngas pass through a series of reactors to perform the water gas shift reaction (WGS) in which CO is converted into CO_2 and hydrogen with the existence of steam. Effectiveness of this operation reduced the concentration of CO to almost zero at the exit stream of low temperature shift reactor (LTS).

When the air to water to fuel A/W/F ratio was set at 1:2.5:4, the outlet temperature of ATR reactor obtained was 371.1 °C which is higher than the temperature at which auto-thermal reaction can be self-activated (that is 350 °C), which implies that the reaction can proceed as designed. The effluent of the ATR was then cooled to 204.4 °C, 93.33 °C, 37.78 °C and 58.60 °C by passing it through high temperature shift (HTS) reactor, LTS reactor and PrOx reactor respectively which was lower than the outlet temperature of ATR to prevent reversible reaction. There is a slight increase from the inlet temperature (37.78 °C) of the PrOx reactor to the outlet (58.6 °C), this is attributed to the exothermic nature of the PrOx reaction [13].

Under the temperature profile conditions, the product conversion in the fuel processor is presented in Table 2. The feed stream into the ATR which contains 52.23% of methane, 13.67% of O_2, 33.69% of H_2O and 0.41% of H_2 was converted to produce 60.58% of hydrogen, 24.14% of CO, 7.39% of H_2O and 1.86% of O_2. Higher production of H_2 in this study when compared with 40% of H_2 reported by Kamaruddin [13] can be attributed to the ratio of Air to Fuel that increase the yield of H_2 in the auto-thermal reactor.

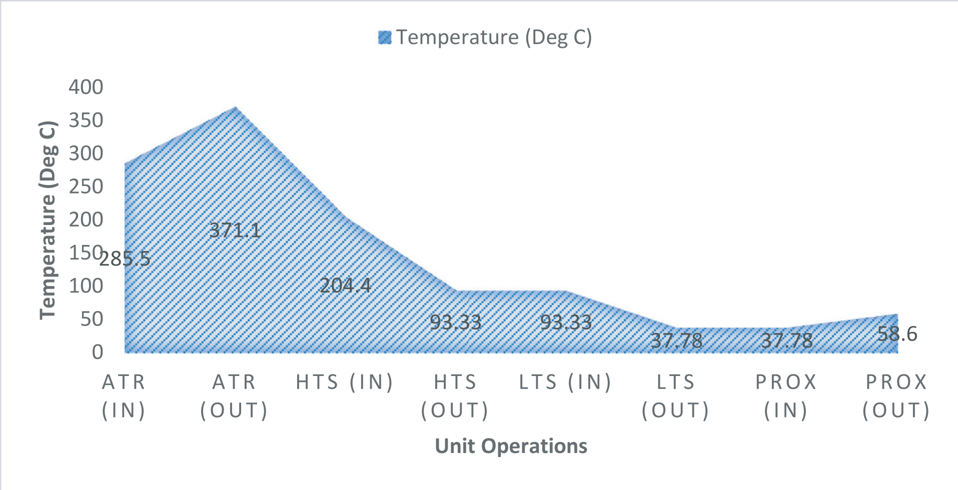

Fig. 4. Temperature profile of unit operation.

Table 2
Bio-methane conversion at the exit of all reactors.

Unit equipment	Inlet streams		Outlet streams	
	Composition	%	Composition	%
ATR	CH_4	52.23	H_2	60.58
	O_2	13.67	CO	24.14
	H_2O	33.69	H_2O	7.39
	H_2	0.14	O_2	1.86
			CH_4	6.03
HTS	H_2	60.58	H_2	69.47
	CO	24.14	CH_4	5.25
	H_2O	7.39	CO_2	16.79
	O_2	1.86	H_2O	2.68
	CH_4	6.03	CO	4.20
			O_2	1.62
LTS	H_2	69.47	H_2	70.45
	CH_4	5.25	CH_4	5.08
	CO_2	16.79	CO_2	19.49
	H_2O	2.68	H_2O	2.60
	CO	4.20	CO	0.81
	O_2	1.62	O_2	1.56

Table 3
Comparison of exit composition of all reactors with previous study.

	Unit equipment	Components%						
		CH_4	O_2	H_2O	CO_2	CO	H_2	N_2
Current work (biomethane)	ATR	6.03	1.86	7.39	–	24.14	60.58	–
	HTS	5.25	1.62	2.68	16.79	4.20	69.47	–
	LTS	5.08	1.56	2.6	19.79	0.81	70.45	–
	PrOx	0	0	0	0.01	0	99.99	–
Previous work (propane)	ATR	0	0	13.93	10.00	5.44	40.12	30.50
	HTS	0	0	12.25	11.69	3.75	41.81	30.50
	LTS	0	0	8.54	15.39	0.5	45.51	30.50
	PrOx	0	0	9.04	15.25	0	44.37	31.34

Also, WGS reactor convert CO into CO_2 and more hydrogen in the presence of steam. Therefore, the CO percentage decreased from 24.14% (inlet) to 0.81% (outlet), while the CO_2 percentage increased from 16.79% to 19.4% and Hydrogen from 60.58% to 70.45%. Air was injected at the PrOx reactor to oxidize the remnant CO to CO_2.

Table 3 presents the composition of components at the exit of all reactors simulated this study as compared with previous study [12]. The present study indicated non-presence of N_2 in the streams because it has been separated from air before channelling it to ATR thereby making H_2 the only input into fuel cell based on PEMFC requirement.

Table 4
Analysis of energy flow in Biogas equipment.

Equipment	Material streams energy (kJ/h)		Energy streams (KW)		Energy (KW)	
	In	Out	In	Out	Required	Produced
W-100	−3,871,744.76	−3,871,744.76	0	0	0	0
SP-100	−3,871,744.76	−3,871,744.76	0	0	0	0
V-100	−3,097,395.81	−3,097,395.81	0	0	0	0
MX-100	−36,105,598.07	−36,105,598.07	0	0	0	0
MX-101	−43,728,603,183	−43,728,603,183	0	0	0	0
E-100	−42,427,945,394	−43,428,833,508	–	278,024.5	–	278,024.5
P-100	−43,728,603,183	−43,725,763,517	788.8	–	788.8	–
DIG-1	−36,105,598.07	−66,385,937.89	−8411.21	–	−8411.21	–
DIG-2	−10,554,997.84	−14,685,530.57	−1147.37	–	−1147.37	–
MX-102	−60,216,824.79	−60,216,824.79	0	0	0	0
C-100	−60,216,824.79	−58,848,494.17	380.09	–	380.09	–
X-100	−43,784,612,011	−43,784,612,011	0	0	0	0
S-100	−8,400,601.8	−8,400,601.8	0	0	0	0
S-101	−43,776,211,409	−42,757,831,205	339,141.87	56,339.83	282,802.04	–

Comparison with the previous study as presented in Table 3 showed a higher hydrogen conversion at the exit of LTS. The improvement can be linked to easy reforming conditions of methane to rich hydrogen as compared to propane and optimization of operating parameters conducted.

Fuel processor system efficiency

With the air to water to fuel A/W/F ratio (1:2.5:4), the calculated fuel processor system efficiency obtained was 83.5% which has a direct effect on the net electrical efficiency of the system. This value is slightly lower than 85% and 87% respectively reported by Olgun et al. (2004) and Kamaruddin et al., [13]. The variation in the results may be attributed to the different processing approach employed to determine the efficiency, steam reforming process was used in the reported works different from auto-thermal reforming utilized in this study. However, Ersoz et al., [7] presented a lower fuel processor efficiency of 73.5% using auto-thermal reforming, the low value of the efficiency may be connected to the choice of parameters and water to fuel ratio.

Energy analysis and thermodynamic efficiencies

The result of energy analysis for the selected configurations is presented in Table 3.3. The reference state for temperature and pressure was taken to be 298.15 K and 1atm respectively as reported by Kotas (1995). The thermodynamic fluid packages employed were Acid gas PP and Extended NRTL-Ideal. Data obtained from simulation include thermodynamic parameters such as temperature, pressure, specific enthalpy and entropy, stream flow rate, composition and those related to economic analysis such as equipment size, and utility requirement. The thermodynamic parameters were used to carry out energy and exergy analysis to determine thermodynamic performance in terms of energetic and exergetic efficiency and lost work.

Energy analysis of biogas process

Results contained in Table 4 revealed that the energy content at the outlet streams have higher values when compared to the energy in the inlet streams, this happens in equipment that are energy producing like cooler (E-100) and both digesters. Having higher energy content in the inlet streams than the outlet streams indicates that energy was transferred to the system which must have been supplied either through material streams or energy streams (thermal or heat) known as utility streams, electrical or work grade energy as depicted in pump, compressor and columns reboiler which are energy requiring equipment (Cengel and Boles, 2010).

Equipment with zero energy streams indicated that they neither require nor produce energy like mixers and water tanks. Therefore, it is enough to state that the process is energy requiring since majority of the equipment requires energy and few produces energy and the magnitude of energy required is much when net balance is conducted.

The results of energy conservation for unit equipment presented in Table 4 revealed that the regenerator (S-101) is the unit that is observed to have the highest energy consumption, because it requires about 96% of the total energy requirement from an external supply source. This is due to the fact that the reboiler at the bottom does the highest work to regenerate the used DEA and recycle it back to the system. On the other hands, the Cooler (E-100) produces the highest energy of 278,024.5 KW, which signify availability of energy in excess of what is required around X-100 (to perform component separation) and it could also be utilized somewhere else in the process to save cost and reduce energy pollution. This energy integration ultimately improved energy efficiency and economic performance.

Table 5
Analysis of energy flow in fuel processor and fuel cell equipment.

Equipment	Material streams energy (kJ/h)		Energy streams (KW)		Energy (KW)	
	In	Out	In	Out	Required	Produced
C-101	−52,061.95	1,429,406.27	411.52	–	411.52	–
X-101	1,429,406.27	1,429,406.27	5.68E-14	–	5.68E-14	–
T-101	303,774.89	303,774.89	0	0	0	0
MIX-100	−27,289,337.01	−27,289,337.01	0	0	0	0
W-101	−84,927,887.1	−84,927,887.1	0	0	0	0
P-100	−84,927,887.1	−84,923,572.18	1.2	–	1.2	–
E-101	−84,923,572.18	−68,336,864.52	4607.42	–	4607.42	–
T-102	−68,336,864.52	−68,336,864.52	0	0	0	0
ATR	−27,289,337.01	−16,458,743.16	3008.5	–	3008.5	–
HT Shift	−31,570,763.97	−38,904,390.18	−2037.12	–	−2037.12	–
LT Shift	−52,673,153.35	−53,707,587.7	287.33	–	287.33	–
X-102	−44,432,441.09	−89,207,666.49	0	0	0	0
Anode	342,784.3	342,784.3	0	0	0	0
Cathode	342,784.3	464,294.3	0	0	0	0
Combustor	396,144.6	464,294.3	0	0	0	0

Energy analysis of fuel processor and fuel cell

The values presented in Table 5 shows the available energy across each equipment unit and it revealed that all the equipment are energy requiring and none produces energy. All equipment except the shift reactors, PrOx, cathode and combustor have higher energy value in their inlet streams. Higher energy content in the inlet streams than the exit streams is an indication that transfer of net energy was to the system and must have been supplied via external utility streams [5]. Critical evaluation shows that the steam generator (E-101) requires the highest form of external energy taking about 73% of total energy requirement in fuel processor plant and the pump requires the least energy supply of 1.1986 kW as a result of its electrical energy demand.

It is important to mention that the total energy required by the biogas plant is 274,412.35 kW and total energy produced is 278,024.5 kW and total energy required by the fuel processor and fuel cell is 6279.145 kW, taking a net balance revealed that the energy produced which is close to the energy required can actually be used to balance the energy requirement to save cost and reduce energy wastage which ultimately produce a conservative system and the purpose of sustainability was achieved.

It is worthy of note that energy balance of the process is efficient based on energy requirement and energy produced, which testify to the principle of thermodynamics first law in the selected process route. It is also important to acknowledge that satisfaction of energy conservation principle does not require the status of energy quality and its direction of flow. Even though, conservation principle of energy is fulfilled, the possibility of the process to proceed as expected depends on whether the process is in the direction of increasing entropy [17]. Careful observation of the result shows that the process satisfied the condition of the first law of thermodynamic which is energy conservation principle, indicating good energy balance across the respective units. Despite revealing energy performance, yet this analysis is limited in knowing whether the process can take place or otherwise (Cengel and Boles, 2010; [17]). Therefore, a further assessment is considered, which is exergy analysis.

Exergy analysis

The results in Table 6 presents the exergy flow into the equipment and out, in the biogas process configurations. The result revealed that, in equipment such as water tanks (W-100), component splitter (SP-100) and separator (S-100), total exergy change was zero which is due to the fixed components composition passing through the units. This explains further that there was no chemical reaction or mixing effect high enough to cause composition changes (Suleiman et al., 2014). Notably, decrease in exergy along the process is an indication that the processes take place as they proceeded in the direction of entropy generation (irreversibility).

The exergy changes around some equipment like pump (P-100) and MX-102 in the process is an indication that there were composition changes that resulted into exergy lost, consumption and conversion to other form(s) like heat (thermal) energy [4]. The change in exergy is highest in Cooler (E-100) and next is the regenerator (S-101) followed by the amine contactor (X-100) which also confirm the area of highest exergy destruction in a similar work presented by Suleiman et al. [19]. This shows that the change that took place in the composition favoured the proportion of components that have high standard chemical exergy and high flow rate associated with the outlet streams. The positive values of net exergy suggest that the entire system is behaving as an exergy sink which is lost or reused for the process to proceed.

The results shown in Table 7 revealed the exergy flow in and out of the equipment in fuel processor and fuel cell configurations. The results indicated that the change in exergy content of equipment such as Tee (T-101) and Water tank (W-101) was Zero, this is as a result of the fixed components composition passing through the units. This further indicates that no chemical reaction occurred or mixing effect substantial enough to cause composition changes. The change in the

Table 6
Analysis of exergy flow in biogas equipment.

Equipment	Exergy in (kJ/h)	Exergy out (kJ/h)	Exergy loss (kJ/h)
W-100	59.62	59.62	0
SP-100	59.62	59.62	0
V-100	47.7	47.66	−0.04
MX-100	−213,577.14	−207,056.05	6521.09
MX-101	27,792,515.98	27,984,865.44	192,349.46
DIG-1	−207,056.05	−112,248.29	94,807.77
DIG-2	−127,774.54	−111,137.12	16,637.42
MX-102	17,600.19	17,507.7	−2.49
C-100	17,507.7	1,157,591.35	1,140,083.65
X-100	−126,531,699.3	−115,480,068.8	11,051,630.53
P-100	27,984,865.44	−127,689,290.6	−55,674,156.1
E-100	−337,987,604.8	27,900,128.94	365,887,733.8
S-100	790,333.12	790,333.12	0
S-101	−116,270,401.9	−336,619,158.9	−220,348,757

Table 7
Analysis of exergy flow in fuel processor and fuel cell equipment.

Equipment	Exergy in (kJ/h)	Exergy out (kJ/h)	Exergy loss (kJ/h)
C-101	456,977.23	1,695,001.03	1,238,023.80
X-101	1,695,001.03	1,834,887.55	139,886.52
T-101	386,341.85	386,341.85	0
MIX-100	2,468,809.04	1,704,450.76	−764,358.28
W-101	1849.3	1849.3	0
P-100	1849.3	5131.14	3281.84
E-101	5131.14	5,183,836.55	5,178,705.41
T-102	5,183,836.55	5,183,836.55	−1.49012E-08
ATR	1,704,450.76	1,623,404.67	−81,046.1
HT Shift	3,778,902.68	1,623,404.67	−2,155,498.01
LT Shift	1,649,387.69	1,152,605.24	−496,782.45
X-102	1,151,838.82	1,960,933.89	809,095.061
Anode	1,404,906.5	1,404,906.5	0
Cathode	1,404,906.5	6,941,238.62	5,536,332.13
Combustion	6,912,002.01	6,918,861.70	6859.69

exergy of all equipment irrespective of its magnitude, indicates that no process exists without exergy being consumed, which further collaborate reasons why 100% exergy efficient cannot be achieved in a system.

The exergy lost in the equipment, result from changes in composition that have been transformed to another form of energy particularly, the thermal energy (heat). Significantly, decrease in exergy along the process is a reflection that the processes took place in the direction of irreversibility as they proceed. The exergy change around the equipment points to composition changes that results into exergy lost, consumption or conversion to other form(s) of energy mainly heat energy [5]. Further analysis shows that the change in exergy is apparently high in Cathode and compressor (C-101). This is as a result of greater changes in composition in these units and the reaction that occurred between hydrogen and oxygen to form water at the cathodes which liberate a huge energy.

The change in exergy flow presented shows that majority of the overall net change is positive. However, the observable exergy change with negative values are found around MIX-100, ATR, HT shift and LT shift reactors which indicates that the change in composition that took place is in favour of the proportion of components that have high standard chemical exergy and outlet streams associated high flow rate. The positive values of net exergy suggest that the entire system is behaving as an exergy sink in which is lost or reused for the process to proceed. Moreover, the outcome around various units revealed that the resultant effect was as a result of the generation of entropy along the process which led to exergy losses in substantial quantities at various stages in the process which also points out the fact that the process ability to generate required work is fading downstream in the direction of greater entropy generation. The performance of these equipment can be improved upon by ensuring minimal composition changes, proper choice of heat transfer media in exchangers and use of most appropriate and optimal operating parameters [10].

Thermodynamic efficiencies

The thermal efficiencies of major components in the systems along with auxiliary units' efficiencies determine the net electrical efficiency of the PEM fuel cell system. Also, to increase the overall efficiency of the system, combined heat power (CHP) approach system was employed. The thermal (energy) efficiencies of the Biogas Plant, ATR, HTS, LTS, and PrOx reactors are represented as η_{Biogas}, η_{ATR}, η_{HTS}, η_{LTS}, and η_{PrOx}, respectively. They are expressed as the ratio of the heating values of the outlet streams to the inlet streams. The heating value of the streams were calculated by multiplying the lower heating

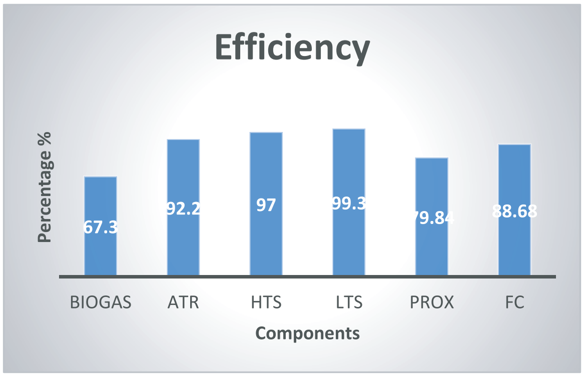

Fig. 5. Efficiency of major components and subsystems.

Table 8
Lost work and exergetic efficiencies of major equipment.

Equipment	Exergy in (kJ/h)	Exergy out (kJ/h)	Lost work	Exergetic efficiency
MX-100	−213,577.14	−207,056.05	6521.09	96.95
DIG-1	−207,056.05	−112,248.29	94,807.77	54.21
DIG-2	−127,774.54	−111,137.12	16,637.42	86.98
C-100	17,507.69	1,157,591.35	1,140,083.65	83.53
X-100	−126,531,699.3	−115,480,068.8	11,051,630.53	91.27
S-101	−116,270,401.9	−336,619,158.9	−220,348,757	34.55
C-101	456,977.23	1,695,001.03	1,238,023.80	87.45
MIX-100	2,468,809.04	1,704,450.76	−764,358.28	69.04
P-100	1849.3	5131.14	3281.84	83.24
E-101	5131.14	5,183,836.55	5,178,705.41	31.25
ATR	1,704,450.76	1,623,404.67	−81,046.1	95.25
HT Shift	3,778,902.68	1,623,404.67	−2,155,498.01	42.96
LT Shift	1,649,387.69	1,152,605.24	−496,782.45	69.88

value (LHV) with their corresponding molar flow rate. The net electrical efficiency of the overall PEM fuel cell system covers all the efficiencies namely: Biogas Plant η_{Biogas}, Fuel Processor η_5 and the Fuel Cell η_6.

The result shown in Fig. 5 indicates that the LTS reactor has the highest energetic efficiency performance because its steam requirement is minimal, followed by HTS reactor and then the autothermal reformer.

The high efficiency in the LTS reactor may be ascribed to high-grade thermal energy utilized, whose thermal properties encourages low irreversibility rate and adequate use of its available useful energy [5]. The good efficiency obtained around the fuel processor may be linked to the energy quality and low pressure drops along the process. The efficiency of PrOx is a bit lower because the LHV of the gas stream decreases sharply after the PrOx exit prior to the entrance of the PEM fuel cell due to the extraction of other gases like CH_4, CO_2 and remnant condensed water from the hydrocarbon fuel gases leaving only the hydrogen gas according to the fuel cell requirements [15]. The biogas plant is observed to have the least efficiency. The calculation of cumulative efficiency represents the results of component efficiencies and the final net electrical efficiency for the three sub-systems. The product of the calculated component efficiencies gives the final net electrical efficiency (η_6) of 42.32%.

With the combined-Heat-Power (CHP) approach, the efficiency of the Fuel Cell increased to 89.2%. A net power production of 1.45 MW is realized from the Fuel Cell by multiplying the LHV of hydrogen at the anode with the electrochemical efficiency (hydrogen utilization efficiency) of 0.6.

Lost work and exergetic efficiencies of major equipment

The result of irreversibility (lost work) and exergetic efficiency obtained for the entire process configurations is presented in Table 8. The units associated with higher irreversibilities are the Steam generator (E-101), regenerator (S-101), Digester-1 and the Shift Reactors as presented in Table 8. Maximum exergy destruction is detected in steam generator, followed by the regenerator. The high loss of exergy in the regenerator is as a result of exergy consumed for the regeneration process and lost due to the inherent irreversibility nature of the regenerator which is applicable to any process occurring in similar manner, as reflected in the steam generator. [19]. The exergy losses in those units is linked primarily to irreversibilities caused by mixing, pressure drop, fluid friction, use of thermal energy and chemisorptions process in the case of contactor.

The equipment that is observed with the least efficiency is the regenerator, this is due to the two heat exchangers attached (condenser and reboiler) as its major source of irreversibility in addition to pressure drop and effect of fluid resistance [17]. It is important to note that the assessment is not limited to identification of the least efficient process route

Table 9
Comparative efficiency (present and previous works).

Parameters	Current work	Previous works			
	(bio-methane)	A (methane)	B (propane)	C (gasoline)	D (gasoline)
H_2 yield%	60.58	44	41.62	–	–
FP efficiency	83.5	87.7	83.14	73.5	85.5
Electrical Eff	42.32	–	–	36.8	41

A – Kamaruddin et al., [13].
B – Henry Insiong. (2006).
C – Ersoz et al., [7].
D – Olgun et al., (2004).

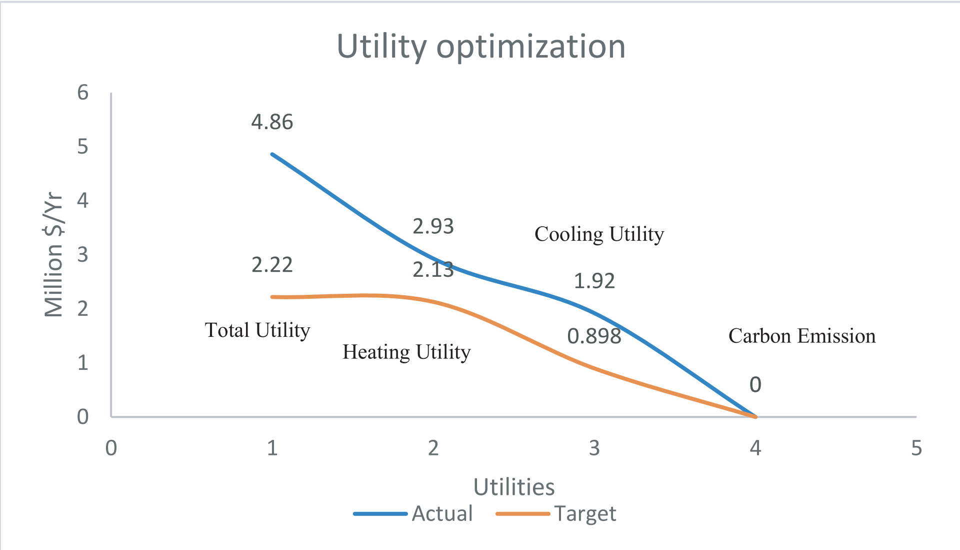

Fig. 6. Utility requirement.

alone, but also in the identification of equipment that require major concern to remedy inefficiency and improve performance of the system.

In addition to the irreversibility, the equipment exergetic efficiency results for the process are presented in Table 8. Critical observation of the result revealed that the compressor, contactor, ATR, pump and mixers have high exergetic efficiency. High exergetic efficiency in equipment like pump and compressor is as a result of high-grade energy (electricity) that is used to power them.

Critical analysis based on the performance of each unit reveals that process equipment are linked with irreversibility and the work potential is dropping at varying rate downstream. This reduction in work potential downstream of the systems implies that the recognized principle which state that energy has quality and proceed in a particular direction (the second law of thermodynamic) is fulfilled. The satisfaction of this fundamental requirement and that of the energy conservation principle is a proof that the process configuration is sufficient to proceed at the stipulated optimal simulation conditions to achieve the set objectives.

Table 9 presents the comparative efficiency of this study with previous works. The previous works were selected for comparison based on the fact that the same software (HYSYS) was adopted. Comparison of this study with reported works showed an improvement in hydrogen yield, a good fuel processing efficiency and a higher electrical efficiency was also achieved

An important reason for this improvement can be associated with sensitivity analysis conducted on simulation to identify the optimum operating condition and the equation of state employed to model the thermodynamics and predicts phase behaviour of components with wide range of volatility.

Economic analysis

Economic analysis of the successfully converged fuel cell process was carried out, which consists of the energy saving cost, equipment costs, utility cost and the economic evaluation.

The energy saving cost evaluation result carried out for the selected configurations is presented in Fig. 6. The figure shows the actual utilities and the target. The target was achieved after the optimization of the process. It revealed the reduction in

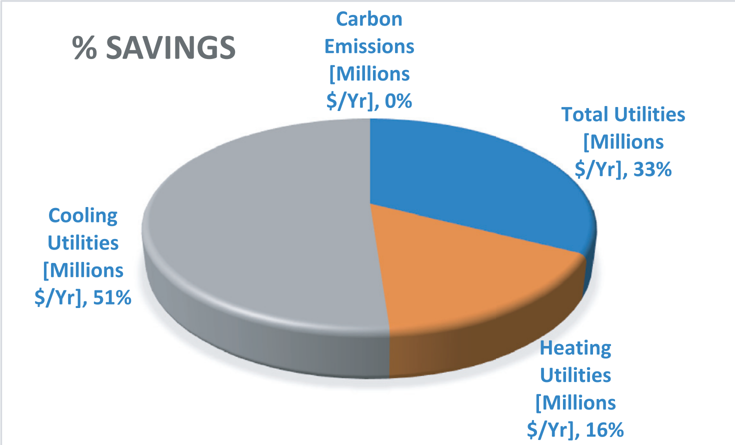

Fig. 7. Cost percentage savings.

Table 10
Economic summary.

Name	Summary
Total Capital Cost [USD]	381,235
Total Operating Cost [USD/Year]	1,131,240
Total Raw Materials Cost [USD/Year]	0
Total Product Sales [USD/Year]	0
Total Utilities Cost [USD/Year]	1,020,680
Rate of Return [Percent/'Year]	20
Equipment Cost [USD]	100,060.78
Total Installed Cost [USD]	191,877

total utility consumption and the cost saved from $4.864 million/yr to $2.22million/yr. This shows the importance of energy and exergy analysis on process configuration. The total utilities combine both the heating and cooling utilities. The Figure also shows the actual heating utilities requirement and the target, which include LP steam, HP steam, MP steam and Hot oil that is used to increase the temperature of process streams. It is interesting to note that a total of $ 0.8 million/year of heating media is saved constituting of about 16% of the total savings as shown in Fig. 7. On the other hand, a huge amount of savings is also realized from cooling utilities as shown in Fig. 7. The cooling utilities include water and air. A total savings of $1.022 million/yr is realized constituting about 51% of the total savings.

The available savings of total utility obtained per year is $2.6 million constituting 33% of actual savings and Zero carbon emission. The summary of the cost analysis evaluated around equipment is presented in Table 10. The estimation of the cost is based on the cost index of 2016 provided by Aspen Econs. The table shows that raw material cost is zero because it is assumed that the plant is situated in the area of abundant human waste. The electricity generated from the fuel cell plant is estimated to be approximately 1.45 MW using a CHP approach.

A rate of return of 20% is realized which means, in 5 years, the total investment can be recovered and profit will start accruing after the period of rate of return (ROR). The payback period of 5 years is an indication that the investment in these processes is safe and profitable; and the profit is realized after 5 years. The general overview of the processes based on economic and thermodynamics performance shows that the investment is worthwhile and indeed waste can be turned to wealth.

Conclusions

The process route required to generate 1.45 MW from human waste using PEMFC has been identified in this study. The three stages involved were; biogas production, hydrogen production and fuel cell application. A successfully converged simulation was attained at hydrogen production rate of 358 kgmole/h, 330 K, 4.8 bar and 99% purity from the PrOx exit stream. Optimization of the process revealed a total utility saving of 26.37 Million/yr and 83.5% fuel processor efficiency. Energy balance revealed that the total energy required is 274,412.35 kW and total energy produced is 284,303.645 kW, which can actually be used to balance the energy requirement to save cost and reduce energy wastage, a net electrical efficiency of 42.32% was realized. Assessment of exergy indicates that the units of high irreversibilities are the steam generator, regenerator and Digester-1 and shift reactors. Maximum exergy destruction was detected in the regenerator which also remain the least efficient equipment. Economic analysis provides a rate of return of 20% and profit is realized after 5 years, which is an indication that the investment is safe and profitable. The general overview of the processes based on economic and thermodynamics performance shows that the investment is worthwhile and indeed waste can be turned to wealth.

Declaration of Competing Interest

The authors have declared that there was no conflicts of interest among authors.

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