Sustainable Energy Conversion in Multigeneration Power Plant Configuration for Climate Change Mitigation

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Kurzfassung

Die wachsenden Bedenken hinsichtlich der Energiesicherheit, der ökologischen Nachhaltigkeit und der negativen Auswirkungen der Verbrennung fossiler Brennstoffe haben den Bedarf an innovativen und nachhaltigen Energielösungen verstärkt. In diesem Zusammenhang wird in dieser Studie ein biomassebetriebenes integriertes System vorgeschlagen. Das System umfasste eine anaerobe Vergärung, Vergasung, einen Protonenaustauschmembran-Elektrolyseur (PEME), einen Sabatier-Reaktor, eine Festoxidbrennstoffzelle (SOFC), eine Gasturbine, eine Dampfturbine und einen Organischen Rankine-Kreislauf (ORC). Biomasse-Rohstoffe in Form von tierischen Abfällen dienen als Input im Fermenter und werden in Biogas und Ernterückstände im Vergaser in Aufbereitetes Synthesegas umgewandelt. Synthesegas und Methan aus dem Biokonvertierungsprozess werden einem SOFC-GT-Topping-Zyklus mit Wärmerückgewinnungs-Bottom-Zyklen von Dampfturbinen und organischen Rankine-Zyklen zugeführt. Um den Beitrag des ORC zur Anlagenleistung zu verbessern, wird eine Auswahl des besten ORC-Arbeitsmediums aus sechs vorausgewählten Kandidaten (MM, MDM, Cyclopentan, Cyclohexan, R1233zd(E) und durchgeführt. Das vorgeschlagene System wird unter thermoökonomischen, R600a) exergoökonomischen und erweiterten exergoökonomischen Gesichtspunkten in der Software Engineering Equation Solver (EES) bewertet. Es wird Wasserstoff von 0,0023 kg/s mit einem PEME-Wirkungsgrad von 73,73 % gewonnen und zur Aufwertung des Synthesegases von niedrigem Heizwert von 3,85 auf 33,48 MJ/kg weiterverwendet. Es wird eine parametrische Analyse durchgeführt, um die Auswirkung der Konstruktionsparameter auf die Leistung der Anlage sicherzustellen, und es wird gezeigt, dass ein niedriges Druckverhältnis, eine hohe Stromdichte und eine hohe Turbineneinlasstemperatur für eine hohe Energie- und Exergieeffizienz, und niedrige Gestehungs und Produktkosten günstig sind. Eine auf der erweiterten exergoökonomischen Analyse basierende Analyse des Verbesserungspotenzials deutete auf eine Möglichkeit zur Rückgewinnung von 32 % der gesamten in der Anlage zerstörten Exergie hin. Abschließend wird eine multikriterielle Optimierung mithilfe eines multiobjektiven genetischen Algorithmus (MOGA) in MATLAB durchgeführt, um die Exergieeffizienz zu maximieren und die Stromgestehungskosten und die spezifischen Produktkosten zu minimieren. Nach den Ergebnissen der tri-objektiven Optimierung am optimalen Punkt kann die Anlage einen Energie- und Exergiewirkungsgrad von 55,07 % bzw. 45,08 % erreichen. Die Gesamtleistung beträgt 9,10 MW, bei Stromgestehungskosten von 109,02 \$/MWh. Die spezifischen Produktkosten werden um 16,8 % auf 22,52 \$/GJ minimiert. Der Kapitalwert der Konfiguration wird optimal erhöht und die Gewinnschwelle um 16,6 % auf 5,43 Jahre gesenkt. Bei der stufenweisen CO₂-Emission wird eine Reduzierung um 16,7 % erzielt, der Exergie-Nachhaltigkeitsindex der Anlage steigen jedoch von 1,58 auf 1,82.

Abstract

The escalating concerns over energy security, environmental sustainability, and the adverse impacts of fossil fuel combustion have intensified the need for innovative and sustainable energy solutions. In this context, a biomass-driven integrated system is proposed in this study. The system comprises of anaerobic digestion, gasification, proton exchange membrane electrolyzer (PEME), Sabatier reactor, solid oxide fuel cell (SOFC), a gas turbine, steam turbine, and organic Rankine cycle (ORC). Biomass feedstocks in the form of animal waste serves as input in the digester and is converted to biogas, and crop residue is converted to syngas in the gasifier. Upgraded syngas and methane from the bio-conversion process is fed to a SOFC-GT topping cycle, with heat recovery bottoming cycles of steam turbine and organic Rankine cycles. To improve the contribution of the ORC to the plant performance, a selection of the best ORC working fluid from six preselected candidates (MM, MDM, cyclopentane, cyclohexane, R1233zd(E), and R600a) is carried out. The proposed system is assessed from thermo-economic, exergo-economic and enhanced exergoeconomic viewpoints in Engineering Equation Solver (EES) software. Hydrogen of 0.0023 kg/s with PEME efficiency of 73.73 % is obtained and further used in upgrading the syngas from low heating value of 3.85 to 33.48 MJ/kg. Parametric analysis was performed to ascertain the effect of design parameters on the plant's performance, and it is indicated that low pressure ratio, high current density, and high turbine inlet temperature are favorable for high energy and exergy efficiency, low levelized cost and low cost of product. An improvement potential analysis based on the enhanced exergo-economic analysis indicates a possibility of recovering 32 % of the total exergy destroyed in the plant. Lastly, a multi-criteria optimization is performed using a multiobjective genetic algorithm (MOGA) in MATLAB to maximize exergy efficiency, minimize the levelized cost of electricity, and the specific cost of product. According to the results of the triobjective optimization at the optimum point, the plant can attain energy and exergy efficiencies of 55.07 % and 45.08 %, respectively. The total power output is 9.10 MW, with a levelized cost of electricity of 109.02 \$/MWh. The specific cost of product is minimized by 16.8 % to 22.52 \$/GJ. The net present value of the configuration is increased at optimum, and the break-even point reduced by 16.6 % to 5.43 years. A reduction of 16.7 % is obtained for the levelized emission of CO₂, however, the plant's exergy sustainability index increases from 1.58 to 1.82.

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Symbols and abbreviations

Symbols	Name	Unit
Α	Activation area	m ²
Ėx	Exergy rate	kW
е	Specific exergy	kW/kg
F	Faraday constant	C/mol
G	Gibbs free energy	kJ
g	Specific Gibbs free energy	kJ/kmol
h	Specific enthalpy	kJ/kg
Ι	Current	А
j	Current density	A/m ²
Κ	Equilibrium constant	-
'n	Mass flow rate	kg/s
Ν	Number of cells	-
'n	Molar flow rate	kmol/s
Р	Pressure	kPa
Ż	Rate of heat transfer	kW
S	Specific entropy	kJ/kgK
r	Pressure ratio	-
Т	Temperature	Κ
V	Voltage	V
Ŵ	Power	kW
Ζ	Cost rate	\$

Subscript and superscripts

0	Ambient condition
a	Anode

act	Activation
са	Cathode
ch	Chemical
conc	Concentration
cr	Critical
Ε	Electrolyzer
eff	Effective
ex	Exergy
FC	Fuel cell
Net	Net value
ph	Physical
ohm	Ohmic
orc	Organic Rankine cycle
<i>PEM</i> E	Proton exchange membrane electrolyzer
sofc	Solid oxide fuel cell
st	Steam turbine
th	Thermal

Greek symbols

α	Number of electrons	no.
β	Exergy ratio	-
η	Efficiency	%
ρ	Power density	kW/cm ²
σ	PEME conductivity	$\Omega^{-1}m^{-1}$
λ	Water content at anode and	cathode
μ	Chemical potential	
θ	Pressure factor of ORC	-

Abbreviations

AC	Air compressor
ACC	Annual capital cost
AD	Anaerobic digestion

BEP	Breakeven point
BICFCC	Biomass integrated co-fired combined cycle
BIPFCC	Biomass integrated post-fired combined cycle
CH4	Methane
СО	Carbon monoxide
CO ₂	Carbon dioxide
COx	Oxides of carbon
CRF	Capital recovery factor
EES	Engineering equation solver
EFCC	Externally fired combustion chamber
ER	Equivalence ratio
ESI	Exergy sustanaibility index
FC	Feel compressor
GT	Gas turbine
H2	Hydrogen
HCCI	Homogeneous charge compression ignition engine
HHV	Higher heating value
HRSG	Heat recovery steam generator
HX	Heat exchanger
LCOE	Levelized cost of electricity
LE	Levelized emission
LHV	Lower heating value
LINMAP	Linear programming for multidimensional analysis of preference
MCFC	Molten carbonate fuel cell
MDM	Octamethyltrisiloxane
MED	Multi-effect desalination
MM	Hetamethyldisiloxane
MOEA	Multi-objective evolutionary algorithm
MOGA	Multi-objective genetic algorithm
MSW	Municipal solid waste
MTH	Methanation

NOx	Oxides of nitrogen
NPV	Net present value
NSGA	Non-dominated sorting genetic algorithm
ORC	Organic Rankine cycle
R123	2,2-Dichloro-1,1,1-trifluoroethane
R1233zd(E)	1-Chloro-3,3,3-trifluoropropene
R141b	1,1-Dichloro-1-fluoroethane
R600a	Isobutane
SNG	Synthetic natural gas
SOFC	Solid oxide fuel cell
ST	Steam turbine
TEE	Total effective electricity
TOPSIS	Technique for order of preference by similarity to ideal solution
WTE	Waste-to-energy

1 Introduction

1.1 Background

Energy systems are crucial in driving almost all practical systems and are a necessity for the quality of life[1]. Despite the major role of energy in societal growth and development, major adverse results of human activities in power generation from fossil fuel resources are increased greenhouse gas (GHG) emissions, waste heat, pollution, and global warming[2]. These challenges are induced by population growth and the need for more energy[3]. Reliance on fossil fuels contributes to economic uncertainties attributed to energy price volatility, with the likelihood of rising prices and scarcity of fuel supplies. However, efforts to mitigate these challenges require the development of energy-efficient systems powered by clean energy resources otherwise conventional energy systems must be replaced or greatly enhanced before they can meet the multifaceted and interdisciplinary needs of this century[4].

Addressing some of the issues associated with the use of fossil fuels for energy production, energy resources must be diversified to include less expensive and more environmentally benign resources. Furthermore, upgrading low-energy dense fuel from renewable energy resources such as biomass has also been strongly encouraged as a potential option for attaining significant reductions in energy costs, low environmental impact, and energy savings[5]. Hence, sustainable energy conversion is crucial for reducing greenhouse gas emissions, mitigating climate change, and achieving a more sustainable energy future[6],[7]. Sustainable energy conversion refers to the transforming of renewable energy sources into usable energy forms in a way that is environmentally friendly, efficient, and economically viable. This includes the use of technologies such as solar photovoltaic, wind turbines, hydroelectric power, geothermal power, and biomass energy to convert natural resources into electricity, heat, and mechanical energy[6]. The approach created by sustainable energy development is expected to set up a balance between economic, social, and environmental considerations of energy conversion systems to meet current energy

needs without affecting future demands. Hence, with continued research, development, and investment, a sustainable energy future is possible[8].

In recent years, there has been a significant increase in the use of sustainable energy sources globally. According to the International Energy Agency (IEA)[9], renewable energy accounted for approximately 29 % of the world's electricity production in 2020, up from 27 % in 2019. This increase is due to the declining costs of renewable energy technologies, favorable government policies, and increased public awareness of the need for sustainable energy sources. Despite the increase in renewable energy adoption, the world's total energy demand continues to rise. The majority of this energy demand is projected to come from developing countries, where the population is growing, and economies are expanding[10].

This is evident in Nigeria, a developing country located in Sub-Saharan Africa, having a population of over 200 million and abundant in sources of renewable energy still facing the challenges of meeting energy demands for power generation[11]. With increasing population and the continuous yearning for developmental growth, Nigeria's demand for sustainable energy is very high and unabated with diminishing available energy infrastructure grossly inadequate to meet energy demands in rural areas[12]. This coupled with the nation's soaring GHG emissions with records of 25% (98.22 MtCO2e) between 1990 to 2014, has emphasized the importance of sustainable energy systems, currently viewed as one crucial tool for improving human lives and standard of living especially for those in the rural areas of the nation[13]. In view of overcoming these challenges, Nigeria has pledged to reduce its emissions by 20 % in 2030 as against business as usual via improving energy efficiency by 20 % with at least 13 MW of renewable power provision to rural areas with no access to the power grid[14].

Globally, there is room for the development and assessment of sustainable solutions that make use of one or more environmentally friendly inputs. This will pave the way for the production of a variety of expected outputs to achieve the goal of sustainable energy conversion, with a transition from conventional to sustainable energy.

1.2 Problem statement and justification of the study

The escalating concerns over energy security, environmental sustainability, and the adverse impacts of fossil fuel combustion have intensified the need for innovative and sustainable energy solutions[15]. In this context, the concept of a biofueled multigeneration plant has gained prominence as a potential remedy to address multiple challenges simultaneously. A biofueled multigeneration plant is an integrated energy system that combines the production of biofuels with electricity, heat, and other valuable products, aiming to achieve enhanced resource efficiency, reduced greenhouse gas emissions, and greater energy independence[6]. Such plant allows for the integration of several methods of energy conversion ranging from biochemical (anaerobic digestion), thermochemical (gasification, and methanation), electro-chemical (fuel cell and electrolyzer), and as well as power cycles (gas turbine, steam turbines, and Organic Rankine cycle)[16]. However, the successful implementation of such a complex and multifaceted system is riddled with various challenges that demand meticulous attention. One of the foremost challenges is the development of a robust feedstock supply chain that can ensure a steady and sufficient supply of biomass resources. The feedstock procurement involves numerous intricacies, including resource availability, transportation logistics, and competition with food and other industries. Efficiently managing these factors while adhering to sustainability criteria remains a significant hurdle that needs to be addressed. Remote and rural localities seeking development in terms of energy supply and host agrarian facilities capable of supplying sustainable amounts of biomass can be considered as areas of interest in meeting this goal[11], [17], [18].

Furthermore, biomass conversion into high-quality fuels is faced with technical difficulties such as waste heat generation and upgrading low-grade fuels[19], [20]. The selection and optimization of appropriate conversion technologies, such as biochemical and thermochemical processes, require comprehensive research to enhance conversion efficiency and product yields. Overcoming issues resulting from feedstock variability, process integration, and reactor design are pivotal to achieving cost-effective biofuel production in the multigeneration plant. Another significant issue that needs to be resolved is integration. Process compatibility, heat exchanger design, and efficient energy distribution systems must all be carefully considered. This will allow for diverse processes within the multigeneration plant, such as power generation, heat recovery, and biofuel production, to be integrated seamlessly[6]. Achieving a high level of process integration improves overall efficiency while also reducing the plant's environmental impact. For instance, Ogorure et al.[21] in a previous study, proposed a 5.2 MW integrated multigeneration plant to address the power demand of an agrarian facility. They integrated anaerobic digestion and gasification of biomass in a plant with solid oxide fuel cell, gas turbine, steam turbine, organic Rankine cycle (ORC) for power generation and cooling purposes. Despite the high efficiency reported, the study did not address improving

the low energy density of the syngas from gasification, optimization of the power plant, and utilizing the excess power from the plant.

Other major challenges are the economic viability and environmental sustainability of multigenerational plants. Financial risks can be significant due to the up-front capital expenditures needed to set up such sophisticated systems as well as the unpredictability of feedstock, and biofuel market demand[22]. For determining the long-term viability of these plants and attracting required investments, it is essential to consider precise economic modeling that can account for the full value chain, from feedstock supply to product distribution. Environmental sustainability is at the core of the biofueled multigeneration plant concept. Nevertheless, ensuring that the net environmental benefits are realized requires an assessment that accounts for the entire supply chain, and energy production processes and balances the environmental advantages of reduced plant emissions such as CO_2 , CO, NO_x emissions with potential negative impacts[23].

Lastly, the development of an efficient and sustainable biofueled integrated plant presents a multifaceted problem that covers technical, economic, environmental, and policy domains. Finding solutions to the challenges associated with feedstock availability, conversion technologies, integration, economics, and sustainability will be essential in achieving the full potential of this innovative energy concept. The successful resolution of these challenges will not only contribute to a diversified and cleaner energy mix, it will also encourage economic growth and environmental stewardship on a global scale.

1.3 Aim and Objectives

The study aims to conduct thermo-economic and optimization analysis for optimal performance and minimum cost of an integrated thermal plant configuration for a tropical region and to determine possible plant component modifications. The plant consists of a hybrid biomass conversion unit with a syngas upgrade and an integrated power plant unit. The primary objectives of this study are as follows:

- i. To develop, in a tropical area, an integrated biomass conversion system based on the anaerobic digestion (AD) and gasification of agricultural waste for the generation of power, and hydrogen production (used in syngas upgrade).
- ii. To analyze the overall system performance of integrating different thermodynamic and electro-chemical systems of anaerobic digestion and gasification, methanation, organic

Rankine cycle (ORC), Brayton cycle, Rankine cycle, solid oxide fuel cell (SOFC), proton exchange membrane electrolyzer(PEME).

- iii. Determination of the benefits of employing the integrated plant from a thermodynamic, economic, and environmental point of view.
- iv. To investigate the impact of selected operating parameters, including the choice of ORC working fluid on the system outputs.
- v. Perform a multi-objective optimization of the power plant to improve the integrated plant performance with respect to net power output, levelized cost of electricity and specific cost of products.

1.4 Scope of study

With a population of 1.78 million, the chosen location in this study is southeast of Rivers State in southern Nigeria. The estimated power consumption in this region is 240 GWh, which is hardly available. However, a significant amount of biowaste from agricultural activities is produced in this region, which should be sufficient to support energy recovery through biowaste conversion. To estimate the necessary biowaste flow input to the biomass section of the plant, realistic values for the annual biowaste generation in each locality are needed. However, due to a lack of census data, accurately reported data may not always be available, necessitating an estimation based on previously reported data. Analytical investigations are predominantly carried out in this research. Thermodynamic, electro-chemical, and economic principle-based computational models were used in the computation. Data obtained from previous studies of a hybrid bio-conversion plant were used to validate the biomass conversion from raw biomass to fuel. Multi-criteria optimization was conducted using a multi-objective genetic algorithm (MOGA) and implemented in a MATLAB-EES computational interface.

1.5 Research methodology

The first and second laws of thermodynamics were applied in developing computational models used to analyze the plant configuration components and the proposed performance of the system. Thermodynamic, economic, exergo-economic, and environmental analysis were performed in the Engineering Equation Solver (EES) software. Simulations were performed in a Microsoft Excel environment and optimization integrating multi-objective genetic algorithm in MATLAB software, with models in EES was carried also out.

1.6 Thesis Outline

This study is structured into 5 chapters, and Chapters 2 to 5 are outlined as follows:

Chapter 2 introduces the reader to the hybrid biomass-fueled multigeneration concepts with its main components and operation. Firstly, a review of biomass conversion methods is presented, which is preceded by a biofuel upgrade. In addition, a review of previous studies on the integration of several energy conversion techniques from fuel to useful energy forms related to the objectives of this study is presented.

In Chapter 3, a detailed description of the integrated plant, as well as thermo-economic, conventional and enhanced exergo-economic modeling is presented in other to address the objectives of the study. Furthermore, the modeling of the optimization of the power plant with the application of MOGA and LINMAP and its performance considering different ORC working fluids, with varying plant variables were presented.

Chapter 4 contains the results of the thermo-economic, conventional, and enhanced exergoeconomic assessment and optimization method presented in Chapter 3. Model validation of the gasification process, PEME, and SOFC, as well as simulation of key operating parameters are presented in this chapter.

The overall summary and conclusions are summarized in Chapter 5 along with the thesis contributions to knowledge and further studies recommendations.

2 Literature Review

2.1 Overview

In this chapter, relevant developments and studies in biomass conversion techniques, anaerobic digestion, gasification, syngas upgrade, and the thermodynamic performance assessment and optimization of thermal power plants of interest to this work are presented. Research works on biomass to power generation and other useful products using proton electron membrane electrolyzer, Solid oxide fuel cell, gas and steam turbine cycles, organic Rankine cycle (ORC) power plants, multi-generation energy systems and multi-criteria decision analysis are also reviewed. This chapter concludes with a summary of the reviewed literature and the gap that this study intends to fill.

2.2 Biomass conversion: Anaerobic digestion, gasification, fuel density, and upgrade

Biomass obtained from organic matter such as agricultural residues, plants, and forestry byproducts, is known to possess significant potential for energy generation[24], [25]. This resource is considered a renewable energy source due to its reliance on the continuous growth and replenishment of biomass feedstocks. Such feedstock includes energy crops (switchgrass and miscanthus), agricultural residues (corn stover, rice husks, cassava peels, palm effluent, and fibre), forestry residues, municipal waste and organic waste from animals[26], [27], [28]. The diverse nature of biomass feedstock ensures a steady supply of raw materials which can be transformed into various forms of energy such as biofuels, electricity, and heat, and hence, its versatility and an option for sustainable energy. Biomass possesses the potential to significantly contribute to global energy generation. It is estimated that up to one-third of the world's total energy can be provided by biomass. This includes heat for industrial processes and residential heating, as well as energy generation[18], [29]. Biomass energy can be integrated into various sectors to enhance overall efficiency. For instance, combined heat and power (CHP) systems allow simultaneous production of electricity and useful heat, maximizing energy utilization. Furthermore, it can also be co-fired in existing power plants, reducing greenhouse gas emissions and providing a transition toward cleaner energy[30]. With proper management and technological advancements, biomass can play a significant role in reducing greenhouse gas emissions and transitioning towards a cleaner and more sustainable energy future.

2.3 Biochemical and thermochemical conversion of biomass

Biomass can be utilized through biochemical and thermochemical conversion methods[31]. In Figure 2.1, a breakdown of the conversion methods of biomass is presented. Anaerobic digestion, a biochemical method, uses microorganisms to break down organic matter in the absence of oxygen, producing biogas that can be used for electricity and heat[31].



Figure 2.1: Biomass conversion through biochemical and thermochemical routes[31] Biogas is composed of about 55 % methane (CH₄) and 45 % carbon dioxide (CO₂), with high calorific value[32]. Aerobic digestion, also a biochemical method, occurs in the presence of oxygen

to produce CO₂, and digestate in solid form. The digestion processes employ the use of a digester[31]. Pyrolysis involves heating biomass feedstock in the absence of oxygen to produce bio-oil, syngas, and biochar[33], [34]. In combustion, the biomass is burned directly to produce heat, which can be directly used or converted to electricity via steam turbines[35].

Gasification converts biomass with air or steam as catalysts into combustible products of syngas which can be used for heat and power generation or further biofuel refined into biofuels[36], [37]. Syngas composition includes hydrogen (H₂), carbon monoxide (CO), and lower molecular hydrocarbons such as CH₄[38]. Due to the amount of hydrogen it contains, the fuel is suitably applied in fuel cells, and multigeneration system configuration[29]. Gasifiers such as the entrained flow, fluidized bed and fixed bed gasifiers are used in carrying out the gasification process. The downdraft and updraft are two most considered gasifier designs[31]. Liquefaction of solid biomass into liquid fuel can be accomplished through pyrolysis, gasification, or hydrothermal processes. The biomass is transformed into an oily liquid by exposing it to water at temperatures of 300 – 350 °C and high pressures of 12 - 20 MPa over an extended length of time[39]. The process has attracted little interest due to complex reactor and fuel-feeding system design and high cost implications[40].

The concepts of anaerobic digestion and gasification has gained widespread attention. In the conversion of biomass to useful forms owing to distinct advantages ranging from cost, biomass type, product gas, energy demand, and environmental impact and have been reported in literature. Imeni *et al.*[41] carried out techno-economic analysis on the anaerobic digestion of manure from livestock and whey cheese. They considered the biogas production rate and levelized cost of electricity as profitability determinants. Their results showed that mono-digestion attracted higher cost compared to co-digestion of manure with 30 % whey cheese, with the return of the investment less than 10 years. A theoretical model of anaerobic digestion was proposed by Achinas and Euverink[42] to predict the biogas amount of agricultural waste. They applied the elemental composition of biomass from the modified Buswell and Hatfield model to estimate the amount of theoretical biochemical methane potential of the biomass. According to Li *et al.*[43] Co-digestion and pretreatment are predominant activities that greatly enhance the product of anaerobic digestion of biomass. They investigated the co-digestion of cow manure with other substrates to biogas. Riau *et al.*[44] considered the anaerobic co-digestion of catch crop species of ryegrass, forage rape and

black oat, with dairy manure for high biomass yield of biomethane. Their results indicated an estimated 35 - 48 % increase in estimated, methane output compared to single digestion of dairy manure. Ma *et al.*[45] carried out a meta-analysis on methane yields during anaerobic co-digestion of animal waste with other feedstocks. High methane yield was reported for co-digestion of animal waste with cow manure having the best improvement with optimum substrate containing carbon/nitrogen ratio of 26 to 34. Allesina *et al.*[46] modeled the coupling of gasification and anaerobic digestion processes for maize as fuel in a power plant. Maize grains were fed to a digester and maize stover were gasified to obtain products of biogas and syngas. An improved energy yield of 39% was obtained when compared to the conventional standalone digester. Zou *et al.*[47] proposed an enhanced anaerobic digestion of biomass through coupled microbial fuel cell. They reported improved biogas and methane yield with accelerated substrate degradation at a suitable voltage of 0.9 V.

Li *et al.*[48] developed a mathematical model in predicting the fates of carbon C, nitrogen N, and phosphorus P, in anaerobic digestion of pig manure and the optimization of biogas product. Good accuracy and reasonable outcome of C, N, and P were exhibited. Also, simulation of CH₄ production demonstrated the benefit of pretreatment of the waste to improve the amount of CH₄ produced. Chen *et al.*[49] studied the air gasification of digestate in a downdraft gasifier at temperature ranges of 600 - 800 °C and equivalence ratio (ER) of 0.25 - 0.30. their results showed zero ash slagging at temperatures of 800 °C and ER of 0.25 - 0.30. Also, gasification at high temperatures with medium ER favored the syngas yield, resulting in high cold gas efficiency of 67.01 % and LHV of 4.78MJ/Nm³. Galvagno *et al.*[50] analyzed the gasification of citrus peels coupled with a 120 kW (DC) solid oxide fuel cell unit. Heat from the fuel cell was used to preheat the raw biomass in a waste-heat recovery scheme. Due to water gas shift reactions, a subsequent decrease in CO and high CO₂ amounts was reported. They attained a high H₂:CO₂ ratio at a high steam-to-biomass ratio. Said *et al.*[51] optimized the gasification of syngas from Napier grass via stoichiometric equilibrium. They reported an optimum temperature of 1123 K and ER of 0.21 for high yields of H₂ in the syngas.

Despite the advantages attributed to the conversion of biomass through gasification and anaerobic digestion as shown in the above literature, the energy density of the fuel gas is low and requires upgrading to meet the standards of natural gas.

2.4 Hybrid conversion with syngas upgrade

Hybrid biomass conversion involves the integration of one or more biomass conversion techniques for the purpose of improving product quality. For instance, notwithstanding the gains associated with syngas from gasification, the fuel density is low compared to natural gas[52]. This can be improved by converting the CO₂, and CO to CH₄ in a secondary downstream reactor or Sabatier reactor in a process referred to as methanation[53]. In order to attain a successful methanation process, hydrogen feed of 4 times the amount of CO₂ is required. The system however requires a source of hydrogen, such as an electrolyzer[54], [55]. The methanation process can be used directly in the conversion of CO and CO₂ obtained from other processes, or incorporated into a system that utilizes biomass of fossil fuels. A general integrated configuration for synthetic natural gas (SNG) production from biomass or carbon dioxide is presented in Figure 2.2. Anaerobic digestion and gasification are utilized in producing fuel gas from biomass. The plant includes parts for methanation, condensation, and heat recovery in addition to portions for electrolysis-based hydrogen production.



Figure 2.2: Hybrid biomass conversion to synthetic natural gas with methanation

In selecting a suitable electrolyzer, the Proton exchange membrane (PEM) electrolyzers are electrochemical devices essential for producing clean and sustainable fuels like synthetic natural gas, also known as syngas[56]. By transforming syngas into more valuable fuels through hydrogen

(H₂) production, these electrolyzers play a crucial role in upgrading syngas and lowering the carbon footprint of many industrial processes. PEME are a form of water electrolysis technology that converts water into hydrogen and oxygen gases through the passage of an electric current[57]. They are made up of a solid polymer electrolyte membrane that selectively permits protons (H⁺ ions) to be transported while obstructing the movement of electrons[58]. With this configuration production of hydrogen is for syngas upgrade can be efficient and controlled.

Syngas is an important intermediate product in many industrial processes, including coal gasification, biomass gasification, and syngas-based chemical synthesis. PEM electrolyzers are particularly suited for processing dynamic syngas as they work at temperatures that are relatively low $(60 - 80^{\circ}C)$ in comparison to other electrolysis technologies[59]. This leads to improved energy efficiency and quicker response times. These electrolyzers are easily scalable to accommodate the syngas production capacities of a range of applications, from small-scale dispersed systems to substantial industrial facilities[60]. PEM electrolyzers have become a versatile and effective technology for syngas upgrade procedures, enabling the creation of synthetic gases that are cleaner and more valuable. PEM electrolyzers are anticipated to play a significant part in reaching sustainability goals by upgrading syngas and helping to the creation of a cleaner, greener energy landscape as the world continues to move towards cleaner energy sources and lower carbon emissions.

In reality, the system integration can be more complex, however, the integrated system can allow for improvements. This includes heat recovery employed to generate power or heat and may be applied to improve the hydrogen production from the electrolyzer depending on the plant's requirements[52]. In another scenario, the plant can be tasked with creating high-quality electricity for the grid as multigeneration is more effective than producing fuel for only the purpose of generating some of the required electricity. Ogorure *et al.*[61] conducted a study on waste heat conversion for syngas upgrade. A hybrid biomass conversion system with integrated ORC-PEME and various ORC working fluids was examined. Maximal energy efficiency of 19.23% was recorded in the ORC configuration with a regenerator using R1233zd(E) as working fluid. Adnan *et al.*[62] identified the combination of electrochemical reduction of CO₂ were converted into chemical products (i.e., CO or MeOH) through an electrolyzer, and electricity from the gasification process. Variation of the flow rate of the gasifying agent (O₂ equivalence ratio between 0.36 and 1.00) provided significant changes in the syngas composition (H₂: 28 – 65 %; CO: 25 – 43 %) and heating value (12 - 30 MJ/kg). The cooling system had the largest contribution to the exergy and a net negative CO₂ emission ranging from - 0.09 to - 0.17 kg CO₂/GJ at O₂ equivalence ratio of 0.36 was obtained for the configuration.

Clausen et al.[63] carried out the thermodynamic modeling and analysis of biomass to syngas system from manure. Their results highlighted the potential increase in energy efficiency and total yield when compared to a conventional stand-alone anaerobic digester. An increase from 49 % for a stand-alone anaerobic digester to 138 % for the conversion of digestate to gas by thermal gasification, and syngas upgrade the syngas to bio-SNG by electrolysis was reported. This was attributed to the gasifying agent, a two-stage gasifier with tar-free product, and the application of an electrolyzer for the syngas upgrade. Becker *et al.*[64] explored thermodynamic and kinetic models in analyzing the methanation process for the conversion of hydrogen and carbon dioxide to SNG. Hydrogen was supplied by a low-temperature PEME unit. The overall plant efficiency was reported as 78.1 % of LHV and 81 % of HHV. Furthermore, the electrolyzer technique of hydrogen production was estimated to provide LHV efficiency of 66 % for grid power generation with costs of up to 66 \$/MWh. He[65] strategically integrated the processes of gasification and anaerobic digestion in the conversion of biomass for biofuel production with the aim of reducing production costs. Bio-syngas product with estimated 44% biomass conversion was attained. Moreover, an 18.6 % reduction in overall cost was reported with the conversion technique. However, this study did not cover exergy, exergo-economic and environmental assessment of the plant. Yao et al.[66] investigated a hybrid biomass conversion technique with a combined anaerobic digestion and gasification process. AD was applied in converting yard waste to biogas and woody biomass with effluent from the AD was converted to syngas via gasification. Optimal energy of 70 % was reported for the combined methods. Harvanto et al.[67] studied the thermodynamic limits of a syngas upgrading process through minimization of Gibbs free energy over a temperature range of 400 - 1300 K. Further conditions considered were pressures of 0.1 - 100 K. 1 MPa, and several carbon-to-steam ratios. The study revealed that between pressure ranges of 0.1 -0.3 atm applied at temperatures less than 1000 K, high H₂ content can be observed in the products. Andrea and Yahya[68] proposed a power-to-gas technique integrating an electrolyzer, gasifier, and Sabatier reactor for syngas upgrade. Under thermodynamic analysis, energy and exergy efficiencies in the range of 55 - 80 % and 35 - 40 %, respectively, were obtained. Their results stated a peak revenue of 0.22 \$/kWh on a dry biomass basis, with the best operational performance under conditions of feeding the methanation unit with the entire gas mix as that of the products.

Gao *et al.*[69] analyzed the methanation reactions of oxides of carbon (CO and CO₂) using the method of minimization of Gibbs free energy. They reported that high CH₄ yield can be obtained from the methanation of CO under high pressures, low temperatures, and high H₂:CO ratios. Also, the methanation of CO₂ was identified as relatively difficult under the same conditions as CO with high CO₂ content in the product.

According to Kuo et al. [70], the role of waste-to-energy conversion technologies in generating renewable energy and finding solutions to environmental challenges cannot be undermined. They developed a WTE power plant with feedstock of refuse-derived fuel, integrated with plasma gasifier, solid oxide fuel cell (SOFC), and combined heat and power unit. The results indicated that preprocessing with torrefaction improved syngas quality, with a negative effect on the return of investment due to high energy consumption by the plasma gasification and torrefaction. Towards improving the quality of the product of biomass conversion technologies for better technical, economic, social, and environmental performance indicators, hybrid biomass conversion, and biofuel upgrading are notable options as revealed in open literature. Zhang et al.[71] proposed an integrated anaerobic digestion and gasification conversion system to determine the potential of food waste for hydrogen and methane co-production. With the application of thermal-equilibrium model, a higher hydrogen content of 28.9 % with a high moisture content of 55 wt-% was generated by gasification when the value of ER is 0.35. A calorific value of 5.59 MJ/Nm³ for syngas at the conditions of 60 wt-% of moisture content and 0.35 of ER was obtained. The overall energy performance of the showed that the combined conversion significantly improved the total energy output in the form of electricity and heat.

2.5 Thermo-economics and environmental analysis of multigeneration configurations

Engineering and economics apply thermo-economic analysis approach to evaluate and improve the efficiency and cost-effectiveness of energy systems and processes[72]. In order to assess the trade-offs between energy efficiency, capital costs, running costs, and environmental implications, it blends thermodynamics, economics, and engineering principles[73]. Making essential decisions concerning the design, operation, and optimization of energy systems, such as power plants, industrial processes, and renewable energy technologies, requires a multidisciplinary approach[72], [74]. In thermo-economic analysis, firstly a thorough thermodynamic examination

of the energy and exergy flows within a system is carried out. Energy input and output are quantified in energy analysis, whereas energy quality is evaluated in exergy analysis by considering the availability and utility of the energy. These assessments aid in locating system losses and inefficiencies [75]. Thermo-economic analysis integrates the results of thermodynamic analysis with cost modeling to evaluate the economic performance of an energy system. It considers factors such as the cost of components, materials, energy sources, and labor, as well as the expected revenue or benefits from the system's operation[76]. Furthermore, the analysis employs various efficiency and performance metrics, including the first-law efficiency (energy efficiency), secondlaw efficiency (exergy efficiency), and economic metrics such as the levelized cost of electricity (LCOE)[77]. These metrics provide a comprehensive view of how efficiently an energy system converts resources (e.g., fuel or renewable inputs) into useful energy and at what cost. Environmental factors, such as greenhouse gas emissions and other pollutants, are increasingly considered in thermo-economic analysis[78]. This enables decision-makers to evaluate the environmental impact of energy systems and make decisions that are in line with regulatory and sustainability objectives. Numerous energy systems and processes, such as multigeneration plants, industrial processes (such as chemical manufacturing), and renewable energy technologies (such as biomass, wind and solar farms), are subjected to thermo-economic analysis.

Although the use of biofuels for power generation has a long history, recent improvements in energy conversion systems have made it possible to use these fuels in new systems that integrate several power cycles. Hence the development of the concept of biofueled multigeneration systems. By making the best use of available resources locally, multigenerational plants can be a very useful alternative for meeting needs[79]. These plants enable the fusion of multiple technologies, including fuel cells, gas turbines, steam turbines, organic Rankine cycle, refrigerating systems, hybrid, combined, multistage, and cascaded systems. This method can be applied to plants that have the capacity to produce large amounts of thermal energy, such as gasification or the thermochemical conversion of biomass[80].

Solid Oxide Fuel Cells (SOFC) have attracted substantial interest in the field of energy production and integrated plants due to their high efficiency, low emissions, and adaptability[81]. Such plants have a huge potential for improving energy efficiency and minimizing environmental impact by incorporating SOFCs. The chemical energy of the fuel is immediately converted into electricity and heat via SOFCs, which are electrochemical devices. They run at high temperatures (usually between 600 and 1000 °C), which facilitates quick electrochemical reactions and efficient fuel

consumption[81–83]. The anode and cathode materials in these cells are porous, and the solid electrolyte is commonly yttria-stabilized zirconia (YSZ)[81]. To increase energy efficiency in a multigenerational plant, SOFCs can be integrated in a variety of ways. As part of a hybrid system, the SOFC stack is frequently coupled with a gas turbine or a micro gas turbine[84]. The turbine is driven by the SOFC's high-temperature exhaust gases to produce more electricity, and any unused heat can be utilized in combined heat and power applications. The overall energy efficiency of multigenerational plants based on SOFCs is high, frequently surpassing standalone plants. The combined production of useable heat and electricity leads to this efficiency, which lowers primary energy use and greenhouse gas emissions[85]. Pollutant emissions such as NOx and COx are reduced in part because there is no combustion taking place inside the SOFC stack[86]. The fuel flexibility of SOFCs is well recognized. They can run on several hydrocarbon fuels, such as hydrogen, biogas, and natural gas, making them useful for a range of applications and easily adaptable to available fuel sources[81]. There is a lot of potential for meeting energy needs more effectively and sustainably by integrating SOFCs into integrated facilities. These systems are likely to be crucial in the shift to cleaner and more sustainable energy sources as technology advances and becomes more commercially viable. SOFCs that are included into multigeneration plants offer an appealing option for satisfying power and heat demands with high efficiency and no environmental effect. They play a crucial role in the shift to a more efficient and sustainable energy system due to their fuel flexibility, and capacity to function in a variety of applications. Future advancements and wider implementation of SOFC technology are likely to be the outcome of ongoing research and development.

Additionally, multigeneration plants possess the advantage of utilizing low-grade energy from its thermal processes[79]. This gives room for further generation of power through the application of the organic Rankine cycles (ORC). The ORC is a thermodynamic procedure that has attracted a lot of interest as a reliable and effective way to produce electricity from low-temperature heat sources[87]. The selection of working fluids is important in defining the cycle's performance, efficiency, and environmental impact, which are key elements influencing the success of the ORC[88]. While working fluids in the ORC are specifically chosen for their applicability in low-temperature heat recovery applications, they perform similar functions to steam in conventional Rankine cycles. These fluids are necessary for efficient heat transfer and the transformation of heat energy into mechanical work that can be used to produce electricity. The efficiency, operating temperature range, environmental impact, and safety issues of the ORC are all impacted by the

working fluid choice, making it crucial[89]. The Organic Rankine Cycle (ORC) working fluid categorization is crucial for maximizing the efficiency of ORC systems because different working fluids have varying thermodynamic features and characteristics that make them suited for diverse applications. These fluids include refrigerants, organic fluids, and high-temperature fluids such as siloxanes which have been studied by researchers for the improvement of the ORC performance[88,90,91].

Studies from literature indicate the potency of biomass-based multigeneration configurations with biogas and syngas for global climate change and environmental decarbonization. Wu et al. [29] modeled a power and heat generation plant consisting of a biomass gasifier, SOFC, homogeneous charge compression ignition engine (HCCI), and waste heat recovery unit. A comprehensive evaluation based on thermodynamic and environmental principles were performed with results revealing the gasifier as the largest contributor to exergy destruction having a relative exergy destruction of 21.5 %. The HCCI engine, on the contrary, contributed more to the exergy destruction and less power, compared to the SOFC. High exergo-economic factor was associated with the SOFC due to its high investment cost. Emissions of CO₂ were estimated to range from 0.119 - 0.139 t/GJ, indicating low environmental impacts for the proposed plant. A biogas multigeneration plant with biogas generated from the anaerobic digestion of sewage sludge and useful outputs of power, heat, freshwater, and hydrogen from the integration of the PEME was evaluated by Safari and Dincer[4]. Using a gas turbine (GT) and ORC unit with a net power of 2.17 MW, the plant's overall energy and energy efficiencies were 63.6 % and 40 %, respectively. A waste-to-energy multigeneration plant featuring an anaerobic digester and gasifier was examined by Ogorure et al. [21]. They provided an example of how to successfully integrate multiple thermodynamic cycles using bio-syngas. They did not, however, take into account the thermodynamic analysis of the biomass to bio-syngas process. Moharamian et al.[92] assessed three biomass and biomass-natural gas integrated plants of 4 MW capacity. The plants consisted of an externally fired combustion chamber (EFCC) which burns only syngas from gasification, a biomass-integrated co-fired combined cycle (BICFCC), and biomass integrated post-fired combined cycle (BIPFCC) that utilized a mix of syngas and fossil fuel. This was considered based on the high flue gas temperature attained in the combustion process to drive the gas turbine. An ORC bottoming cycle was considered for five working fluids of R141b, R123, n-Pentane, HFE7000, and water in the analysis. The exergy efficiency of the BIPFCC was reported higher in all working ORC fluids and lower total unit product cost in the EFCC under exergo-economic analysis. Gholizadeh[93] carried out a thermodynamic and thermo-economic analysis of a biogasfueled modified gas turbine coupled with an organic Rankine cycle to ascertain the performance and cost of the plant and the results are compared with the basis coupled system. The results indicated the combustion chamber and the recovery heat exchanger had the highest contribution to the total exergy destruction rate. Net output electricity of 1368 kW, with thermal efficiency, exergy efficiency and total product cost of 41.83 %, 38.91 %, and 17.2 \$/GJ, respectively, was obtained for the configuration. Further results revealed that both efficiencies of systems could be improved by reducing the air compressor pressure ratio and steam turbine inlet pressure. The total product cost of the plant can be minimized with a high gas turbine inlet temperature and low air compressor pressure ratio.

A multigeneration plant that produced hydrogen and ammonia using syngas from solar and biomass sources was examined by Tukenmez *et al.*[94]. Energy and energy efficiencies of 5.876 % and 55.64 %, respectively, with 20 MW of net power and a 0.0855 kg/s hydrogen generation rate were achieved. Soleymani *et al.*[95] analyzed an integrated SOFC-GT topping cycle with a biogas reforming cycle. Flue gases exiting the topping unit was used in the reforming cycle to produce hydrogen. The thermodynamic analysis of the configuration yielded a net power, exergy efficiency, and hydrogen flow rate of 2.72 MW, 64.65 %, and 0.07453 kg/s. In the afterburner, a significant amount of exergy destruction of 26 % of the total exergy destruction rate was reported.

Holagh *et al.*[96] suggested a combined system including a desalination unit, gas turbines, SOFC, biomass burner, ORC, and refrigeration cycle. A desalinated water flow rate of 0.96 kg/s, net power of 4.4 MW and cooling capacity of 0.16 MW, was obtained by an exergo-economic and environmental study. Further results of the study include a hydrogen production rate of 0.00155 kg/s, a total cost rate of product of 223 \$/hr, costs per unit exergy of 11.28 \$/GJ, and an estimated CO₂ emission of 10.79 kmol/MWh. Shariatzadeh *et al.*[97] modelled and optimized a hybrid SOFC system fed by biogas obtained from hospital waste. The plant consisted of a 50 kW tubular SOFC combined with a chiller, combustion chamber and heat recovery steam generator. Their results showed increasing current density had a positive effect on the power outlet of the cell and attained maximum point before a negative effect on the cell energy production after this point. Also, high temperature at a constant current density increases the cell power which is a positive effect of the system temperature growth on current density and efficiency. With single objective optimization using genetic algorithm, the proposed plant can be afforded in the long term and will yield an annual net profit of US \$ 874,200.

Mahmoudi and Khani[98] applied thermodynamic and thermo-economic principles to determine the product costs and the order of significance of system components of an integrated SOFC-GT with domestic water heater plant. The exergy efficiency of the plant was maximized at a compressor pressure ratio of 9.49 and a current density of 1642 A/m². However, high pressure ratios were not favorable for the economic performance of the plant. In addition, the sum of the unit costs of the products is minimized at a maximum current density of 1523 A/m². Furthermore, an increase in the gas turbine inlet temperature or steam-to-carbon ratio deteriorates both the thermodynamic and economic performance of the system.

Emadi et al.[99] studied the prospect of a dual-loop ORC system in increasing the performance of an integrated SOFC-GT plant. They applied an artificial neural network (ANN) and multi-objective optimization method in the investigation of 20 ORC working fluids. Using R601 and Ethane as working fluids in the top and bottom cycles, respectively, the best results of 51.3 % overall system exergy efficiency and net electrical power of 1.04 MW were reported. The levelized cost of electricity (LCOE) for the SOFC-GT-ORC configuration was 33.2 \$/MWh, which was 12.9 % and 73.9 % less than SOFC-GT and SOFC cycles, respectively. Sevinchan et al.[100] proposed a biomass-fueled multigeneration plant of 1.078 MW net power, a heat capacity of 0.198 MW and 87.54 kW cooling load. The energy and exergy efficiencies increased with increasing useful outputs. The highest exergy destruction rate of 65 % was recorded in the combustion chamber, and low contributions of 9.2 % and 14.3 % recorded in the evaporator of ORC and biomass digester. Li et al. [101] carried out an exergy and environmental analysis of a biomass trigeneration system with solar panel coupling. The major components of the system were a gasifier, solar collector, combustion engine, and an absorption and liquid desiccant unit. Power was generated from the combustion of syngas and exit flue gas was utilized in a bottom cycle of absorption refrigeration for cooling and heat exchanger for hot water. A total exergy efficiency of 19.21 % was obtained, with decreased biomass consumption through the utilization of solar energy. The highest exergy destruction was from the gasifier, combustion chamber and gas/water heat exchanger. A reduction in CO₂ emissions was indicated with the combination of biomass and solar energy input resources. According to Leonzio[102] multigeneration processes can simultaneously produce several products and allow for the reduction of greenhouse emissions and more energy savings, compared to conventional systems. In his study, a biogas plant composed of an absorption heat pump, mechanical compression heat pump, steam plant and thermal recovery heat unit was analyzed. The plant generated electrical power, heat capacity and cooling capacity of 925 kW, 2523 kW, and

473 kW, respectively. However, due to a negative net present value and a payback period of 14 years, the plant was considered infeasible.

Malik et al.[103] proposed a multigeneration plant combining energy generating sources of biomass combustion and a geothermal resource. Both sources of energy were employed in an ORC, absorption cooling unit, Linde-Hampson liquefaction cycle, a drying unit and water heater system. Despite high exergy destruction recorded in the combustion chamber and boiler, the net energy and exergy efficiencies of the configuration were 56.5 % and 20.3 %, respectively. Ahmadi et al.[104] analyzed a multigeneration system for power generation, hydrogen production, and cooling purposes fueled by the combustion of biomass. An integration of ORC, PEME, absorption refrigeration, and domestic water heater were utilized in the plant configuration. The results revealed the combustion chamber and ORC evaporator as the major sources of the total irreversibility, with the pinch point temperatures having high effect on the system performance. Furthermore, the potential for the reduction of CO₂ emission was high for the plant configuration when compared to a conventional power generation plant. Bamisile et al.[105] studied a biogaspowered multigeneration configuration for power, heating, and cooling through the integration of reheating and regeneration in two steam cycles. They reported an improvement in the plant performance in terms of energy and exergy efficiencies from 43.96 % and 33.34 % to 64 % and 34.51 %, respectively, when several useful outputs were produced. He et al.[106] proposed a biomass-driven cogeneration plant comprising a gasifier, gas turbine, Stirling engine, and carbon dioxide cycle coupled with a domestic water heater. Feedstocks which include paper, wood, paddy husk, and municipal solid waste were used in the gasifier as the input fuel. The plant was analyzed from thermo-economic, and environmental viewpoints, with multi-objective optimization using genetic algorithm in MATLAB to obtain optimum operating parameters. The results indicated municipal solid waste as the input biomass with the highest exergy efficiency of 41.36 % and the lowest CO₂ emission of 0.9021 t/MWh. From the optimization analysis, a maximum exergy efficiency of 42.03 % was obtained with MSW. Also, the minimum achievable specific cost of product was 10.94 \$/GJ with paddy husk as input fuel.

An integrated system analyzed based on the principles of thermodynamics was reported by Anvari *et al.*[107]. They presented exergo-economic and thermodynamic viability of integrating gas turbines, with HRSG, absorption refrigeration, and regenerative ORC for power and heating. According to Chitsaz *et al.*[108], application of a primary driver's waste heat to sustain a subsystem serves a means of increasing the overall efficiency of a system. This allows for the generation of

useful energy from waste heat utilization. They modeled a trigeneration system with SOFC, absorption refrigeration and heat exchanger under exergo-economic principles. The results showed that an increase in current density increases the unit cost of electricity. The main exergy destruction occurred in the heat exchanger, SOFC, and combustion chamber. A maximum unit cost of product was obtained as 34.2 \$/GJ and a minimum of 26.5 \$/GJ. High thermodynamic performance and reduced CO₂ emissions were reported by Yilmaz et al.[109] as characteristics of biogas multigeneration systems. Based on their findings, a 400 °C increase in the GT inlet temperature resulted in a mass flow rate of 0.04 kg/s in hydrogen production and an electricity generation of 15 MW from an initial 7.5 MW. Georgousopoulos et al.[110] studied the performance of ORC configurations under three heat scenarios of air separation unit, carbon capture, and syngas cooling unit using pure refrigerant working fluids, and zeotropic mixtures. Under various conditions of ORC working fluids, they demonstrated the thermodynamic and techno-economic evaluation of a waste heat recovery ORC in a biomass-integrated gasification combined cycle. The syngas cooling scenario yielded the highest plant efficiency improvement of 2.86 percent, with the best economic performance of 35.42 – 35.67 €/MWh LCOE and a payback period of 5.7 – 5.8 years. Furthermore, the economic advantage of zeotropic mixtures over pure fluids was negligible. Mosaffa et al.[111] assessed a solar-based biogas-steam reformer with hydrogen and methanol production under thermo-economic principles. The gas turbine cycle and reformer possessed the largest exergy destruction rates of 140 MW and 134 MW respectively. In another study, Yunus et al.[112] considered a biomass-based integrated plant with products of power, hydrogen, hot water and fresh water. Syngas from biomass gasification was used to fuel the power plant integrated with a combined ORC-ejector refrigeration unit. Energy and exergy efficiencies of the plant were 56.17 % and 52.83 %, respectively. In order to assess the thermodynamic and electrochemical properties, Pirkandi et al.[113] studied the performance of nine configurations of the steam cycle portion of the hybrid SOFC-GT-ST system. Their findings indicate that utilizing ST in the bottoming cycle increased power generation by 200 % and 15 % for GT and SOFC-GT cycles, respectively. Habibollahzade et al. [114] proposed four multigeneration systems based on integrating syngaspowered SOFC/GT system for electricity, cooling, heating, and fuel production from CO₂ capture. They reported that the current density of the SOFC is a highly influential parameter as it led to a 7 % increase on exergy efficiency when adjusted. Furthermore, air compressor pressure ratio was effective in the increase of fuel production rate, and a 5 % reduction of CO2 emissions was attained by CO₂ recycling ratio.
2.6 Advanced exergo-economic assessment

The essence of an advanced exergy-based analyses in the design of energy-conversion systems is to provide further details to engineers to better understand and improve the design. Such analysis can significantly help overcome the limitations of a conventional analysis revealing the actual potential for a system component improvement and the level of interactions between components of the system[115].

These limitations are attributed to technological constraints resulting from either the material costs or availability and manufacturing methods. Hence, an enhanced exergy analysis involves splitting the exergy destruction of each component of the plant into unavoidable and avoidable parts, and in further cases, endogenous and exogenous parts. This approach entails limiting the exergy efficiency to a maximum value for any component irrespective of the investment cost and the efficiency is obtained at the point where the investment cost becomes extremely large, or mathematically infinite[116]. The unavoidable exergy destruction with respect to the product is obtained at this point. In order to obtain maximum efficiency, key thermodynamic parameters of the component are appropriately selected based on practical terms. Also, the unavoidable investment cost is determined under inefficient conditions that are practically unattainable due to the high cost of fuel associated that can result in inefficient solutions.

These approaches, which can also be combined provide additional, and valuable information for improving the overall performance of the plant by revealing where thermodynamic inefficiencies can be limited[117]. Several studies have shown the application of advanced exergy analyses to various energy conversion systems. You *et al.*[118] presented an assessment of a combined power, heating, and refrigerating system with multi-effect desalination (MED) based on an SOFC-micro gas turbine plant. Through the application of conventional and advanced exergoeconomic analysis, the conventional exergy analysis indicated the largest sources of irreversibility in the system were the after-burner, SOFC and MED, with contributions of 20.08 %, 12.99 %, 12.91 %, respectively. In the advanced analysis, results revealed the inverter, MGT and air compressor had the most potentials to minimize exergy destructions. Also, the enhanced exergy efficiency of each component in the advanced analysis was higher than the conventional analysis due to the reduction in the exergy losses. The low exergo-economic factors of the after-burner, and Preheaters from the conventional and advanced, showed these components as key components to be optimized for reduction in the exergy destruction cost rates.

Li *et al.*[119] identified exergy-based analysis as a great method for understanding and improving energy conversion processes. They performed an advanced exergy analysis on an ash agglomerating fluidized bed gasification process. With a total exergy destruction of 4.67 MW. the analysis indicated that 54.18% of this value was avoidable. The gasifier had the largest potential for reducing exergy losses and with gasification temperature range of 1173 - 1273 K, the total exergy destruction and avoidable exergy destruction decreases. Moharamain *et al.*[120] proposed an integrated co-fired combined power plant with biomass gasification and natural gas with hydrogen production. The product hydrogen was re-introduced into the combustion chamber to increase the performance efficiency and reduce fossil fuel consumption. Results from the enhanced exergy and exergo-economic analyses identified that hydrogen injection resulted in 67 % lower fossil fuel consumption and 19 % lower CO₂ emissions. Thermal and exergy efficiencies were reduced by 37 % and 39 %, respectively. The total exergy destruction was largely affected by the hydrogen injection, as the exergy destruction cost as a result of hydrogen injection into the combustion chamber of the co-fired combined cycle. The total unit product cost for the combined cycle with hydrogen injection was 18 % lower than when hydrogen injection was not available.

Soltani *et al.*[121] reported an enhanced exergy analysis of an externally-fired power plant coupled with biomass gasification. Their results indicated a weak interaction between plant components as the endogenous exergy destruction was higher compared to the exogenous value within each plant component. Also, due to the higher unavoidable exergy destruction in the components, performance improvement was highly limited. The heat exchanger, the steam turbine, the combustion chamber and the condenser were identified as components with merit priority for modifications from the advanced exergy analysis. However, only the gasifier and the combustion chamber were identified as priority components by the conventional exergy analysis.

Hamedi *et al.*[122] applied the advanced exergy analysis to determine major contributors to the energy-intensiveness of a light olefins plant in south Iran. Results from the plant simulation in ASPEN HYSYS software indicated that only 15.36% of the exergy destruction is avoidable in the cracking furnace despite its contribution of 55.3% of the total exergy destruction of the plant. Further results showed that consideration of prespecified costs and environmental impacts for energy streams or internal material resulting from the conversion process yields significant error in identifying the improvement priorities with respect to economic and environmental aspects. Liao et al.[123] proposed alternative ORC combined systems of Simple Supercritical Carbon dioxide ORC (RS-

sCO₂-ORC) for moderate-to-low temperature waste heat recovery from flue gas using energy and exergy analysis. Advanced exergy analysis involving the splitting of exergy destruction into avoidable and unavoidable portions was applied to obtain detailed information on plant components inefficiency and the potential for system improvement. The results showed an optimal compression ratio of 1.8 and 2.2 for the S-sCO2-ORC and RSsCO2-ORC systems, respectively. Based on the advanced exergy analysis, modifications of the turbine and condenser could improve efficiency of the ORC system the endogenous exergy rate is higher than exogenous exergy in all the system components. Muhammad et al.[124] demonstrated the effectiveness of the advanced exergy analysis on an energy conversion system to maximize the energy and exergy savings. They designed and optimized a CO₂ capture/liquefaction system using a heat pump with modifications of R717 and R290. The conventional exergy analysis indicated 43.76 % of input exergy is destroyed. However, the advanced exergy analysis revealed an avoidable exergy destruction of 48.85% and 51.20% of the total exergy destruction for R290 and R717 modifications, respectively. Also, the optimized electrical power consumed by the proposed system through advanced exergy analysis is 15.5 % lower than that of the initial system. Idrissa and Boulama[125] carried out advanced exergy analysis on a combined Brayton power cycle. The combustion chamber was identified as the component with the largest exergy destruction and also a large portion of the irreversibility was unavoidable. The irreversibility in the turbines and both compressors were mainly avoidable. Sensitivity analysis indicated a low-pressure ratio and combustion temperature of 1000 K – 1600 K significantly improved the overall plant performance.

Ustaoglu *et al.*[126] evaluated the performance of an ORC-based bio-waste-powered cogeneration system using conventional and advanced exergy approaches. They considered eight working fluids including wet (methanol, water) and dry (isopentane, n-pentane, n-octane, n-heptane), isentropic (R141b, R123). Maximum power generation, process heat, thermal and exergy efficiencies of 0.1213 MW, 0.9182 MW, 81.36 %, and 40.76 %, respectively, were obtained with R141b as working fluid the ORC-CHP configuration. The conventional exergy analysis identified the turbine, evaporator, and process heater as the most critical components. However, the advanced exergy analysis indicated only 14.5 % of the total exergy destruction, can be recovered for system improvements.

Song *et al.*[127] investigated the improvement potential and limitation of the improvability of a SOFC system combined with a kinetic-based modeling pre-reformer. The plant performance was based on conventional and advanced exergy analyses. The stack was identified by the conventional

exergy analysis as the major contributor to the exergy destructions with the exergy destructions of component stack 0.40 MW. Whereas the enhanced analysis showed that 48.39% of the total exergy destruction is avoidable from the stack and the heat exchanger having the largest potential for improvement as a result of the large avoidable endogenous exergy destructions. Oyekale *et al.*[128] applied the modified exergy approach to a 0.63 MW hybrid solar-biomass organic Rankine cycle (ORC) cogeneration plant. They investigated sources of thermo-economic setbacks in the plant and best approach to improving the configuration. The results indicated over 50 % of the total irreversibility rates can be prevented in nearly all of the components of the plant. Through the optimization of internal operations of the plant components, above 60 % of irreversibility cost rates can be avoided.

2.6 **Optimization**

Pressures on engineering designs have increased as a result of rising global competition and the drive for better, more effective processes. This coupled with growing reservations about the environment, safety, and other considerations are driving interest in manufacturing higher-quality goods at low cost[129]. From the standpoint of society, creating a system that only completes the stated objective is no longer sufficient. Hence it is important to optimize processes such that one or more quantity of interest is minimized or maximized, thereby improving the design[130]. Optimization is a significant technique in energy engineering in the determination of the optimal performance of a system [76]. Questions bothering energy, exergy, economic, and environmental issues can be answered simultaneously through the formulation of an appropriate optimization model for a system[131]. In formulating the model, first a system boundary is defined where all subsystems that affects the performance of the system are included. This is followed by defining the quantity of interest, also referred to as objective function, which can be based on efficiencies, levelized costs, costs associated with exergy destruction, and environmental impact of the plant[132]. Multi-objective optimization is the consideration of more than one objective function to obtain the optimal solution [132], [133]. Elements relevant to the formulation of the optimization problem are decision variables that characterize the design options adequately. It is expected that such variables include essential variables that could affect the plant performance, as well as the cost-effectiveness. In thermal systems, many decision variables are constrained due to operating principles or conservation laws, thereby restricting the values employed in searching for an optimum[83], [75]. Optimization methods can be categorized as classical for unconstrained minimum or maximum of differentiable and continuous functions. Classical optimization involves the application of differential calculus to find the optimum solution. Numerical optimization methods are mathematical optimization techniques specifically defined towards the objective. This includes linear, non-linear, and quadratic programming, stochastic, dynamic, combinatorial, integer programming, and evolutionary algorithm (fuzzy logic, artificial neural network, and genetic)[134]. Genetic algorithm utilizes a search method based on evolutionary processes in biology including reproduction, selection and mutation[135]. This method can as well be adapted for several objective functions with competing objectives i.e. multi-objective genetic algorithm (MOGA). Unlike traditional genetic algorithms, which only have one goal, MOGA uses several objective functions to calculate a population's fitness. Finding a set of optimal solutions that satisfactorily strike a balance between the various objectives is the aim of MOGA. To achieve this, a set of solutions known as the Pareto front is generated, covering the whole space of generated solutions[134]. The optimal trade-off between the different goals is represented by the Pareto front. It is impossible to improve one Pareto front solution without also improving the other [132]. The different objectives in MOGA are used to evaluate population members, and the most adept members are selected based on their ability to balance conflicting objectives. The selected individuals are then recombined through crossover and mutation operations to form a new population. Before the Pareto front is reached, this process needs to go through several generations. Every point on the Pareto front represents an optimal solution for the combined optimization. The selection of a specific option from the pareto front is achieved through a decision method meant to balance the relative importance of the objectives [136]. Such methods include linear programming for multidimensional analysis of preference (LINMAP), and technique for order of preference by similarity to ideal solution (TOPSIS)[137], [138].

The application of the aforementioned optimization methods to biofuel multigeneration plants has been reported in open literature. Taheri *et al.*[139] carried out a multi-objective optimization to produce power, cooling, natural gas, and hydrogen from a biomass-based multigeneration plant. The plant was made up of an absorption refrigeration unit, a gasifier, a combined gas and steam turbine, a PEME, cascading Rankine cycles, and a liquid natural gas subsystem. The optimization results using the genetic algorithm yielded an exergy efficiency of 39.023 % and a total product cost rate of 1107 \$/hr. Ahmadi *et al.*[140] performed a thermo-economic analysis and multi-objective optimization of a biomass-fueled plant integrating a biomass combustion, ORC, double-

effect absorption chiller, heat exchanger, PEME for hydrogen production, water heater for hot water and desalination system. The total cost rate and exergy efficiencies were optimized using non-dominated sorting genetic algorithm. Al-Rashed and Alfrand[141] optimized a combined GT with supercritical CO₂ system driven by biogas generated from anaerobic digestion. A multicriteria optimization of the system produced a 24.6 % improvement of the total product cost for configuration with inlet cooling of CO₂ against the configuration without inlet cooling. Detchusananard *et al.*[142] carried out a thermo-economic analysis of a wood biomass gasification with solid oxide fuel cell, for production of power and heat. Multi-objective optimization approach was applied in improving the plant performance using robust Pareto solutions. The integrated plant showed improved heat integration with maximum exergy efficiency of 61.2 % and minimized levelized cost of electricity.

Rashidi and Khorshidi[133] presented an optimization scheme of a biomass gasification based multigeneration plant for cooling, power generation and desalination. They employed evolutionary algorithm methods in optimizing the total cost rate and exergy efficiency of the plant. Balafkandeh et al.[143] conducted a multi-objective optimization on a biomass-fueled multigeneration plant integrated with an absorption refrigeration system, a gas turbine, and a supercritical CO₂ unit. The system performed better, achieving 47.8 % more energy efficiency and a unit product cost of 5.436 \$/GJ using the biogas fuel from digestion compared to syngas from gasification. According to Khalid *et al.* [16] optimization can help overcome the challenges of individual systems. They considered a solar-biomass hybrid multigeneration system with levelized cost of electricity obtained as 117 \$/MWh and energy and exergy efficiencies of 91.0 % and 34.9 %, respectively. The system comprised of two ORC configurations, solar unit, gas turbine, and vapour absorption refrigeration, and gas turbine. Furthermore, Ahmadi et al.[144] applied genetic algorithm to obtain optimum performance of a 39 MW multigeneration plant with SOFC-GT and thermal desalination system with multiple-effects. From the exergo-economic analysis, they established that with high current density in SOFC, the exergy efficiency increased to the maximum limit and accordingly the LCOE showed reverse behavior. Furthermore, increase of the operation temperature of SOFC, the exergy efficiency increases and LCOE reduces. Optimization of the system using genetic algorithms revealed that at optimal point, the exergy efficiency is 63.5% and LCOE is 6.43 cent\$/kWh. Sadeghi et al.[145] optimized a trigeneration plant from thermo-economic view point using genetic algorithm. The optimization focused on improving the exergy efficiency and cost of product and under optimal points the exergy efficiency and total product cost were obtained as 48.25 % and 25.94 % respectively. Mehrabadi and Boyaghchi [78] studied an integrated biomass system consisting of a gasifier, SOFC-GT unit, thermoelectric generator, MSF-BR unit, and CO₂ capture unit proposed for power and freshwater production. NSGA II was applied in the optimization of four objective functions of energy and exergy efficiencies, and total cost and environmental impact rates of the plant. They reported optimal energy and exergy values of 48.93% and 45.2 %, respectively, with low total cost of 2.39\$/GJ, and minimum levelized CO₂ emission of 0.138 ton/MWh. Behzadi et al. [146] considered a biomass-based solid oxide fuel cell for power generation, cooling, and fresh water production that is connected with a gas turbine, a reverse osmosis desalination unit, and a double-effect absorption chiller. Gasification of MSW was utilized in the biomass conversion process. A multi-objective optimization of the proposed plant resulted in exergy efficiency and total product unit cost of 38.16% and 69.47 \$/GJ, respectively. According to scatter distribution of decision variables, the gas turbine pressure ratio, stack temperature differential, and ratio of CO₂ recycling were the most sensitive characteristics and should be minimized for optimum plant performance. Guo et al.[147] applied genetic algorithm and TOPSIS to improve the efficiencies and economic costs of a combined SOFC-GT system as an alternative for distributed power and heat generation. Karimi et al.[3] examined a gasifier-SOFC-GT-ORC plant fuelled by rice straw as feedstock for a biomass-based heat and power plant. They reported the SOFC current density had the largest impact on the efficiency and cost rate of the plant. Through multi-objective optimization, an optimum exergy efficiency of 35.1 %, cost rate of 10.2 \$/hr, and high irreversibility in the gasifier were reported. Nazari et al.[148] proposed a biomass-based power system combined with cooling and heating in an externally fired gas turbine, an ORC cycle, an absorption chiller, and a solar pre-heater. In the proposed configuration, solar thermal energy was used to preheat the process air for the combustion process and before heating to specified turbine inlet condition by the flue gases from bagasse fuel. The base case evaluation with pressure ratio and turbine inlet temperature of 4.84 and 875 °C, respectively, showed that the proposed plant can obtain energy and exergy efficiencies of 55.56 % and 20.38%, with product cost rate is 26.4 \$/hr. Also, heating, and cooling rate capacity of were 187.6 kW, 202.8 kW, and 47.48 kW, respectively. A multi-objective multi-verse optimization algorithm was utilized to plot the Pareto-frontier of the objective function. The optimum solution indicated that the studied configuration could attain exergy efficiency of 22.20 %, with a product cost rate is 24.86 \$/hr. Mamaghani et al. [149] suggested an MCFC-GT-ORC integrated plant for power generation under energy, exergy, economic and environmental analysis. They performed a MOGA optimization to

improve the plants' exergy efficiency and minimize the total cost rate of the integrated plant. From their optimization results, increase in interest rate, corresponds with high exergy efficiency and total cost rate; and an increase of 25 % in the unit cost of fuel raises the total cost rate by 3% in the low exergy efficiency zone. An evolutionary based method was presented by Toffolo and Lazzaretto[150] in the design and optimization of the exergy and economic parameters of a thermal system. The multi-objective evolutionary algorithm (MOEA) revealed best trade-off solutions of the two competing objectives for the CGAM cogeneration configuration. Fakhari *et al.*[151] investigated the performance and cost-effectiveness of a biomass plant with gasifier, PEM fuel cell, multi-effect desalination unit and a series two-stage ORC. Tri-objective optimization using genetic algorithm was performed with different zeotropic mixtures in the ORC revealing an optimal exergy efficiency of 23.43 % and minimum total cost rate of 64.91 \$/hr. They reported R601a-C2Butene showed superiority from thermodynamic and economic views. The scatter distribution of the decision variables revealed that with optimal values in a distributed domain, moisture content and current density are not a sensible variable.

From the review of literature, the application of optimization techniques is essential in the design and modelling of energy systems as improvements in the systems performance indicators, as well as other areas of possible concern can be revealed. Hence in this study, a MOGA is adopted in other to optimize the integrated power plant.

2.7 Summary

Although, there is rising interest in the utilization of integrated plants for the generation of power, and other products, when possible, especially renewable fueled designs, there is the need to ensure that these designs meet required demands in terms of cost, efficiencies, and sustainable development goals. Therefore, it is very imperative for investors and policy makers to have suitable information on the performance and cost implications of such integrated systems. The literature shows that biomass fueled systems are highly promoted and important in power generation, and other products such as heating, cooling, clean water and hydrogen production. Under thermo-economic principles, a hybrid biomass conversion system that combines gasification, methanation, and anaerobic digestion has not yet been studied. Furthermore, consumption of hydrogen production for future use. Considering the amount of heat available during the cooling of syngas from gasification,

the integration of an ORC to the biomass conversion unit is also an area open for investigation. Hence, several ORC working fluids are studied to attain optimum performance of the configuration. In other to utilize the energy of biofuel from the biomass conversion process for power generation, a SOFC-GT-ST-ORC integrated plant was introduced. Optimization using MOGA to address minimization of the levelized cost and maximize exergy efficiency of the power plant will be considered.

3 Methodology

3.1 Preamble

The mathematical modeling of the thermodynamic, economic, environmental, and exergoeconomic performance characteristics of the proposed biomass integrated plant configuration, is presented in this chapter. Operating data from reviewed literature was used to validate the mathematical models. The plant is located in the tropic region of Nigeria on lat. 4.45 °N and long. 6.5 °E, however, the technology can be applied across regions with similar environmental conditions.

3.2 Plant description

The biomass conversion process adopted in this study highlights biomass conversion to power through biochemical and thermochemical conversion with syngas upgrade. This is summarized in Figure 3.1. The upgraded syngas is further used in the plant generate power and improve the power plant output.



Figure 3.1: Integrated biomass conversion system for power generation

The schematic diagram of the proposed plant is shown in Figure 3.1. The plant is made of a hybrid biomass conversion unit and an integrated power generation unit. Gasification and anaerobic digestion are employed for feedstock conversion to gaseous fuel in the hybrid biomass conversion unit. Livestock waste of high moisture content is fed to the digester and undergoes a sequence of processes aided by microorganisms to produce biogas. Feedstocks of crop waste are converted to synthesis gas through the gasifier at high temperatures. This syngas is cooled in a syngas heat exchanger (HXS), cleaned up, and fed to a methanation unit (MTH) for an upgrade. Steam recovered from the syngas cooling is converted to power through an organic Rankine cycle configuration (ORC I). A proton exchange membrane electrolyzer (PEME) is employed to produce hydrogen for the methanation process, where the final product is upgraded synthetic natural gas. Clean biogas from the process of digestion is supplied to an integrated solid oxide fuel cell (SOFC) and gas turbine (GT) top cycle to produce power. A portion of the power produced by the electrolysis of the air and gaseous fuel mixture (streams 29 and 33) in the SOFC is sent to the PEME. In a combustion chamber (CC), extra fuel from the methanation process and leftover fuel from the SOFC process are burned. Utilizing the high-temperature gas produced during combustion, the GT expands to produce power. The high-temperature gas stream (36) that exits the GT moves through the preheaters of fuel and air (HX1 and HX2), enhancing the temperature of the feeds to the SOFC through streams 29 and 33. The flue gas that leaves the topping cycle (stream 38) is used in a steam turbine (ST) plant to generate additional power. Owing to the high thermal energy of the flue gas exiting the heat recovery steam generator (HRSG) through stream 39, it is fed to an organic Rankine cycle (ORC II) unit where it is converted to power. The net power output is a sum total of power obtained from the power cycles in the configuration.

3.3 Assumptions

The following general assumptions were made in modelling the integrated plant [141], [152]:

- i. Operating conditions are under steady state and steady flow throughout the system.
- ii. Ambient pressure of 101.325 kPa, and temperature of 298.15 K.
- iii. Negligible changes are assumed for the kinetic, potential energy, and exergy.
- iv. 21 % oxygen and 79 % nitrogen make up the composition of air.



Figure 3.2: Schematic diagram of the combined biomass conversion unit and power sections in the integrated biofueled power plant

- v. For the gasification unit, a downdraft gasifier is assumed.
- vi. The plant assumes that every gas stream is under ideal gas condition.
- vii. Minimal heat losses during the gasification and methanation processes, as well as in the SOFC-GT system.
- viii. A maximum operating pressure of $0.9 P_{cr}$ for ORC.

3.4 Thermodynamic modeling

Thermodynamic modeling is primarily based on the first and second laws of thermodynamics, and employs the use of mass, energy, and exergy balance equations as stated below[148], [29]:

Mass balance:
$$\sum \dot{m}_{in} = \sum \dot{m}_{out}$$
 (3.1)

Energy balance:
$$\sum \dot{m}_{in}h_{in} + \dot{Q} = \sum \dot{m}_{out}h_{out} + \dot{W}$$
 (3.2)

Exergy balance:
$$\sum \dot{E} x_{in,k} = \sum \dot{E} x_{out,k} + \dot{E} x_{D,k}$$
 (3.3)

where $\dot{m}(kg/s)$ represents the mass flow rate, h(kJ/kg) is the specific enthalpy of the stream, $\dot{Q}(kW)$ and $\dot{W}(kW)$ are the heat and power transfer rates, respectively. $\dot{E}x$ represents the exergy function which is evaluated as a product of mass flow rate and specific exergy. Neglecting elevation and velocity variation which are relevant to the kinetic and potential exergy components, specific exergy is expressed in terms of chemical and physical exergy as [100]

$$e = e^{ch} + e^{ph} \tag{3.4}$$

where e^{ch} and e^{ph} of the *i*-th composition are evaluated as

$$e_i^{ch} = \sum_{i=1}^n y_i e_{0,i}^{ch} + RT_0 \sum_{i=1}^n y_i \ln(y_i)$$
(3.5)

$$e^{ph} = (h_k - h_0) - T_0(s_k - s_0)$$
(3.6)

where $e_{0,i}^{ch}$ and y_i are the standard chemical exergy and mole fraction of the *i*-th species; h_k, h_0, T_0, s_k, s_0 are the specific enthalpy, temperature and specific entropy of the stream and environment, respectively, for the *k*-th component.

3.4.1 Gasifier

To produce premium synthesis gas, the biomass downdraft gasifier is run at temperatures as high as 1023 K [21]. With air as gasifying agent, the general gasification reaction in the gasifier is given as[153], [154]:

$$CH_a O_b N_c S_d + w H_2 O + n_{Air} Air \rightarrow$$

$$n_{CO}CO + n_{CO_2}CO_2 + n_{CH_4}CH_4 + n_{H_2O}H_2O + n_{H_2}H_2 + n_{H_2S}H_2S + n_{N_2}N_2$$
(3.7)

where $CH_aO_bN_cS_d$, w, and n_i are the chemical formula of the crop waste in terms of carbon, moisture content and amount in mole of the *i*-th species specified in the equation. The gasification equation was evaluated using the recommended methodologies by Athari *et al.*[153]. The moisture content was evaluated as

$$w = \frac{\bar{M}_{bio} \times MC}{\bar{M}_{H_2O}(1 - 0.01MC)}$$
(3.8)

The values of n_i are determined by applying mass and energy balances to the entire reaction in Equation (3.7) and with respect to the equilibrium constants for the shift reaction and methane formation reaction presented in Equations (3.9) and (3.10), respectively.

$$K_1: CO + H_2O \leftrightarrow H_2 + CO_2 \tag{3.9}$$

$$K_2: \mathcal{C} + 2H_2 \quad \leftrightarrow \quad \mathcal{C}H_4 \tag{3.10}$$

The shift process and methane formation equilibrium constants are written as follows:

$$K_1 = \frac{n_{H_2} n_{CO_2}}{n_{CO} n_{H_2O}} \tag{3.11}$$

$$K_2 = \frac{n_{CH_4}}{{n_{H_2}}^2} \tag{3.12}$$

The equilibrium constants, K_1 and K_2 are expressed in terms of Gibbs functions for the shift reaction and methane formation, respectively, as [46]

$$-\frac{\Delta g_i^0}{\bar{R} T_3} = \ln K_i; \ (i = 1, 2)$$
(3.13)

$$\Delta g_i^0 = (h - Ts)_{product} - (h - Ts)_{reactant}$$
(3.14)

A chemical balance for the component of the syngas is obtained from the generic gasification equation along with the two equilibrium constant equations to solve for the specific molar amount of the syngas components.

Carbon:	$n_{CO} + n_{CO_2} + n_{CH_A} =$		(3.1	5)	
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Hydrogen: n	$n_{CH_A} + n_{H_2O} + n_{H_2} + n_{H_2}$	$n_{H_2S} = a + 2w$	(3.16)
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Oxygen:
$$n_{CO} + n_{CO_2} + n_{H_2O} = b + w + n_{Air}$$
 (3.17*a*)

- Nitrogen: $n_{N_2} = c + n_{Air}$ (3.17b)
- Sulphur: $n_{H_2S} = d$ (3.17c)

Energy balance for the gasification reaction, under no heat loss at gasification temperature is given as:

$$\bar{h}_{f,bio}^{0} + w\bar{h}_{f,H_{2}O}^{0} = n_{CO}(\bar{h}_{f,CO}^{0} + \Delta\bar{h}_{CO}) + n_{CO_{2}}(\bar{h}_{f,CO_{2}}^{0} + \Delta\bar{h}_{CO_{2}}) + n_{CH_{4}}(\bar{h}_{f,CH_{4}}^{0} + \Delta\bar{h}_{CH_{4}}) + n_{H_{2}O}(\bar{h}_{f,H_{2}O}^{0} + \Delta\bar{h}_{H_{2}O}) + n_{H_{2}}(\bar{h}_{f,H_{2}}^{0} + \Delta\bar{h}_{H_{2}}) + n_{H_{2}S}(\bar{h}_{f,H_{2}S}^{0} + \Delta\bar{h}_{H_{2}S}) + n_{N_{2}}(\bar{h}_{f,N_{2}}^{0} + \Delta\bar{h}_{N_{2}})$$

$$(3.18)$$

where $\Delta \bar{h}$ is the specific enthalpy difference between gasification temperature and reference temperature, and $\bar{h}_{f,bio}^0$ is the enthalpy of formation of the biomass obtained from its heating value.

3.4.2 Digester

The anaerobic digestion process of the animal waste is expressed in a chemical reaction as

 $C_x H_a O_b N_c S_d + n_{H_2O} H_2 O \rightarrow n_{CO_2} CO_2 + n_{CH_4} CH_4 + n_{H_2S} H_2 S + n_{NH_4} NH_4$ (3.19) where $C_x H_a O_b N_c S_d$ represents the animal waste formula. *x*, *a*, *b*, *c*, *d* are known number of atoms of *C*, *H*, *O*, *N*, and *S*, well-defined using waste ultimate analysis as in Table 3.5. The modified Buswell approach according to Ogorure *et al.*[21] is employed in calculating approximately the composition of biogas in the waste.

3.4.3 Methanation

Carbon dioxide and carbon monoxide from the gasification process are converted to synthetic natural gas (SNG) in the methanation reactor by means of two exothermic reactions with hydrogen gas[52]:

Sabatier/reforming reaction:	$CO_2 + 4H_2 \leftrightarrow CH_4 + 2H_2O$	(3.20)

CO methanation:
$$CO + 3H_2 \leftrightarrow CH_4 + H_2O$$
 (3.21)

Using reactants from the gasification product, the molar composition of the SNG is calculated using the Gibbs free energy of chemical equilibrium method. The calculation of thermodynamic equilibrium in chemical reactions, using Gibbs free energy minimization, sheds light on the yield and selectivity of thermodynamically stable products. This depends on influencing factors, and kinetic obstacles can be found by contrasting the findings of experiments. The equation of the total Gibbs free energy is written as [69], [61]:

$$G = \sum_{i=1}^{M} n_i \mu_i = \sum_{i=1}^{M} n_i \mu_i^o + RT \sum n_i \ln\left(\frac{y_i P}{P_0}\right)$$
(3.22)

Lagrange multipliers λ_i are introduced for each species *i* under mass balance constraints, such that

$$\mu_i + \sum_{j=1}^{K} \lambda_j n_{ij} = 0 \tag{3.23}$$

Combining both equations yields

$$\mu_i^o + RT \sum_{j=1}^K \lambda_j n_{ij} = 0$$
(3.24)

where μ , is the chemical potential. Equation (3.24) is written with respect to the unknown species to set up a system of linear equations. These equations are solved simultaneously until an approximation of the minimum free energy converges. At the point of convergence, molar composition of the products of the methanation process are obtained[61].

The low heating value (LHV) per kg of the gas from the digestion, gasification and methanation processes is obtained with respect to the mass fraction of the *i*-th component of the product gas:

$$LHV_{fuel} = \sum y_i LHV_i \tag{3.25}$$

3.4.4 Proton exchange membrane electrolyzer

The Proton exchange membrane electrolyzer (PEME) splits water into hydrogen and oxygen through electrical power, and its operation is like a fuel cell operating in opposite manner. The main reactions(anode, cathode, and overall) involved in the production of hydrogen in the PEME are obtained as [96]:

Anode:
$$H_2 O_{(liq)} \to 2H^+ + 2e^- + \frac{1}{2}O_{2(gas)}$$
 (3.26)

Cathode: $2H^+ + 2e^- \rightarrow H_{2(gas)}$ (3.27)

Overall:
$$H_2 O_{(liq)} + \dot{W}_{PEM} \to H_{2(gas)} + \frac{1}{2} O_{2(gas)}$$
 (3.28)

Additionally, the molar rate balance of each electrolyzer cell for the reactions of water, hydrogen, oxygen, are evaluated as

$$\dot{n}_{H_2,22} = \frac{N_E j_E A_E}{2F}$$
(3.29)

$$\dot{n}_{O_2,21} = \frac{1}{2} \dot{n}_{H_2,22} \tag{3.30}$$

$$\dot{n}_{H_20,20} = 2\dot{n}_{H_20,rxt} + \dot{n}_{O_2,21} + \dot{n}_{H_20,21}$$
(3.31)

$$\dot{n}_{H_20,rxt} = \dot{n}_{H_2,22} \tag{3.32}$$

where \dot{n}_i for the *i*-th specie in Equations (3.29 – 3.32) is the molar conversion rate, j_E , A_E , N_E , are the current density, active surface area, and number of cells of the PEME.

With the electrical power as an input parameter, coupled with its relationship with N_E , j_E , A_E , the amount of hydrogen required to satisfy the methanation reaction can be obtained theoretically. The input power is obtained as

$$\dot{W}_{PEM} = N_E I_E V_E \tag{3.33}$$

where I_E , and V_E are the current and net voltage of the electrolyzer, respectively, and are given as

$$I_E = j_E A_E \tag{3.34}$$

$$V_E = E_{Nernst,E} + V_{act} + V_{Ohm} + V_{conc}$$
(3.35)

where $E_{Nernst,E}$ is minimum voltage applied across the PEME. V_{act} , V_{Ohm} , and V_{conc} , are the activation loss, ohmic loss and concentration loss calculated according to the methods of Saeed and Warkozek[59]. These are over potential (voltage losses) that must be compensated by increasing the voltage value for the reactions to occur and are evaluated as follows:

$$E_{Nernst,E} = 1.229 - 8.5 \times 10^{-4} (T_E - T_o)$$
(3.36)

The activation loss, V_{act} which occurs in the anode and cathode, is a result of the slow reaction rate on the electrode surface and can be obtained as a summation of the anode and cathode components using the Butler–Volmer equation[155]:

$$V_{Act,a} = \frac{\bar{R}T}{F} \sinh^{-1}\left(\frac{j_E}{2j_{o,a}}\right); \ j_{o,a} = j_{o,a}^{ref} \exp\left(\frac{-E_{Act,a}}{\bar{R}T_E}\right)$$
(3.37)

$$V_{Act,c} = \frac{\bar{R}T_E}{F} \sinh^{-1}\left(\frac{j_E}{2j_{o,c}}\right); \ j_{o,c} = j_{o,c}^{ref} \exp\left(\frac{-E_{Act,c}}{\bar{R}}\right)$$
(3.38)

where j_E , and j_o are the limiting current density and exchange current densities of the electrodes. The ohmic loss which results from a resistance to the flow of either electrons or ions through the material of the electrodes/electrolyte and the various interconnections is given as [57]:

$$V_{Ohm} = j_E R_E \tag{3.39}$$

where R_E is the ohmic resistance which depends on the thickness, and conductivity of the electrolyzer membrane, $\sigma[\lambda(x)]$, and can be calculated as [96];

$$R_E = \int_0^1 \frac{dx}{\sigma[\lambda(x)]}$$
(3.40)

$$\lambda(x) = \frac{\lambda_a - \lambda_c}{L} x + \lambda_c \tag{3.41}$$

$$\sigma[\lambda(x)] = 0.5139 - 0.326 \exp\left[1268\left(\frac{1}{303} - \frac{1}{T_E}\right)\right]$$
(3.42)

The concentration loss occurs due to fluctuations in the concentration of hydrogen at the electrodes surfaces during the reaction. In Equation (3.43), j_l is the limiting current density of the electrolyzer.

$$V_{conc} = \frac{\bar{R}T_E}{\alpha 2F} \left(1 - \frac{j_E}{j_l}\right) \tag{3.43}$$

The efficiency of the electrolyzer based on the net voltage, and lower heating value of hydrogen are [59],[57]:

$$\eta_{E,V_E} = \frac{1.48}{V_E} \tag{3.44}$$

$$\eta_{E,LHV} = \frac{LHV_{H_2}}{\dot{W}_{PEM}} \tag{3.45}$$

3.4.5 Solid oxide fuel cell

The solid oxide fuel cell (SOFC) potential to use hydrogen and carbon monoxide as fuel at the same time is a crucial and distinctive quality. Methane and carbon monoxide are produced as a fuel mixture inside the fuel cell through direct internal reforming. The fuel cell model in this work is based on a tubular design. The key chemical reactions in the SOFC are[96]:

Reforming:
$$CH_4 + H_2 0 \rightarrow CO + 3H_2$$
 (\dot{x}) (3.46)

Shift:
$$CO + H_2O \to CO_2 + H_2$$
 (ý) (3.47)

Electrochemical:
$$H_2 + \frac{1}{2}O_2 \rightarrow H_2O$$
 (*ż*) (3.48)

where \dot{x} , y, \dot{z} , are the gas molar conversion rates for the reforming, shift, and electrochemical reactions, respectively. With the application of already defined air and fuel utilization factors, UF_{air} and UF_f , respectively, and the molar balance of the reacting gases, the mass balance, between the inlet and exit of the SOFC can be evaluated with respect to the molar conversion rates \dot{x} , \dot{y} and \dot{z} using the shift, electrochemical and reforming reactions as follows [96], [113]:

$$\dot{n}_{CH_4,30} = \dot{n}_{CH_4,29} - \dot{x}_{29} \tag{3.49}$$

$$\dot{n}_{H_2,30} = \dot{n}_{H_2,29} + 3x_{29} - \dot{y}_{29} - \dot{z}_{29} \tag{3.50}$$

$$\dot{n}_{C0,30} = \dot{x}_{29} - \dot{y}_{29} \tag{3.51}$$

$$\dot{n}_{CO_2,30} = \dot{n}_{CO_2,29} + \dot{y}_{29} \tag{3.52}$$

$$\dot{n}_{H_20,30} = \dot{n}_{H_20,29} - \dot{x}_{29} - \dot{y}_{29} + \dot{z}_{29} \tag{3.53}$$

$$K_{s} = \frac{\dot{n}_{CO_{2},30} \times \dot{n}_{H_{2},30}}{\dot{n}_{CO,30} \times \dot{n}_{H_{2},0,30}}$$
(3.54)

$$K_{r} = \frac{\dot{n}_{CO,30} \times \dot{n}_{H_{2},30}^{3}}{\left(\dot{n}_{CH_{4},30} \times \dot{n}_{H_{2}O,30}\right) \left(\frac{P_{29}}{\dot{n}_{29}}\right)^{2}}$$
(3.55)

$$z_{29} = UF_{fuel} (\dot{n}_{H_2,29} + 3\dot{x}_{29} + \dot{y}_{29})$$
(3.56)

$$\dot{n}_{O_2,34} = \dot{n}_{O_2,33} - \frac{\dot{z}_{29}}{2} \tag{3.57}$$

$$UF_{air} = \frac{\dot{z}_{29}}{2\dot{n}_{O_2,33}} \tag{3.58}$$

where K_s and K_r are equilibrium constants of shift and reforming reactions. Both parameters are evaluated in terms of Gibbs free energy. UF_{fuel} and UF_{air} are the fuel and air consumption factors, respectively. The direct current power generated by the SOFC is calculated as

$$\dot{W}_{FC,DC} = N_{FC} I_{FC} V_{FC} \tag{3.59}$$

$$I_{FC} = j_{FC} A_{FC} \tag{3.60}$$

$$\dot{W}_{FC,AC} = \eta_{in\nu} \dot{W}_{FC,DC} \tag{3.61}$$

$$\rho = \frac{\dot{W}_{FC,AC}}{A_{FC}} \tag{3.62}$$

where $\dot{W}_{FC,DC}$, $\dot{W}_{FC,AC}$, η_{inv} are the DC power, AC power, and inverter efficiency, respectively; N_{FC} is the number of cells and V_{FC} is the net voltage. Equation (3.62) is the SOFC power density which is a measure of how much electrode active area is needed for peak performance[112].

The open circuit voltage of the fuel cell is not achievable due to irreversibilities brought on by activation, concentration, and ohmic over potential; hence, these factors must be taken into consideration to determine the net voltage[96]. The net voltage is obtained as

$$V_{FC} = E_{Nernst,FC} - (V_{act} + V_{Ohm} + V_{conc})$$
(3.63)

where $E_{Nernst,FC}$, $V_{Act,a}$, $V_{Act,c}$, V_{Ohm} , V_{conc} , are the Nernst or open circuit voltage, activation, ohmic and concentration voltages, respectively. These parameters are evaluated as follows:

$$E_{Nernst,FC} = -\frac{\Delta g^o}{2F} + \frac{\bar{R}T_{FC}}{2F} \ln\left(\frac{a_{H_2,30} \cdot a_{O_2,34}^{\frac{1}{2}}}{a_{H_2,0,30}}\right)$$
(3.64)

where change in molar Gibbs free energy of is evaluated with respect to H_2O , H_2 and O_2 as

$$\Delta g^{o} = g^{o}_{H_{2}O} - g^{o}_{H_{2}} - \frac{1}{2}g^{o}_{O_{2}}$$
(3.65)

$$g_i^o = (h - T_{FC}.s^o)_i; \quad i = H_2 0, H_2, 0_2$$
(3.66)

$$a_i = \frac{y_i P_{30}}{P_o} \tag{3.67}$$

where a_i is the activity factor of species *i* in electrochemical reaction, P_{30} , and P_o are the cells operating pressure and atmospheric pressure, respectively.

The activation losses are evaluated in the same method as that of the electrolyzer in equations (3.37) and (3.38), whereas the ohmic loss, and concentration loss are obtained as[96]:

$$V_{Ohm} = j_{FC} \left(R_C + \sum \rho_i L_i \right) \tag{3.68}$$

where $\sum \rho_i L_i$ is expressed as

$$\rho_a = \left(9.5 \times \frac{10^7}{T_{FC}} \exp\left(-\frac{1150}{T_{FC}}\right)\right)^{-1}$$
(3.69)

$$\rho_c = \left(4.2 \times \frac{10^7}{T_{FC}} \exp\left(-\frac{1200}{T_{FC}}\right)\right)^{-1}$$
(3.70)

$$\rho_{el} = \left(3.34 \times 10^4 \exp\left(-\frac{10300}{T_{FC}}\right)\right)^{-1} \tag{3.71}$$

$$\rho_{int} = \left(9.3 \times \frac{10^6}{T_{FC}} \exp\left(-\frac{1150}{T_{FC}}\right)\right)^{-1}$$
(3.72)

where L, ρ_a , ρ_c , ρ_{el} , ρ_{int} , are the thickness of fuel cell layer, resistivities of anode cathode electrolyte and interconnect, respectively.

$$V_{conc} = V_{conc,a} + V_{conc,c} \tag{3.73}$$

$$V_{conc,a} = \frac{\bar{R}T_{FC}}{2F} \left(\ln\left(1 + \frac{P_{H_2}j_{FC}}{P_{H_2O}j_{a,s}}\right) - \ln\left(1 - \frac{j_{FC}}{j_{a,s}}\right) \right)$$
(3.74)

$$V_{conc,c} = -\left(\frac{\bar{R}T_{FC}}{4F}\ln\left(1 - \frac{j_{FC}}{j_{c,s}}\right)\right)$$
(3.75)

$$j_{a,s} = \frac{2F \cdot P_{H_2} \cdot D_{a,eff}}{\bar{R}T_{FC} \cdot L_a}$$
(3.76)

$$j_{c,s} = \frac{4F.P_{O_2}.D_{c,eff}}{\left(\left(\frac{P_{33} - P_{O_2,33}}{P_{33}}\right)\bar{R}T_{FC}.L_c\right)}$$
(3.77)

where $j_{a,s}$ (A/m²), $j_{c,s}$ (A/m²), $D_{a,eff}$ (m²/s), $D_{c,eff}$ (m²/s) are the limiting current densities and effective gaseous diffusivity through anode and cathode, respectively.

3.4.6 Organic Rankine cycle and choice of working fluid

In the selection of working fluids for ORC systems, two key considerations are achieving high performance and attaining optimum use of heat sources [110]. In order to accomplish these goals, the thermophysical characteristics of the working fluid as well as the quality of the heat source are essential. On the other hand, attention is also drawn to their effects on the environment, including flammability, toxicity, global warming potential (GWP), and ozone depletion potential (ODP) [35], [88]. As a result, six ORC working fluids as listed in Table 3.1which include siloxanes (hetamethyldisiloxane, MM, and octamethyltrisiloxane, MDM), hydrocarbons (cyclopentane and cyclohexane), and refrigerants (R1233zd(E) and R600a) are selected for this study and distributed among the three groups. These fluids are expected to maximize the performance of the ORC.

Table 3.1: Thermophysical, safety, and environmental properties of selected ORC working fluids [99], [88]

Parameter	MM	MDM	Cyclopentane	Cyclohexane	R1233zd(E)	R600a
T_{cr} (K)	518.75	564.09	511.72	553.64	439.6	408
P_{cr} (bar)	19.3	14.1	45.7	40.7	35.7	36.3
ODP	n.a	n.a	0	0	0.00024	0
GWP	n.a	n.a	0	low	7	4
ASHREA safety	n.a	n.a	n.a.	A3	Al	A3
group						

3.4.7 Heat exchangers

The energy balance across the heat exchangers with interaction of hot and cold streams due to local temperature difference is given as

$$\dot{Q}_{k} = [\dot{m}_{in}(h_{in} - h_{exit})]_{hot} = [\dot{m}_{in}(h_{exit} - h_{in})]_{cold}$$
(3.78)

where \dot{Q}_k , \dot{m}_{in} are the rate of heat transfer, mass flow rate of working fluid at inlet of the component. In the case of a regenerator, where there is heat recovery from the hot stream exiting the turbine to preheat the working fluid prior to entering an evaporator and improve plant efficiency, as in ORC I and ORC II, the enthalpies of the inlet and outlet streams of the regenerator are evaluated as

$$\Delta h_{hot} = \Delta h_{cold} \tag{3.79}$$

$$h_{exit,cold} = h_{in,cold} + \eta_{reg} (h_{exit,hot} - h_{in,cold})$$
(3.79a)

where $\Delta h_i(kJ/kg)$ is the change in enthalpy of the hot and cold fluids, and η_{reg} is the efficiency oft he regenerator.

3.4.8 Solution Pumps

The pump aids the circulation of the working fluid across the plant via change in pressure of the working fluid, and its power consumption, $\dot{W}_{P,i}$ (i = ORC-P1, ORC-P2, S-P) is given as

$$\dot{W}_{P,i} = \frac{\dot{m}_{in}(h_{exit} - h_{in})}{\eta_P}$$
(3.80)

where η_P , h_{exit} , h_{in} are the isentropic efficiency of the pump, and specific enthalpy of working fluids at exit and inlet, respectively. The working fluid exit pressure is evaluated in terms of a pressure factor,

$$P_{exit} = \theta P_{in} \tag{3.81}$$

3.4.9 Compressor

The power necessary to drive the compressor is can be evaluated as

$$\dot{W}_{C,i} = \dot{m}_{in}(h_{exit} - h_{in}) = \dot{m}_{in}c_{p,i}(T_{exit} - T_{in})$$
(3.82)

where $\dot{W}_{C,i}$ is the compression power needed to drive the compressor (i = A-COMP, F-COMP), $c_{p,i}$ is the isobaric specific heat capacity of the respective component fluid. The exit temperature T_{exit} is obtained as

$$T_{exit} = T_{in} \left(1 + \frac{1}{\eta_{C,i}} \left(1 - \left(\frac{P_{exit}}{P_{in}}\right)^{\frac{\kappa-1}{\kappa}} \right) \right)$$
(3.83)

where $\eta_{C,i}$, P_{exit} , P_{in} are the isentropic efficiency, exit and inlet pressures of the compressor.

3.4.10 Power turbine

The actual turbine powers are given as

$$\dot{W}_T = \eta_T \dot{m}_{in} (h_{in} - h_{exit}) \tag{3.84}$$

where η_T , is the efficiency of the turbine under isentropic condition. However, in the case of the gas turbine, GT, the power generated is given as

$$\dot{W}_{T,GT} = \dot{m}_{in}(h_{in} - h_{exit}) = \dot{m}_{in}c_{P,fg}(T_{35} - T_{36})$$
(3.85)

 $c_{P,fg}$ is the isobaric specific heat of flue gas, and T_{35} , T_{36} are the inlet and exit temperatures of the flue gas. T_{36} is expressed in terms of the turbine isentropic efficiency and exponent κ [156]

$$T_{36} = T_{36} \left(1 - \eta_{GT} \left(1 - \left(\frac{P_{35}}{P_{36}} \right)^{\frac{1-\kappa}{\kappa}} \right) \right)$$
(3.86)

3.4.11 Combustion chamber

The energy balance across the combustion chamber was evaluated according to Ogorure et al.[21] as

$$\dot{m}_{f,25}LHV_{f,25} + \dot{m}_{f,30}LHV_{f,30} + \dot{m}_{34}h_{34} = \dot{m}_{fg}c_{P,fg}T_{fg} + \dot{Q}_{in}$$
(3.87)

$$\dot{m}_{fg} = \dot{m}_{f,25} + \dot{m}_{f,30} + \dot{m}_{34} \tag{3.88}$$

$$\dot{Q}_{in} = \dot{m}_{waste} LHV_{waste} \eta_{CC} \tag{3.89}$$

where $LHV_f(kJ/kg)$, $\dot{m}_{fg}(kg/s)$, $\dot{Q}_{in}(kW)$ are the lower heating value of fuel, mass flow rate of flue gas and heat of combustion, respectively.

3.4.12 Exergy analysis

The specific chemical exergy of the feedstock is obtained according to the relationship[157]

$$e_{waste}^{ch} = \beta_{waste} LHV_{waste} \tag{3.90}$$

$$\beta_{waste} = \frac{1.0414 + 0.0177 \left(\frac{H}{C}\right) - 0.3328 \left(1 + 0.0537 \left(\frac{H}{C}\right) + 0.0493 \left(\frac{N}{C}\right)\right)}{1 - 1.4021 \left(\frac{H}{C}\right)}$$
(3.91)

$$LHV_{waste} = 4.187(81C + 300H - 26(O - S) - 6(9H + MC))$$
(3.92)

where β_{waste} is the proportion of chemical exergy to the LHV of biomass (crop waste and animal waste).

The exergy balance across the plant components of the integrated plant is presented in Table 3.2.

 Table 3.2:
 Exergy balance equations for components of the integrated plant

Plant component	Exergy equation
Gasifier	$\vec{E}x_{D,G} = \vec{E}x_1 + \vec{E}x_2 - \vec{E}x_3$
P1	$\vec{E}x_{D,P1} = \vec{W}_{P1} + \vec{E}x_8 - \vec{E}x_9$
HXS	$\vec{Ex}_{D,HXS} = \vec{Ex}_3 - \vec{Ex}_4 + \vec{Ex}_9 - \vec{Ex}_{10}$

SC	$\vec{Ex}_{D,SC} = \vec{Ex}_4 - \vec{Ex}_{4a} + \vec{Ex}_5$
ORC-P1	$\dot{Ex}_{D,P2} = \dot{W}_{P2} + \dot{Ex}_{12} - \dot{Ex}_{13}$
REG	$\dot{Ex}_{D,REG} = \dot{Ex}_{13} - \dot{Ex}_{14} + \dot{Ex}_{16} - \dot{Ex}_{17}$
EVP	$\dot{Ex}_{D,EVP} = \dot{Ex}_{10} - \dot{Ex}_{11} + \dot{Ex}_{14} - \dot{Ex}_{15}$
ORC-T	$\dot{Ex}_{D,ORCT} = \dot{Ex}_{15} - \dot{Ex}_{16} - \dot{W}_{T1}$
COND	$\dot{Ex}_{D,COND} = \dot{Ex}_{17} - \dot{Ex}_{12} + \dot{Ex}_{18} - \dot{Ex}_{19}$
PEME	$\dot{Ex}_{D,PEME} = \dot{W}_{PEME} + \dot{Ex}_{20} - \dot{Ex}_{21} - \dot{Ex}_{22}$
Valve	$\vec{Ex}_{D,V} = \vec{Ex}_5 + \vec{Ex}_{22} - \vec{Ex}_{25}$
MTH	$\dot{Ex}_{D,MTH} = \dot{Ex}_{24} - \dot{Ex}_{25} - \dot{Ex}_{26}$
Digester	$\dot{Ex}_{D,AD} = \dot{Ex}_6 - \dot{Ex}_7 - \dot{Ex}_{23}$
BC	$\dot{Ex}_{D,BC} = \dot{Ex}_7 - \dot{Ex}_{7a} - \dot{Ex}_{27}$
F-COMP	$\dot{Ex}_{D,FCOMP} = \dot{W}_{F,k} + \dot{Ex}_{27} - \dot{Ex}_{28}$
HX1	$\dot{Ex}_{D,HX1} = \dot{Ex}_{28} - \dot{Ex}_{29} + \dot{Ex}_{36} - \dot{Ex}_{37}$
A-COMP	$\dot{Ex}_{D,ACOMP} = \dot{W}_{A,k} + \dot{Ex}_{31} - \dot{Ex}_{32}$
HX2	$\dot{Ex}_{D,HX2} = \dot{Ex}_{32} - \dot{Ex}_{33} + \dot{Ex}_{37} - \dot{Ex}_{38}$
SOFC	$\vec{Ex}_{D,SOFC} = \vec{Ex}_{29} - \vec{Ex}_{30} + \vec{Ex}_{33} - \vec{Ex}_{34}$
CC	$\vec{Ex}_{D,CC} = \vec{Ex}_{30} + \vec{Ex}_{34} - \vec{Ex}_{35}$
GT	$\dot{Ex}_{D,Gt} = \dot{Ex}_{35} - \dot{Ex}_{36} - \dot{W}_{Gt}$
HRSG	$\vec{Ex}_{D,HRSG} = \vec{Ex}_{38} - \vec{Ex}_{39} + \vec{Ex}_{42} - \vec{Ex}_{43}$
S Pump	$\dot{Ex}_{D,P3} = \dot{W}_{P3} + \dot{Ex}_{41} - \dot{Ex}_{42}$
S-T	$\dot{Ex}_{D,St} = \dot{Ex}_{43} - \dot{Ex}_{44} - \dot{W}_{St}$
S-COND	$\vec{Ex}_{D,CONDst} = \vec{Ex}_{44} - \vec{Ex}_{41} + \vec{Ex}_{45} - \vec{Ex}_{46}$
ORC-P2	$\vec{Ex}_{D,ORCP2} = \vec{W}_{ORCP2} + \vec{Ex}_{47} - \vec{Ex}_{48}$
REG 2	$\vec{Ex}_{D,REG2} = \vec{Ex}_{48} - \vec{Ex}_{49} + \vec{Ex}_{51} - \vec{Ex}_{52}$
EVP2	$\vec{Ex}_{D,EVP2} = \vec{Ex}_{39} - \vec{Ex}_{40} + \vec{Ex}_{49} - \vec{Ex}_{50}$
ORC-T2	$\vec{Ex}_{D,ORC-T2} = \vec{Ex}_{50} - \vec{Ex}_{51} - \vec{W}_{ORC-T2}$
ORC-COND	$\dot{Ex}_{D,ORC-COND} = \dot{Ex}_{52} - \dot{Ex}_{47} + \dot{Ex}_{53} - \dot{Ex}_{54}$
Total Exergy	$\dot{Ex}_{D,total} = \sum_{k} \dot{Ex}_{D,k}$

3.4.13 Thermodynamic performance indicators

The energy efficiency measures the useful energy exiting a system with respect to the input energy into the system. The plant's performance based on thermal efficiency of a thermal cycle is defined as

$$\eta_{th} = \frac{\dot{W}_{net}}{\dot{Q}_{in}} = \frac{\dot{W}_{SOFC-GT} + \dot{W}_{ST} + \dot{W}_{ORCI} + \dot{W}_{ORCII}}{\dot{Q}_{in}}$$
(3.93)

The exergy efficiency which is also known as the second law efficiency assesses the effectiveness of the energy resource utilization. It is defined as the ratio of the exergy output in the product to the exergy input. The exergy efficiency of the plant is expressed as

$$\eta_{ex} = \frac{\dot{W}_{net}}{\dot{E}x_{in}} = \frac{\dot{W}_{net}}{\dot{E}x_{waste}}$$
(3.94)

On component basis, the exergy efficiencies of the components of the integrated can be categorized as follows

For work-producing devices the exergetic efficiency is expressed as

$$\eta_{ex,k} = \frac{W_k}{Ex_{in} - Ex_{exit}} \tag{3.94a}$$

The exergetic efficiency for work-consuming devices is

$$\eta_{ex,k} = \frac{Ex_{in} - Ex_{exit}}{W_k} \tag{3.94b}$$

For heat exchanger, the exergetic efficiency is expressed as

$$\eta_{ex,k} = \frac{\vec{E}x_{cold,exit} - \vec{E}x_{cold,in}}{\vec{E}x_{hot,in} - \vec{E}x_{hot,exit}}$$
(3.94c)

3.5 Economic analysis

This section presents models to evaluate the equipment cost and parameters of economic merit of the integrated plant.

A major economic parameter that depicts the cost effectiveness of the multigeneration plant is the levelized cost of electricity (*LCOE*). This parameter is expressed as the proportion of annualized cost of a plant, to effective electricity generated in kWh by the plant and is obtained as[158], [159]: $ACC = Z_{TCI} \times CRF$ (3.95)

$$CRF = \frac{i_r (1+i_r)^n}{(1+i_r)^n - 1}$$
(3.96)

$$LCOE = \frac{ACC (\$/yr)}{TEE (kWh)}$$
(3.97)

where ACC, Z_{TCI} , TEE, CRF, i_r , and n are annual capital cost, total capital investment cost, total effective electricity, capital recovery factor, interest rate and operating life of the plant, respectively.

$$Z_{TCI} = Z_f (Z_{TM} + Z_{Aux}) \tag{3.98}$$

$$Z_{TM} = 1.18 \sum Z_{BM} \tag{3.99}$$

$$Z_{BM} = Z_P^0 F_{BM} \tag{3.100}$$

$$Z_{Aux} = 0.3 \sum_{k=1}^{n_{eq}} Z^o_{BM,k}$$
(3.101)

$$\log_{10}(Z_P^0) = K_1 + K_2 \log_{10} Q + K_3 (\log_{10} Q)^2$$
(3.102)

$$f_{BM} = B_1 + B_2 f_M f_P \tag{3.103}$$

$$\log_{10}(f_P) = C_1 + C_2 \log_{10} P_g + C_3 (\log_{10} P_g)^2$$
(3.104)

$$Z_f = \frac{CEPCI_{chosen year}}{CEPCI_{ref year}}$$
(3.105)

where *CEPCI* is the chemical engineering plant cost index. The reference year according to Turton *et al.*[158] is 2011 and the chosen year is 2020. Z_{TM} , Z_{BM} , Z_P^o , $Z_{BM,k}^o$, and Z_{Aux} are the total module cost, bare module cost, bare module cost of equipment *k* without pressure and material correction factors, and auxiliary cost, respectively. *Q* and P_g are the capacity of components and K_1 , K_2 , K_3 , B_1 , B_2 , C_1 , C_2 , C_3 , f_M , f_P are correction factors of cost price, listed in Table 3.6. The scaling factor, Z_f was introduced in other to modify the equipment cost from a specific reference year to the chosen year. For heat exchangers (HXS, EVP, COND, HX) where area is the defined capacity, the surface area of heat transfer is given as

$$A_k = \frac{Q_k}{U_k LMTD_k} \tag{3.106}$$

where U_k is the overall heat transfer coefficient; $LMTD_k$ represent the logarithmic mean temperature difference, evaluated as[160]

$$LMTD_{k} = \frac{\Delta T_{A} - \Delta T_{B}}{\ln\left(\frac{\Delta T_{A}}{\Delta T_{B}}\right)}$$
(3.107)

A and B represents two ends where the hot and cold stream enter and exit, respectively. The break-even point, *BEP* and net present value, *NPV* of the plant were evaluated as

$$BEP = \frac{Z_{TCI}}{R_{el}}$$
(3.108)

$$NPV = -Z_{TCI} + R_{el} (CRF)$$
(3.109)

$$R_{el} = C_{el}(TEE) \tag{3.110}$$

where, R_{el} is the annual revenue of electricity, and C_{el} is the unit price of electricity.

3.6 Conventional and enhanced exergo-economic analysis of the integrated plant

3.6.1 Conventional and enhanced exergo-economic analysis

The objective of the exergo-economic analysis is to evaluate the economic viability of the proposed multigeneration power plant. The goal of an exergo-economic analysis of a system is to identify the mechanisms by which costs are formed and to calculate the exergy-based unit costs of product streams based on exergy and economic principles[161]. For each exergy transfer, a cost rate must be evaluated. These cost rates are related via a cost rate balance such that the sum of cost rates of all inlet streams and initial capital cost of component k, equals the sum of cost rates associated with the exit streams. This is represented as follows[162]:

$$\sum_{i=1}^{n} (\dot{C}_{i})_{in} + \dot{Z}_{k} = \sum_{j=1}^{m} (\dot{C}_{j})_{exit}$$
(3.111)

where \dot{C}_i is the cost rate associated with the exergy stream *i* and can be expressed in terms of the unit cost of exergy stream as:

$$\dot{C}_i = c_i \dot{E} x_i \tag{3.112}$$

The fuel and product exergy streams are defined according to the component, where the source consumed in generating the product represents the fuel. The exergy balance expressed in terms of these parameters is:

$$\dot{E}x_{F,k} = \dot{E}x_{D,k} + \dot{E}x_{P,k} \tag{3.113}$$

The cost rate of exergy destruction expressed in terms of the cost rate of fuel and exergy destroyed in any component, is obtained as:

$$\dot{C}_{D,k} = \left(c_F \dot{E} x_D\right)_k \tag{3.114}$$

The exergo-economic factor which is a measure of the cost effectiveness of the system is evaluated as:

$$f_k = \frac{\dot{Z}_k}{\dot{Z}_k + \dot{C}_{D,k}}$$
(3.115)

The application of Equations (3.110 - 3.113) to calculate the unknown exergy streams costs, cost balance for the plant components and necessary auxiliary equations are listed in Table 3.3 Table 3.3: Cost balance and auxiliary equations for components of the integrated plant

Component	Cost balance equation	Auxiliary equation
Gasifier	$\dot{C}_1 + \dot{C}_2 + \dot{Z}_{gasifier} = \dot{C}_3$	$c_1 = 0; c_2 = 1.1 (\$/GJ)$
P1	$\dot{C}_8 + \dot{C}_{P1} + \dot{Z}_{P1} = \dot{C}_9$	$c_8 = 0$; $c_{P1} = c_{ORC,T}$
HXS	$\dot{C}_3 + \dot{C}_9 + \dot{Z}_{HXS} = \dot{C}_{10} + \dot{C}_4$	$c_{3} = c_{4}$
SC	$\dot{C}_4 + \dot{Z}_{SC} = \dot{C}_{4a} + \dot{C}_5$	$c_{4a} = c_5$
ORC P1	$\dot{C}_{12} + \dot{C}_{ORC,P1} + \dot{Z}_{ORC,P1} = \dot{C}_{13}$	$c_{ORC,P1} = c_{ORC,T}$
REG	$\dot{C}_{16} + \dot{C}_{13} + \dot{Z}_{REG} = \dot{C}_{14} + \dot{C}_{17}$	$c_{16} = c_{17}$
EVP	$\dot{C}_{10} + \dot{C}_{14} + \dot{Z}_{EVP1} = \dot{C}_{11} + \dot{C}_{15}$	$c_{10} = c_{11}$
ORC T	$\dot{C}_{15} + \dot{Z}_{ORC.T1} = \dot{C}_{16} + \dot{C}_{ORC.T1}$	$c_{15} = c_{16}$
COND	$\dot{C}_{17} + \dot{C}_{18} + \dot{Z}_{COND1} = \dot{C}_{12} + \dot{C}_{19}$	$c_{17} = c_{12}$; $c_{18} = 0$
PEME	$\dot{C}_{20} + \dot{C}_{PEME} + \dot{Z}_{PEME} = \dot{C}_{21} + \dot{C}_{22}$	$c_{PEME} = c_{SOFC}; c_{21} =$
		$c_{22}; c_{20} = 0$
Valve	$\dot{C}_5 + \dot{C}_{22} = \dot{C}_{24}$	
MTH	$\dot{C}_{24} + \dot{Z}_{AC} = \dot{C}_{25} + \dot{C}_{26}$	$c_{25} = c_{26}$
Digester	$\dot{C}_6 + \dot{Z}_{digester} = \dot{C}_7 + \dot{C}_{23}$	$c_6 = 1.1 (\text{GJ}); c_7 = c_{23}$
BC	$\dot{C}_7 + \dot{Z}_{BC} = \dot{C}_{7a} + \dot{C}_{27}$	$c_{7a} = c_{27}$
F-COMP	$\dot{C}_{27} + \dot{C}_{FC} + \dot{Z}_{FC} = \dot{C}_{28}$	$c_{FC} = c_{GT}$
HX1	$\dot{C}_{32} + \dot{C}_{37} + \dot{Z}_{AHX} = \dot{C}_{33} + \dot{C}_{38}$	$c_{37} = c_{38}$
A-COMP	$\dot{C}_{31} + \dot{C}_{AC} + \dot{Z}_{AC} = \dot{C}_{32}$	$c_{AC} = c_{GT}; c_{31} = 0$
HX2	$\dot{C}_{28} + \dot{C}_{36} + \dot{Z}_{FHX} = \dot{C}_{29} + \dot{C}_{37}$	$c_{36} = c_{37}$

SOFC	$\dot{C}_{29} + \dot{C}_{33} + \dot{Z}_{SOFC} = \dot{C}_{30} + \dot{C}_{34} + \dot{C}_{SOFC}$	$c_{SOFC} = c_{30} = c_{34}$
CC	$\dot{C}_{30} + \dot{C}_{34} + \dot{Z}_{CC} = \dot{C}_{35}$	
GT	$\dot{C}_{35} + \dot{Z}_{GT} = \dot{C}_{36} + \dot{C}_{GT}$	$c_{35} = c_{36}$
HRSG	$\dot{C}_{38} + \dot{C}_{42} + \dot{Z}_{HRSG} = \dot{C}_{39} + \dot{C}_{43}$	$c_{38} = c_{39};$
S-P	$\dot{C}_{41} + \dot{C}_{p3} + \dot{Z}_{SP} = \dot{C}_{42}$	$c_{SP} = c_{ST}$
S-T	$\dot{C}_{43} + \dot{Z}_{ST} = \dot{C}_{44} + \dot{C}_{ST}$	$c_{43} = c_{44}$
S-COND	$\dot{C}_{44} + \dot{C}_{45} + \dot{Z}_{COND2} = \dot{C}_{41} + \dot{C}_{46}$	$c_{44} = c_{41}; c_{45} = 0$
ORC-P2	$\dot{C}_{47} + \dot{C}_{p4} + \dot{Z}_{ORC,P2} = \dot{C}_{48}$	$c_{ORCP2} = c_{ORC,T2}$
EVP 2	$\dot{C}_{39} + \dot{C}_{49} + \dot{Z}_{EVP2} = \dot{C}_{40} + \dot{C}_{50}$	$c_{39} = c_{40};$
REG 2	$\dot{C}_{51} + \dot{C}_{48} + \dot{Z}_{Reg2} = \dot{C}_{49} + \dot{C}_{52}$	$c_{51} = c_{52};$
ORC-T2	$\dot{C}_{50} + \dot{Z}_{ORC.T2} = \dot{C}_{51} + \dot{C}_{ORC.T2}$	$c_{50} = c_{51};$
ORC COND	$\dot{C}_{52} + \dot{C}_{53} + \dot{Z}_{COND2} = \dot{C}_{47} + \dot{C}_{54}$	$c_{47} = c_{52}$; $c_{53} = 0$

The unit cost of exergy of the integrated plant is calculated as

$$c_{P,Total} = \frac{\sum \dot{Z}_k + \dot{C}_f}{\dot{W}_{net}}$$
(3.116)

3.6.2 Enhanced exergo-economic assessment

In the search for optimization potentials in plant components, the fuel-product concept in relation to exergy and exergo-economic analysis serves as a useful tool in achieving this purpose. These tools aide in the identification of components irreversibilities and inefficiencies in the processes of energy conversion. However, these methods provide a view that the irreversibilities in the component may be recovered. Due to design constraints, a portion of the irreversibility is intrinsic and is unavoidable. With this insight, considerations of the unavoidable and avoidable exergy destruction in energy components can estimate real improvement potentials, thus allowing the focus of optimization on the avoidable portions.

$$\dot{E}x_{D,k}^{UN} = \dot{E}x_{P,k}^{r} \left(\frac{\dot{E}x_{D}}{\dot{E}x_{P}}\right)_{k}^{UN}$$
(3.117)

where $\vec{E}x_{P,k}^r$ is the exergy of product from component k under real conditions. $\left(\frac{\vec{E}x_D}{\vec{E}x_P}\right)_k^{UN}$ is the unavoidable exergy destruction per unit of product exergy obtained under maximum thermodynamic parameters for which the efficiency of the component reaches its achievable

maximum. The avoidable part of the exergy destroyed is obtained as the difference between the exergy destroyed within the component and unavoidable portion,

$$\dot{Ex}_{D,k}^{AN} = \dot{Ex}_{D,k} - \dot{Ex}_{D,k}^{UN}$$
(3.118)

Under unavoidable conditions, exergy efficiency can be obtained as:

$$\eta_{ex}^* = \frac{\vec{E}x_{P,k}^r}{\vec{E}x_{f,k}^r - \vec{E}x_{D,K}^{UN}}$$
(3.119)

Under similar circumstances, the cost per unit product exergy under unavoidable conditions $\left(\frac{\dot{z}}{Ex_P}\right)_k^{UN}$ is evaluated under extremely inefficient conditions of the component, such that high fuel costs are associated with the component. The cost rates associated with the exergy destruction, $\dot{C}_{D,k}^{UN}$, and investment cost under unavoidable conditions, \dot{Z}_k^{UN} are obtained as:

$$\dot{C}_{D,k}^{UN} = c_{f,k} \dot{E} x_{D,k}^{UN}$$
(3.120)

$$\dot{Z}_{k}^{UN} = \dot{E}x_{P,k}^{r} \left(\frac{\dot{Z}}{\dot{E}x_{P}}\right)_{k}^{UN}$$
(3.121)

The avoidable costs are obtained by difference between the total cost rates and unavoidable cost rates:

$$\dot{C}_{D,k}^{AV} = \dot{C}_{D,k} - \dot{C}_{D,k}^{UN}$$
(3.122)

$$\dot{Z}_{D,k}^{AV} = \dot{Z}_k - \dot{Z}_{D,k}^{UN} \tag{3.123}$$

Hence, the exergo-economic factor under these concepts is obtained as:

$$f_k^* = \frac{\dot{Z}_{D,k}^{AV}}{\dot{C}_{D,k}^{AV} + \dot{Z}_{D,k}^{AV}}$$
(3.124)

There are several advantages to dealing with needless exergy destruction and avoidable costs: The sum of avoidable cost rates $\dot{C}_{D,k}^{AV} + \dot{Z}_{D,k}^{AV}$ characterizes the possibility for decreasing costs associated with the kth component under design conditions significantly better than the sum of total cost rates utilized from the initial design conditions. Similarly, the modified exergetic efficiency η_{ex}^* characterizes the potential for exergy savings associated with the kth component better than the efficiency under base conditions. Furthermore, the modified exergetic efficiency can be used to compare the performance of similar components.

3.7 Environmental assessment

The environmental considerations play a major role in the evaluation of innovative power cycles and CO_2 has the high negative impact regarding the global GHG emissions[156]. In this work, the CO_2 emission per unit of power metric is employed to examine the environmental effects of the proposed configuration. The amount of CO_2 emission of the plant for complete combustion can be estimated as

$$\dot{m}_{CO_2} = 44.01 y_{CO_2} \left(\frac{\dot{m}_{fuel}}{M_{fuel}}\right)$$
(3.125)

where y_{CO_2} and M_{fuel} are the mole fraction of carbon in the fuel and the molar weight of the fuel, respectively.

The levelized CO₂ emission evaluated with respect to the net power output is given as [78]:

$$LE_{CO_2} = 3600 \frac{\dot{m}_{CO_2}}{\dot{W}_{Net}} (kg/MWh)$$
(3.126)

The sustainability performance of the plant was assessed by the exergy sustainability index (*ESI*) which is a measure of the exergy of fuel and consumption of efficiency or following parameters:

$$ESI = \frac{1}{1 - \eta_{ex}} \tag{3.127}$$

3.8 Multi-objective optimization

In the optimization analysis of the plant, the exergy efficiency, LCOE, and specific cost of exergy are the objective functions considered. Two cases of optimization are carried out with the first optimization focused on only the exergy efficiency and the LCOE, and the second case based on optimizing all three objectives. The optimization models are defined as:

Case 1: Dual objective optimization

$$Max\{\eta_{ex}, -LCOE\}$$
(3.128)

Case 2: Tri-objective optimization

$$Max\{\eta_{ex}, -LCOE, -c_{P,Total}\}$$
(3.129)

Variable	Range	
r_P	$5 \le r_P \le 10.5 (-)$	
Ĵsofc	$5000 \le j_{SOFC} \le 10000 \ (A/m^2)$	
T_{35}	$1200 \le T_{35} \le 1350$ (K)	
P_{43}	$5500 \le P_{43} \le 10000 (\text{kPa})$	
Δpp_{St}	$10 \leq \Delta p p_{St} \leq 50 (\mathrm{K})$	
$oldsymbol{ heta}$	$0.10 \le \ \theta \ \le 0.9 \ (-)$	
Δpp_{ORCII}	$5 \leq \Delta p p_{ORCII} \leq 15 (K)$	
Δsh_{ORCII}	$2 \leq \Delta sh_{ORCII} \leq 15 (K)$	

Table 3.4: Domain of decision variables for MOGA optimization

An ideal solution for the combined optimization is represented by each point on the Pareto curve. LINMAP, a decision-making technique designed to weigh the relative significance of the objectives was used for the optimal solution. The LINMAP method was applied by first acquiring a non-dimensional Euclidean set of the objective function values in the Pareto front as [132]:

Step 1. Setup a normalized objective matrix with m rows and n columns as follows

$$f_{ij}^{*} = \frac{f_{ij}}{\sqrt{\Sigma(f_{ij})^{2}}}$$
(3.130)

where f_{ij}^* is the non-dimensional solution, f_{ij} is the *i*th solution in the Pareto front, *j* is the number of objectives.

Step 2. Obtain a weighted normalized objective matrix multiplying each column by its weight, w_j $v_{ij} = f_{ij}^* \times w_j$ (3.131)

Step 3. The positive ideal solution, A+, is determined as follows. First, locate the best value of each objective from the column of the objective matrix. This implies that for a maximization objective, the largest value within the column is the best value, and the best value for a minimization objective, is the smallest value in the column. These are mathematically obtained as

$$A^{+} = \{ (Max(v_{ij})|j \in J), (Min(v_{ij})|j \in J^{*})|i \in 1, 2, ..., m \}$$

= $\{ v_{1}^{+}, v_{2}^{+}, ..., v_{j}^{+}, ..., v_{n}^{+} \}$ (3.132)

where *J* is the set of maximization objectives and J^* is the set of minimization objectives, from the total set of $\{1, 2, 3, 4, ..., n\}$.

Step 4. Calculate the Euclidean distance between each solution and the ideal solution: Distance to positive ideal d_i

$$d_{i} = \sqrt{\sum_{j=1}^{m} (v_{ij} - v_{i}^{+})^{2}}$$
(3.133)

The optimal solution having the shortest spatial distance is the recommended objective.

3.9 Solution flowchart

In this analysis, a major portion of the equations require thermodynamic properties for evaluation. The Engineering Equation Solver (EES) was used to solve thermodynamic problems involving the plant components. However, MATLAB software was applied for the MOGA, alongside EES. The characteristics of MOGA used in the optimization process are expressed in Table 3.6. A flow chart for the integration of EES and MATLAB for the optimization methodology is presented in Figure 3.3. Overall, the optimization process is as follows:

- i. The first attempt of thermodynamic, economic, and exergo-economic analysis in EES is performed using input data of parameters listed in Table 3.5.
- Evaluation of objective functions in EES using randomly generated data imported from MATLAB.
- iii. If stop criterion of the optimization is satisfied, optimization is stopped, and results are reported with a Pareto plot in MATLAB.
- iv. Else, obtain new set of random value of decision variables as a data vector generated by genetic algorithm and exported to EES.
- v. Continue process until stop criterion is achieved and end optimization.



Figure 3.3: Flowchart of the multi-objective optimization integrating MOGA in MATLAB with EES

3.10 Boundary condition

The plant specifications taken from the literature are compiled in Table 3.5. Ultimate analysis of the biomass feedstock is sourced from Ogorure *et al.*[21]. Cattle, poultry, and pig waste make up the animal waste category, whereas rice husk, palm fruit effluent, cassava peels, and maize stover, make up the crop waste category. Models according to Saeed and Warkozek [59], and Holagh *et al.*[96] were employed in the PEME analysis. ORC input parameters such as pinch point temperature for the evaporator, and condenser; and operating pressures, were obtained from Emadi *et al.*[99]. To optimize the output of energy sources, the turbines are designed with a turbine efficiency of more than 80 %. The choice of a tubular SOFC model specification according to

Holagh *et al.*[96] is adopted. The selection of subsystems is based on the quality of resource available to each subsystem and the temperature requirements to match system temperature and pressure for successful integration. The pinch point method is used to calculate the temperature of the hot and cold streams in subsystems with heat exchangers.

Table 3.5: Input data [4], [21], [69], [96], [109], [141], [153]

Ambient temperature	298 K
Ambient pressure	101.325 kPa
Ultimate analysis of animal waste	44.26 % C 5.95 % H 32.69 % O 5.66 % N 1.21 % S
Ultimate analysis of crop waste	46.45 % C 5.40 % H 38.69 % O 0.64 % N 0.03 % S
Gasification temperature	1023 K
Gasifier pressure	101.325 kPa
Digester temperature	345 K
РЕМЕ	
Temperature	353 K
Pressure	101.325 kPa
Current density	1000 A/m^2
Activation area	0.01 m^2
Amount of water(anode)	14
Amount of water(cathode)	10
Wall thickness	100 µm
Activation power(anode)	76000 J/kmol
Activation power(cathode)	18000 J/kmol
Pre-exponential indicator(anode)	$1.7 \times 10^5 \text{ A/m}^2$
Pre-exponential indicator(cathode)	$4.6 \text{ x } 10^3 \text{ A/m}^2$
Faraday constant	96486 C/mol
Electrons	2 (-)
ORC	
Pressure	$0.9 P_{cr}$
Base case working fluid	MM
Turbine isentropic efficiency	84 %
Pump efficiency	85 %
Evaporator pinch point	10 K
Pinch point in condenser	5 K
Regenerator pinch point	4 K
Superheat	10 K
Methanation	
Presure	1000 kPa
Temperature	500 K
SOFC	
Current density	5500 A/m ²
Operating temperature	1023 K
Activation area	0.08 m^2

Fuel conversion factor	0.85 (-)
Air factor	0.15 (-)
Exchange current density(anode)	6500 A/m^2
Exchange current density(cathode)	2500 A/m ²
Anode thickness	0.05 x 10 ⁻² m
Cathode thickness	0.005 x 10 ⁻² m
Thickness of electrolyte	0.001 x 10 ⁻² m
Thickness of interconnect	0.03 x 10 ⁻² m
Effective gaseous diffusivity(anode)	$0.2 \ge 10^{-4} \text{ m}^2/\text{s}$
Effective gaseous diffusivity(cathode)	$0.05 \text{ x } 10^{-4} \text{ m}^2/\text{s}$
No. Cells	10000
Steam - carbon ratio	2.5 (-)
Inverter	95 %
Gas turbine	
Turbine efficiency	85 %
Compressor	85 %
Pressure factor	10.5 (-)
Steam Turbine	
Turbine inlet pressure	6000 kPa
Pinch point temperature	15 K
Heat Exchanger	
Heat transfer co-efficient	$1.1 \text{ kW/m}^2\text{K}$
Economic	
Plant life	20 years
Interest rate	12 %
Annual operation	8000 hrs
Maintenance factor	1.06 (-)

Table 3.6: Assumptions for unavoidable condition[120,128]

Component	Exergy	Cost								
Gasifier	$T_{03} = 1023 \text{ K}$	$T_{03} = 923 \text{ K}$								
PEM	$T_{21} = 353 \text{ K}$	$T_{21} = 383 \text{ K}$								
SOFC	$j = 7000 \text{ A/m}^2$	$j = 5000 \text{ A/m}^2$								
Compressor	$\eta_c = 95 \%$	$\eta_{\mathcal{C}} = 70 \%$								
Turbine	$\eta_T = 95 \%$	$\eta_T = 70 \%$								
CC	$T_{35} = 1373 \text{K}$	<i>T</i> ₃₅ =1123 K								
HRSG	10	30 K								
EVP	4 K	15 K								
Pump	$\eta_P = 95 \%$	$\eta_P = 70 \%$								
	<i>K</i> ₁	<i>K</i> ₂	<i>K</i> ₃	B_1	<i>B</i> ₂	<i>C</i> ₁	C_2	<i>C</i> ₃	f_M	f_{BM}
-------------	-----------------------	-----------------------	-----------------------	-------	-----------------------	-----------------------	-----------	-----------------------	-------	----------
Pump	3.3892	0.0536	0.1538	1.89	1.35	-0.3935	0.3957	-0.00226	1.55	-
ORC-T	2.6259	1.4398	-0.1776							3.5
HRVG	4.8306	-0.8509	0.3187	1.63	1.66	0.03881	-0.011272	0.08183	1.8	
Regenerator	2.7652	-0.7282	0.0783	1.74	1.55				1.25	1.25
Condenser	4.8306	-0.8509	0.3187	1.63	1.66	0.03881	-0.011272	0.08183	1.8	
MTH	4.1052	0.5320	-0.0005	L=5[m], D =	0.95[m]				4
Compressor	2.2898	1.3604	-0.1027							2.8
HX fuel	3.3444	-0.2745	-0.0472	1.74	1.55				1.25	1.25
HX Air	3.3444	-0.2745	-0.0472	1.74	1.55				1.25	1.25
Gas Turbine	-21.77	13.2175	-1.5279							3.5
HRSG	4.8306	-0.8509	0.3187	1.63	1.66	0.03881	-0.011272	0.08183	1.8	
ST	2.6259	1.4398	-0.1776							3.5

Table 3.7: Parameters in plant component cost estimation [21], [96], [158], [159]

Gasifier $1600(3600\dot{m}_{biomass})^{0.67}.Z_f$	
Digester $350000 \left(\frac{\dot{v}\tau}{21000m^3}\right)^{0.75} . Z_f$	
SOFC $A_{act}N_{FC}(2.96T_{SOFC} - 1907).Z_f$	
Inverter $100000 \left(\frac{\dot{W}_{SOFC}}{500}\right)^{0.7} . Z_f$	
PEME $1000\dot{W}_{PEME}.Z_f$	

 $CEPCI_{2020} = 596.2, CEPCI_{2011} = 397 [158]; C_{el} = 150 (\$/MWhr)$

4 Results and Discussion

4.1 Preamble

The results obtained from the analysis of the plant using the models and input data in Chapter 3 are presented in this chapter. Validation of models of the gasification process, solid oxide fuel cell (SOFC), and proton exchange membrane electrolyzer (PEME) are presented in section 4.2. The products of the biomass conversion, thermo-economic, and exergo-economic analysis, simulation and optimization results are in sections 4.3 - 4.8.

4.2 Model validation

To verify the reliability of the models, data from the literature was compared with the outcomes of the analysis conducted on the gasification, SOFC, and PEME sub-units. Table 4.1 shows the comparison of the dry syngas composition from the gasification results of wood with those reported by Athari *et al.*[153] and Zainal *et al.*[154]. The maximum deviation from the results is the composition of methane in the syngas.

Composition (%) This study		Experimental data	Zainal <i>et al</i> .	Athari <i>et</i>	
		Zainal et al.[153]	[153]	<i>al</i> . [154]	
<i>N</i> ₂	42.75	42.31	46.68	48.7	
H_2	24.45	15.23	21.61	18.01	
<i>CO</i> ₂	12.61	16.42	12.01	13.84	
СО	20.19	23.04	19.61	18.77	
CH ₄	0.00	1.58	0.64	0.68	
<i>0</i> ₂	0	1.42	0	0	
Calorific value (MJ/m ³)	4.57	4.85	4.72	-	

Table 4.1: Validation of gasification model with wood at a temperature 1073 (K)

The root mean square error of the study and experimental data is 4.33, and between the study, gasification models from Zainal *et al.* [153], and Athari *et al.* [154] are 2.03 and 3.67, respectively.

Using the variations in electrolyzer voltage as shown in Table 4.2 the validity of the PEME data was compared to the study by Saeed and Warkozek[59]. With the cathode and anode at atmospheric pressure, and temperature of 70 °C, maximum error of 10 % was observed with low current density of 6000 A/m². Under current densities of 8000 – 10000 A/m² error values were within agreeable range.

		V_{PEME} (V)	
$j_E (A/m^2)$	This study	Saeed and Warkozek [59]	Error(%)
6000	1.7	1.89	10.0
8000	1.88	1.92	2.08
10000	2.063	1.95	5.79

Table 4.2: Validation of proton exchange membrane electrolyzer model

The validation of the tubular SOFC model is presented in Table 4.3 for the cell voltage, V_{SOFC} , and power density, ρ_{SOFC} , compared with the work of Holagh *et al.*[96]. A minimal error range of 0.6 - 2.4 % for the V_{SOFC} and 2.8 - 5.1 % for ρ_{SOFC} demonstrate strong agreement with the findings in literature as they fall within the permitted range.

	V_{SOFC} (V)			$ ho_{SOFC} (W/m^2)$			
j_{SOFC} (A/m ²)	Present	Holagh <i>et</i>	Error (%)	Present	Holagh	Error	
	work	al.[96]		work	<i>et al.</i> [96]	(%)	
2000	0.779	0.790	1.4	0.163	0.158	3.2	
3000	0.707	0.711	0.6	0.222	0.216	2.8	
4000	0.636	0.644	1.2	0.266	0.253	5.1	
5000	0.566	0.560	1.1	0.296	0.288	2.8	
6000	0.498	0.510	2.4	0.312	0.301	3.7	

Table 4.3: SOFC validation with related studies

4.3 **Products of biomass conversion**

In Table 4.4, the gas composition from the gasification of crop waste, and anaerobic digestion of animal waste is presented. The composition of syngas on wet basis indicates the presence of 0.01 % of hydrogen sulphide equivalent to 100 ppmv. This is due to the sulphur content in the feedstock and can alter the methanation process by reacting with the catalyst, leading to its deactivation and prevention of CO/CO₂ methanation. Through proper gas cleaning techniques such as chemical absorption, and wet scrubbing the H₂S can be removed from the syngas prior to methanation. A low heating value (LHV) of 3.85 MJ/kg is obtained for the syngas and can be attributed to the gasification temperature and the gasifying agent of air. Hence, a high N₂/C ratio,

Syngas composition	Composition (%, wet basis)	Composition (%, dry basis)
<i>N</i> ₂	46.62	49.87
H_2	17.49	18.72
<i>CO</i> ₂	10.86	11.63
СО	18.48	19.77
CH ₄	0.00	0.00
H_2S	0.01	-
<i>H</i> ₂ <i>0</i>	6.54	-
Nitrogen/Carbon (-)	1.59	
LHV (MJ/kg)	3.85	
Biogas composition	Composition (%, wet basis)	Composition (%, dry basis)
CH ₄	46.24	51.79
<i>CO</i> ₂	43.05	48.21
NH ₄	9.79	-
H_2S	0.92	-
LHV (MJ/kg)	14.07	

Table 4.4: Gas composition from the conversion of biomass

where the N_2 acts as a diluent to the syngas reducing its heating value. Furthermore, the composition of the biogas on dry basis shows reasonable amount of CH₄ and CO₂ from the

conversion of animal waste as standard raw biogas from anaerobic digestion can attain a range of 50 - 65 % CH₄, with CO₂ of about 50 - 35 % [42].

4.4 Results of thermo-economic analysis

Table 4.5 displays the thermodynamic and exergo-economic properties of the plant at various points. At each stage of the plant, these include the pressure, temperature, mass flow rate, cost rate, and specific energy cost. The reference conditions for each state point are assumed to be at a pressure, and temperature of 298.2 K and 101.3 kPa, respectively.

S/No	P(kPa)	T(K)	ṁ(kg/s)	$\dot{Ex}(kW)$	Ċ(\$/hr)	c(\$/GJ)
1	101.3	298.2	0.0308	0	0	0
2	101.3	298.2	0.0138	275.7	1.09	1.10
3	101.3	1023	0.0446	211	1.48	1.94
4	101.3	480	0.0446	174.1	1.22	1.94
4a	101.3	330	0.0259	1.468	0.03	6.26
5	101.3	330	0.0187	170.7	3.85	6.26
6	101.3	298.2	1.2980	19938	78.95	1.10
7	101.3	328.2	1.2080	17970	73.19	1.13
7a	101.3	300	0.8678	374.1	1.58	1.17
8	15.33	327.6	0.0129	0.0724	0	0
9	4000	327.9	0.0129	0.1254	1.89	4161
10	3880	561.9	0.0129	13.59	170.03	3477
11	3764	561.9	0.0129	6.305	78.91	3477
12	9.374	308.2	0.0526	0.0102	0.25	6867
13	1745	308.8	0.0526	0.1327	4.41	9243
14	1745	429.9	0.0526	2.645	78.08	8200
15	1745	532	0.0526	9.439	233.35	6867
16	9.374	459.2	0.0526	3.419	84.53	6867
17	9.374	312.8	0.0526	0.4421	10.93	6867
18	101.3	298.2	0.3488	0	0	0
19	101.3	306.5	0.3488	0.1689	12.94	21270
20	101.3	343	0.0523	0.3014	0	0
21	101.3	343	0.0395	12.21	0.8176	18.6
22	101.3	343	0.0128	278.7	18.6552	18.6
23	101.3	328.2	0.0907	1853	7.5492	1.13
24	101.3	345	0.0211	444.7	22.55	14.09
25	100	500	0.0108	376.7	25.04	18.46
26	100	500	0.0103	2.604	0.1731	18.46
27	101.3	298.2	0.3398	17594	74.268	1.17
28	1064	554.9	0.3398	17785	96.3	1.50

Table 4.5: Thermodynamic and exergo-economic properties of the integrated plant

29	1042	785.5	0.3398	17955	102.744	1.59
30	1022	1023	0.708	12411	305.568	6.84
31	101.3	298	10.58	0	0	0
32	1064	637.7	10.58	3370	363.24	29.94
33	1042	785.5	10.58	4331	399.6	25.63
34	1022	1023	10.17	6294	154.98	6.84
35	1001	1273	10.89	13613	486.36	9.93
36	108.2	826.2	10.89	3612	129.096	9.93
37	106	811.5	10.89	3433	122.69	9.93
38	103.9	711.5	10.89	2415	86.29	9.93
39	101.8	528.3	10.89	903.9	32.31	9.93
40	99.8	506.5	10.89	738.9	26.41	9.93
41	15	327.1	1.02	5.552	0.3356	16.79
42	6000	327.6	1.02	11.83	1.3460	31.6
43	6000	653.9	1.02	1229	74.27	16.79
44	15	327.1	1.02	188.4	11.39	16.79
45	101.3	298.2	83.03	0	0	0
46	101.3	311.5	83.03	44.8	12.09	74.97
47	9.374	308.2	1.019	0.1976	0.017	23.67
48	1745	308.8	1.019	2.568	0.73	78.66
49	1745	419.1	1.019	43.87	4.93	31.22
50	1745	522	1.019	170.3	14.51	23.67
51	9.374	446.5	1.019	57.76	4.92	23.67
52	9.374	312.8	1.019	8.556	0.73	23.67
53	101.3	298.2	19.64	0	0	0
54	101.3	304.6	19.64	1.859	1.51	225.30

4.5 Results of biomass conversion and upgrade

The base case analysis focused on obtaining the net power, efficiencies, and levelized cost of electricity (LCOE) using the integration of the power plant's subsystems and bio-conversion unit (BCU). Based on Table 4.6, the overall energy and exergy performance attained values of 44.93 % and 36.78 %, respectively. Combining hydrogen from the PEME at a flow rate of 0.0023 kg/s improved the LHV of syngas from gasification from 3.85 MJ/kg to 33.48 MJ/kg. With a thermal efficiency of 30.04 %, ORC-I produces 5.19 kW of the net power. Additionally, in the base condition, the GT, ST, and ORC-II net power outputs were determined to be 3.856 MW, 0.89 MW, and 0.098 MW, respectively. The SOFC net power of 2.58 MW accounts for 34 % of the net power of the integrated plant. Following the power consumption of 0.45 MW by the PEME, a net power of 7.42 MW was obtained. With the power plant capacity, it can attain a generation of 59 GWh

which is approximately a quarter of the estimated power consumed in the locality with a per capita consumption of 135 kWh. For the configuration, the LCOE is 130.5 \$/MWh. In comparison, this is higher than the average LCOE from bioenergy in selected regions as determined by the International Renewable Energy Agency which include 58 \$/MWh for India, 60 \$/MWh for China, 88 \$/MWh for Europe, and 77 \$/MWh for the rest of the world. However, it is within the range of the LCOE of 250 \$/MWh for bioenergy electricity in regions with inexpensive feedstocks[9]. Furthermore, a BEP of 6.5 years and NPV of 1.135 G\$ is obtained. The low BEP is favourable for the configuration, considering the high NPV. The exergy sustainability index (ESI) for the plant is 1.582 which is expected to be above 1 (as high as possible) for renewable energy systems[12]. The specific cost of product and levelized emission were obtained 27.01 \$/GJ and 0.883 kg/MWh, respectively.

Parameter	Value
LHV of waste (MJ/kg)	17.697
Mass flow rate of feedstock (kg/s)	0.9821
Exergy of biomass (MJ/kg)	20.18
LHV of syngas (MJ/kg)	3.85
LHV of biogas, CH4 (MJ/kg)	15.22
LHV upgraded syngas (MJ/kg)	33.48
	5 101
Net power ORC-I (KW)	5.191
Electrolyzer net voltage(V)	2.01
PEME input power (MW)	0.45
Mass flow rate of hydrogen(kg/s)	0.0023
Efficiency of electrolyzer (%)	73.73
Net voltage SOFC(V)	0.59
SOFC power output (MW)	2.58
SOFC efficiency (%)	40%
Net power (MW) of GT	3.856
Steam Turbine power (MW)	0.89

Table 4.6: Thermo-economic assessment under base condition of the multigeneration power plant

Efficiency of ST (%)	30.10
Steam Turbine Exergy Efficiency (%)	36.81
Net power ORC-II(MW)	0.098
Thermal efficiency ORC-II (%)	30.07
Overall power (MW)	7.42
Energy efficiency of the integrated plant (%)	44.93
Exergy efficiency of the integrated plant (%)	36.78
LCOE(\$/MWh)	130.5
NPV (G\$)	1.135
BEP (yr)	6.5
$c_{P,Total}(\$/GJ)$	27.01
LEco2 (kg/MWh)	0.883
ESI (-)	1.582

4.6 Results of the exergo-economic and environmental analysis

The exergo-economic assessment parameters of the integrated plant components are presented in Table 4.7. The combustion chamber (CC) and gas turbine (GT) returned higher exergy destruction values of 5469 kW and 1810 kW, respectively, whereas fluid pumps (P1) and (ORC-P1) have the lowest values of exergy destruction compared to other components. Furthermore, the lowest exergy efficiencies of 26.75 % and 24.04 % were obtained from condensers of the steam turbine (S-COND) and ORC-II unit (ORC-COND), respectively, which can be attributed to heat losses due to the difference in temperature between the fluids. The syngas clean-up unit (SC), CC and GT have the highest total costs of 167.92 \$/hr, 135.58 \$/hr, and 399.92 \$/hr, respectively, implying the importance of these components from the exergo-economic assessment. The low exergo-economic factors of 0.58 %, and 0.68 % in the regenerator (Reg) of ORC-1, and CC, respectively, indicate that the cost associated with the exergy destruction is higher than the initial investment cost and these components should be points of interest in minimizing the losses associated with exergy.

Component	$\dot{Ex}_D(kW)$	$\eta_k(\%)$	$c_f(\$/GJ)$	$\dot{C}_D(\$/hr)$	$\dot{Z}_k(\$/hr)$	$\dot{C}_D + \dot{Z}_k$	$f_k(\%)$
Carifian		76.54	1 10	0.00	0.00	(1/nr)	(0.00
Gasiner	64.67	76.54	1.10	0.26	0.38	0.64	60.00
PI	0.0084	86.36	7681	0.23	0.18	0.41	44.01
HXS	23.45	88.89	1.94	0.16	2.66	2.82	94.19
SC	1.9240	98.89	1.94	0.02	167.90	167.92	99.99
ORC-P1	0.0207	85.52	7681	0.57	0.34	0.91	37.12
Reg	0.4645	86.92	6867	11.48	0.07	11.55	0.58
EVP	0.4865	97.00	3477	6.09	64.15	70.24	91.33
ORC-T	0.6247	89.62	6867	15.44	0.36	15.80	2.27
COND	0.2630	40.51	6867	6.50	2.26	8.76	25.76
PEME	159.40	64.60	6.84	3.92	8.40	12.32	68.14
Valve	4.6470	98.97	13.91	0.23	0.05	0.28	17.42
MTH	65.42	85.29	14.09	3.32	2.66	5.98	44.50
Digester	115.40	99.42	1.10	0.46	1.79	2.25	79.65
BC	2.1550	99.99	1.13	0.01	2.66	2.67	99.67
F-COMP	4.2620	97.82	23.49	0.36	5.49	5.85	93.84
HX1	8.3640	99.96	9.93	0.30	0.07	0.37	19.67
A-COMP	320.60	91.31	23.49	27.12	51.08	78.20	65.33
HX2	56.45	99.17	9.93	2.02	0.07	2.09	3.50
SOFC	867.40	96.11	5.38	16.81	25.00	41.81	59.79
CC	5469	70.76	6.84	134.65	0.93	135.58	0.68
GT	1810	81.90	9.93	64.69	335.23	399.92	83.82
HRSG	294.10	87.88	9.93	10.51	18.93	29.44	64.29
S-P	0.9935	86.35	25.65	0.09	0.29	0.38	75.65
S-T	144.10	86.15	16.79	8.71	19.87	28.58	69.53
S-COND	138	26.75	16.79	8.34	1.04	9.38	11.07
ORC-P2	0.4015	85.51	42.62	0.06	0.29	0.35	82.23
Reg2	7.9080	86.89	23.67	0.67	0.01	0.68	1.56
EVP2	38.53	95.93	9.93	1.38	3.69	5.07	72.80
ORC-T2	11.98	89.35	23.67	1.02	5.84	6.86	85.13
ORC-COND	6.4990	24.04	23.67	0.55	0.80	1.35	58.97

Table 4.7: Conventional exergo-economic assessment parameters of the plant

According to Table 4.8, the modified exergy efficiencies of each component predicted by the enhanced method are higher than those of the conventional method in Table 4.7. The realistic definition of efficiency from this perspective indicates the beneficial interpretation across the components. For example, the PEME, air compressor (A-COMP), combustion chamber (CC), gas turbine (GT) and HRSG, and steam turbine (S-T) have higher exergy efficiencies in the advanced approach compared to the corresponding values in the conventional method.

Component	$\dot{Ex}_p(kW)$	$\left(\frac{\vec{E}x_D}{D}\right)^{UN}$	$\dot{Ex}_{D,k}^{UN}$	$\dot{Ex}_{D,k}^{AV}$	$\dot{C}_{D,k}^{UN}$	$\dot{C}^{AV}_{D,k}$	η_{ex}^{*} (%)
		$\left(\frac{E}{E}x_{P}\right)_{k}$	(kW)	(kW)	(\$/hr)	(\$/hr)	
Gasifier	211.03	0.2721	57.42	7.25	0.2274	0.0287	96.68
P1	0.052979	0.0281	0.0015	0.01	0.0411	0.1904	88.5
HXS	187.68	0.1158	21.73	1.72	0.152	0.0121	99.09
SC	172.18	0.0084	1.45	0.47	0.0102	0.0033	99.73
ORC-P1	0.1225	0.0298	0.0037	0.02	0.1011	0.4725	87.76
Reg	3.09	0.1427	0.4409	0.02	10.8904	0.5927	99.23
EVP	15.79	0.0147	0.2321	0.26	2.8881	3.2015	98.4
ORC-T	5.4	0.021	0.1134	0.51	2.8068	12.6365	91.35
COND	0.1791	1.2435	0.2227	0.04	5.5057	0.996	81.63
PEME	290.9	0.5215	151.70	7.69	3.7353	0.1892	97.43
Valve	444.75	0.0102	4.54	0.11	0.227	0.0057	99.97
MTH	379.28	0.1218	46.20	19.23	2.343	0.9754	95.17
Digester	19822.6	0.0056	111.01	4.73	0.4383	0.0187	99.98
BC	17967.85	0.000087	1.56	0.60	0.0064	0.0024	99.99
F-COMP	191.04	0.0146	2.79	1.47	0.236	0.1245	99.24
HX1	21388.64	0.00004	0.8555	7.46	0.0322	0.2667	99.97
A-COMP	3369.4	0.0349	117.59	202.88	9.9547	17.1566	94.32
HX2	6746.55	0.0046	31.03	25.29	1.1138	0.9038	99.63
SOFC	21418.6	0.035	749.65	117.32	14.5384	2.2739	99.46
CC	13236	0.2509	3320.91	2148.49	81.7523	52.8968	86.04
GT	8191	0.1752	1435.06	375.24	51.2795	13.4113	95.62
HRSG	2132.73	0.1228	261.90	32.13	9.3629	1.1485	98.52
S-P	6.28	0.0281	0.1765	0.82	0.0163	0.0754	88.49
S-T	896.5	0.0282	25.28	118.83	1.5275	7.1825	88.3
S-COND	50.4	2.7187	137.02	0.98	8.2821	0.0591	98.1
ORC-P2	2.3695	0.0298	0.0706	0.33	0.0108	0.0508	87.75
Reg2	52.42	0.1418	7.43	0.47	0.6335	0.0403	99.11
EVP2	909.24	0.0195	17.73	20.84	0.6322	0.7449	97.76
ORC-T2	100.56	0.0214	2.15	9.83	0.1833	0.8375	91.1
ORC-COND	2.057	1.7112	3.52	2.98	0.2999	0.2539	40.85
Total			6509.79	3108.02	209.23	116.75	

Table 4.8: Enhanced exergo-economic assessment of the integrated plant

Furthermore, the avoidable cost rates indicate cost prospects in each plant component and overall opportunity for cost reduction in the integrated plant. Theoretically, a total avoidable cost rate of 116.75 \$/hr of the total cost rate of 325.98 \$/hr associated with the destruction of exergy can be avoided. Plant components with major contributions to this value include A-COMP, CC, GT, S-T, and ORC-T. In order to achieve an overall economic improvement of the plant, these components are to be prioritized, amongst others.

A comparison of the exergo-economic factors of each component from the conventional and enhanced analysis in Table 4.9 shows lower exergo-economic factors in the conventional method for the Reg, EVP, COND, PEME, CC, HRSG, S-COND, and Reg2, however, the reverse is observed in other plant components. This is due to the high unavoidable cost associated with the exergy destruction. With the application of enhanced exergo-economic analysis, an insight to reach cost improvement and optimized thermodynamic performance of the plant is provided for the design process for effective decision making. This corrects the impression created by the conventional approach on the possibility of recovering the total exergy destroyed in the plant.

Component	$\dot{Z}_k(\$/hr)$	$\dot{Z}_k^{UN}(\$/hr)$	$\dot{Z}_k^{AV}(\$/hr)$	$\dot{C}^{AV}_{D,k}(\$/hr)$	$f_{k}^{*}(\%)$	$f_k(\%)$
Gasifier	0.3841	0.3692	0.0149	0.0287	34.17	60.00
P1	0.1819	0.1769	0.005	0.1904	2.56	44.01
HXS	2.6604	2.5114	0.149	0.012	92.55	94.19
SC	167.904	80.28	87.624	0.0033	99.99	99.99
ORC-P1	0.3385	0.293	0.0455	0.4725	8.78	37.12
Reg	0.067	0.0213	0.0457	0.5926	7.16	0.58
EVP	64.15	7.1358	57.0142	3.2015	94.68	91.33
ORC-T	0.3583	0.3053	0.053	12.64	0.42	2.27
COND	2.2561	1.3048	0.9513	0.996	48.85	25.76
PEME	8.3952	0.8549	7.5403	0.1892	97.55	68.14
Valve	0.0491	0.0479	0.0012	0.0057	17.39	17.42
MTH	2.6604	2.491	0.1694	0.9754	14.80	44.50
Digester	1.7885	0.1789	1.6096	0.0187	98.85	79.65
BC	2.6604	2.66	0.0004	0.0024	14.29	99.67
F-COMP	5.4936	5.43	0.0636	0.1245	33.81	93.84
HX1	0.07319	0.0696	0.00359	0.2667	1.33	19.67
A-COMP	51.08	50.67	0.41	17.16	2.33	65.33

Table 4.9: Cost rates and exergo-economic factor of enhanced exergo-economic assessment

-						
HX2	0.07319	0.0654	0.00779	0.9038	0.86	3.50
SOFC	24.99	19.81	5.18	2.27	69.53	59.79
CC	0.9256	0.3594	0.5662	52.9	1.06	0.68
GT	335.23	269.47	65.76	13.41	83.06	83.82
HRSG	18.93	3	15.93	1.15	93.27	64.29
S-P	0.285	0.0302	0.2548	0.0754	77.17	75.65
S-T	19.87	2.04	17.83	7.18	71.29	69.53
S-COND	1.0382	0.1085	0.9297	0.0591	94.02	11.07
ORC-P2	0.285	0.1531	0.1319	0.0508	72.20	82.23
Reg2	0.0106	0.0051	0.0055	0.0403	12.01	1.56
EVP2	3.6864	1.6231	2.0633	0.7449	73.47	72.80
ORC-T2	5.8428	5.2028	0.64	0.8375	43.32	85.13
ORC-COND	0.796	0.4663	0.3297	0.2538	56.50	58.97

4.7 Simulation

Figure 4.1 - 4.10 presents the results of a sensitivity analysis that aims to the reaction of the performance of the plant to variations in design parameters. This indicates how the SOFC current density, SOFC-GT pressure ratio, combustion temperature, steam turbine inlet pressure, and six ORC working fluids in ORC I and II affect net power, energy efficiency, and LCOE.

The product fraction of the methanation of syngas at equilibrium calculated with Gibbs free energy of minimization method at pressure of 100 kPa in Figure 4.1. The input feed with a stoichiometric H_2 : CO₂ molar ratio of over 4:1 resulted in products of mainly CH₄ and H₂O at low-temperature ranges of 423 – 673 K. Increasing the temperature above 673 K resulted in an increase of the by-products of CO as a result of the reversed water gas shift reaction. Also, the amount of H₂ and CO₂ unreacted increased, with a decrease in CH₄. Based on the exothermic nature of the CO₂ methanation reaction, high temperatures are unfavourable to achieve high methane content in the product gas. With temperatures above 773 K, a maximum fraction of CO₂ is attained before decreasing due to the dominance of the reversed water gas shift reaction. Similar trends were reported by Becker *et al.* [64], and Gao *et al.*[69]. According to Gao *et al.* (Gao *et al.*, 2012b), to achieve a high CH₄ yield at 1 atm the reaction temperature is expected not to surpass 527 K. However, due to considerable kinetic challenges, the reduction of CO₂ to CH₄ by hydrogen is

difficult to achieve. Hence, to hydrogenate CO₂ to CH₄, a significantly active catalyst is needed, specifically at low temperatures.



Figure 4.1: Composition of products of methanation of CO and CO₂ from syngas

The effect of the amount of hydrogen from the PEME with power consumption is shown in Figure 4.2(a). Increasing amount of hydrogen resulted in an increase in input power demand of the PEME due to corresponding increase in the number of cells required to increase the amount of hydrogen produced. The hydrogen in the feed composition of product gas in the methanation process affects the methanation process. Increasing the amount of hydrogen through the PEME integration will enhance the methanation yield at low temperatures. In Figure 4.2(b), the mole fraction of CH₄ in the upgraded syngas, as a function of H₂ content in the methanation reaction at temperatures of 400 – 800 K. High fractions of CH₄ are observed at low temperatures just as in Figure 4.1 compared to temperatures above 700K. The fraction of CH₄ increased with increasing flow rate of H₂, and attained a maximum before decreasing as H₂ content is further increased. A major cause for this is that the reaction attained equilibrium and reverses, coupled with the dominance of other reactions involved in the methanation process.



Figure 4.2: Effect of hydrogen flow rate on PEME power and molar fraction of methane

In Figure 4.3, the voltage efficiency and net voltage are plotted against the current density, jPEME, of the PEME. The modeling results agree well with the study of Saeed and Warkozek [59], supporting the validity of the present simulation. It is found that the cell potential increases rapidly when current density is less than 300 A/m². When jPEME exceeds 300 A/m², the cell potential increases directly with jPEME. The reverse can be seen with the voltage efficiency of the PEME. At low current density, the overall efficiency is relatively high due to low energy consumption. However, at high current density, higher energy consumption per unit of hydrogen produced occurs with a sharp drop in efficiency. This implies high current densities reduces the performance of the electrolyzer despite high net voltage which results in more power consumption by the PEME, and at moderate current densities, efficiency decreases slightly but remains reasonable.



Figure 4.3: Variation of electrolyzer current density *jPEME* in relation to net voltage and efficiency

Figure 4.4(a) shows the variation of the current density with net voltage, voltage loss, and power density of the SOFC. The current density has significant impact on net voltage since it results in a significant drop in voltage as it increases. This is due to rising voltage losses in the SOFC that are caused by the interaction of activation, ohmic, and concentration losses. The power density of the SOFC increases as current density increases attains a maximum of 3.47 kW/cm² at 8000 A/m² before declining with increasing current density. According to Karimi *et al.*[3], this is due to material deterioration and high thermal losses associated with the SOFC at high current density values. Irrespective of these effects on the SOFC, in Figure 4.4(b), high current densities tend to improve the net power of the plant, \dot{W}_{Net} to an optimal value of 7.68 MW and minimum LCOE of 126.3 \$/MWh at j_{SOFC} of 8000 A/m². This has a direct effect on \dot{W}_{Net} according to Equation (3.91) and is proportionate to variations in the SOFC power output.

The effect of SOFC current density on the specific cost of product and levelized emission is shown in Figure 4.4(c). Increasing SOFC current density led to decreasing specific cost of product and



Figure 4.4(a): Effect of SOFC current density, j_{SOFC} on net voltage, voltage loss, and SOFC power density.



Figure 4.4(b): Effect of SOFC current density, j_{SOFC} on net power and LCOE of the integrated plant



Figure 4.4(c): Effect of SOFC current density, j_{SOFC} on specific cost of product and levelized emission of the integrated plant.

levelized emission up to minimum values of 26.08 \$/GJ and 0.85 kg/MWh, respectively, before reversing. The SOFC is a major component of the topping cycle and high current density value directly increases its power output resulting in low specific cost of product and levelized emission. However, exceeding the maximum current density results in depletion of the electrodes and high heat losses, reducing its efficiency and power output, as well as increasing specific cost and levelized emission.

In Figure 4.5(a), it is demonstrated that \dot{W}_{Net} and η_{ex} first increased to optimal values of 7.9 MW and 39.5%, respectively, before decreasing as the compression pressure ratio of the topping cycle, r_P , is increased. This can be explained by a rise in the power produced by the GT relative to the power used by the compressors when the r_P value is low. The high compression ratio is compromised by the efficiency and necessitates high compression work. Moreover, minimum LCOE occurred at low r_P , with subsequent high values of LCOE, low \dot{W}_{Net} and η_{ex} at high r_P .



Figure 4.5(a): Impact of compression pressure ratio of topping cycle on the plant performance



Figure 4.5(b): Effect of compression pressure ratio on specific cost of product and levelized emission

The effect of pressure ratio on the specific cost of product and levelized emission of the plant is presented in Figure 4.5(b). It was observed that operating the plant under high pressure ratio resulted in high specific cost of product and levelized emission, whereas, under low pressure ratio ranges of 4 - 6, low specific cost of product and levelized emission is attained. This is due to the low power output of the plant under high pressure ratio as the air compression process requires high power input thereby reducing the net power output of the plant. Furthermore, low net power output results in low efficiency with more fuel required to produce the same amount of power, resulting in increased emissions per unit of power generated.

Figure 4.6(a) illustrates how the performance of the plant is affected by the GT inlet temperature. The net power, \dot{W}_{Net} and exergy efficiency, η_{ex} of the plant are enhanced when the GT inlet temperature is raised from 1150 K to 1550 K. This is attributed to the high enthalpy drop rate of the gas turbine, thereby increasing its electrical power output by 36.8 %, and corresponding increase in the net power and exergy efficiency. Additionally, the rise in corresponding net power caused by an increase in the GT inlet temperature also led to a decrease in LCOE. Nevertheless, because of the metallurgical characteristics of the turbine blades, there are restrictions on high turbine inlet temperatures.



Figure 4.6(a): Impact of turbine inlet temperature on exergy efficiency, net power, and levelized cost of electricity

In Figure 4.6(b), the impact of the GT inlet temperature on levelized emission and specific cost of product indicates that increasing combustion temperature from 1150 K to 1550 K resulted in decreasing levelized emission and specific cost of product, respectively. Under high combustion temperatures, there is accelerated fuel-air mixing resulting in much better combustion performance and high net power output, hence, the corresponding reduction in specific cost and levelized emission of the plant.



Figure 4.6(b): Effect of turbine inlet temperature on levelized emission and specific cost of product

The thermal efficiency, $\eta_{th,St}$, and power output, $\dot{W}_{St,Net}$, of the steam turbine unit are plotted against the steam turbine inlet pressure, P_{43} , in Figure 4.7(a). A rise in P_{43} results in a corresponding drop in $\eta_{St,ex}$ from 37.40 % to 32.29 %; however, a decrease in steam turbine power output from 0.90 MW to 0.78 MW is observed. This is explained by the fact that, at high ST inlet pressures, the specific volume of steam decreases irrespective of increase in enthalpy change of the steam turbine. However, Figure 4.7(b) shows that the net power of the plant increased from 7.41 MW to 7.45 MW by 0.5 %. The increase in the integrated plants net power is a contribution from the other power cycles in the plant. A noticeable rise in the LCOE of the plant with increasing steam turbine pressure is observed and is a result of the contributions of the ST unit to the integrated plant LCOE. High contributions to the LCOE would result from a low power output of the ST unit.



Figure 4.7(a): Impact of steam turbine inlet pressure on the net power and thermal efficiency of ST unit



Figure 4.7(b): Variation of steam turbine inlet pressure against net power and levelized cost of electricity of the integrated plant



Figure 4.7(c): Effect of steam turbine inlet pressure on specific cost of product and levelized emissions

The effect of the steam turbine inlet temperature on the specific cost of product and levelized emission of the plant is presented in Figure 4.7(c). High steam turbine inlet pressures yielded low specific cost of product and corresponding high levelized emission. This is due to the reduction in the power of the steam turbine under high inlet temperature conditions, thereby reducing its contribution to the net power of the plant. However, the variation between the highest and lowest values of the specific cost of product and levelized emission is low as expected for low-grade power generating systems.

Figure 4.8 shows the operation of ORC-I under six working fluids and their effects on the overall power of the integrated plant. For the siloxane, a net power and thermal efficiency of 5.19 kW and 28.82 % was achieved with MM, and 5.78 kW and 33.09 % with MDM. With cyclopentane and cyclohexane organic fluids, the power output was 4.69 kW and 5.47 kW, respectively, with thermal efficiencies of 26.84 % and 31.03 %. Power outputs of 3.33 kW and 2.69 kW, and efficiencies of 19.04 % and 15.38 %, respectively, were obtained from refrigerants R1233zd(E) and R600a. In comparison to hydrocarbons and refrigerants, the siloxanes offered better power and efficiency. The distinction in the working fluids thermophysical characteristics and the amount of input to ORC-I are responsible for the variations in the results. However, there is no discernible change in the overall power of the integrated plant considering the impact of the power output of ORC-I.



ORC-I Working fluid

Figure 4.8: Net power of the integrated plant with thermal efficiency and the net power of ORC- I with various working fluids

The performance of ORC-II under constant evaporator pressure, pinch point temperature and degree of superheat is shown in Figure 4.9. Power outputs obtained from using MM and MDM as working fluids were 0.262 MW and 0.108 MW, respectively. With hydrocarbons, cyclopentane and cyclohexane, 0.306 MW and 0.214 MW, were obtained, respectively. In contrast, power outputs of 0.329 MW and 0.272 MW were produced by refrigerants R1233zd(E) and R600a, respectively. However, changes in the integrated plant's net power and LCOE with respect to the different working fluids are then visible in Figure 4.10. For the integrated plant using MDM as the working fluid in ORC-II, a net power of 7.43 MW and an LCOE of 132.3 \$/MWh were obtained. With MM as the working fluid, this improves to 7.59 MW with an LCOE of 130.59 \$/MWh. Cyclohexane produced 7.54 MW and 129.3 \$/MWh, respectively, while cyclopentane produced power outputs of 7.63 MW with an LCOE of 128.1 \$/MWh. The plant produced 7.60 MW of power with an LCOE of 128.7 \$/MWh using R600a, and 7.66 MW of net power and 127.9 \$/MWh when R1233zd(E) was considered. In comparison to refrigerants and siloxanes, the hydrocarbon fluids



ORC-II Working fluid Figure 4.9: ORC-II Power and efficiency under different working fluids



Figure 4.10: Performance of ORC-II with overall power and levelized cost of the plant

produced more power output, even though high efficiency was observed with the high temperature fluids of siloxanes. These differences suggest that not every working fluid choice may be appropriate for the high evaporator pressure. For the ORC-II to produce the best results, an optimization of the input parameters is necessary.



Figure 4.11: Improvement potential of the exergy destroyed in the integrated plant using unavailable (UN) and available (AV) exergy concept.





The improvement potential of the integrated plant is presented in Figure 4.11 and 4.12. In Figure 4.11, the available exergy for improvement of the plant performance is 32 %, with an unavailable exergy destruction of 68 %. This implies that up to 3.07 MW of the initial exergy destruction can be utilized for the improvement of the plant performance. Components with high contribution to

the unavailable and available exergy are presented in Figure 4.12. The results indicate that the unavoidable portion of the exergy destruction within the SOFC, HRSG, CC, and GT is greater than the avoidable part. Additionally, a sizable portion of the rate of exergy destruction in the air compressor (A-Comp), steam turbine (S-T), ORCII turbine (ORC-T2), and ORCII evaporator (EVP2) is avoidable. This indicates more potential for minimizing the exergy losses to improve the performance cycle.

4.8 **Optimization results**

4.8.1 Case 1: Dual objective optimization

Based on the results of the sensitivity analysis, the power plant was optimized using a multiobjective genetic optimization method. As demonstrated in Figure 4.13, the optimization process produced a set of ideal values that were then displayed on Pareto curve with the dual goals of minimizing the LCOE and maximizing exergy efficiency. The two points A and B, which are located on the top-left and bottom-right of the graph, respectively, represent the maximum exergy efficiency and the minimum cost on the Pareto curve. Additionally, these points are consistent with each single objective function's optimal points. According to the results displayed in Table 4.10, the exergy efficiency of the proposed plant can be increased by 22.05 %, from 36.78 % to 44.89 % using R1233zd(E) in ORC-II. This also results in a decrease in LCOE from 130.6 \$/MWh base case to 109.10 \$/MWh and a significant improvement in overall power from 7.42 MW to 9.06 MW. The low-pressure ratio ranges from 5.32 to 5.44 of the topping cycle allowed for a high GT net power output of 4.72 MW to 4.82 MW. This is consistent with Figure 4.5(a), which shows that the low-pressure ratios of 4.5 to 5.5 were within the range at which plants performed at their best. Because of the decrease in pressure ratio, maximum current densities in the SOFC between 7222 A/m² and 7460 A/m² increased the net power of the device by a maximum of 2 %. Considering that high temperatures enhance ideal pressure ratios, the 1341 K flue gases temperature into the gas turbine played a role in its high-power output. The net power of the ST unit increased by 31.46 % from 0.89 kW to 1.17 MW at a pressure of 5838.81 kPa and a pinch point of 16.72 K. The optimized results also showed that the net power of ORC-II increased with R600a, going from 0.098 MW to 0.467 MW. The output of R1233zd(E) was 0.443 MW. Using cyclopentane, cyclohexane, MM, and MDM as working fluids, the following power outputs were achieved: 0.424 MW, 0.429 MW, 0.381 MW, and 0.324 MW. From the optimization analysis, the high-power output of the second ORC using R600a as the working fluid resulted from the optimized pressure factor and superheating.

Decision		Base	Working fluids in ORC-II					
variable	Unit	case	MM	MDM	Cyclopentane	Cyclohexane	R1233zd(E)	R600a
r_P	_	10.5	5.43	5.32	5.44	5.42	5.32	5.38
j _{sofc}	A/m^2	5500	7460.26	7333.22	7334.58	7222.56	7244.20	7318.27
T_{35}	Κ	1273	1334.82	1331.94	1329.04	1333.35	1341.5	1329.52
P ₄₃	kPa	6000	5312.77	6058.39	5979.17	6159.30	5838.81	6828.47
Δpp_{st}	Κ	15	14.38	17.55	13.97	21.69	16.72	17.67
θ	—	0.9	0.2532	0.2056	0.3041	0.2470	0.4156	0.6814
Δpp_{ORCII}	Κ	10	5.14	7.78	6.35	6.02	5.04	5.43
Δsh_{ORCII}	Κ	10	4.33	4.98	4.75	5.64	6.54	7.23
Objective	a							
	S M337	2 59	264	2.62	264	2 628	2.62	2 626
VV _{SOFC}		2.30	2.04	2.05	2.04	2.030	2.05	2.030
W_{GT}	IVI W	3.850	4./89	4.//	4.76	4./8	4.82	4.72
W_{St}	MW	0.89	1.17	1.13	1.13	1.097	1.17	1.087
Ŵ _{orcii}	MW	0.098	0.381	0.324	0.424	0.429	0.443	0.467
\dot{W}_{Net}	MW	7.42	8.98	8.85	8.96	8.94	9.06	8.91
η_{th}	%	44.93	54.37	53.59	54.20	54.12	54.84	53.94
η_{ex}	%	36.78	44.5	43.87	44.37	44.30	44.89	44.15
LCOE	\$/MWh	130.60	109.50	110.60	109.90	110.0	109.10	110.60
NPV	G\$	1.133	1.384	1.364	1.380	1.378	1.397	1.373
BEP	yr	6.51	5.45	5.51	5.47	5.49	5.43	5.05
LEco ₂	kg/MWh	0.883	0.724	0.735	0.727	0.728	0.718	0.730
ESI	-	1.58	1.80	1.78	1.80	1.80	1.82	1.79

Table 4.10: Values of the objective function and decision variables with various ORC-II working fluids from the optimal point A for Case 1.

4.8.2 Case 2: Tri-objective optimization

A tri-objective optimization that aims to maximize exergy efficiency, while minimizing the specific cost of product and the levelized cost of electricity is implemented with MOGA. As displayed in Figure 4.14, a Pareto front curve containing the three objective functions was created from a set of optimal values. Points A and B on the 3-D graph represent the maximum exergy efficiency and the minimum costs, respectively, on the Pareto front curve. For the single objective functions of exergy

efficiency, specific cost of product and the levelized cost of electricity, their optimal values are displayed at points A and B.



Figure 4.13(a): Optimal solution for the integrated plant with MM as working fluids in ORC-II



Figure 4.13(b): Optimal solution for the integrated plant with MDM as working fluids in ORC-II



Figure 4.13(c): Optimal solution for the integrated plant with Cyclopentane as working fluids in ORC-II



Figure 4.13(d): Optimal solution for the integrated plant with Cyclohexane as working fluids in ORC-II



Figure 4.13(e): Optimal solution for the integrated plant with R1233zd(E) as working fluids in ORC-II



Figure 4.13(f): Optimal solution for the integrated plant with R600a as working fluids in ORC-II

The outcome of the tri-objective optimization displayed in Table 4.11 implies a 22.7 % increase in the exergy efficiency of the plant from 36.78 % to 45.08 % with R1233zd(E) in ORCII. Furthermore, a decrease in the LCOE to 109.02 \$/MWh from an initial value of 130.60 \$/MWh is revealed. The specific cost of product, $c_{P,Total}$, was minimized from 27.01 \$/GJ to 22.52 \$/GJ, as well as the net power which increased from 7.42 MW to 9.10 MW. Additionally, the LE_{CO2} reduced by 18 % from 0.8835 kg/MWh to 0.715 kg/MWh and the ESI increased to 1.82 (-) due to increase in the exergy efficiency of the plant.

		D						
Decision	Base			Working fluids in ORC-II				
variable	Unit	case	MM	MDM	Cyclopentane	Cyclohexane	R1233zd(E)	R600a
r_P	—	10.5	5.33	5.56	5.45	5.44	5.31	5.43
<i>j_{sofc}</i>	A/m^2	5500	7219.75	7357.85	7382.89	7106.70	7375.92	7389.15
T_{35}	Κ	1273	1332.83	1341.83	1337.45	1334.03	1347.82	1326.64
P_{43}	kPa	6000	6475.24	6513.21	6547.47	6328.71	6486.50	5566.65
Δpp_{st}	Κ	15	10.12	20.25	13.11	17.31	18.44	13.54
θ	_	0.9	0.2722	0.2723	0.3221	0.2309	0.3887	0.5285
Δpp_{ORCII}	Κ	10	5.05	8.08	5.96	6.03	4.94	7.16
Δsh_{ORCII}	Κ	10	4.27	5.16	6.85	4.98	6.02	5.10
Objective	S							
₩ _{sofc}	MW	2.58	2.63	2.65	2.64	2.64	2.63	2.64
\dot{W}_{GT}	MW	3.856	4.77	4.84	4.80	4.813	4.86	4.72
Ŵst	MW	0.89	1.15	1.10	1.14	1.11	1.14	1.14
₩ _{ORCII}	MW	0.098	0.409	0.297	0.449	0.427	0.470	0.390
Ŵ _{Net}	MW	7.42	8.97	8.89	9.03	8.99	9.10	8.89
η_{th}	%	44.93	54.29	53.78	54.64	54.22	55.07	53.78
η_{ex}	%	36.78	44.44	44.02	44.73	44.38	45.08	44.03
LCOE	\$/MWh	130.60	109.60	111.30	109.60	110.10	109.02	110.50
C _{P,Total}	\$/GJ	27.01	22.62	22.97	22.62	22.74	22.52	22.82
NPV	G\$	1.133	1.382	1.368	1.391	1.380	1.403	1.369
BEP	yr	6.51	5.46	5.54	5.46	5.84	5.43	5.50
LE _{CO2}	kg/MWh	0.883	0.725	0.732	0.721	0.726	0.715	0.732
ESI	-	1.58	1.80	1.79	1.81	1.80	1.82	1.79

Table 4.11: Objective function and decision variables with working fluids in ORC-II at optimal point A from tri-objective optimization in Case 2.



Figure 4.14(a): Pareto optimal solution of the tri-objective optimization with MM as working fluid in ORC-II



Figure 4.14(b): Pareto optimal solution of the tri-objective optimization with MDM as working fluid in ORC-II



Figure 4.14(c): Pareto optimal solution of the tri-objective optimization with Cyclopentane as working fluid in ORC-II



Figure 4.14(d): Pareto optimal solution of the tri-objective optimization with Cyclohexane as working fluid in ORC-II



Figure 4.14(e): Pareto optimal solution of the tri-objective optimization with R1233zd(E) as working fluid in ORC-II



Figure 4.14(f): Pareto optimal solution of the tri-objective optimization with R600a as working fluid in ORC-II

In order to provide better understanding of the variation of the decision variables, scattered distribution of the decision variables is shown in Figures 4.15 - 4.22. The results show that pressure ratio, SOFC current density, and the gas turbine inlet temperature tend towards the lower range. This implies that a reduction in these parameters leads to better optimization results of both objective functions in the multi-objective optimization.



Figure 4.15: Scattered distribution of optimum decision variables for SOFC-GT pressure ratio



Figure 4.16: Scattered distribution of optimum decision variables for SOFC current density



Figure 4.17: Scattered distribution of optimum decision variables for Gas turbine inlet temperature



Figure 4.18: Scattered distribution of optimum decision variables for Steam turbine inlet pressure
ORC II pressure factor, pinch point and degree of superheat in Figure 4.20 - 4.22, indicates scattered distributions within their allowable domain, implying that these parameters possess important effects on the trade-off between the exergy efficiency and levelized cost of electricity. The design parameters generated within their maximum domain show that they do not imply a conflict between the objective functions, and increasing such parameters leads to improvement in both objective functions.



Figure 4.19: Scattered distribution of optimum decision variables for Steam turbine pinch point temperature



Figure 4.20: Scattered distribution of optimum decision variables for ORC-II Pressure factor



Figure 4.21: Scattered distribution of optimum decision variables for ORC-II Pinch point temperature



Figure 4.22: Scattered distribution of optimum decision variables for ORC-II Degree of Superheat



Figure 4.23(a): Percentage distribution of exergy destruction rate in components of the plant before optimization



Figure 4.23(b): Percentage distribution of exergy destruction rate in components of the plant after optimization using R1233zd(E) as working fluid in ORC-II

Figure 4.23 summarizes the exergy destruction results on a plant component basis for the proposed system. In Figure 4.23(a), the combustion chamber (CC) accounts for 57 % of the total energy destroyed, with the SOFC-GT unit contributing the most at 89 %. The contribution of the SOFC-GT unit drops to 80% in Figure 4.23(b), which is derived from Point A in Figure 8(e). After optimization, the largest contribution of 37 % emanates from the SOFC due to high thermal losses associated with the fuel cell at high values of current density. Despite this challenge, the net power of the plant and efficiencies of the plant are improved to optimal values, in agreement with Figure 4.4(a). Furthermore, CC dropped to 11 % and GT increased to 25 %. The heat recovery vapour evaporator in ORC-II (EVP2) and HRSG both were increased to 5 %, respectively. These adjustments suggest potential interactions between the parameters of the subsystem to achieve peak performance.



Figure 4.24(a): Plant component contribution to the total cost associated with destruction exergy for the base case



Figure 4.24(b): Plant component contribution to the total cost associated with destruction exergy for optimum case with R1233zd(E)

The contribution of plant components to the total cost of exergy destruction for the base, and optimum cases is shown in Figure 4.24(a) and 4.24(b), respectively. In both cases, SOFC-GT contributed the largest portion to the total cost associated with destruction of exergy irrespective of the reduction from 75 % to 62 % contribution. On component basis, the contribution of the combustion chamber (CC) reduced from 41 % to 8 % after optimization, whereas the solid oxide fuel cell (SOFC) increases from 5 % to 21 %. The increase in contribution of the SOFC component is due to the increase in the current density of the cell as high current densities can result in material deterioration and high thermal defects as well as high thermal energy interaction with the environment.



ORC-II Working Fluid

Figure 4.25: Specific total cost of product and total cost of exergy destroyed for base case and optimized cases in ORC-II

The specific total cost of product, and total cost of exergy destruction for the base case with MM as working fluid in the ORC plants and the optimized cases with different ORC working fluids is presented in Figure 4.25. From initial costs of 27.01 \$/GJ in the base case using MM*, the specific total cost of product was improved by 16.68 % after the optimization with R1233zd(E) as working fluid in ORC-II, whereas total cost of exergy destruction, was reduced by 25.17 % to 244.77 \$/hr. The optimization case with MDM as working fluid returned the lowest total cost of exergy destruction of 231.02 \$/hr, with a specific total cost of product 22.97 \$/GJ. However, with consideration of other factors such as net power output, efficiency and LCOE, as in Table 4.11, this was not selected as the optimum case for the plant.



Figure 4.26: Subsystem cost distribution on capital cost for the base and optimum cases using MM and R1233zd(E) working fluid



Figure 4.27: Subsystems share of levelized cost of electricity of the base and optimum cases using MM and R1233zd(E) working fluid

The subsystems cost distribution on capital cost for the base and optimum cases are presented in Figure 4.26. Except for ORC-II, there is very little difference between the base case and optimal cases for the BCU, SOFC-GT, and ST. This indicates that there is probably little cost impact on the subsystems from an increase in the capital cost of the integrated plants brought on by an increase in plant capacity. The cost share of the various plant sections on the levelized cost of electricity for the base case and optimal scenarios with MM and R1233zd(E) is displayed in Figure 4.27. Following optimization using MM and R1233zd(E), the cost shares related to the BCU, SOFC-GT, and ST units are decreased. On the other hand, the cost share increased in ORC-II in the optimized condition. Therefore, the reduction of the total levelized cost of electricity in Figure 4.24(b) and the increased power outputs and exergy efficiency from the optimized cases, as indicated in Table 4.9, demonstrate the benefits of optimization.

Lastly, Table 4.12 provides a comparison between the findings of this study and those of other research with integrated subsystems in the literature. With a maximum power and energy efficiency of 9.05 MW and 44.87 %, respectively, the suggested configuration in this study is competitive with other studies with multiple outputs ranging from 4 MW to 16 MW and 42 % to 59 %. The integrated plant can produce a balanced, positive energy source based on the comparison.

Author	Year	Configuration	Output
Balafkandeh et	2019	Anaerobic digestion,	Exergy efficiency of 47.8 %,
<i>al.</i> [143]		gasification, GT,	Specific cost of product
		Supercritical CO ₂ ,	5.44 \$/GJ.
		Absorption cooling	
Yilmaz et al.[109]	2019	Biomass gasification,	Net power 15 MW, 0.072 kg/s
		GT, PEME, Kalina	hydrogen mass flow rate,
		cycle, reverse osmosis	Thermal efficiency of 63.84 %
		unit, Absorption	and exergy efficiency 59.26 %,
		refrigeration cycle, heat	Cooling capacity of 4.36 MW.
		pump and dryer.	Heat rate of 5 MW
			Exergy cost rate of 2000 \$/s

Table 4.12: Comparison between the results of this study and other biomass integrated subsystems

Holagh <i>et al</i> .[96]	2020	CH4 fueled SOFC-GT, biomass combustion, ORC, desalination, PEME.	Net power 4.4 MW, and 71 % SOFC contribution. Cooling potential of 0.16 MW. 0.00155 kg/s mass flow rate of hydrogen, 0.96 kg/s of H ₂ O. Thermal efficiency of 77.6 % and Exergy efficiency 47.1 %, Cost per unit exergy 11.28 \$/GJ.
Al-Rashed and Afrand[141]	2021	Biomass, digestion, GT, Supercritical CO ₂ , and Absorption refrigeration cycle.	Total power of 16.5 MW, Exergetic efficiency 56.69 %, Specific cost of product 5.65 \$/GJ.
He et al.[106]	2022	Biomass gasification, Stirling cycle, GT, Water heater, and Subcritical CO ₂ cycle.	Net Power of 8.9 MW. Energy efficiency of 71.13 % Exergy efficiency 42.03 %, Heat capacity of 4.24 MW, Specific cost rate of 10.94 \$/GJ.
Present work	2023	Combined biowastes, Gasification and anaerobic digestion, methanation, PEME, SOFC-GT-ST-ORC	Overall power output of 9.10 MW and 29 % contribution from SOFC, 0.0023 kg/s mass flow rate of hydrogen. Improved LHV(syngas) of 33.48 MJ/kg Energy efficiency 55.07 % and Exergy efficiency 45.08 %, LCOE of 109.02 \$/hr, Specific cost of product 22.52 \$/GJ

5 Summary and outlook

5.1 Summary

In this work, an integrated biomass-fueled configuration designed to meet energy need has been assessed under thermo-economic principles. The goal was to develop a biomass-to-power conversion scheme with a syngas upgrade and conduct thermo-economic and optimization analysis for optimal performance and minimal cost under the environmental conditions of a tropical region. The objectives defined to achieve this aim in Chapter 1 are stated as

- i. To develop, in a tropical area, an integrated biomass conversion system based on the anaerobic digestion (AD) and gasification of agricultural waste for the generation of power, and hydrogen production (used in syngas upgrade).
- To analyze the overall system performance of integrating different thermodynamic and electro-chemical systems of anaerobic digestion and gasification, methanation, organic Rankine cycle (ORC), Brayton cycle, Rankine cycle, solid oxide fuel cell (SOFC), proton exchange membrane electrolyzer (PEME).
- iii. Determination of the benefits of employing the integrated plant from a thermodynamic, economic, and environmental point of view.
- iv. To investigate the impact of selected operating parameters, including the choice of ORC working fluid on the system outputs.
- v. Perform a multi-objective optimization of the power plant to improve the integrated plant performance with respect to net power output, levelized cost of electricity and specific cost of products.

Concerning the first objective, a hybrid biomass conversion system to fuel, and combined power cycles for power generation was developed in Chapter 3. The hybrid conversion system consisted of anaerobic digester, and gasifier for the conversion of animal and crop wastes into biogas and syngas, respectively. Waste heat from the cooling of syngas was converted in an ORC configuration for power generation, which was optimized by considering six different working fluids of MM. MDM, Cyclopentane, Cyclohexane, R1233zd(E), and R600a. A Sabatier reactor

was introduced for an upgrade to synthetic natural gas quality using hydrogen from a proton exchange membrane electrolyzer. Furthermore, the biogas from the digester and the upgraded synthetic gas were utilized in an integration of power cycles of solid oxide fuel cell, a gas turbine, steam turbine unit, and a second ORC for power generation.

To determine the performance of the plant, measured in terms of increasing exergy efficiency, decreasing levelized cost of electricity and specific cost of product, sustainability index, net present value, and levelized emission, thermo-economic, exergo-economic and environmental models were applied and the results, with parametric studies are reported in Chapter 4. Furthermore, a parametric analysis was performed to assess the influence of design parameters on the overall plant performance and its improvement potential.

Lastly, multi-objective optimization was applied to the combined power plant using a genetic algorithm. Using the LINMAP decision-making method, the optimal objectives are traded off to determine the desired working conditions.

5.2 Conclusion

The following succinctly describes the findings in this work:

- i. Using a methanation process, 0.0028 kg/s of hydrogen produced with a 0.45 MW power input from the SOFC was utilized to upgrade the syngas.
- ii. With MDM as the working fluid in ORC-I, a maximum power output of 5.78 kW is achieved.
- iii. Refrigerants R1233zd(E) and R600a, generated a net power of 0.472 MW and outperformed siloxanes (MM and MDM) and hydrocarbons (cyclopentane and cyclohexane) as working fluids in ORC-II. Nonetheless, the net power and LCOE of the integrated plant were used to determine the R1233zd(E) working fluid selection.
- iv. Parametric investigations showed that at low temperature ranges between 423 673 K, products of CH₄ and H₂O can be achieved in the methanation process. The low-pressure ratio in the compression process in the topping cycle resulted in high net power output and exergy efficiency, LCOE, levelized emission and high sustainability index. High current density values in the SOFC are favorable for high plant performance. With increasing turbine inlet temperature, low LCOE, specific cost of product and levelized emission are

obtained. Nonetheless, the reverse was observed for the net power output and exergy efficiency of the plant.

- v. The results of the conventional and enhanced exergo-economic analyses, respectively showed that the combustion chamber is a major component for improvement, with 39 % of its exergy destruction available for recovery. The enhanced exergo-economic analyses further revealed that 32 % of the total exergy destruction and 36 % of the associated cost is avoidable.
- vi. The net power produced by the SOFC, GT, ST, and ORC-II were increased to 2.63 MW,
 4.86 MW, 1.14 MW, and 0.472 MW, respectively, after optimization. This adds up to a total power of 9.10 MW for the integrated plant.
- vii. At the system's optimal points, the overall energy and energy efficiencies were found to be 55.07 % and 45.08 %, respectively.
- viii. Following optimization, the LCOE was reduced from an initial 130.6 \$/MWh to 109.02 \$/MWh.
- ix. The specific cost of product was reduced from 27.01 \$/GJ for the base case to 22.52 \$/GJ with the tri-objective optimization.
- x. Net present value of 1.403 G\$ with breakeven point of 5.43 years after optimization was obtained using R1233zd(E) as working fluid in ORC-II.
- xi. The levelized CO₂ emission, and exergy sustainability index of the plants improved from initial values of 0.883 kg/MWh, and 1.58 to 0.715 kg/MWh, and 1.82, respectively.

5.3 Thesis contribution

- i. The conceptual design of a hybrid biomass conversion unit integrating a digester, gasifier, proton exchange electrolyzer, Sabatier unit, and organic Rankine cycle.
- ii. Development of an integrated biomass-fueled configuration for power generation, with increased plant performance and low environmental impact.
- iii. Identification of improvement potentials of the plant through advanced exergo-economic assessments and optimization with genetic algorithm.

5.4 Outlook

The following recommendations are suggested for future work

- i. Off-design analysis should be considered for the power plant configuration in order to ascertain the utilization of the plant under universal conditions.
- ii. Component based optimization should be carried out to minimize the cost contributions of components to the levelized cost as well as reduce losses related to exergy.
- iii. ORC-II may be fully utilized for provision of power to the PEME in a future study as this may reduce the cost associated with the destruction of exergy in the BCU.
- iv. The consideration of a fuel cell/electrolyzer such as a solid oxide fuel cell and solid oxide electrolyzer (SOFC-SOEC) can be considered in the topping cycle of SOFC-GT and the results compared to that of the present configuration. This can provide possible answers to questions on the power consumption of the PEME, amount of hydrogen produced for methanation and system efficiency.
- v. The production of other products other than electricity should be considered to enhance the performance of the plant. Considering the locality of the proposed plant is an agro-facility, cooling products and process heat generation can be considered in place of power generation to meet other demands for thermal energy.

5 Zusammenfassung und ausblick

5.1 Zusammenfassung

In dieser Arbeit wurde eine integrierte, mit Biomasse betriebene Konfiguration zur Deckung des Energiebedarfs nach thermoökonomischen Grundsätzen bewertet. Ziel war es, ein System zur Umwandlung von Biomasse in Energie mit einem Syngas-Upgrade zu entwickeln und eine thermoökonomische und Optimierungsanalyse für optimale Leistung und minimale Kosten unter den Umweltbedingungen einer tropischen Region durchzuführen. Die zur Erreichung dieses Ziels in Kapitel 1 festgelegten Ziele lauten wie folgt

- Entwicklung eines integrierten Biomassekonversionssystems in einem tropischen Gebiet, das auf der anaeroben Vergärung (AD) und der Vergasung von landwirtschaftlichen Abfällen zur Stromerzeugung und zur Wasserstoffproduktion (für die Synthesegasaufbereitung) basiert.
- Analyse der Gesamtsystemleistung der Integration verschiedener thermodynamischer und elektrochemischer Systeme der anaeroben Vergärung und Vergasung, der Methanisierung, des organischen Rankine-Zyklus (ORC), des Brayton-Zyklus, des Rankine-Zyklus, der Festoxidbrennstoffzelle (SOFC) und des Protonenaustauschmembran-Elektrolyseurs (PEME).
- iii. Bestimmung der Vorteile des Einsatzes der integrierten Anlage aus thermodynamischer, wirtschaftlicher und ökologischer Sicht.
- iv. Untersuchung der Auswirkungen ausgewählter Betriebsparameter, einschließlich der Wahl des ORC-Arbeitsmediums, auf die Systemleistung.
- Durchführung einer Mehrzieloptimierung des Kraftwerks, um die Leistung der integrierten Anlage im Hinblick auf die Nettoleistung, die Stromgestehungskosten und die spezifischen Kosten der Produkte zu verbessern.

Im Hinblick auf das erste Ziel wurde in Kapitel 3 ein hybrides System zur Umwandlung von Biomasse in Kraftstoffe und kombinierte Kraft-Wärme-Kopplung zur Stromerzeugung entwickelt. Das hybride Umwandlungssystem bestand aus einem anaeroben Fermenter und einem Vergaser für die Umwandlung von tierischen und pflanzlichen Abfällen in Biogas bzw. Synthesegas. Die Abwärme aus der Abkühlung des Synthesegases wurde in einer ORC-Konfiguration zur Stromerzeugung umgewandelt, die unter Berücksichtigung von sechs verschiedenen MM-Arbeitsflüssigkeiten optimiert wurde. MDM, Cyclopentan, Cyclohexan, R1233zd(E), und R600a. Ein Sabatier-Reaktor wurde für eine Aufrüstung auf synthetische Erdgasqualität unter Verwendung von Wasserstoff aus einem Protonenaustauschmembran-Elektrolyseur eingeführt. Darüber hinaus wurden das Biogas aus dem Fermenter und das aufbereitete synthetische Gas in einem integrierten Energiekreislauf aus einer Festoxid-Brennstoffzelle, einer Gasturbine, einer Dampfturbineneinheit und einem zweiten ORC zur Stromerzeugung verwendet.

Zur Bestimmung der Leistung der Anlage, gemessen an der Steigerung des Exergie-Wirkungsgrades, der Senkung der Stromgestehungskosten und der spezifischen Produktkosten, des Nachhaltigkeitsindexes, des Kapitalwerts und der Emissionen, wurden thermoökonomische, exergoökonomische und ökologische Modelle angewandt, deren Ergebnisse zusammen mit Parameterstudien in Kapitel 4 dargestellt werden. Darüber hinaus wurde eine parametrische Analyse durchgeführt, um den Einfluss der Auslegungsparameter auf die Gesamtleistung der Anlage und ihr Verbesserungspotenzial zu bewerten.

Schließlich wurde für das Kombikraftwerk eine Mehrzieloptimierung mit Hilfe eines genetischen Algorithmus durchgeführt. Mit Hilfe der LINMAP-Entscheidungsmethode werden die optimalen Ziele gegeneinander abgewogen, um die gewünschten Arbeitsbedingungen zu bestimmen.

5.2 Schlussfolgerung

Es folgt eine kurze und prägnante Beschreibung der Ergebnisse dieser Arbeit:

- Mit Hilfe eines Methanisierungsprozesses wurden 0,0028 kg/s Wasserstoff, der von der SOFC mit einer Leistungsaufnahme von 0,45 MW erzeugt wurde, zur Aufbereitung des Synthesegases verwendet.
- ii. Mit MDM als Arbeitsmittel im ORC-I wird eine maximale Leistung von 5,78 kW erreicht.
- iii. Die Kältemittel R1233zd(E) und R600a erzeugten eine Nettoleistung von 0,472 MW und übertrafen die Siloxane (MM und MDM) und Kohlenwasserstoffe (Cyclopentan und Cyclohexan) als Arbeitsmittel in ORC-II. Nichtsdestotrotz wurden die Nettoleistung und die LCOE der integrierten Anlage verwendet, um die Auswahl des Arbeitsmittels R1233zd(E) zu bestimmen.

- iv. Parametrische Studien haben gezeigt, dass bei niedrigen Temperaturbereichen zwischen 423 und 673 K die Produkte CH4 und H2O im Methanisierungsprozess gewonnen werden können. Das niedrige Druckverhältnis beim Verdichtungsprozess im Topping-Zyklus führte zu einem hohen Nettoenergie- und Exergie-Wirkungsgrad, LCOE, nivellierten Emissionen und einem hohen Nachhaltigkeitsindex. Hohe Stromdichtewerte in der SOFC sind vorteilhaft für eine hohe Systemleistung. Mit steigender Turbineneintrittstemperatur werden niedrige LCOE, spezifische Produktkosten und gleichbleibende Emissionen erreicht.
- v. Die Ergebnisse der konventionellen und der erweiterten exergoökonomischen Analyse zeigten, dass die Brennkammer eine wichtige Komponente für Verbesserungen ist, wobei 39 % der Exergievernichtung für eine Rückgewinnung zur Verfügung stehen. Die erweiterten exergoökonomischen Analysen ergaben ferner, dass 32 % der gesamten Exergievernichtung und 36 % der damit verbundenen Kosten vermeidbar sind.
- vi. Die von SOFC, GT, ST und ORC-II erzeugte Nettoleistung konnte nach der Optimierung auf 2,63 MW, 4,86 MW, 1,14 MW bzw. 0,472 MW gesteigert werden. Dies ergibt eine Gesamtleistung von 9,10 MW für die integrierte Anlage.
- vii. An den optimalen Punkten des Systems wurden Gesamtenergie- und Energiewirkungsgrade von 55,07 % bzw. 45,08 % ermittelt.
- viii. Nach der Optimierung konnten die Stromgestehungskosten von ursprünglich 130,6 \$/MWh auf 109,02 \$/MWh gesenkt werden.
- ix. Die spezifischen Produktkosten wurden durch die Drei-Ziel-Optimierung von 27,01 \$/GJ im Basisfall auf 22,52 \$/GJ gesenkt.
- x. Ein Kapitalwert von 1,403 G\$ mit einem Break-even-Punkt von 5,43 Jahren nach der Optimierung wurde bei Verwendung von R1233zd(E) als Arbeitsmedium in ORC-II erreicht.
- xi. Die nivellierten CO₂-Emissionen und der Exergie-Nachhaltigkeitsindex der Anlagen verbesserten sich von anfänglichen Werten von 0,883 kg/MWh und 1,58 auf 0,715 kg/MWh bzw. 1,82.

5.3 Beitrag der Dissertation

- Konzeptionelles Design einer hybriden Biomasseumwandlungseinheit, die einen Fermenter, einen Vergaser, einen Protonenaustauschelektrolyseur, eine Sabatier-Einheit und einen organischen Rankine-Zyklus integriert.
- Entwicklung einer integrierten biomassebetriebenen Konfiguration zur Stromerzeugung mit erhöhter Anlagenleistung und geringer Umweltbelastung.
- iii. Identifizierung von Verbesserungspotenzialen der Anlage durch erweiterte exergoökonomische Bewertungen und Optimierung mit genetischen Algorithmen.

5.4 Ausblick

Für zukünftige Arbeiten werden folgende Empfehlungen vorgeschlagen:

- i. Für die Kraftwerkskonfiguration sollte eine Off-Design-Analyse in Betracht gezogen werden, um die Anlagenauslastung unter universellen Bedingungen sicherzustellen.
- Eine komponentenbasierte Optimierung sollte durchgeführt werden, um den Kostenbeitrag der Komponenten zu den nivellierten Kosten zu minimieren und die mit Exergie verbundenen Verluste zu reduzieren.
- iii. ORC-II kann in einer zukünftigen Studie vollständig zur Stromversorgung des PEME genutzt werden, da es die mit der Exergievernichtung in der BCU verbundenen Kosten reduzieren kann.
- iv. Die Berücksichtigung einer Brennstoffzelle/eines Elektrolyseurs wie einer Festoxidbrennstoffzelle und eines Festoxidelektrolyseurs (SOFC-SOEC) kann im Topping-Zyklus von SOFC-GT in Betracht gezogen und die Ergebnisse mit denen der aktuellen Konfiguration verglichen werden. Dies kann mögliche Antworten auf Fragen zum Stromverbrauch des PEME und der Menge des für die Methanisierung und Systemleistung erzeugten Wasserstoffs liefern.
- v. Die Produktion anderer Produkte als Elektrizität sollte in Betracht gezogen werden, um die Leistung der Anlage zu verbessern. Da es sich bei der geplanten Anlage um einen landwirtschaftlichen Betrieb handelt, kommt anstelle der Stromproduktion auch die Produktion von Kälteprodukten und Prozesswärme zur Deckung des sonstigen Wärmeenergiebedarfs in Frage.

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List of Publications

- 1. **Oreva Joe Ogorure**, Florian Heberle, Dieter Brüggemann, Thermodynamic analysis of a combined organic Rankine cycle (ORC) with proton exchange membrane electrolyzer (PEME) in an integrated biomass conversion system, 2021, ORC2021 Proceedings, Munich
- 2. **Ogorure OJ**, Heberle F, Brüggemann D. Thermo-economic analysis and multi-criteria optimization of an integrated biomass-to-energy power plant. Renew Energy 2024; 224:120112. https://doi.org/10.1016/j.renene.2024.120112