



University of Sheffield

Feasibility Assessment of Municipal Waste Management System for Energy and Fuel Generation

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DECLARATION

The candidate confirms that the work submitted is his own and that appropriate credit has been given where reference has been made to the work of others.

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RESEARCH OUTPUTS

List of publications:

1. Q. A. M. O. Arifianti, S. Michailos, M. F. Rojas-Michaga, K. Rabea, K. J. Hughes, Lin Ma, D. Ingham, M. Pourkashanian, **“Conceptual design and thermodynamic investigation of novel energy and fuel generation systems from municipal waste coupled with carbon capture and storage”**. *Energy Nexus*, p.100460. (<https://doi.org/10.1016/j.nexus.2025.100460>).
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ABSTRACT

The transition to net zero requires integrated waste-to-energy (WtE) pathways that can simultaneously improve waste management, strengthen energy security, and mitigate carbon emissions. Plastic and food waste have different properties and therefore require different conversion technologies: plastic waste can be converted through plasma gasification, whereas food waste can be treated through anaerobic digestion. Integrating these technologies provides opportunities to produce multiple energy products while avoiding the environmental impacts associated with conventional waste disposal. However, the technical, economic, and environmental implications of integrating these waste conversion technologies with carbon capture and storage (CCS) remain insufficiently understood, particularly in developing economies. Therefore, this thesis investigates integrated plasma gasification, anaerobic digestion, and CCS systems to produce electricity, cooling, heating, liquefied biomethane, and hydrogen, using Indonesia as the study context.

Process models were developed in Aspen Plus[®] using a consistent plastic and food waste feedstock. The systems were evaluated through energy and exergy analyses, techno-economic analysis (TEA), and life cycle assessment (LCA). The investigated configurations included combined cooling, heating, and power (CCHP), liquefied biomethane production, and hydrogen co-production under no-CCS, pre- and post- combustion CCS conditions.

The results demonstrate the technical and economic feasibility and environmental potential of the integrated systems. CCHP-only configurations produced more electricity, heating, and cooling, while liquefied biomethane configurations achieved higher energy and exergy efficiencies. CCS increased production costs but substantially reduced global warming potential, with some configurations achieving net-negative emissions through biogenic CO₂ storage. Separate waste processing achieved the highest hydrogen yield, thermodynamic efficiencies, and lowest hydrogen cost, whereas co-gasification provided the greatest emissions reduction. However, current Indonesian incentives remain insufficient to ensure economic competitiveness. Pilot-scale validation, cost reductions, and stronger policy and

financial support are therefore required to enable commercial deployment in developing economies.

Keywords: Waste-to-energy; Plasma gasification; Anaerobic digestion; Carbon capture and storage; CCHP; Liquid biomethane; Hydrogen; Techno-economic analysis; Life cycle assessment

NOMENCLATURE

AD	anaerobic digester
AFR	air fuel ratio
ARC	absorption refrigeration chiller
BL	blowdown loss
CAPEX	capital expenditure
CC	combustion chamber
CCGT	combined-cycle gas turbine
CCHP	combined cooling heating power
CCS	carbon capture storage
CEPCI	chemical engineering plant cost index
COC	cycles of concentration
COP	coefficient of performance
DCFA	discounted cash flow analysis
DGO	direct and general overhead
DL	drift loss
EL	evaporation loss
ELECNRTL	electrolyte non-random two-liquid
ER	equivalence ratio
FCI	fixed capital investment
FOM	fixed operating, and maintenance costs
GHSV	gas hourly space velocity
GT	gas turbine
HP	high-pressure
HPWS	high Pressure water scrubber
HRSG	heat recovery steam generator
HTR	hot temperature reactor
HTWGSR	high temperature water gas shift reactor
HTZ	hot temperature zone
IC	Indirect cost
IDC	installed direct costs
IP	intermediate pressure
IRR	internal rate of return
ITF WC	insurance, tax and financing working capital
LCA	life cycle assessment
LCOE	levelized cost of electricity
LCOH	levelized cost of hydrogen
LCOWT	levelized cost of waste treated
LCObM	levelized cost of liquid biomethane
LHV	lower heating value
LP	low-pressure
LS	labour and supervision
LSL	lean solvent loading

LTR	low temperature reactor
LTWGSR	low temperature water gas shift reactor
LTZ	low temperature zone
Li-Br	lithium bromide
MCFC	molten-carbonate fuel cell
ML	methane liquefier
MLM	maintenance labour and materials
MSW	municipal solid waste
MW	megawatt
NIDC	non-installed direct costs
NPV	net present value
NRTL	non-random two-liquid
OPEX	operating expenditure
PEC	purchased equipment cost
PENG-ROB	Peng-Robinson
PG	plasma gasifier
PRBM	Peng–Robinson Boston–Mathias
PSA	pressure swing adsorption
RSL	rich solvent loading
SAMR	steam to air mass ratio
SFR	steam to feedstock ratio
SOFC	solid oxide fuel cells
SRD	specific reboiler duty
SRK	the Soave–Redlich–Kwon
ST	steam turbine
T&S	transportation and storage
TCR	total capital requirement
TDC	total direct costs
VC	variable cost
WC	working capital
WGS	water gas shift
WGSR	water gas shift reaction
XFHE	cross flow heat exchanger
wt	weight

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1 INTRODUCTION

This chapter introduces the fundamental aspects of the proposed research and comprises several sections, which describe the current status and challenges of the energy sector and the development of municipal waste utilization. Lastly, also the main and specific objectives of this research work are displayed.

1.1 Overview of global waste generation and energy demand challenges

Municipal solid waste (MSW) generation is on track to surge from 2.3 billion tonnes in 2023 to roughly 3.8 billion tonnes by 2050 [1]. This growth is accompanied by variations in waste composition: lower-income countries produce proportionally more food and green waste, while higher-income nations discard larger volumes of paper, cardboard, and plastics [2]. Food loss and waste alone exacerbate the problem, an estimated 14 percent of food is lost before it even reaches retail outlets [3], and a further 17 percent is wasted at the retail and consumer stages [4]. When food ends up in landfills, its decomposition emits methane, a greenhouse gas many times more potent than CO₂.

Plastic waste presents a parallel challenge. In 2020, the world generated some 360 million tonnes of plastic refuse, a figure expected to climb to 500 million tonnes by 2030. Despite numerous recycling efforts, only about 9 percent of all plastic is actually recycled; the remainder, approximately 70 percent, accumulates in landfills or the natural environments [5], [6]. Landfilled waste can lead to groundwater contamination, slope failures at disposal sites, flooding, and methane emissions [7]. In fact, landfill operations are responsible for an estimated 16 percent of global greenhouse-gas emissions, making them the second-largest source of anthropogenic methane [8]. Uncontrolled disposal also creates breeding grounds for disease vectors, such as mosquitoes in discarded tires and containers, raising the risk of illnesses like dengue fever [9]. Moreover, visible waste accumulation can undermine local economies by deterring tourism [10]. Together, these trends underscore the urgent need for more effective waste-management strategies that prioritize waste reduction, material recovery, and the capture of greenhouse-gas emissions from landfill sites.

Global energy demand is likewise on an upward trajectory, with forecasts anticipating an increase of 11 to 18 percent by 2050, driven primarily by population growth and economic development in emerging markets [11]. Continued reliance on fossil fuels to meet this growing demand risks exhausting finite reserves, particularly if extraction outpaces natural replenishment [12]. Today, about 86 percent of global primary energy needs are supplied by oil, gas, and coal [13], with oil alone accounting for 36 percent and gas and coal contributing 27 percent and 23 percent, respectively. Renewables and other alternative sources make up only the remaining 14 percent. These figures underscore the urgency of diversifying our energy mix and accelerating the adoption of non-fossil alternatives.

1.2 Waste-to-Energy (WtE) as a dual solution

UNEP's latest data [1] highlight the urgent need to shift toward a zero-waste paradigm, not only to curb pollution and greenhouse-gas emissions from discarded materials but also to safeguard public health and the environment. Concurrently, rising energy demand and the unsustainable pace of fossil-fuel consumption make the exploration of alternative energy sources imperative. Waste-to-Energy (WtE) technologies present a compelling dual solution: municipal solid waste, with a net calorific value of roughly 10–17 MJ/kg, can be converted into useful energy form, such as heat, electricity, or biofuels, while simultaneously reducing the volume of refuse destined for landfills [14], [15]. By capturing the latent energy in waste streams, WtE aligns with circular-economy principles and helps diversify the energy mix away from fossil fuels. Moreover, these practices support the United Nations' Sustainable Development Goal 11, making cities and human settlements inclusive, safe, resilient, and sustainable by improving air quality and strengthening municipal waste management.

1.3 WtE pathways and technologies

Energy recovery from MSW is a critical component of sustainable waste management, enabling the production of valuable outputs such as electricity, heat, and fuels. The selection of appropriate conversion technologies depends fundamentally on the physicochemical characteristics of the waste. Accordingly,

these technologies are broadly categorized into two principal pathways: biochemical and thermochemical conversion. This classification facilitates the optimization of energy extraction efficiency and environmental performance. Moreover, effective pre-treatment of waste is essential to enhance feedstock quality and ensure optimal operation of the conversion processes. The accompanying Figure 1-1 delineates a comprehensive mapping of the waste conversion framework, integrating waste types, pre-treatment strategies, and corresponding conversion technologies.

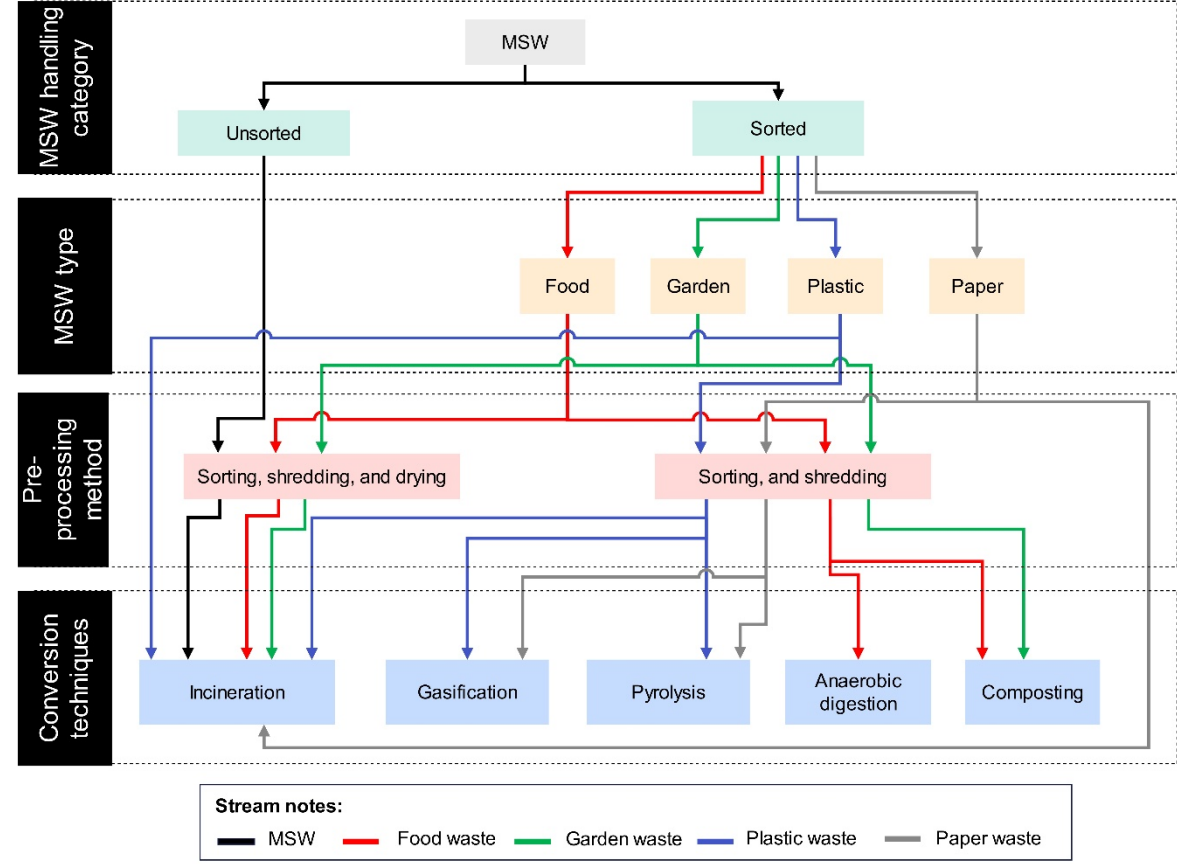


Figure 1-1. Identification of suitable waste-to-energy technologies for various feedstocks [16], [17].

Among the various waste-to-energy technologies, incineration remains the most mature and widely adopted method. Operating at temperatures between 800°C and 1000°C with excess air supply, incineration effectively combusts the feedstock [18], [19]. According to Figure 1-1, incineration is particularly well-suited for processing unsorted MSW, untreated plastic and paper waste, as well as pre-treated food and garden waste. Its key advantages include substantial volume reduction up to 80–90% and mass reduction of 70–80% [20], alongside complete disinfection and

energy recovery [21]. The high-temperature flue gases generated during incineration can be harnessed to produce steam for power generation via steam turbines.

However, incineration faces significant challenges due to the emission of harmful pollutants, notably dioxins and acidic gases such as SO_2 , NO_x , and HCl [22]. Additionally, corrosion caused by HCl can degrade equipment and reduce overall energy efficiency [23]. These environmental and operational limitations constrain the widespread adoption of incineration and drive ongoing research towards more energy-efficient and environmentally sustainable MSW-to-energy technologies.

In contrast to incineration, pyrolysis processes feedstock in the absence of oxygen at temperatures ranging from 300 to 900°C. This is a well-established technology, historically utilized for charcoal production from wood thousands of years ago. Pyrolysis is recognized for its versatility, as it can produce three distinct types of products: solid char, liquid pyrolysis oil, and combustible gases (syngas) [24]. By adjusting operating parameters such as temperature and heating rate, the proportions of these products can be controlled to meet specific needs [24]. The oxygen-deficient environment within a pyrolysis reactor inhibits the formation of harmful compounds such as dioxins and furans, rendering pyrolysis more environmentally friendly compared to conventional incineration. Additionally, pyrolysis plants generate significantly lower noise pollution than typical incineration facilities. Various reactor types are employed for the pyrolysis of different waste materials, including fixed-bed reactors, batch and semi-batch reactors, rotary kilns, fluidized bed reactors, and microwave-assisted reactors. Additionally, innovative technologies such as plasma reactors and solar reactors are being explored to enhance pyrolysis performance and efficiency.

Gasification operates under sub-stoichiometric conditions, where the supply of air is deliberately kept below the level needed for full combustion, typically at temperatures between 550°C and 1600°C [21]. Because it relies on partial oxidation, gasification is especially effective when the feedstock is a homogeneous, carbon-rich material with a high heating value [25]; in the case of MSW, this means that plastics and paper, once sorted and shredded, make ideal inputs. However, the heterogeneous nature of raw MSW necessitates pre-treatment steps such as

moisture reduction and particle-size homogenization to boost calorific value and overall process efficiency [21], [26], [27]. Compared with other thermal treatments like combustion and pyrolysis, gasification generally achieves higher energy recovery, but large-scale deployment has so far been confined to regions—mostly in developed countries—where landfill space is at a premium [28], [29]. In contrast, efforts to introduce gasification in many developing countries are hampered by uneven waste composition, variable particle sizes, and elevated moisture levels, all of which undermine reliable operation and performance.

Anaerobic digestion harnesses bacteria to convert a wide range of organic wastes, such as food scraps, agricultural residues, and the organic fraction of municipal solid waste into biogas, a renewable fuel, and nutrient-rich digestate [30]. Food waste, in particular, is well suited to this process because its high carbon content and rapid biodegradability promote efficient microbial activity [31], [32]. Unlike thermal treatments like incineration, pyrolysis, or gasification, anaerobic digestion operates at lower temperatures, requires less external energy input, and occupies a smaller space [33]. The process can be carried out under mesophilic conditions, typically around 37°C, or thermophilic conditions, between 50°C and 55°C; most commercial systems favour the mesophilic range due to its greater stability and reduced heating demands [34].

1.4 Useful output generated from waste to energy pathways

There is increasing interest in integrating multiple energy-conversion cycles to extract diverse forms of useful energy from a single fuel source, thereby boosting overall efficiency and reducing greenhouse-gas emissions. At its simplest, this approach takes the form of combined heat and power (CHP), or cogeneration, which captures waste heat, often 60–80% of the primary energy that would otherwise be lost in conventional power generation and repurposes it for applications such as space heating or domestic hot water [35]. By contrast, traditional thermal and electrical systems suffer substantial heat losses that drag down their overall efficiency. Building on CHP, combined cooling, heating, and power (CCHP) systems further expand the utility of by-product heat by driving thermally activated equipment, such as absorption or adsorption chillers for space

or process cooling while simultaneously generating heat and electricity [35]. These chillers can substitute for electric air-conditioning, further reducing electrical demand and peak loads. Although many CCHP installations today still rely on fossil fuels, growing research is investigating cleaner thermal inputs, ranging from solar thermal and geothermal to biomass and biogas, to create multigeneration plants that are more flexible, lower-carbon, and better aligned with future sustainability goals [36].

In addition to CHP and CCHP, multigeneration systems can also include on-site fuel production, a critical strategy for reducing our global reliance on fossil fuels and mitigating climate change. Biofuels, such as bioethanol, biodiesel, and particularly biomethane, offer significant reductions in pollutants like SO_x , PM, NO_x , and greenhouse-gas emissions [37]. Biomethane is produced by anaerobic digestion of organic waste, which generates biogas rich in CH_4 and CO_2 . This biogas is then upgraded in order to remove CO_2 and other impurities and to yield a final product that is typically 95–97 mol% methane and 1–3 mol% carbon dioxide [38], [39]. Because its gas composition closely matches that of natural gas, biomethane can serve as an alternative fuel across a wide range of applications, including transportation, power generation, and residential uses such as space heating, hot water, and gas-stove cooking [39].

When pipelines are unavailable, biomethane is often transported in liquefied form, which boasts a much higher energy density. Liquid biomethane can be produced either by cryogenic upgrading, where purified gas is directly cooled to cryogenic temperatures or by conventional upgrading (e.g., water scrubbing, chemical scrubbing, pressure swing adsorption (PSA), or membrane separation) followed by small-scale liquefaction [39]. At atmospheric pressure, liquefaction compresses biomethane approximately 600-fold, enabling long-distance delivery [40], [41]. However, the cryogenic refrigeration cycles and high-pressure expansion work required for biomethane liquefaction are more energy-intensive than those used for conventional LNG, resulting in higher power consumption and operational cost. Indeed, a large-scale plant (100 MW LHV) requires about 12.4 MW of power for liquefaction [42], while small-scale dual-expander systems consume roughly 1.45 kWh/kg of LBM produced [43], and conventional or cryogenic setups demand

between 0.8 and 1.8 kWh per Nm³ of biogas [44].

Alongside biofuels, transitioning to a hydrogen-based economy represents another crucial pathway to achieving net-zero CO₂ targets. Although hydrogen production has been studied since the nineteenth century, processes relying on fossil feedstocks, such as steam methane reforming, have dominated due to their lower costs and operational simplicity [45]. Today, hydrogen can be synthesized via electrolysis, steam methane reforming, gasification, thermochemical and photochemical methods, and biochemical or biological routes, with energy inputs sourced from renewables, nuclear power, or fossil fuels and feedstocks ranging from water and natural gas to coal and biomass. To clarify the varying carbon footprints of these pathways, a color-coded classification system categorizes hydrogen types by production process, energy input, and associated emissions. Table 1-1 summarizes this system, highlighting each “color” of hydrogen’s sustainability profile.

Figure 1-2a illustrates the production costs of various “colours” of hydrogen under current market conditions. At present, gray, black, and brown hydrogen, synthesized via conventional, fossil-fuel pathways, remain the cheapest options [46], [47]. However, as carbon taxes on CO₂ emissions rise, the relative expense of these fossil-derived hydrogen types will climb, improving the economic competitiveness of turquoise, blue, and green hydrogen [46], [47]. Pink hydrogen, produced via water electrolysis powered by nuclear energy, can already achieve economic viability and price stability, although its cost is heavily influenced by the capital outlay of the nuclear plant itself [46], [47]. Finally, orange hydrogen, generated by electrolyzers running on grid electricity, sees its price fluctuate in tandem with a country’s energy mix and prevailing electricity rates [48].

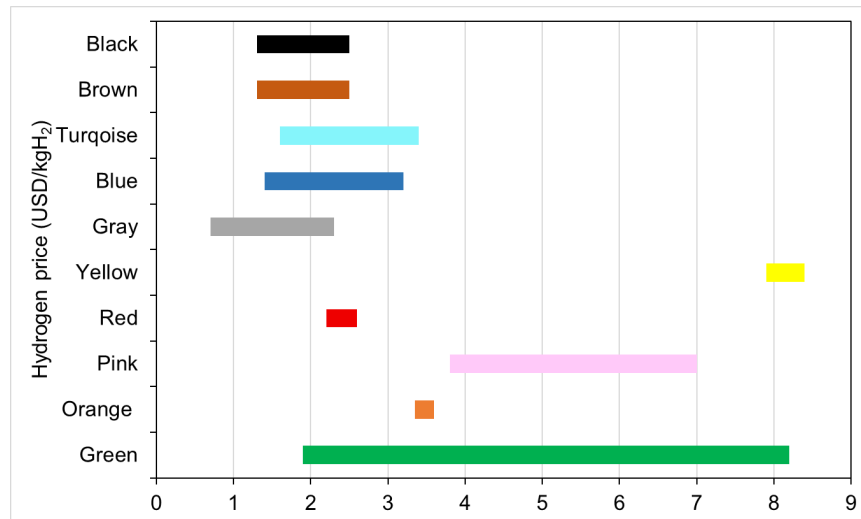
Table 1-1. Classification of hydrogen types based on energy source, production method, and emission profile [48].

Type (color code)	Primary energy input	Hydrogen feedstock	Production technology	Emission characteristic
Green	Renewable energy	Water	Electrolysis	Zero direct CO ₂ emissions

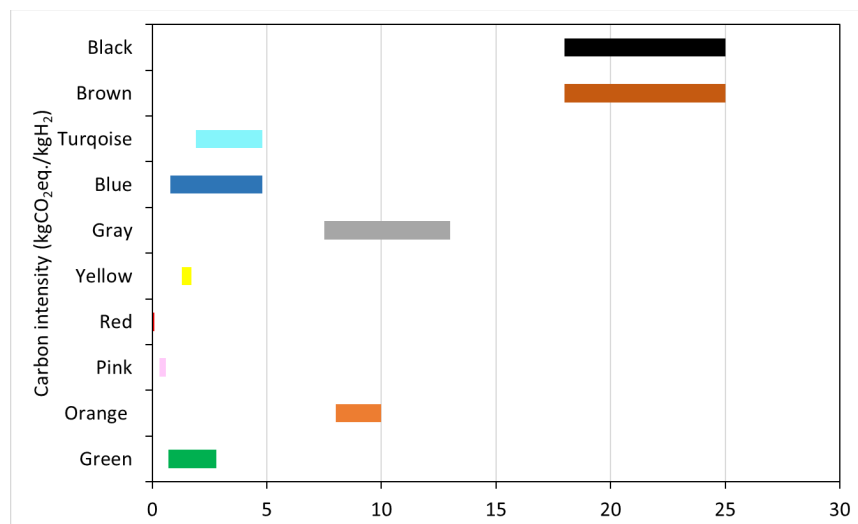
Orange	Mixed electricity grid	Water, CO ₂ , and iron-rich geological formations	Diagenesis	Emissions vary with electricity source
Pink	Nuclear	Water	Electrolysis	Zero direct CO ₂ emissions
Red	Nuclear	Water	Thermolysis / Thermochemical	Zero direct CO ₂ emissions
Yellow	Solar	Water	Electrolysis	Zero direct CO ₂ emissions
Gray	Fossil-fuel based electricity	Natural gas	Steam methane reforming	High CO ₂ emissions
Blue	Fossil-fuel based electricity	Natural gas/ biomass	Steam methane reforming	Low CO ₂ emissions
Turquoise	Fossil-fuel based electricity	Natural gas	Pyrolysis	No direct CO ₂ emissions
Brown	Brown coal (lignite)	Lignite coal / Biomass	Gasification	High CO ₂ emissions
Black	Black coal, (bituminous coal)	Bituminous coal	Gasification	High CO ₂ emissions

Figure 1-2b compares the lifecycle carbon intensity of these hydrogen variants. Pink and red hydrogen have almost zero direct CO₂ emissions in production, though they carry additional environmental considerations, including radioactive waste management, decommissioning, and accident risk, associated with nuclear technology [47]. Yellow hydrogen, derived from solar-driven electrolysis, also boasts very low operational emissions, with its main impacts stemming from solar-farm construction and water use. Green hydrogen, produced by renewable-powered electrolysis, approaches a zero-carbon pathway, although its full lifecycle footprint

depends on the embodied emissions of wind turbines, hydro installations, or photovoltaic panels [49], [50]. Blue hydrogen's carbon intensity spans a broad range depending on the specific reforming technology employed and the fraction of CO₂ captured and sequestered [46], [48]. At the opposite end of the spectrum, black and brown hydrogen, produced from coal without carbon capture, exhibit the highest CO₂ intensities of all the colour categories [47].



(a)



(b)

Figure 1-2. Comparison of hydrogen price range (USD/kg H₂) in (a) and carbon intensity (kgCO₂ eq./kgH₂) in (b) for different types of hydrogen classified by color.

1.5 Reduction of GHG emissions in waste to energy plants

A landmark agreement was reached at the 2024 UN Climate Change Conference (COP-29) in Azerbaijan, under which developed nations committed to mobilizing at least \$300 billion per year by 2035 to support developing countries in cutting emissions and protecting vulnerable communities. Yet even this historic pledge will fall short of the scale required to reach net-zero energy emissions by 2050 unless CCS capacity is dramatically expanded. According to the International Energy Agency [51] and the Carbon Sequestration Leadership Forum [52], global CCS deployment must be increased by ten- to fifteen-fold over current levels by 2030 and by roughly one hundred-fold relative to the 2020 baseline (40 Mt CO₂/yr) by mid-century. This expansion could enable the removal of up to 20 percent of global greenhouse-gas emissions [51].

The economic implications of this scale-up are substantial. The Intergovernmental Panel on Climate Change has estimated that, in the absence of CCS technologies, the cost of mitigation efforts could rise by 138 percent by 2100 [53]. As of mid-2024, fifty operational CCS facilities capture some 51 Mt CO₂ annually, and an additional 578 projects under development could add capacity for 365 Mt CO₂ more [54]. To finance such an ambitious transition, carbon-pricing revenues, the redirection of existing fossil-fuel subsidies, and private-sector investment have been identified as the principal funding sources [55]. Under current proposals, CO₂ prices would need to climb to approximately \$130 per tonne in advanced economies and \$90 in emerging markets by 2030, rising further to \$250 and \$200 per tonne, respectively, by 2050 [51].

Biomass energy with carbon capture and storage (BECCS) refers to the process of capturing CO₂ emitted during biomass combustion and permanently storing it in geological formations [56]. By sequestering more CO₂ than is released, BECCS delivers net-negative emissions rather than merely achieving net zero [57]. Although municipal waste includes significant volumes of plastic, the thermal conversion of plastics via pyrolysis or gasification simply releases fossil-derived carbon. Moreover, additional CO₂ is produced when the resulting fuels are eventually combusted for heat or power. To align waste-to-energy processes with

net-zero goals, these thermal technologies must be paired with CCS, so that all CO₂ generated is captured and stored. Only through the integration of CCS at every stage of waste conversion, decarbonization of energy generation from waste can be achieved.

1.6 Motivation

MSW management and energy security are increasingly important challenges worldwide. Rising population and consumption levels have led to a continuous increase in waste generation. At the same time, growing energy demand and the environmental consequences of fossil fuel dependence have intensified interest in alternative and more sustainable energy systems. In this context, WtE plants have gained considerable attention as they offer a dual benefit: reducing the volume of waste sent to landfill while simultaneously recovering useful energy. Despite their widespread implementation, WtE continues to face challenges related to overall efficiency, environmental performance, and effective system integration. Consequently, the development of more integrated and efficient waste-based energy systems remains an important research topic with broad international relevance.

Within this broader context, Indonesia provides a relevant case study due to its growing energy demand, substantial MSW generation and ongoing development of WtE infrastructure. Indonesia is located on the geographic coordinates of 6°08' N latitude to 11°15' S latitude, and from 94°45' E to 141°05' E longitude. As an archipelago country, Indonesia consists of five large islands, namely Java, Sumatera, Kalimantan, Sulawesi, and Papua, and also approximately 30 smaller islands (Figure 1-3). The group of the islands is situated at a crossroads between Asia and Australia and lies between the Indian and Pacific oceans. The territory covers a land area of 1,826,440 sq km and a water area of 93,000 sq km. Indonesia is intersected by the earth's equator which results in a constant twelve hours of day and night for the whole year. As a tropical country, Indonesia has an average temperature of 28°C and a high relative humidity ranging from 70 to 90%.



Figure 1-3. Map of Indonesia [58].

Energy demand in Indonesia continues to grow, with fossil fuels remaining the dominant supply for electricity, transportation, and household cooking. Cooling demand alone is projected to rise by approximately 35 percent in buildings by 2030 [59]. At the same time, domestic fossil-fuel reserves are dwindling: proven crude-oil stocks fell from 7.31 billion barrels in 2015 to just 4.17 billion barrels by 2020, while coal and natural-gas reserves stood at 38.87 million tonnes and 62.39 trillion cubic feet, respectively, that same year [60]. These pressures highlight the need for alternative and more efficient energy systems.



Figure 1-4. The composition of waste in Indonesia [61].

Indonesia also faces a significant waste management challenge. With a population

of 273.87 million in 2021, the country generated 17.67 million tonnes of municipal solid waste (MSW) in the same year [61], [62]. Of that amount of waste, household waste occupies the highest quantity with a percentage of 40.9%. It is also worth noticing that the proportion of waste composition in Indonesia is dominated by food waste at 28.4%. Plastic waste is found to be the second-highest with a ratio of 16%, whereas woody waste's share of composition is 12.4%, followed closely by papers at 12.2%. The detailed composition of another sort of waste can be observed in Figure 1-4.

With the continuous increase of population in Indonesia which has resulted in a massive amount of waste being generated, MSW has become an intricate problem since it leads to increased methane emission, expansion of landfill areas, and affects people's health. In order to cope with the escalation of waste, Indonesia's government has executed several waste management practices. One of the waste disposal treatments is converting the waste into energy, such as producing refused-derived fuel and adopting various waste conversion technologies, such as biodigester, incinerator, pyrolysis, landfill emission gas, and gasification. The largest waste power plant in Indonesia generates electricity with a capacity of 11 MW which consists of 2 MW produced from the utilization of the landfill gas and 9 MW generated from gasification and is fueled by up to 1000 tonnes of municipal solid waste a day [63].

In the current operations of this 11 MW plant, the landfill gas and gasification conversion routes are implemented in separates areas. Landfill gas is combusted in a boiler to produce steam, which drives a turbine for electricity generation, while in the gasification route, the produced syngas is also used to convert water into steam, which is subsequently expanded through a steam turbine to produce electricity. However, the plant has not been equipped with a waste heat recovery system, and as a result, a considerable amount of residual thermal energy is released into the environment. The Indonesian government plans to build another 11 waste power plants with a capacity range between 9 to 38 MW.

Against this background, Indonesia represents a relevant example of a wider global issue: although WtE systems are increasingly adopted as a solution to both waste management and energy supply, their performance is often constrained by

insufficient heat recovery and limited overall efficiency. This creates strong motivation to investigate more integrated waste-based energy systems that can maximize resource utilization and improve technoeconomic and environmental performance.

1.7 Research aims and objectives of the thesis

The motivation for this study stems from the urgent need to implement effective waste management practices that not only reduce the volume of waste generated but also protect the environment from the harmful impacts of waste-related pollution. Among the various types of municipal solid waste, plastic and food waste are the most prevalent across the globe. The application of plasma gasification for converting waste into syngas has been recognized as a promising solution, owing to its extremely high operating temperatures that facilitate the complete decomposition of waste into their fundamental molecular components.

Despite its potential, a review of the current literature reveals a lack of studies focusing on the combined utilization of both food and plastic waste, either via plasma gasification alone or in conjunction with anaerobic digestion. Furthermore, existing plasma gasification systems discussed in the literature are typically configured for single-output production, which limits the efficiency and energy recovery of the process. In contrast, the current study explores multiple-output generation, which allows for more effective utilization of the available energy within the system.

In the context of global climate commitments, the integration of carbon capture and storage (CCS) technologies has become essential to reduce emissions from waste-to-energy systems. However, a comparative evaluation of two CCS approaches: pre-combustion and post-combustion using monoethanolamine (MEA) absorption within the framework of plasma gasification remains unexplored. Additionally, while most published studies assume a CO₂ capture rate of 90%, this research investigates a higher rate of 95% which acknowledges that increasing the capture rate demands more heat and may therefore impact overall system efficiency and net energy production.

Although both techno-economic assessment (TEA) and life cycle assessment (LCA) are critical tools for evaluating sustainability, their integration into a combined assessment framework remains limited in existing studies. To address this gap, the current study leverages established process modelling data to conduct a comprehensive TEA and LCA investigation on the conversion of food and plastic waste into various energy products.

Methodologically, the thesis progresses in a structured way. It begins with process modelling of the proposed system in Aspen Plus, followed by thermodynamic assessment to evaluate mass, energy, and exergy performance. The analysis is then extended through TEA and LCA to determine economic viability and environmental impacts. Finally, the framework is further expanded through system extension to investigate alternative product pathways, thereby demonstrating the wider applicability of the proposed methodology across different WtE configurations.

This research is organised into three case studies:

- Research Case 1 examines the conceptual design of conversion of food and plastic waste into combined cooling, heating, and power (CCHP) and liquid methane, evaluated in detail thermodynamically across three configurations: no CCS, integrated with pre-combustion CCS, and post-combustion CCS.
- Research Case 2 extends the evaluation of Research Case 1 by conducting TEA and LCA of the proposed system.
- Research Case 3 further expands the framework to investigate hydrogen production from the same waste streams, integrated pre-combustion CCS, and post-combustion CCS and evaluated using combined TEA and LCA.

To achieve the main goal of this research, the following specific objectives have been defined:

1. To develop process models in Aspen Plus for the conversion of plastic and food waste into CCHP, liquid methane, and hydrogen.
2. To evaluate the thermodynamic performance of the proposed systems in terms of mass, energy, and exergy efficiencies.
3. To assess the effects of alternative CCS configurations, including pre-

- combustion and post-combustion capture, on overall system performance.
4. To estimate key economic indicators, including CAPEX, OPEX, LCOE, LCOLBM, LCOWT, and LCOH using deterministic inputs and uncertainty analysis.
 5. To quantify the environmental impacts of the proposed systems, with particular emphasis on GWP.
 6. To establish a unified performance database that supports system comparison, feasibility evaluation, and future decision-making for WtE development.

1.8 Thesis structure

The content of this thesis is divided into seven chapters, each addressing a specific aspect of the study, as detailed below:

Chapter 1 introduces the research by outlining its relevance and the rationale for focusing on the conversion of municipal waste into energy. It also presents the main aim and objectives of the study and provides a brief overview of the thesis structure.

Chapter 2 outlines the key concepts necessary to understand the working principles of the components involved in the scenarios studied. It also offers a critical review of relevant techno-economic and environmental studies from the literature. The chapter further identifies existing research gaps and highlights the importance of the present study.

Chapter 3 describes the process modelling approach using Aspen Plus and presents the general methodology applied for the thermodynamic, economic, and environmental assessments. Scenario-specific methodologies are detailed separately in the corresponding chapters.

Chapter 4 presents the detailed thermodynamic evaluation of Research Case 1, which involves converting municipal waste (food and plastic waste) into CCHP and liquid methane. The chapter investigates the impact of integrating CCS on thermodynamic performance indicators, including overall energy and exergy

efficiency, across two scenarios: a system producing only CCHP, and a system producing both CCHP and liquid methane. It also analyses the carbon balance for each scenario. The findings are based on the author's published work titled *"Conceptual design and thermodynamic investigation of novel energy and fuel generation systems from municipal waste coupled with carbon capture and storage"* [64].

Chapter 5 provides the technical, economic and environmental evaluation of Research Case 1, continuing the analysis of the same WtE pathway described in Chapter 4. It explores the effects of CCS integration on the system's technical, economic, and environmental performance for two different scenarios. Additionally, the chapter examines the potential influence of policy-driven economic incentives linked to the main product of each scenario: electricity in Scenario 1 and liquid methane in Scenario 2 with considerations based on the estimated global warming potential (GWP). The findings in this chapter are drawn from the author's publication titled *"Economic and Life Cycle Assessment of Novel Energy and Fuel Generation Systems from Municipal Waste through Plasma Gasification and Anaerobic Digestion Coupled with Carbon Capture and Storage"* [65].

Chapter 6 presents the thermodynamic, economic, and environmental assessment of Research Case 2, which focuses on the conversion of municipal waste into hydrogen through three integrated conversion pathways. This chapter evaluates the effects of two CCS scenarios on system performance, namely, pre-CCS, and post-CCS. In Case 1, food and plastic waste are ideally separated, allowing food waste to be processed via anaerobic digestion and plastic waste via plasma gasification. In Case 2, there is no separation between the two waste streams, and both are treated together using plasma gasification. Case 3 builds upon Case 1 by integrating a steam methane reformer (SMR) to enhance hydrogen production from the biogas generated during anaerobic digestion. Additionally, this chapter explores the potential impact of policy-driven economic incentives for hydrogen, based on its estimated global warming potential (GWP). A portion of the results from this chapter was presented at the American Society Mechanical Engineering (ASME) conference in the technical paper titled *"Sustainable Hydrogen Production From Plastic Waste via Plasma Gasification and Carbon Capture Integration: A*

Combined Thermodynamic and Economic Assessment” [66]. In addition, the findings have been published in a journal paper entitled *“Hydrogen and CCHP generation from waste via combined plasma gasification and anaerobic digestion with CCS: A techno-economic and life cycle assessment”* [67].

Chapter 7 presents the overall conclusions and summarizes the key findings of the thesis. It also discusses the main limitations of the study and offers recommendations for future research directions.

2 LITERATURE REVIEW

2.1 Waste to Energy (WtE) technologies

Energy generated from waste can be carried out either by thermochemical conversion or a biochemical process. The thermochemical process releases directly the stored energy within the biomass as heat via combustion/ co-firing or converts it into syngas via gasification. In contrast, biochemical conversion relies on the action of microorganisms decompose organic materials into biofuel. The following subsections provide a more detailed explanation of the fundamental concept of plasma gasification and anaerobic digestion, which represent the waste to energy technologies applied in the present study.

2.1.1 Plasma gasification

The gasification process comprises four key stages: drying, pyrolysis, oxidation, and reduction [68]. The primary product is synthesis gas (syngas), a mixture predominantly composed of hydrogen, carbon monoxide, and short-chain hydrocarbons, which can be further utilized for power generation, liquid fuel synthesis, hydrogen production, and chemical manufacturing [28]. The reactions occur in the biomass and MSW gasification can be seen in Table 2-1. As shown, the reactions of 7, 8, 9, 11, and 15 are commonly considered as basic gasification process. The quality of the produced syngas is influenced chiefly by the operating temperature and the choice of gasifying agents, which may include air, steam, or oxygen-enriched air [69]. Raw syngas typically contains impurities such as tar, sulfides, alkali metals, chlorides, and particulates, necessitating appropriate cleaning prior to downstream applications [20], [70]. Gasification technologies are classified based on the heat source into autothermal and allothermal types [71]. Autothermal gasification relies on partial oxidation of the feedstock itself to supply the required heat, as seen in air gasification [68]. In contrast, allothermal gasification uses an external heat source, such as a plasma torch in plasma gasification, to provide the necessary thermal energy [21].

Table 2-1. Chemical reactions in biomass and MSW gasification [72], [73].

Main reaction:				
$CH_xO_y + aO_2 + wH_2O + 3.76aN_2$ $= n_1H_2 + n_2CO + n_3CO_2 + n_4H_2O + n_5CH_4 + n_6N_2 + n_7C(char) + tar$				
Oxidation reaction:				
1	$C + 1/2 O_2 \rightarrow CO$	-111 MJ/kmol	Carbon oxidation	partial
2	$CO + 1/2 O_2 \rightarrow CO_2$	-283 MJ/kmol	CO oxidation	
3	$C + O_2 \rightarrow CO_2$	-394 MJ/kmol	Carbon oxidation	
4	$H_2 + 1/2 O_2 \rightarrow H_2O$	-242 MJ/kmol	Hydrogen oxidation	
5	$CH_4 + 1/2 O_2 \rightarrow CO + 2H_2$	-36 MJ/kmol	CH4 oxidation	partial
6	$CH_4 + 2O_2 \rightarrow CO_2 + 2H_2O$	-803 MJ/kmol	Oxidation	
Gasification reactions involving steam:				
7	$CH_4 + H_2O \rightarrow CO + 3H_2$	206 MJ/kmol	Steam reforming	methane
8	$CO + H_2O \rightarrow CO_2 + H_2$	-41 MJ/kmol	Water-gas reaction	shift
9	$C + H_2O \rightarrow CO + H_2$	131 MJ/kmol	Water-gas reaction	
10	$C_nH_m + nH_2O \rightarrow nCO + (n + m/2)H_2$	Endothermic	Steam reforming	
Gasification reactions involving hydrogen:				
11	$C + 2H_2 \rightarrow CH_4$	-75 MJ/kmol	Hydrogasification	
12	$CO + 3H_2 \rightarrow CH_4 + H_2O$	-227 MJ/kmol	Methanation	
13	$2CO + 2H_2 \rightarrow CH_4 + CO_2$	247 MJ/kmol	Methanation	
14	$CO_2 + 4H_2 \rightarrow CH_4 + 2H_2O$	165 MJ/kmol	Methanation	
Gasification reactions involving carbon dioxide:				
15	$C + CO_2 \rightarrow 2CO$	172 MJ/kmol	Boudouard reaction	
16	$C_nH_m + nCO_2 \rightarrow 2nCO + m/2 H_2$	Endothermic	Dry reforming	
Decomposition reactions of tars and hydrocarbons:				
17	$pC_xH_y \rightarrow qC_nH_m + rH_2$	Endothermic	Dehydrogenation	
18	$C_nH_m \rightarrow nC + m/2 H_2$	Endothermic	Carbonization	

Plasma gasification operates at extremely high temperatures under oxygen-starved (sub-stoichiometric) conditions, thermally decomposing waste into simple species and converting nearly all carbon to fuel gas [74]. Among waste-conversion options, it most closely approximates “pure” gasification because heat is supplied externally by a controllable plasma torch rather than by combustion [75]. The organic fraction of the feed becomes a high-calorific-value syngas suitable for use as a fuel or chemical feedstock, while the inorganic fraction vitrifies into a non-leachable slag that can be safely disposed of or repurposed as construction material [76]. Owing to the severe thermal environment and rapid quenching, the reactor exit gas is comparatively clean, and no bottom ash is produced [77]. Environmental performance is typically superior to incineration, with substantially lower air-emission levels and reduced slag-leachate toxicity [78].

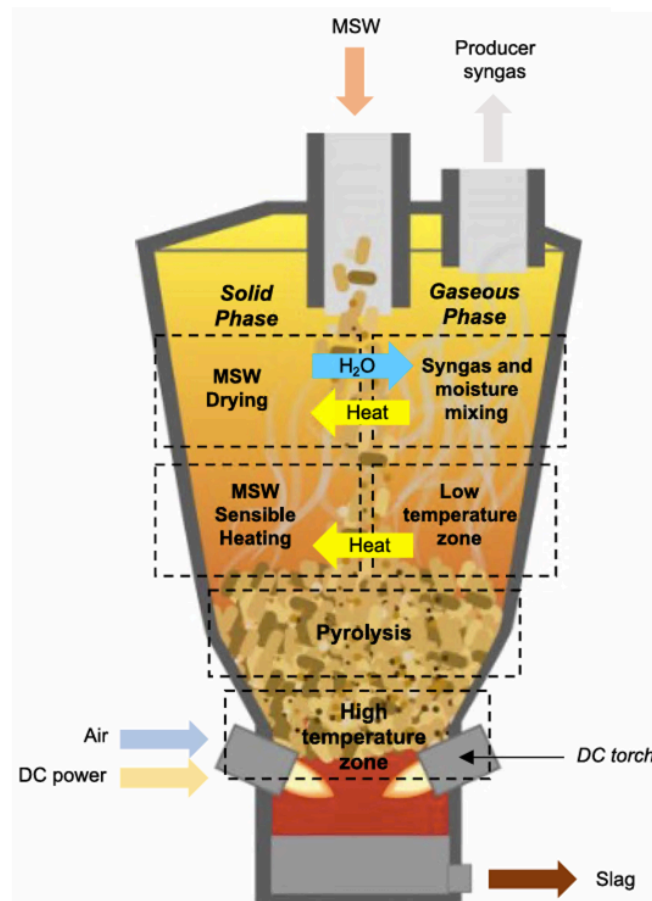


Figure 2-1. Schematic of a plasma gasification reactor showing process zones and flows [79].

Figure 2-1 schematically illustrates a plasma gasification reactor with its process zones and flow paths. Plasma gasification proceeds through four stages [75]. Pre-

heat/drying ($\sim 110^{\circ}\text{C}$) brings incoming feedstock into contact with hot syngas thus the moisture evaporates [73], [76], [80]. Devolatilization (pyrolysis, $\sim 300\text{--}700^{\circ}\text{C}$) follows, where the dried material breaks down into volatiles and char under radiation and convection from the plasma core [81], [82]. Next comes the gasification/reforming zone ($\sim 800\text{--}1,500^{\circ}\text{C}$), where char is converted via reactions with steam and/or limited O_2 [83]. High-temperature steam activates reactions (7–10), as summarized in the Table 2-1 [73]. The sequence ends in the melt/vitrification zone, a high-temperature region (slag formation $\sim 1,200^{\circ}\text{C}$) locally exposed to the plasma core ($>3,000^{\circ}\text{C}$), where inorganic components fuse into a non-leachable slag suitable for tapping, casting, or water quenching [84], [85]. Downstream, the product gas is quenched within seconds to a few hundred degrees and then sent to heat-recovery and cleanup units, which suppress dioxin/furan re-formation and prepare the syngas for use [77].

Plasma gasification employs an externally supplied thermal source, a plasma torch, to convert waste with minimal combustion [77]. The torch issues a gas jet at temperatures up to $5,000^{\circ}\text{C}$ or higher, providing precisely controllable energy that elevates reactor temperatures enough to dissociate chemical bonds and generate abundant reactive species (radicals, electrons, ions, and electronically excited molecules) [76], [85], [86], [87]. The intense radiative flux accelerates reaction kinetics and enables pathways that are difficult to achieve at conventional temperatures [88], [89]. Owing to the steep thermal gradients in the reactor, the product stream can be rapidly quenched, thereby minimizing the reformation of persistent organic pollutants such as dioxins and furans [88], [90]. The principal drawback of this thermal plasma is its high electricity demand [76]. This challenge may be mitigated through energy integration strategies, which can help reduce overall energy intensity and enhance system performance. Another potential approach to reducing this demand is to operate at a lower gasification temperature. However, lowering the operating temperature may require additional process steps, such as catalytic tar reforming or cracking and optimization of the gasifying-agent ratio, in order to maintain syngas quality suitable for downstream applications [91], [92]. These additional requirements can introduce further challenges, including greater tar-management demands, more complex process control, catalyst handling or recovery, and the possible need for extra gas-cleaning units.

A typical plasma gasification system comprises the following components:

1. **Feedstock preparation and feeding:** size reduction (e.g., shredders/grinders), drying, and metered feeding (often via lock-hoppers) into the reactor [93], [94].
2. **Plasma gasifier (reactor):** a refractory-lined chamber equipped with one or more plasma torches that supply the required heat to sustain gasification under oxygen-starved conditions [95].
3. **Syngas conditioning and cleanup:** quench and heat-recovery (e.g., waste-heat boiler), particulate removal (cyclone, ceramic or fabric filters), acid-gas/trace-contaminant removal (wet or dry scrubbing), and final polishing before use in power or chemical applications [93], [96].
4. **Slag handling:** tapping of molten inorganic material, followed by casting (ingots/bricks) or water-quenching to produce non-leachable granules for reuse or disposal [93].
5. **Plasma torches and power supply:** water-cooled torches with appropriate process-gas services, coupled to high-current, tightly regulated power supplies and power-conditioning equipment to maintain stable arc operation under extreme thermal loads [87].
6. **Monitoring and control system:** instrumentation for temperature, pressure, flow, and gas composition, with Programmable Logic Controllers (PLCs) and Distributed Control Systems (DCS) automation of feed rate, torch power, and slag removal to ensure safe, optimal operation [97], [98].

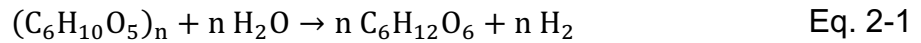
2.1.2 Anaerobic digestion

Anaerobic digestion involves the decomposition of organic materials in the absence of oxygen to yield biogas. It provides several benefits, such as less energy consumption, low quantity of biological solid waste generation, production of combustible gas to replace fossil fuel, and odorless final products [99]. The different substrates can have different characteristics, such as different content of carbohydrates and proteins, and thus lead to different mechanisms during the process that occur which affects the amount of methane production [99].

There are four essential steps required for AD to occur, namely hydrolysis, acidogenesis, acetogenesis, and methanogenesis. All these processes can be described as follows [100]:

Hydrolysis

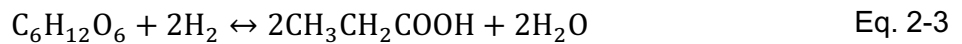
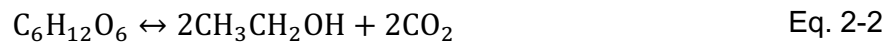
Hydrolysis, as the first stage in the AD process, cleaves the chemical bond with the addition of water. The interaction between water molecules with pairs of cations and anions can change pH levels and leads to H-O bond cleavage. It is often characterized as a relatively slow reaction step that can limit the rate of the entire AD process, especially when solid waste substrates are fed into the digester. The hydrolysis reaction can be expressed by following equations:



The above equation shows the hydrolysis of cellulose, with the products being glucose and hydrogen. During hydrolysis, the anaerobic organism is employed to break the chemical bonds of the insoluble cellulose in H₂O in order to form soluble compounds.

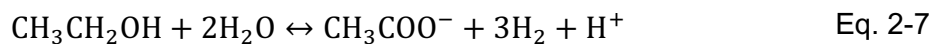
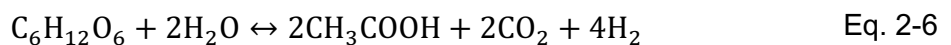
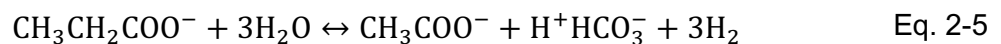
Acidogenesis

The yielded products from the first phase are fermented into volatile fatty acids by the use of acidogenic bacteria. Eq. 2-2 to Eq. 2-4 denote the sequential reactions taking place during acidogenesis.



Acetogenesis

The fermented products formed in the acidogenesis stage are converted into acetate, CO₂, and H₂ with the help of acetogenic bacteria. The following reactions are involved in the phase of acetogenesis:



Methanogenesis

The bacteria called methanogens are responsible for the transformation of CH_3COOH and H_2 into CO_2 and CH_4 . Methanogenesis is able to significantly reduce the organic pollution load, thus this stage can be considered an efficient way to remove carbon pollution. Several reactions occur during the methanogenic stage as represented by the following equation.



Through four important steps above, AD generates various gases with different compositions, as shown in Table 2-2. CH_4 which is known as the largest compound of natural gas, also contributes as the primary flammable constituent in the raw biogas and varies depending upon the source [100]. G Sailer et al. [101] summarized the CH_4 potentials from numerous feedstocks, such as sewage sludge (210–300 TWh), food waste (880–1100 TWh), manure (2600–3800 TWh), and crops (3080–3920 TWh).

Table 2-2. Typical composition of raw biogas [100].

Gas	Composition (%)
CH_4	50-57
CO_2	25-50
N_2	0-10
H_2	0-1
H_2S	0-3
O_2	0-2

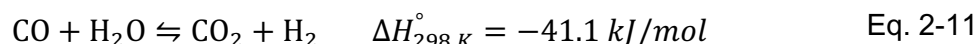
In addition, the temperature of the AD process plays a paramount role in the generation of CH_4 . There are two ranges of temperature to keep the anaerobic microorganisms surviving and growing during the digestion process, namely mesophilic (20–45°C) and thermophilic temperatures (49–57°C). It is recommended that the AD operates under thermophilic since the optimum composition of CH_4 can be reached [102], [103]. The temperature of AD can be achieved and maintained

using an internal or external heating system [104]. The internal heating system uses steam or hot water as the heat transfer medium flowing through the installed heating pipes on the digester walls to extract heat from the biogas boiler, power cycle, and effluent (digestate) [104]. Meanwhile, the external heating system built separately from the AD installation allows the substrates to absorb heat from various external heat sources such as solar energy, geothermal, steam or hot water, and electric energy before being pumped into the reactor [104]. The external heat sources are required when the internal heat sources are inadequate to support the AD process [104]. Compared to AD, gasification has greater efficiency due to the shorter reaction time required [105]. In addition, the thermochemical approach can decompose the material which cannot be broken down via the biological approach.

2.2 Syngas and biogas conditioner

2.2.1 Water gas shift reactor (WGSR)

WGSR is employed to convert carbon monoxide (CO) in product gas, in the presence of steam, into carbon dioxide (CO₂) and hydrogen (H₂) [106]. It is a redox-type reaction, represented as [106]:



This reversible reaction is moderately exothermic and has been known since its discovery by the Italian physicist Felice Fontana in 1780 [107]. It was later adopted industrially in syngas upgrading, most notably as part of the Haber–Bosch process for ammonia synthesis [108]. The WGSR also plays a central role in steam reforming [109], [110], partial oxidation [111], [112], gasification [113], [114], methanol and dimethyl ether synthesis [115], [116], the Fischer–Tropsch process [117], and ironmaking in blast furnaces [118], [119]. Furthermore, it enhances the performance of proton exchange membrane fuel cells (PEMFCs), since CO in the anode feed can poison the electrodes and degrade cell performance [120], [121]. By lowering CO concentrations to as little as 0.1–0.3%, the WGSR simultaneously improves fuel cell durability while increasing hydrogen availability [122].

Although the reaction is thermodynamically favoured whenever CO and H₂O coexist, kinetic barriers limit its progress at low temperatures [106]. To overcome this limitation, catalysts are introduced to reduce activation energy and enable efficient conversion [106]. Depending on temperature and catalyst type, the WGSR is commonly divided into two stages: the high-temperature shift (HTS) reaction, typically operating at 350–500°C with Fe–Cr based catalysts, and the low-temperature shift (LTS) reaction, occurring at 150–250°C with Cu–Zn based catalysts [123]. The HTS stage ensures rapid CO conversion at elevated temperatures, while the LTS stage further drives the equilibrium forward at lower temperatures, enriching the product stream in H₂ and CO₂ [106]. The power-law reaction rate expressions used in the WGSR model can be characterized as follows:

HTWGSR (catalyst: Fe-Cr):

$$r = 4.56 \exp\left(\frac{-88}{RT}\right) \times \frac{(P_{CO})^{0.9} (P_{H_2O})^{0.31}}{(P_{CO_2})^{0.156} (P_{H_2})^{0.05}} \times (1 - \beta) \quad [124] \text{ (Eq.2-12)}$$

LTWGSR (catalyst: CuO/ ZnO/ Al₂O₃):

$$r = 4.785 \exp\left(\frac{-34.983}{RT}\right) \times \frac{(P_{CO})^{0.854} (P_{H_2O})^{1.99}}{(P_{CO_2})^{1.926} (P_{H_2})^{0.573}} \times (1 - \beta) \quad [125] \text{ (Eq.2-13)}$$

where r is reaction rate, R is the universal gas constant, T is the reaction temperature, P_{CO} , P_{H_2O} , P_{CO_2} , and P_{H_2} are the partial pressure of CO, H₂O, CO₂, and H₂, respectively, and β is the equilibrium approach factor which can be calculated using the following equation:

$$\beta = \frac{[CO_2][H_2]}{[CO][H_2O]K_p}, \quad K_p = \exp\left(\frac{4577.8}{T} - 4.33\right) \quad [106] \text{ (Eq.2-14)}$$

where $[CO_2]$, $[H_2]$, $[CO]$, and $[H_2O]$ are the concentrations or partial pressures of carbon dioxide, hydrogen, carbon monoxide, and steam, respectively, and K_p is the equilibrium constant of the reaction.

2.2.2 High pressure water scrubber (HPWS)

The water scrubbing (WS) process is widely applied for upgrading biogas by selectively removing carbon dioxide (CO_2) from the raw gas mixture. Biogas derived from anaerobic digestion is composed primarily of methane (CH_4) and CO_2 , with trace quantities of hydrogen sulfide (H_2S), water vapour, hydrogen (H_2), nitrogen (N_2), and oxygen (O_2). For utilization as a vehicle fuel or for injection into natural gas grids, the methane content of biogas, typically in the range of 50–70%, must be enriched to at least 97% [126]. Because the high CO_2 fraction lowers the calorific value of raw biogas, upgrading is essential to improve its energy density, stability, and economic viability [127].

High-pressure water scrubbing (HPWS), a variant of WS based on physical absorption, is one of the most established and commercially implemented upgrading techniques [127], [128]. In this process, CO_2 and other soluble impurities are absorbed into pressurized water, while the methane-rich gas exits as the upgraded product. HPWS has been extensively developed and deployed in several European countries, where it is recognized as a mature, reliable, and relatively simple technology [126]. Moreover, it is often described as an environmentally friendly method since it does not rely on chemical solvents; instead, it uses only water as the absorbent medium [127].

Despite these advantages, HPWS also presents limitations. The solubility of CO_2 in water is relatively low, requiring large volumes of water to achieve effective separation [129]. This high-water demand represents the main drawback of technology, necessitating careful consideration of water availability, recycling, and treatment in large-scale applications. Nevertheless, its feasibility may be improved through effective water recovery and reuse strategies within the overall process design.

2.3 Energy conversion technologies

2.3.1 Combined cycle gas turbine (CCGT)

The gas produced from plasma gasification and/or anaerobic digestion can be utilized to drive a combined cycle gas turbine (CCGT) for electricity generation. A

CCGT system consists of three main components: the gas turbine (GT), the heat recovery steam generator (HRSG), and the steam turbine (ST). In this configuration, the syngas is first directed to the GT to produce electricity, while the hot exhaust gases from the GT are subsequently used in the HRSG to generate steam, which powers the ST for additional electricity production.

The GT operates based on the Brayton cycle, comprising a compressor, a combustion chamber, and a turbine. For an ideal Brayton cycle, efficiency can be enhanced by increasing the pressure ratio and/or the turbine inlet temperature. The flow path in the GT is as follows: ambient air is compressed to a pressure of approximately 15–27 bar, mixed with syngas from plasma gasification and/or anaerobic digestion, and combusted to produce high-temperature flue gas [130]. This flue gas drives the turbine, converting thermal energy into mechanical shaft power, which in turn drives the generator to produce electricity. The exhaust gas leaving the turbine typically reaches 593–638°C [130].

In the combined cycle configuration, this exhaust heat is recovered in the HRSG to generate steam for the bottoming cycle. The HRSG generally comprises four major sections: preheater, economizer, evaporator, and superheater. In integrated gasification combined cycle (IGCC) systems, HRSGs are often designed with multiple pressure levels: low pressure (LP), intermediate pressure (IP), and high pressure (HP) to provide steam conditions suitable for the three-stage ST [130].

The ST operates according to the Rankine cycle, with the HRSG serving as the heat source in place of a conventional boiler. The cycle begins with feedwater preheating in the preheater and economizer, followed by evaporation in the drum to produce saturated steam, which is then converted to superheated steam in the superheater. As the steam expands through the multi-stage ST, its thermal energy is transformed into mechanical shaft power, which drives the generator to produce electricity. The exhaust steam is finally condensed using a cooling medium. Due to growing concerns regarding thermal pollution of natural water bodies and the conservation of groundwater resources, air cooling via cooling towers has increasingly been adopted as an alternative to water cooling [130].

2.3.2 Absorption refrigeration cooler (ARC)

The thermal energy which is still in the flue gas is utilized to operate absorption refrigeration cooler (ARC). The main difference between absorption refrigeration with the vapor compression refrigeration cycle is the driver of the cycle. Absorption refrigeration uses thermal energy to retrieve the refrigerant vapor from the liquid solution before the refrigerant enters the condenser, while the vapor compression refrigeration cycle requires a mechanical compressor to drive the fluid. Two common working fluids in an absorption cooling system includes ammonia-water (ammonia is the refrigerant and water is the absorbent) and lithium bromide-water (lithium bromide/ LiBr as the absorbent and water as the refrigerant). Other differences are summarized in table below:

Table 2-3. Comparison of compression and absorptive refrigeration.

Parameters	Compression Refrigeration	Absorptive Refrigeration [131]
Type of compression	Mechanical compression	Thermal, absorption loop
Power source	Electrical energy	Heat energy 85-180°C
Coefficient of Performance	2.5-3.5	0.6-1.0

The performance of a LiBr–H₂O absorption refrigeration cooler (ARC) is influenced by several operating parameters, with the generator temperature being the most critical [132]. An increase in generator temperature leads to a higher coefficient of performance (COP). Additionally, raising the solution flow ratio reduces the required generator temperature but results in a lower COP. A larger temperature difference between the condenser and absorber decreases both the COP and the cooling capacity, with the COP being more sensitive to changes in absorber temperature than in condenser temperature.

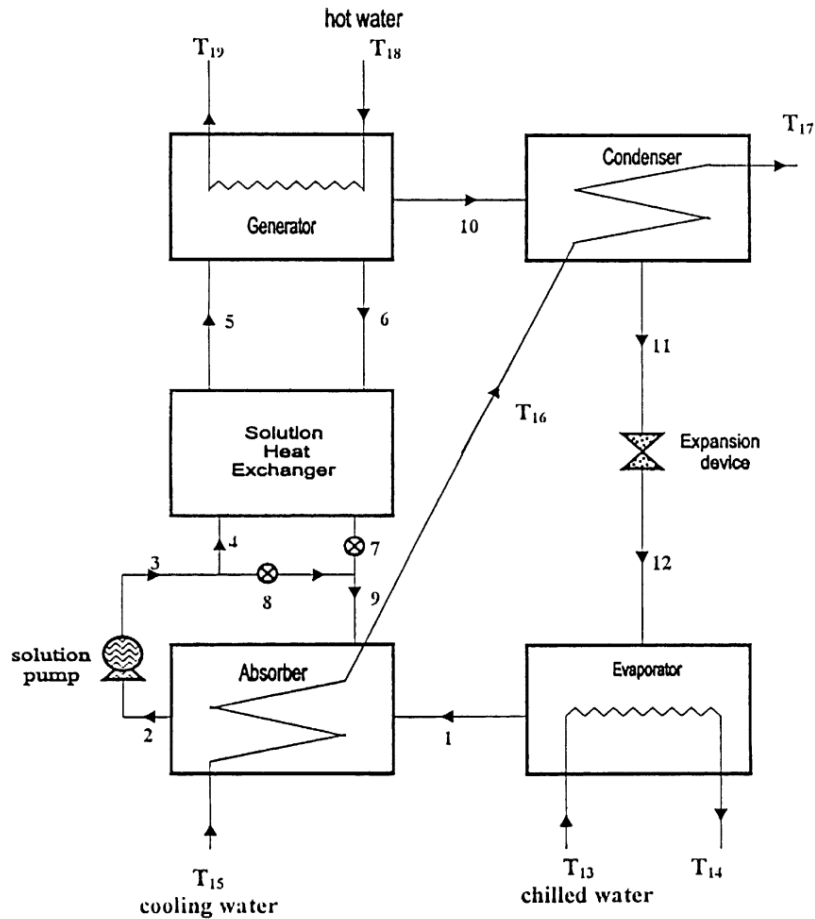


Figure 2-2. Schematic diagram of a single-effect LiBr–H₂O absorption refrigeration cycle [133].

The ARC consists of four main components: an absorber, a generator, a condenser, and an evaporator [133]. The process flow of the ARC is illustrated in Figure 2-2. Heat supplied to the generator boils the weak solution from the absorber, releasing water vapor and leaving behind a LiBr-rich solution. This concentrated solution is returned to the absorber through a throttling valve to maintain the pressure differential between the high- and low-pressure sides of the system. The water vapor produced in the generator is condensed in the condenser and then expanded through an expansion device to the evaporator pressure. In the evaporator, the refrigerant absorbs heat and evaporates, producing the cooling effect. The low-pressure vapor is then absorbed by the LiBr solution in the absorber, releasing heat that is removed by cooling water. A pump circulates the weak solution, with part of it directed to the generator through the solution heat exchanger, while the remainder is mixed with the concentrated solution returning

from the generator. This intermediate solution then flows back to the absorber to complete the cycle.

2.3.3 Steam methane reformer

Steam methane reforming (SMR) is a technology used to produce hydrogen from fossil fuels, in which methane reacts with steam to produce mixed of hydrogen, carbon dioxide, and carbon monoxide. Three reversible reactions occur in steam methane reforming, with the strongly endothermic reforming reactions (Eq. 2-15 and Eq. 2-16) and moderately exothermic water gas shift reaction (Eq. 2-17), in which the reaction can be seen as follows [134]:

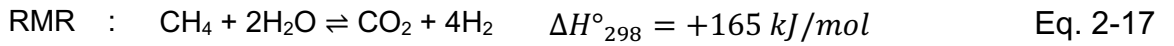
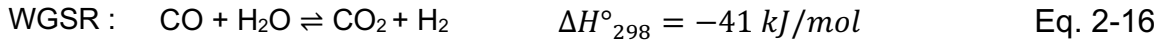
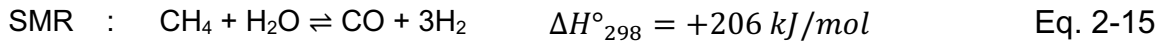


Table 2-4. Kinetic rate expression for reaction in SMR [135].

Name of reaction	Reaction	Kinetic rate expression	
SMR	$\text{CH}_4 + \text{H}_2\text{O} \rightarrow \text{CO} + 3\text{H}_2$	$R_1 = \frac{\frac{k_1}{P_{\text{H}_2}^{2.5}} \left(P_{\text{CH}_4} P_{\text{H}_2\text{O}} - \frac{P_{\text{H}_2}^3 P_{\text{CO}}}{K_{eq,1}} \right)}{[\text{DEN}]^2}$	Eq. 2-18
WGSR	$\text{CO} + \text{H}_2\text{O} \rightarrow \text{CO}_2 + \text{H}_2$	$R_2 = \frac{\frac{k_2}{P_{\text{H}_2}} \left(P_{\text{CO}} P_{\text{H}_2\text{O}} - \frac{P_{\text{H}_2} P_{\text{CO}_2}}{K_{eq,2}} \right)}{[\text{DEN}]^2}$	Eq. 2-19
RMR	$\text{CH}_4 + 2\text{H}_2\text{O} \rightarrow \text{CO}_2 + 4\text{H}_2$	$R_3 = \frac{\frac{k_3}{P_{\text{H}_2}^{3.5}} \left(P_{\text{CH}_4} P_{\text{H}_2\text{O}}^2 - \frac{P_{\text{H}_2}^4 P_{\text{CO}_2}}{K_{eq,3}} \right)}{[\text{DEN}]^2}$	Eq. 2-20
$\text{DEN} = 1 + \frac{F_{\text{H}_2\text{O}}^{\text{in}}}{F_{\text{CH}_4}^{\text{in}}} + \frac{F_{\text{H}_2}^{\text{in}}}{F_{\text{CH}_4}^{\text{in}}} + K_{\text{CO}} P_{\text{CO}} + K_{\text{H}_2} P_{\text{H}_2} + K_{\text{CH}_4} P_{\text{CH}_4} + \frac{K_{\text{H}_2\text{O}} P_{\text{H}_2\text{O}}}{P_{\text{H}_2}}$			Eq. 2-21

The kinetic rate expressions for these reactions are presented in Table 2-4, where k denotes the kinetic constant, with the corresponding expressions for k_1 , k_2 , and k_3 provided in Table 2-5. Similarly, K is the chemical equilibrium constant, with the corresponding expressions for K_1 , K_2 , K_3 , K_{CO} , K_{H_2} , K_{CH_4} and $K_{\text{H}_2\text{O}}$ given in Table 2-5.

Table 2-5. Kinetic expressions, Aspen Plus driving-force coefficients, and adsorption constants used in the reactor models [136].

Variable	Kinetic expression	Aspen coefficient for driving force constant
SMR		
k_1	$4.23 \times 10^{15} \exp\left(\frac{-240.1}{RT}\right)$	Term 1 $A = 33.55$ $B = -28,879$
k_1/K_1	$4.23 \times 10^{15} \exp\left(\frac{-240.1}{RT}\right) / \exp\left(53.093 - \frac{26892}{T}\right)$	Term 2 $A = -19.55$ $B = -1987$
WGSR		
k_2	$1.96 \times 10^6 \exp\left(\frac{-67.13}{RT}\right)$	Term 1 $A = -5.22$ $B = -8074.33$
k_2/K_2	$1.96 \times 10^6 \exp\left(\frac{-67.13}{RT}\right) / \exp\left(-4.0089 + \frac{4340.7}{T}\right)$	Term 2 $A = -1.2111$ $B = -12415.03$
RMR		
k_3	$1.020 \times 10^{15} \exp\left(\frac{-243.9}{RT}\right)$	Term 1 $A = 32.13$ $B = -29336.1$
k_3/K_3	$1.020 \times 10^{15} \exp\left(\frac{-243.9}{RT}\right) / \exp\left(49.085 - \frac{22552}{T}\right)$	Term 2 $A = -16.955$ $B = -6784.1$
Denominator		
Variable	Kinetic expression	Adsorption constants
K_{CO}	$8.23 \times 10^{-5} \exp\left(\frac{70.65}{T}\right)$	$A = -20.92$ $B = 8497.72$
K_{H_2}	$6.12 \times 10^{-9} \exp\left(\frac{82.90}{T}\right)$	$A = -30.42$ $B = 9971.13$
K_{CH_4}	$6.65 \times 10^{-4} \exp\left(\frac{38.28}{T}\right)$	$A = -18.83$ $B = 4604.28$
K_{H_2O}	$1.77 \times 10^5 \exp\left(\frac{-88.68}{T}\right)$	$A = 12.08$ $B = -10666.3$

Since reforming is an endothermic process, it is promoted at high operating

temperatures [134]. As it is also accompanied by an increase in volume, lower pressure further favors the process. Conversely, the WGS reaction is exothermic and therefore benefits from lower temperatures, while pressure variations have little impact on its equilibrium [134]. The addition of excess steam enhances methane conversion but requires additional energy for steam generation [134]. In practical applications, a steam to carbon ratio (S/C), expressed as $\frac{P_{H_2O}}{P_{CH_4}}$, is typically maintained at around 3 [134]. This operating condition not only facilitates methane conversion but also reduces the risk of coke formation [134].

2.3.4 Liquefaction

Stranded gas refers to resources that are uneconomical to develop due to remoteness or small reserve size [137]. Current exploitation options include: (i) volume reduction technologies such as liquefied natural gas (LNG), compressed natural gas (CNG), and natural gas hydrate (NGH) [138], [139], [140], [141], [142], [143], (ii) conversion to liquid fuels and chemicals via gas to liquid (GTL) [144], [145], (iii) conversion to electricity for subsea transmission (GTW) [138], [139], [140], [141], [142], [143], [145]. Among these, LNG has the longest history, with nearly a century of development, and continues to dominate due to rapid growth in liquefaction, regasification, and transport capacity [146]. More recently, small scale LNG technologies have emerged, offering lower investment costs, and operational flexibility, making them attractive for stranded gas development.

The technological advances made in LNG liquefaction for stranded natural gas provide a direct foundation for renewable gas applications. In particular, the same principles can be applied to biogas derived biomethane, which after purification can be liquefied to produce liquid biomethane. Similar to LNG, liquefied biomethane offers significant volume reduction (by approximately 600 times), enabling efficient storage, transport, and distribution. This is especially important for decentralized or small-scale anaerobic digestion facilities located far from gas grids or urban demand centers, where direct injection or pipeline connection is not feasible.

Qyyum et al. [147] investigated nitrogen expander-based liquefaction systems for producing liquid biomethane, highlighting their advantages of low capital cost, operational simplicity, and compact design. Their study confirmed that liquefaction

is a feasible and cost-effective option for transporting upgraded biogas (biomethane) over long distances. Pellegrini et al. [39] explored cryogenic biogas liquefaction and demonstrated that low-temperature processes, particularly dual-pressure cryogenic distillation outperform conventional amine scrubbing, with the cryogenic conditions used in upgrading synergistically supporting liquid biomethane production. Gustafsson and Svensson [148] assessed the environmental performance of LNG, diesel, and liquid biomethane. Their findings indicated that while LNG can increase climate impacts relative to diesel by up to 10%, liquid biomethane significantly reduces emissions compared to both. Using a German electricity mix, liquid biomethane from manure reduced climate change impacts by 45–70% relative to diesel, while liquid biomethane from food waste achieved reductions of 50–75%.

2.3.5 Pressure swing adsorption (PSA)

Hydrogen purification technologies are generally classified into three main categories based on their underlying principles: membrane separation, physical adsorption, and chemical separation. Among these, gas separation by adsorption is recognized as one of the most cost-effective and widely applied methods in industry. Within this category, PSA has been the most extensively developed and commercialized, with applications in hydrogen purification [149], [150], air separation [151], [152], and carbon capture [153], [154].

The PSA process operates by periodically varying the pressure of a gas stream, exploiting differences in the adsorption capacities of specific gases on solid adsorbents. The efficiency of separation depends strongly on the choice of adsorbent material and the configuration of the process. Hydrogen, with its exceptionally low adsorption capacity compared to other common gas molecules such as CO₂, CO, and CH₄, is especially well suited for separation by PSA [155]. As a result, PSA systems are capable of achieving hydrogen purities up to 99.99% from industrial H₂-rich gas mixtures, establishing the technology as the benchmark for large-scale hydrogen purification [156].

Beyond industrial applications, PSA has also been applied to hydrogen purification from waste utilization pathways. For example, Pallozzi et al. [157] evaluated a

prototype biomass gasification system consisting of a dual fluidized bed gasifier, WGSR, desulphurization reactor, and PSA integrated with a micro gas turbine. The system achieved a maximum hydrogen yield of 75.2 gH₂/kg biomass. Similarly, Moneti et al. [158] investigated the influence of operating conditions in a bubbling fluidized bed biomass gasifier equipped with a WGSR and PSA. Their results showed that hydrogen yield increased with higher gasification temperature and steam-to-biomass ratios, though these conditions also raised the overall energy demand. The global efficiency of the process was found to depend strongly on the PSA unit, since its off-gas comprising residual CO, CH₄, and H₂ could be combusted in the dual fluidized bed gasifier to provide additional heat, reducing or eliminating the need for auxiliary fuel.

2.4 Carbon capture and storage (CCS)

CCS technologies applied to combustion-based energy systems are commonly divided into three main approaches: pre-combustion capture, oxy-fuel combustion, and post-combustion capture [159]. The process flow for each approach is illustrated in Figure 2-3. In the pre-combustion route, the fuel is converted to produce a synthesis gas, from which CO₂ is separated before the hydrogen-rich gas is combusted for power generation [159]. In oxy-fuel combustion, air is replaced by a mixture of oxygen and recycled CO₂ as the oxidant. This produces a flue gas consisting mainly of CO₂ and water vapour, which can be condensed and separated relatively easily [159]. A portion of the flue gas is usually recirculated to regulate flame temperature and maintain stable operation. Post-combustion capture, by contrast, removes CO₂ from the flue gas after conventional combustion has occurred. Because it does not require major modifications to the combustion process itself, this option is considered the most straightforward for retrofitting existing facilities [159].

MEA is the most widely applied solvent in both pre- and post-combustion carbon capture processes. Chemically, MEA has a melting point of 10.3°C and a boiling point of approximately 170°C [160]. It is characterized by a relatively high CO₂ absorption capacity, fast reaction kinetics, and high removal efficiency [161]. An additional advantage is its effectiveness in treating gas streams with low CO₂

concentrations [162]. However, MEA also has a notable drawback, namely its limited resistance to thermal and oxidative degradation [163]. Despite this drawback, MEA remains attractive because of its comparatively low management costs [164]. In addition, MEA is well-characterized, non-proprietary, relatively inexpensive, and biodegradable, with established reclamation methods and limited environmental and health impacts, making it a reliable benchmark solvent for CO₂ capture studies [165]. Its extensive documentation also provides a robust baseline for process modelling, techno-economic comparison, and policy-relevant analysis, not because it is superior in all respect, but because it remains the most widely studied and commonly adopted reference solvent in the field.

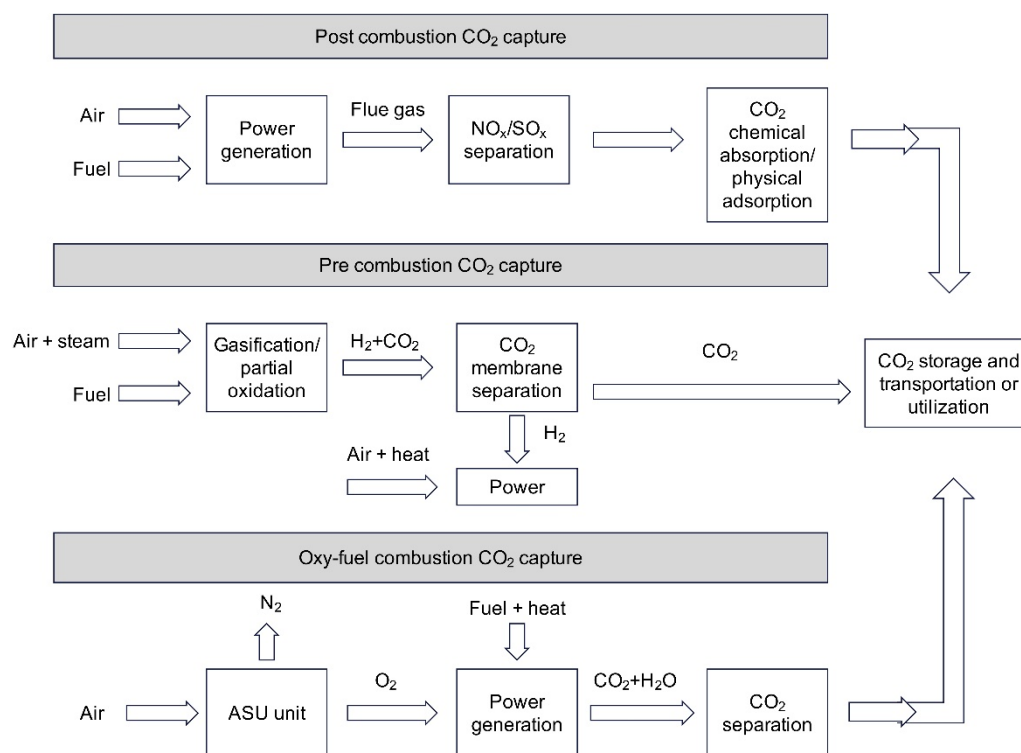
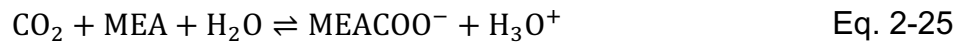


Figure 2-3. Schematic representation of post-combustion, pre-combustion, and oxy-fuel CO₂ capture technologies [159].

In practice, aqueous MEA solutions are typically prepared at concentrations of 15–30 wt.%. Increasing the solvent concentration can reduce the required circulation flow rate, thereby lowering operating costs, but at the expense of increased corrosion risk. In this thesis, a concentration of 35 wt.% MEA is adopted, as it provides a practical balance between minimizing solvent flow requirements and maintaining acceptable operational risks. Furthermore, several previous design

studies have also employed 35 wt.% MEA [165], [166], [167], making it a relevant and widely accepted reference point for this work.

In a conventional MEA-based carbon capture system, the process typically consists of two main columns: the absorber and the stripper (desorber), as illustrated in Figure 2-4. In the absorber, CO₂ from the flue gas is removed by contact with a lean MEA solution through fast and reversible chemical reactions. Once loaded with CO₂, the solvent becomes “rich” and is transferred to the stripper, where heat is applied to release the CO₂ and regenerate the solvent. The regenerated lean solvent is then recycled back to the absorber. To improve energy efficiency, a cross-heat exchanger is employed to preheat the rich solvent using heat recovered from the lean solvent stream. As reported in many works observing the MEA carbon capture, both kinetic and equilibrium reactions are involved. A set including three ionic equilibrium reactions (Eq. 2-22 - Eq. 2-24) and two kinetic reversible reactions involving the CO₂ (Eq. 2-25 and Eq. 2-26) was considered [168].



Increasing the CO₂ capture rate improves environmental performance by enhancing overall abatement but also raises the energy requirement for solvent regeneration. This additional steam demand for the reboiler reduces the net power output of the system. Sanpasertparnich et al. [169] reported that the steam flow requirement increases exponentially with higher capture rates: increasing the capture rate from 70% to 80% requires approximately 2.6% more steam, whereas an increase from 80% to 90% requires about 6.7% more steam. In this thesis, a capture rate of 95% is applied to maximize the climate mitigation benefits of the system, ensuring near-complete CO₂ removal from the flue gas.

Once captured, carbon dioxide is compressed into a liquid state to facilitate transportation [170]. Pipeline networks are the most common means of transport,

with the CO₂ delivered to underground storage facilities for long-term sequestration [171]. In practice, CO₂ is usually transported in a supercritical state. Typical operating conditions for CO₂ pipelines range between 85–150 bar and 13–44°C, which maintain stability and minimise energy losses [170], [172]. Storage options include injection into depleted oil and gas reservoirs, abandoned gas fields, or deep saline aquifers, with onshore depleted reservoirs generally representing the least costly option [173]. The suitability of geological storage is determined by factors such as injection capacity, reservoir design (including offshore enhanced oil recovery projects), displacement of pore fluids, monitoring requirements, and assurance of long-term containment to prevent leakage [173].

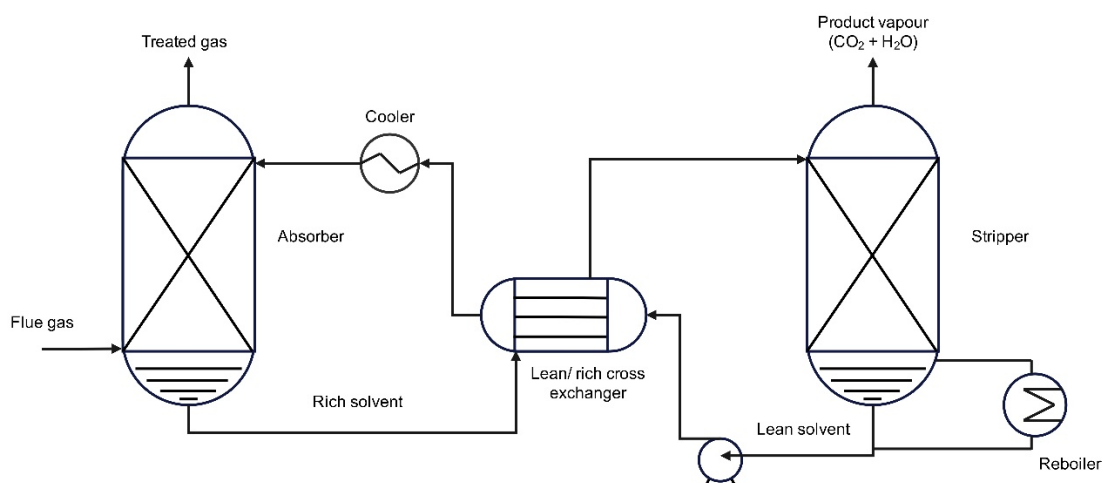


Figure 2-4. Process flow diagram of a conventional MEA-based CO₂ capture system, showing the absorber, stripper, reboiler, cooler, and the lean/rich cross heat exchanger [168].

Because major CO₂ emission sources and suitable storage sites are often geographically distant, the development of extensive pipeline networks is considered unavoidable [170]. The cost of pipeline transport is highly sensitive to both the quantity of CO₂ transported and its phase state [174]. While transportation and storage contribute significantly to the overall economics of carbon capture and storage (CCS), the capture process itself accounts for the majority of costs, approximately 75% of the total, making it the most capital- and energy-intensive component of the CCS chain [175].

Despite being one of the most effective approaches for reducing greenhouse gas emissions, CCS technologies face several limitations. Their main drawback is the high energy demand associated with CO₂ capture and compression, which can reduce plant efficiency by approximately 7-12% [176]. This energy penalty also contributes significantly to the overall cost of CCS deployment [159], [177]. For example, electricity prices may increase by up to 42% in natural gas combined cycle plants and 62% in pulverized supercritical coal plants retrofitted with CCS [159]. Although retrofitting existing plants with CCS is often less expensive than constructing entirely new CCS-equipped facilities [178], substantial technological progress is still required to reduce costs and improve the affordability of large-scale deployment [179].

Nevertheless, captured CO₂ can also provide wider benefits. In addition to enabling negative emissions, CCS and carbon capture and utilization (CCU) can create economic opportunities for industry, such as through carbon credits and certificates, while also supporting broader transitions towards carbon neutrality [57]. In waste-utilization systems, anaerobic digestion is often considered carbon neutral because it primarily emits biogenic CO₂. Gasification can also approach carbon neutrality when biogenic feedstocks are used. However, the integration of CCS can further reduce emissions and improve overall carbon performance. For instance, Cormos et al. [180] evaluated an IGCC power plant co-fuelled with coal and biomass, integrated with CCS and downstream units such as PSA, methanation, and Fischer–Tropsch synthesis. Their study demonstrated the potential to co-produce power, hydrogen, substitute natural gas, and synthetic fuels while achieving substantial CO₂ reductions. Chai et al. [181] investigated the integration of pyrolysis/gasification with CCU using Aspen Plus modelling. Their findings suggested that adding solid carbon materials (e.g., biochar or carbon-based catalysts) and increasing reforming temperatures to 600–700°C could enhance CO₂ conversion. Similarly, Meerman et al. [182] carried out techno-economic analysis of coal- and biomass-based IGCC plants with CCS.

2.5 Conceptual design and thermodynamic assessments of WtE–CCS systems

Plasma gasification has been widely studied as an advanced thermochemical method for converting waste into syngas. It employs high temperature plasma, typically generated by non-transferred arc torches (~4000°C) to break down waste materials into syngas and vitrified slag suitable for construction application [77], [183], [184], [185], [186]. Compared with conventional incineration and gasification, plasma gasification produces cleaner syngas with higher hydrogen and carbon monoxide content. Studies have also demonstrated its superior energy and environmental performance. For example, Cudjoe and Wang [187] showed that PG outperforms incineration of plastic waste across energy, economic, and environmental metrics. Mazzoni and Janajreh [188] examined co-gasification of plastic and MSW in an IPGCC system, reporting the best performance with 30% plastic and 70% MSW, achieving 38% efficiency. Kwon and Im [189] compared thermal and non-thermal plasma gasification in IPGCC, reporting cold gas efficiencies of 60-80% higher than traditional IGCC systems.

While plasma gasification is effective for dry and high calorific feedstocks such as plastics, food waste presents challenge due to its high moisture content (70-90%). For gasification, the feedstock moisture must be below 30% [190], [191]. Anaerobic digestion is therefore considered a more suitable process for converting food waste into biogas [32], [192]. Although plasma gasification and anaerobic digestion are both well studied individually, no studies have been found that integrate the two processes for dual-feedstock conversion.

In parallel, research interest in integrated systems has grown significantly. Multi-output systems capable of producing electricity, heating, cooling, fuels, and even fresh waters are seen as promising pathways for increasing energy efficiency and reducing greenhouse gas emissions [193], [194], [195], [196], [197]. CCHP systems, in particular, have been identified as a flexible option. However, applications of plasma gasification in CCHP remain limited. Only two studies have reported such integration: Zhang et al. [198], who proposed a plasma gasification based system with SOFC, gas turbine, supercritical CO₂ cycle, and absorption refrigeration; and Zhang et al. [199], who investigated a plasma gasification system

using dried food waste with SOFC, GT, CO₂ cycle, and cascade absorption refrigeration. Beyond these, most plasma gasification studies have focused on syngas [76], [86], [200], [201], hydrogen [202], [203], [204], [205], electricity [186], [206], [207], [208], [209], [210], [211], [212], combined hydrogen and electricity [185], [213], [214], CCHP [198], [199] and methanol [215], [216].

Another important consideration is CCS, which is critical for achieving climate mitigation targets [217], [218]. CCS integration into plasma gasification systems has been investigated for various waste feedstocks [185], [198], [205], [219], [220], [221]. However, only two studies explicitly considered plastic waste: one examined post-combustion CCS with 90% efficiency, limited to hydrogen and electricity [185]; the other investigated pre-combustion CCS at the same efficiency, also limited to hydrogen and electricity [221]. To date, no study has compared the pre-combustion and post-combustion CCS configurations at higher capture efficiencies (e.g., 95%), which is essential for identifying the most efficient integration strategy for maximizing carbon mitigation in plasma gasification systems. Furthermore, research combining CCS with dual feedstock plasma gasification and anaerobic digestion systems and linking to multiple energy outputs, remains very limited.

While this study is applied in the context of Indonesia, the system configuration is adaptable and could be implemented in other regions facing similar waste and energy challenge. Related research in different geographical contexts reinforces these gaps. For example, Albizzati et al. [222] evaluated waste treatment strategies in the EU using LCA but focused on conventional methods, such as composting, recycling, and incineration. Rafiquee and Shabbiruddin [223] investigated smart waste management options in India but concentrated on collection and handling rather than energy recovery. Kua et al. [224] assessed anaerobic digestion and pyrolysis of multiple waste streams in Singapore, identifying climate benefits but without CCS and with outputs limited to heat and electricity.

Together, the literature shows that plasma gasification, anaerobic digestion, CCHP systems, and CCS have been studied individually, yet their combined application within a single, multi output waste to energy system remains largely unexplored. From this review, several research gaps can be identified:

1. Lack of integration of plasma gasification for plastic waste with anaerobic

digestion for food waste.

2. Limited research on multi-output systems (electricity, heating, cooling, fuels) incorporating plasma gasification and anaerobic digestion.
3. Few studies examining the role of CCS in dual feedstock waste to energy systems based on plasma gasification integration; existing work is confined to single outputs (hydrogen, electricity).
4. No comparative studies of pre- and post- CCS in plasma gasification systems, particularly at higher capture efficiency.
5. Existing waste management studies in different regions (EU, India, Singapore) mainly focus on conventional methods, infrastructure, or limited energy outputs, with little consideration of advanced integrated WTE systems that include CCS and multi-output generation.
6. Need for comprehensive thermodynamic assessments that maximize energy and exergy efficiency, heat integration, and overall system sustainability.

2.6 Economic and life cycle assessments of WtE–CCS systems

WTE technologies have been increasingly assessed not only for their technical feasibility but also for their economic and environmental performance. Economic assessment and LCA are widely applied tools for evaluating viability and sustainability of energy conversion systems. Economic assessment provides insights into capital investment, operational costs, and profitability, while LCA quantifies environmental impacts across the system lifecycle, including greenhouse gas (GHG) emission, air pollutants, and resource use.

In the context of thermal treatment of waste, incineration has historically dominated, but it faces criticism due to high GHG emissions and air quality concerns [225]. Plasma gasification, by contrast, offers a cleaner thermochemical pathway, producing a low tar syngas suitable for downstream conversion and a vitrified slag by product that can be repurposed in construction [77], [79], [80], [186], [226], [227]. Plasma gasification has been extensively studied for syngas, hydrogen, electricity, and methanol production [93], [185], [187], [206], [221], [228], [229], [230], [231], [232], [233], [234], and many works have also conducted economic and LCA evaluations. These evaluations, however, have largely examined plasma

gasification as a standalone technology, with very limited attention to multi-output or integrated system configurations.

Anaerobic digestion is well established for high moisture food waste [32], [192] and several studies have examined its environmental performance, often highlighting its potential to reduce methane emissions compared to landfill disposal. Yet, economic and LCA studies generally treat plasma gasification and anaerobic digestion separately, with limited research on integrated systems. Combining plasma gasification of plastics with anaerobic digestion of food waste could provide both economic and environmental benefits by valorizing two major streams simultaneously, but this integration has not been comprehensively evaluated in the literature.

CCS integration into WtE systems has been investigated to a limited extent. Two studies considered CCS in plasma gasification of plastic waste: one assessed post-combustion MEA capture with a 90% carbon capture rate [185], and another studied pre-combustion capture at the same efficiency [221]. Both were restricted to single outputs (hydrogen and electricity), and neither included TEA nor LCA evaluations. Moreover, comparisons of pre- and post-combustion CCS configurations in plasma gasification systems remain scarce, despite the importance of such analysis for identifying optimal carbon mitigation pathways.

From the reviewed literature, three clear research gaps can be identified:

1. No study has applied the economic and LCA to an integrated dual feedstock plasma gasification and anaerobic digestion system.
2. Dual-feedstock WtE systems producing multiple energy outputs (such as CCHP and liquid biomethane) remain largely unexplored from an integrated economic and environmental perspectives.
3. CCS integration within economic assessment and LCA frameworks for combined plasma gasification and anaerobic digestion system has not been systematically studied.

2.7 Hydrogen and CCHP generation from waste via combined plasma gasification and anaerobic digestion

Hydrogen is increasingly recognized as a critical energy carrier for advancing decarbonization, with a wide range of production pathways distinguished by their feedstocks, methods, and associated carbon intensities [48], [235]. Green hydrogen from renewable-powered electrolysis is considered the most sustainable option, while blue hydrogen from fossil sources incorporates CCS to lower emissions. Although projections from the IEA indicate substantial cost reductions in low-carbon hydrogen production by 2050, particularly for electrolysis [236], alternative feedstocks, including waste, are gaining attention for their potential to contribute to hydrogen supply.

A variety of waste streams, including food waste [237], [238], sewage sludge [239], [240], medical waste [183], [203], woody biomass [241], [242], plastic waste [243], [244], have been investigated for hydrogen production using both thermochemical and biochemical methods. Plasma gasification has emerged as a promising thermochemical technology due to its ability to operate at extremely high temperatures (1200-5000°C) and achieve near complete feedstock breakdown. Plasma gasification produces a syngas rich in H₂, CO, and CO₂ with low tar and impurity content, while also generating vitrified slag as a stable by product [75], [245], [246], [247]. Reported cold gas efficiencies typically range from 69% to 84%, depending on feedstock, and operational parameters [79], [80], [248].

Several recent studies have examined plasma gasification specifically for hydrogen production, often coupled with purification and CCS. Mallick and Prabu [249] integrated plasma gasification of refused derived fuel with a post combustion CCS unit, achieving energy and exergy efficiencies of 48.3% and 45.9% and a LCOH of 1.01 USD/kg. In a subsequent study [185], they introduced chemical looping reforming, reporting efficiencies of up to 72.4% depending on feedstock. Aich et al. [214] compared different water gas shift reactor designs, showing that a sorption-enhanced configuration substantially lowered hydrogen costs and global warming impacts compared with conventional designs. Yin et al. [203] investigated hydrogen from medical waste via plasma gasification and ionic liquid absorption, achieving an

exergy efficiency of 70.8%, though environmental impacts were not assessed. Zhao et al. [234] analysed hydrogen production from mixed biomass and plastics, reporting 57.8% energy efficiency and CO₂ emission intensity of 234.9 kg/MWh. Zhang et al. [250] added an LCA perspective, showing that over 80% of energy use was concentrated in raw material collection and gasification, with a GWP of 11.01 kgCO₂-eq/kg biomass processed.

While these studies demonstrate the feasibility of plasma gasification for hydrogen production, they have largely focused on hydrogen as a single product. Opportunities to increase system efficiency through multi output configurations that combine plasma gasification with downstream processes, such as coupling hydrogen generation with CCHP, remain underexplored. Moreover, integration of plasma gasification with anaerobic digestion has received little attention, despite the latter's suitability for treating high moisture organic fractions such as food waste. Comprehensive TEA and LCA of such integrated systems, particularly when incorporating CCS, remain scarce.

From the reviewed literature, the research gaps can be identified:

1. Limited exploration of multi-output configurations, with most plasma gasification studies focusing solely on hydrogen rather than integrating co-products such as CCHP.
2. Lack of integration of plasma gasification with anaerobic digestion, despite anaerobic digestion being highly suitable for high moisture organic waste streams.
3. Few comprehensive TEA-LCA studies of plasma gasification-based hydrogen systems, particularly those that consider multiple energy outputs alongside CCS.

3 METHODOLOGY

This section begins with the description of the proposed system and several assessments conducted to understand the behavior of the energy system, such as thermodynamic analysis, thermo-economic analysis, and environmental impact analysis. In addition, the model simulation and validation procedures are explained. The preliminary result is also presented at the end of this chapter.

3.1 Process modelling

Process within the observed energy system is simulated using Aspen Plus. Aspen Plus adopts a flowsheet simulation approach to systematically model each stage. This includes the conversion of the raw material to the final output under steady state conditions. The translation of a process into an Aspen Plus model involves the following steps [251]:

1. Specify the chemical components involved in the process by selecting them from the Aspen Plus databanks.
2. Select appropriate thermodynamic models to represent the physical properties of components and mixtures, which are built into Aspen Plus.
3. Develop the process flowsheet:
 - a. Identify the equipment involved in the process.
 - b. Define the input and output process streams.
 - c. Select suitable equipment models from the Aspen Plus Model Library and drag them onto the flowsheet.
 - d. Label each equipment model (block) and connect the blocks with the process streams.
4. Specify the component flow rates and define the thermodynamic conditions, such as temperature, pressure, and composition for all feed streams.
5. Set the operating conditions for the equipment model (block).

The simulation results from Aspen Plus are presented in stream tables, which provide detailed information on mass flow rate, temperature, pressure, enthalpy, entropy and composition of each stream. These results enable users to perform mass and energy balance calculations, ensuring a comprehensive understanding of the process performance.

Aspen Plus V14 is used in this study to model most of the equipment, while Aspen Plus V12 is specifically used for the CCS plant. This is because the CCS model in this study is based on the MEA Steady-State Model, developed by the DOE Carbon Capture Simulation Initiative (CCSI) [252], which was originally designed using Aspen Plus V12. The input data for the process streams and process units are obtained from a comprehensive literature review. Detailed input data can be found in the chapter where the case study is explained. In addition, selected Aspen Plus model configurations for representative cases are presented in Appendix subsection 9.1 to improve transparency and reproducibility of the simulation methodology.

3.2 Thermodynamic assessment

3.2.1 Energy analysis

Energy analysis is a conventional method of assessing the way energy is used in an operation involving the physical or chemical processing of materials and the transfer and/or conversion of energy. This usually encompasses performing energy balances, which are based on the first law of thermodynamics, and evaluating the energy efficiencies. This balance is applied to determine the reduced waste exergy emissions, such as heat losses, and to promote waste and heat recovery.

Mass Balance Equation

Mass balance is a form of conservation of mass principle frequently employed in engineering analysis. The principle is characterized by a specific region of space called a control volume. An expression for the mass balance in terms of rate for control volumes with several inlets and exits can be expressed as follows:

$$\frac{dm_{cv}}{dt} = \sum_i \dot{m}_i - \sum_e \dot{m}_e \quad \text{Eq. 3-1}$$

where m_{cv} denotes mass within the control volume at time t , and $\dot{m}_i$ and $\dot{m}_o$ are the mass flow rates at the inlet and exit, respectively.

Energy Balance Equation

For a control volume, the energy balance considers all the energy components from the fluid flowing across the control surface (at the state inlet and exit) as follows:

$$\frac{dE_{cv}}{dt} = \dot{Q}_{cv} - \dot{W}_{cv} + \sum_i \dot{m}_i \left(h_i + \frac{V_i^2}{2} + gz_i \right) - \sum_e \dot{m}_e \left(h_e + \frac{V_e^2}{2} + gz_e \right) \quad \text{Eq. 3-2}$$

where $\frac{dE_{cv}}{dt}$ specifies the time rate of change of energy within the control volume, and $\dot{Q}_{cv}$ and $\dot{W}_{cv}$ are the heat transfer rate and work rate, respectively. The terms h , V , g , and z indicate specific enthalpy, velocity, standard acceleration of gravity, and elevation at the inlet i and exit e , respectively.

3.2.2 Exergy analysis

The exergy analysis is performed based on both the first law of thermodynamics and the second law of thermodynamics which can identify the locations of the energy degradation in a process for future operation or technology improvement. In addition, the exergy analysis can measure the quality of heat in a waste stream.

The main aim of the exergy analysis is to determine meaningful (exergy) efficiencies and the causes and true magnitudes of exergy losses as follows.

$$\frac{dE_{cv}}{dt} = \sum_j \dot{E}_{qj} - \dot{W}_{cv} + \sum_i \dot{E}_{fi} - \sum_e \dot{E}_{fe} - \dot{E}_d \quad \text{Eq. 3-3}$$

where

$$\dot{E}_{qj} = \left(1 - \frac{T_0}{T_j} \right) \dot{Q}_j \quad \text{Eq. 3-4}$$

$$\dot{E}_{fi} = \dot{m}_i e_{fi} \quad \text{Eq. 3-5}$$

$$\dot{E}_{fe} = \dot{m}_e e_{fe} \quad \text{Eq. 3-6}$$

and $\frac{dE_{cv}}{dt}$ is the time rate of change of exergy within the control volume. The terms $\dot{E}_{qj}$ specify the rate of exergy transfer accompanying the heat transfer at the rate $\dot{Q}_j$ and the temperature is T_j . $\dot{E}_{fi}$ and $\dot{E}_{fe}$ are the exergy of the flow entering at the inlet i and the exergy of the flow exiting at exit e , respectively, and $\dot{E}_d$ indicates the exergy destruction rate.

3.2.3 The energy and exergy content of the feedstock

Table 3-1 presents the proximate and ultimate analysis of the food waste and mixed plastic waste used in the current study. The chemical composition of food waste and mixed plastic waste is inputted into the anaerobic digestion and plasma gasification models, respectively. It should be noted that both feedstocks were represented using literature-based data rather than direct characterization of site-specific waste samples. For food waste, only the principal components reported in the source were considered, and therefore the full chemical complexity of real food waste may not be completely represented. For mixed plastic waste, the feedstock was represented by a literature-based mixed plastic sample comprising 58.6 wt.% polyethylene (HDPE and LDPE), 26.7 wt.% polypropylene, 8.7 wt.% polystyrene, 5.6 wt.% polyethylene terephthalate, and 0.4 wt.% thermosetting plastics, rather than exact composition of an actual waste stream. Although these assumptions may not fully capture the heterogeneity of real waste streams, they provide a practical and consistent basis for the present simulation study.

Table 3-1. Proximate and ultimate analysis of the feedstock used in the study.

Characteristics		Food Waste [253]	Mixed Plastic [254]
Proximate Analysis (wt.%, as received)			
Moisture		76.47	0.6
Ash		14.81	13.3
Volatiles		83.1	85.7
Fixed Carbon		2.09	0.4
Ultimate Analysis (wt.%, as dry basis)			
Carbon		40.03	79.77
Hydrogen		5.72	15.47
Nitrogen		1.9	2.76
Oxygen		37.08	2
Sulfur		0.46	0
Chlorine		0	-

In order to assess the thermodynamic efficiency of each system, it is necessary to determine the energy and exergy content of all process streams. The energy content of the feedstock can be determined by employing the subsequent

calculation:

$$\dot{E}_{feedstock} = \dot{m}_{feedstock} \cdot LHV_{feedstock} \quad \text{Eq. 3-7}$$

where $\dot{m}_{feedstock}$ is the mass flow rate of feedstock (kg); $LHV_{feedstock}$ is the lower heating value of feedstock which can be obtained from the equation as follows:

$$\begin{aligned} LHV_{feedstock} = & 0.339 \times \%C + 1.029 \times \%H \\ & + 0.109 \times \%S - 0.112 \times \%O \\ & - 0.025 \times \%W \text{ [MJ/kg]} \end{aligned} \quad \text{Eq. 3-8}$$

where C, H, S, O, and W are the mass fraction of carbon, hydrogen, sulfur, oxygen, and water in the dry feedstock as received (%). The properties of each feedstock can be seen from Table 3-1.

For the exergy content of the feedstock, the empirical correlation proposed by Szargut [255] is employed to determine the specific chemical exergy of solid fuels, which is afterwards utilized for the computation of the chemical exergy of the solid waste. The exergy rate of the feedstocks can be mathematically represented as follows:

$$\dot{e}x_{feedstock} = \lambda LHV_{feedstock} \quad \text{Eq. 3-9}$$

$$\dot{E}x_{feedstock} = \dot{m}_{feedstock} \dot{e}x_{feedstock} \quad \text{Eq. 3-10}$$

where $\dot{e}x_{feedstock}$ is the specific chemical exergy of the feedstock (MJ/kg); $\dot{E}x_{feedstock}$ is the exergy content of the feedstock (MW). Here, λ is the statistical correlation that can be calculated based on the ratio of $\frac{O}{C}$:

$$\lambda = 1.0438 + 0.0158 \frac{H}{C} + 0.0813 \frac{O}{C} \quad \text{for } \frac{O}{C} \leq 0.5 \quad \text{Eq. 3-11}$$

$$\lambda = \frac{1.0414 + 0.0177 \left(\frac{H}{C} \right) - 0.3328 \left(\frac{O}{C} \right) \left[1 + 0.0537 \left(\frac{H}{C} \right) \right]}{1 - 0.4021 \left(\frac{O}{C} \right)} \quad \text{for } 0.5 \leq \frac{O}{C} \leq 2 \quad \text{Eq. 3-12}$$

where H, O, C, are the percentage of hydrogen, oxygen, and carbon in the ultimate analysis, respectively.

3.2.4 The energy and exergy content of the generated gas

For the output gas generated through gasification and anaerobic digestion, the

energy content of individual gases can be determined by employing the following equation:

$$\dot{E}_{gas} = \dot{m}_{gas} \cdot LHV_{gas} \quad \text{Eq. 3-13}$$

$$LHV_{gas} = 1.20087 (\%H_2) + 0.10160(\%CO) + 0.49853(\%CH_4) \text{ [MJ/kg]} \quad \text{Eq. 3-14}$$

where $\dot{E}_{gas}$ is the energy content of the generated gas from either the AD or PG (MW); $\dot{m}_{gas}$ is the mass flow rate of the gas; H_2 , CO , and CH_4 are the mass fraction of hydrogen, carbon monoxide, and methane in the product gas (%); LHV_{gas} is the lower heating value of gas (MJ/kg).

The following equations could be used to calculate the exergy content of the generated gas.

$$\dot{e}x_{gas} = \sum_i^n y_i ex_{0,i} + RT_0 \sum_i^n y_i \ln y_i \quad \text{Eq. 3-15}$$

$$\dot{E}x_{gas} = \dot{m}_{gas} \dot{e}x_{gas} \quad \text{Eq. 3-16}$$

Where $\dot{e}x_{gas}$ is the exergy content of the generated gas (kJ/kg); y is the molar fraction of each substance; and ex_0 is the standard molar chemical exergy (kJ/kmol) at 298 K and 1 atm in which the value of each gas composition can be seen in Table 3-2.

Table 3-2. Standard chemical exergy of selected gases [256].

Substance	Standard Chemical Exergy (kJ/kmol)
Carbon dioxide (CO_2)	19,870
Carbon monoxide (CO)	275,100
Hydrogen (H_2)	236,100
Methane (CH_4)	831,650
Nitrogen (N_2)	720
Oxygen (O_2)	3,970
Water vapor (H_2O)	9,500

3.3 Economic assessment

Assessing the financial viability of the observed energy system requires an accurate production costs estimation. This study focuses on evaluating the economic feasibility of different energy systems by determining three key cost metrics: capital expenditure (CAPEX), operational expenditure (OPEX), and levelized cost of the products. Levelized cost metrics considered include levelized cost of electricity (LCOE), liquid biomethane (LCOBM), waste treated (LCOWT), and hydrogen (LCOH).

A discounted cash flow analysis (DCFA) is performed to calculate these levelized costs metrics under the condition that the net present value (NPV) = 0 or equally internal rate of return (IRR) = *discount rate*. Various financial assumptions supporting this analysis, including the pricing of the by-products, are detailed in Table 3-3. The basis of the DCFA model is defined by the equation presented below.

$$\sum_{n=1}^{n=20} \frac{CF_n}{(1 + IRR)^n} = 0 \quad \text{Eq. 3-17}$$

$$CF_n = P_n(1 - t) + D_n t \quad \text{Eq. 3-18}$$

where CF represents the after-tax cash flow for each year, n is the number of years, P denotes gross profits, t is the tax rate, and D refers to the depreciation of plant assets, assumed to occur over 10 years of operation. Eq. 3-17 and Eq. 3-18 constitute a standard DCFA framework and are widely used for assessing the long-term economic viability of engineering projects [257], [258].

Table 3-3. Parameters for conducting the DCFA.

Parameter	Description
Location	Indonesia
Plant life	20 years
Currency	USD
Base year	2023
Plant capacity	546,394 tonnes of waste/year
Discount rate	10%
Tax rate	30%
Construction period	3 years

First 12 months' expenditures	10% of FCI
Next 12 months' expenditures	50% of FCI
Last 12 months' expenditures	40% of FCI
Depreciation method	Straight line
Depreciation period	10 years
Working capital	5% of FCI

3.3.1 CAPEX estimation

The calculation of capital expenditure (CAPEX) follows a bottom-up approach, in which the cost of each component is estimated as a proportion of the total purchased equipment cost (PEC), as shown in Table 3-4. The PEC is determined using economic evaluation software combined with baseline cost equations derived from prior research studies. The Aspen Plus Economic Evaluator is utilized to estimate the PEC for standard process equipment, such as condensers, evaporators, and pumps. However, for equipment involving complex simulation models, such as gasification (PG), anaerobic digestion (AD), the water-gas shift reactor (WGSR), and the carbon capture and storage (CCS) plant, which exceed the limitations of the Aspen Plus database, cost estimations rely on baseline price equations from previous studies. The detailed of the PEC can be seen in the Chapter 5 and 6. To account for discrepancies in equipment capacity and differences in the year of economic analysis, Eq. 3-19 is applied.

Table 3-4. The approach for estimating CAPEX [257].

Expenses	Cost factor
Installed direct costs (IDC)	
Total purchased equipment cost	1.00*PEC
Purchased equipment installation	0.39*PEC
Instrumentation and controls	0.26*PEC
Piping	0.31*PEC
Electrical systems	0.10*PEC
Non-installed direct costs (NIDC)	
Buildings	0.55*PEC
Yard improvements	0.12*PEC
Land	0.06*PEC
Total direct costs (TDC)	IDC + NIDC
Indirect costs (IC)	0.255*PEC
Fixed capital investment (FCI)	TDC+IC
Start-up costs	0.05*FCI

Interest during construction	Estimated
Total capital requirement (TCR)/ CAPEX	E + F + G
Working capital (WC)	0.05 * FCI
Annualised CAPEX (ACAPEX)	$\text{CAPEX} \times \left(\frac{\text{IRR} \times (1 + \text{IRR})^n}{-1 + (1 + \text{IRR})^n} \right)$

$$C = C_0 \left(\frac{S}{S_0} \right)^f \times \left(\frac{I_{\text{current}}}{I_0} \right) \quad \text{Eq. 3-19}$$

where C represents the cost of the unit when operating at its actual capacity (S), while C_0 denotes the base cost of the unit at a specified reference capacity (S_0). The scaling capacity factor (f) varies depending on the type of process equipment. I_{current} and I_0 correspond to the Chemical Engineering Plant Cost Index (CEPCI) for the year of the current study and the reference year from the literature, respectively. Eq. 3-19 is widely used in TEA to transfer literature-based equipment costs to different plant capacities and reference years [257].

3.3.2 OPEX estimation

Operating expenditure (OPEX) encompasses all recurring costs associated with the routine operations of a plant. OPEX is defined as the sum of fixed operating, and maintenance costs (FOM), and variable operating costs (VC). The calculation of OPEX is based on the assumptions outlined in Table 3-5.

Table 3-5. The approach for estimating OPEX [259].

Expenses	Cost
A. Fixed Operating and Maintenance (FOM)	$\sum(\text{A1 to A8})$
(1) Labour [260]	Eq. 3-20*2.0415 USD/h
(2) Supervision	0.25*A(1)
(3) Direct overhead	0.5*(A(1) + A(2))
(4) General overhead	0.5*(A(1) + A(2) + A(3))
(5) Maintenance Labour	0.015*FCI
(6) Maintenance materials	0.015*FCI
(7) Insurance and tax	0.010*FCI
(8) Financing working capital	0.1*WC
B. Variable Cost (VC)	

See the component of VC in Sub Chapter 5.2.2 and 6.2.4.

$$\text{OPEX} = \text{FOM} + \text{VC}$$

The total required labour is estimated using the methodology proposed by Alkhatay and Gerrard [59], as expressed in the equation below:

$$N_{OL} = (6.29 + 31.7P^2 + 0.23N_{np})^{0.5} \quad \text{Eq. 3-20}$$

$$N_{np} = \sum \text{Equipment} \quad \text{Eq. 3-21}$$

where N_{OL} represents the number of operators assigned per shift, P denotes the number of processing stages involving particulate solids handling, such as transportation and distribution, particulate size control, and particulate removal. Additionally, N_{np} corresponds to the number of non-particulate processing stages, which include compression, heating, cooling, mixing, and reaction. Eq. 3-20 and Eq. 3-21 are established engineering approximations commonly applied to estimate labour requirements when plant-specific industrial data are unavailable [59].

Variable costs (VC) are determined by summing the costs associated with feed water, cooling water, waste treatment, MEA solvent, WGS catalyst, and CO₂ transport and storage. The respective prices are detailed in Sub Chapter 5.2.2 and 6.2.4. Since the simulation does not directly provide water loss data from the cooling system, the required makeup water is estimated based on a general assumption. Therefore, the amount of makeup water (MU_{water}) is calculated using the following equation [261]:

$$MU_{water} = EL + DL + BL \quad \text{Eq. 3-22}$$

where EL represents the evaporation loss which occurs when cold and dry air interacts with hot cooling water. The evaporation rate is typically estimated as 1% of the circulation flow for every 10°F (5.56°C) temperature increase between the cooling tower inlet and outlet. To account for climatic variations, a correction factor is applied. In humid climates, a factor of 0.65 may be used, whereas in extremely dry conditions, the factor ranges between 1.0 and 1.2. Under general conditions, a correction factor of 0.85 is considered reasonable. Evaporation loss (EL) can be

estimated using the following equation [261]:

$$EL = \left(0.85 * \left(\frac{1}{100}\right) * \Delta T\right) * \frac{1}{10} * \text{circulation flow} \quad \text{Eq. 3-23}$$

DL represents drift loss, which occurs when water droplets are carried away in the discharge vapors from the cooling tower. It generally falls within the range of 0.1% and 0.2% of the circulation flow, with a conservative estimate of 0.2%.

BL, or blowdown loss, is the process of removing a portion of the concentrated circulating water to control the buildup of dissolved solids, which increases due to evaporation. This process helps prevent scale formation within the system. In a conservative scenario, four cycles of concentration (*COC*) are assumed [262]. *BL* is estimated using the following equation [261]:

$$BL = \frac{EL}{COC - 1} \quad \text{Eq. 3-24}$$

Eq. 3-22, Eq. 3-23, and Eq. 3-24 are standard engineering correlations used to estimate cooling tower makeup water requirements when plant-specific water loss data are not directly available from the simulation.

3.4 Sensitivity and uncertainty analysis for the economic investigation

Sensitivity and uncertainty analyses are conducted to ensure consistency and reliability of the economic evaluation which assists in identifying the most influential input parameters affecting the estimated indicators. A local sensitivity analysis is performed by varying one parameter at a time while keeping all other parameters constant. The summary table of the independent variables along with their respective lower and upper bounds can be seen in Sub Chapter 0 and 6.2.5. These parameters were chosen due to their significant levels of uncertainty. The low and high ranges for PEC were derived from the classification guidelines of AACE International, which apply to plants with low maturity levels, such as waste-to-energy process plants [263]. Specifically, the PEC values were assumed to vary between -30% and +50%.

The low, median, and high estimates for CO₂ transportation and storage costs are

based on the 2016 UK Department of Energy and Climate Change report [264], since Indonesia has not yet implemented CCS technology, with its first CCS project targeted for operation in 2030 [265]. Other parameters are aligned with typical market-relevant factors, such as tax rates and discount rates [259]. The tax rate ranges from a minimum of 0%, accounting for potential tax exemptions, to a maximum of 40%. The discount rate reflects project investment risk, with a lower bound of 8% and an upper bound of 12%. Tipping fee and carbon credit prices are based on Indonesian median values, with $\pm 20\%$ variation applied following previous studies on investment incentives for plasma gasification combined cycle power plants utilizing municipal solid waste [266].

To assess the combined impact of variations in uncertain parameters on the LCOE, LCO_{bM}, LCO_H, and LCO_{WT}, a well-known statistical method, such as Monte Carlo simulation, is employed. The input parameters and their respective value ranges, which reflect plausible variability based on literature and current market conditions, are detailed in Sub Chapter 0 and 6.2.5. Using Matlab, 10,000 simulation trials were performed, each randomly sampling values within these specified ranges according to predefined distributions.

This extensive sampling allowed for the generation of probability distribution curves for LCOE, LCO_{bM}, LCO_H and LCO_{WT}, from which 95% confidence intervals (CI) were determined. These confidence intervals provide a robust measure of uncertainty around the mean cost estimates. Furthermore, key statistical metrics, including the mean, median, and standard deviation (STD), were computed from the simulation results to summarize the central tendency and variability of the cost outcomes.

3.5 Life cycle assessment (LCA)

LCA has become an essential tool as public awareness of environmental issues continues to grow. Consequently, many sectors adopt this procedure to evaluate the environmental impact of their activities. The primary objective of LCA is to identify the most sustainable product, process, or service with minimal impact on human health and the environment [267]. In addition, LCA serves multiple purposes, including supporting environmental assessment, establishing baseline

data for processes, informing public policy, and aiding in product certification [268]. The environmental impacts of the observed energy system are quantified using a cradle-to-gate approach, in accordance with International Organization for Standardization (ISO) 14040 standards. This assessment excludes the distribution, usage, and final disposal stages. The LCA is performed using SimaPro V.9 software and follows a standardized four-step process, which includes goal and scope definition, inventory analysis, impact assessment, and interpretation. The analysis is based on the system boundaries, focusing on input and output flows while excluding the detailed modeling of internal processes.

3.5.1 Goal and scope definition, functional unit

The goal of this LCA study is to assess and compare the environmental sustainability of converting municipal waste (plastic and food waste) into useful energy outputs. In an LCA study, selecting the functional unit (FU) is crucial, as it significantly impacts the results and must align with the study's goal and scope. For this study, the chosen FU is 1 kg of waste, comprising 74% food waste and 26% plastic waste. The analysis assumes a plant lifespan of 20 years.

3.5.2 Life cycle inventory (LCI)

The LCI for this study is constructed using two primary data sources. The first source consists of mass and energy balances derived from process modelling conducted in Aspen Plus. This dataset provides standardized values for the conversion of primary raw materials (plastic and food waste) into useful outputs, measured per kilogram of plastic and food waste. Additionally, it includes data on various pollutants and waste streams generated during the conversion process.

The second data source is the "Ecoinvent-3" database, accessed via SimaPro, which offers comprehensive inventories. The system model follows "allocation, cut-off by classification system" method. Since the designed energy system is intended for implementation in Indonesia, priority is given to Indonesia-specific LCI data. However, if such data is unavailable, inventories from the "Rest of the World" databases are used as substitutes.

For this assessment, the inventory for infrastructure construction is based on

existing data from the Ecoinvent database. The inventory requirements for newly constructed infrastructure are estimated using the six-tenths factor rule, as represented in the following equation [269] [270].

$$IP_2 = IP_1 \left(\frac{CPP_2}{CPP_1} \right)^k \quad \text{Eq. 3-25}$$

In this equation, IP_1 represents the capacity proportion of the original Ecoinvent process plant, while IP_2 denotes the capacity proportion for the corresponding process in the observed system. Similarly, CPP_1 and CPP_2 correspond to the capacities of the original Ecoinvent process plant and the observed system, respectively. The scaling factor (k), typically set to 0.6 [269], accounts for the non-linear relationship between system capacity and proportional requirements.

3.5.3 Life cycle impact assessment (LCIA)

In this study, the Life Cycle Impact Assessment (LCIA) was performed using the ReCiPe midpoint (H) method. This approach enables a detailed analysis of emissions across various stages of energy production by evaluating 18 midpoint impact categories. Given that the system processes two distinct types of carbon waste: fossil-based waste (plastic waste) and biogenic waste (food waste), it is essential to differentiate between these sources when calculating carbon emissions. This distinction is made by determining the ratio of biogenic carbon in biogas (from food waste) to fossil carbon in syngas (from plastic waste). By establishing this ratio, the contributions of biogenic and fossil carbon to total emissions can be accurately assessed. As a result, in the inventory under the "Emissions to air" category, carbon emissions are categorized into "Carbon dioxide, fossil," "Carbon dioxide, biogenic," "Carbon monoxide, fossil," and "Carbon monoxide, biogenic."

In cases involving CCS plants, captured biogenic CO₂ is considered a "negative emission", assigned a characterization factor of minus one under "Carbon dioxide" in the "Emissions to air" category. However, in the observed system, the captured CO₂ is not entirely biogenic. For example, if the captured biogenic carbon ratio is 0.27, the characterization factor becomes minus 0.27. Conversely, for cases with fully captured fossil CO₂, the characterization factor is set to zero.

When the investigated system produces a variety of products, an allocation

technique must be applied to distribute life cycle inventory data and environmental impacts [271], [272]. This approach assigns emissions to products in accordance with their flow characteristics, such as mass, energy content, carbon content, or the economic value [273]. Although all products in the system contribute to energy generation, their energy quality differs, leading to variations in work potential across different outputs [274]. Therefore, this study adopts an exergy-based allocation method to account for these differences. Similar allocation approaches have been utilised in previous studies [275] [276] [277].

4 Conceptual design and thermodynamic investigation of novel energy and fuel generation systems from municipal waste coupled with carbon capture and storage

Abstract

Waste generation and energy demand are increasing and both require innovative energy symbiosis strategies to meet climate targets. Traditional waste-to-energy processes rely on incineration, but more efficient and sustainable solutions are needed. The aim of the study is to investigate for the first time the feasibility of generating cooling, heating, power (CCHP), and liquid biomethane from plastics and food waste integrated with carbon capture and storage (CCS). The system, modelled in Aspen Plus, consists of a plasma gasifier (PG), anaerobic digester (AD), combined cycle gas turbine (CCGT), absorption refrigeration cooler (ARC), and biomethane liquefier. Two scenarios were analyzed: (1) a standalone CCHP system and (2) its integration with liquid biomethane production. Each scenario includes a baseline (without CCS), pre-combustion CCS, and post-combustion CCS, both with a 95% CO₂ capture fraction. Utilising 5 kg/s of plastic and 13.97 kg/s of food waste, the system generates net power (29.76–85.67 MW), cooling (2.72–4.04 MW), heating (13.99–27.87 MW), and 43.26 MW of liquid biomethane. The highest energy and exergy efficiencies achieved are 49.44% and 41.20%, with carbon emissions ranging from 0.008 to 0.247 kgCO₂/kg waste. The findings of this novel study highlight the potential of integrating several energy systems for an effective waste management strategy that can contribute to the provision of several energy vectors while the inclusion of CCS ensures that significant emission reduction can be attained.

Keywords: Waste management, Waste-to-energy, Plasma gasifier, Carbon capture and storage

4.1 Introduction

The transition towards sustainable development has become a global imperative,

as reflected in the United Nations Sustainable Development Goals (SDGs). Among them, Goal 11 emphasizes the creation of sustainable cities and communities, which includes improving waste management and reducing environmental pollution. MSW management is particularly important in developing countries, where urbanisation and population growth contribute to increasing waste generation. Indonesia, the fourth most populous country in the world, generated approximately 33.14 millions tonnes of MSW in 2022 [278], with plastic and food waste identified as the largest contributors to the overall waste stream. Plastic waste accounted for around 28.4%, while food waste contributed 16% of the total waste [278]. These fractions are particularly problematic: food waste releases methane, a potent greenhouse gas, during decomposition, while plastic waste degrades into microplastics that threatened terrestrial and aquatic ecosystems [279], [280].

Waste can be utilized as a valuable energy source due to its relatively high energy content. Plastic waste is characterized by a higher heating value (HHV) ranging from 23.97 to 46.48 MJ/kg [254], [281]. Meanwhile, the lower heating value (LHV) of food waste ranges from 10.54 to 25.32 MJ/kg [282], based on its composition. Addressing these waste streams is therefore central to both environmental protection and sustainable energy production. Different conversion technologies are available to utilise these waste fractions, broadly classified into thermochemical and biochemical processes. Plasma gasification is a thermochemical technology capable of operating at extreme temperatures ($>4000^{\circ}\text{C}$), enabling complete decomposition of waste into clean syngas and vitrified slag [183]. Anaerobic digestion, on the other hand, is well suited for high moisture organic feedstocks, such as food waste, and produces biogas through microbial activity [32], [192]. Together, these processes provide complementary pathways for converting the dominant waste fraction in Indonesia into useful energy carriers.

The integration of energy conversion technologies into multi output energy systems has attracted increasing attention in recent years. CCHP systems, also referred to as trigeneration systems, can deliver multiple useful energy outputs simultaneously, thereby enhancing overall efficiency and reducing resource consumption. In addition, biogas from anaerobic digestion can be upgraded and liquefied into liquid biomethane (LBM), providing a sustainable alternative to liquefied petroleum gas

(LPG) and enabling decentralised energy supply or transport fuel applications. Furthermore, the incorporation of CCS technologies into WtE systems represents a promising approach for deep decarbonisation, aligning with international climate targets set by the Intergovernmental Panel on Climate Change [218] and the International Energy Agency [217].

Despite these advancements, several research gaps remain. Previous studies have predominantly examined plasma gasification or anaerobic digestion individually, focusing on single feedstocks and limited energy outputs, such as syngas [76], [86], [200], [201], hydrogen [202], [203], [204], [205], electricity [186], [206], [207], [208], [209], [210], [211], [212], combined hydrogen and electricity [185], [213], [214], CCHP [198], [199] and methanol [215], [216]. Only a small number of studies have investigated the use of PG within CCHP configurations, and even fewer have considered the simultaneous utilisation of multiple feedstocks. While CCS has been explored in certain plasma gasification-based systems, it has largely been restricted to limited energy outputs, and has not been comprehensively applied to dual feedstock, multi output configurations [185], [198], [205], [219], [220], [221]. Furthermore, studies that include liquid biomethane production alongside CCHP remain absent. To the best of the author's knowledge, no existing work has reported on the coupling of plasma gasification for plastic waste and anaerobic digestion for food waste within an integrated CCHP and liquid biomethane system enhanced with CCS.

This chapter presents the research published in the author's first journal paper, which addresses the identified gaps through the development of a novel WtE configuration. The proposed system integrates plasma gasification of plastic waste and anaerobic digestion of food waste, with outputs directed into a CCHP system and liquid biomethane production, enhanced by pre and post combustion CCS at 95% capture efficiency. The system is scalable and adaptable, and it has the potential to support centralized grid integration in urban areas as well to serve as a decentralized energy source for off-grid regions. The system can be implemented in hot-humid climates as waste heat from gas turbine exhaust drives an ARC and this can reduce electrical cooling demand. At the same time, in colder regions, it can be used for district heating to provide low-carbon thermal energy. Additionally, liquid

biomethane production offers a sustainable alternative to liquefied petroleum gas (LPG). Thus, it can reduce dependence on fossil fuels and facilitate energy security. Beyond energy recovery, the system supports circular economy principles by repurposing slag for construction materials and digestate as agricultural biofertilizers. The proposed conceptual designs have the potential to set the basis for subsequent detailed engineering studies and inform policy making about sustainable waste management routes that can offer deep decarbonization.

The study performed a detailed thermodynamic assessment to evaluate the performance of individual components and the system as a whole, focusing on energy and exergy efficiency, process integration, and emission reduction potential. Beyond energy recovery, the system concept also supports circular economy practices by enabling the reuse of by products such as vitrified slag in construction and digestate as biofertilizer.

4.2 Methodology

4.2.1 System description

An integrated system for generating electricity, cooling, heating, and liquid fuel has been designed for one of Indonesia's largest landfills. The Benowo landfill, located in Surabaya, the country's second most populous city with approximately 2.8 million residents as of the 2020 census, covers an area of about 37.4 hectares [283], [284]. In 2020, the landfill received 2,222.62 tonnes of waste daily, predominantly composed of 54.31% food waste and 19.44% plastic waste, with the remainder consisting of paper, twigs, rubber, glass, and fibres [278]. Currently, the waste-to-energy system at Benowo is limited to electricity generation, leaving significant potential for expansion into other energy outputs [278].

The energy needs of the surrounding community extend beyond electricity to include cooling, heating, and fuel. In a country with a hot-humid climate like Indonesia, the demand for cooling energy is expected to rise significantly [285], [286]. Waste heat recovered from the landfill's power plant offers a sustainable solution to address these demands, such as providing energy for space cooling. Additionally, the landfill's proximity to coastal areas presents an opportunity to

support the region's salted fish industry, one of Indonesia's key export products. The traditional sun-drying process for fish poses hygiene risks, but utilizing recovered waste heat for fish drying can improve product quality and safety.

Moreover, LPG, a widely used fuel in Indonesian households, can be substituted with liquid biomethane derived from food waste, reducing dependence on natural gas and offering a sustainable alternative. Beyond energy recovery, the system's byproducts present additional value. Slag produced by the PG can be used as a construction material, while dried digestate from anaerobic digestion can serve as an organic fertilizer, supporting circular economy principles.

In response to the growing need for sustainable energy solutions derived from waste, this research proposes two scenarios for optimizing energy recovery and reducing environmental impacts through advanced waste-to-energy technologies. The first scenario involves the application of a CCHP system. Meanwhile, the second scenario integrates the CCHP system with liquid biomethane production. Both scenarios assume the production of syngas and biogas, as shown in Figure 4-1 and Figure 4-2. Specifically, syngas is generated by feeding shredded plastic waste to the PG, and biogas is produced from the anaerobic digestion of food waste. To evaluate the impact of carbon emissions, each scenario examines three configurations: a baseline without CCS, one with pre-combustion CCS (pre-CCS), and one with post-combustion CCS (post-CCS). The CO₂ emissions captured from the system are transported to the nearest CO₂ storage site, the Bawean Basin, which has a storage capacity of 1.16 gigatons and is located approximately 150 kilometers from the landfill [287].

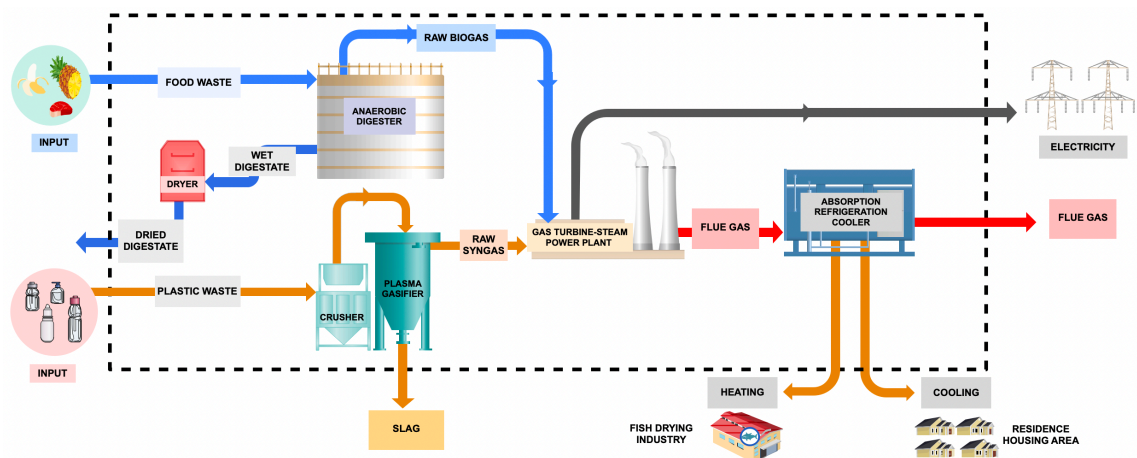
4.2.1.1 Scenario 1 (CCHP)

Figure 4-1a, b, and c are the visual representation of the three configurations of the first scenario. The first configuration as shown in Figure 4-1a is the simplest one and serves as the basis for further comparisons. In this set up, a mixture of raw biogas and syngas drives a combined-cycle gas turbine (CCGT) which utilizes a gas turbine (top cycle) and a steam turbine (bottom cycle) to generate electricity. The residual heat in the flue gas exiting the CCGT unit can be recovered in a single-effect ARC to produce cooling suitable for a housing area. Additionally, the

waste heat generated from the ARC can be used in industrial applications such as fish drying.

The second configuration, illustrated in Figure 4-1b, introduces two additional subsystems, highlighted within the red dotted box. These include a water gas shift (WGS) reactor and a pre-CCS plant. In this setup, the raw syngas is sent to the WGS reactor before it is mixed with the raw biogas. Through the WGS reactor, the CO present in the syngas is transformed into CO₂, hence raising the concentration of CO₂ in the gas stream. The resulting syngas is then mixed with the raw biogas and fed to the pre-CCS plant before combustion in the CCGT.

The final configuration, shown in Figure 4-1c, differs from the others by incorporating a post-combustion carbon capture (post-CCS) plant. Instead of releasing the flue gas directly into the atmosphere, as in the first configuration, the flue gas passes through the post-CCS plant, which is located after the ARC. The post-CCS system captures CO₂ from the flue gas before it is released, as indicated by the red dotted box in Figure 4-1c.



(a)

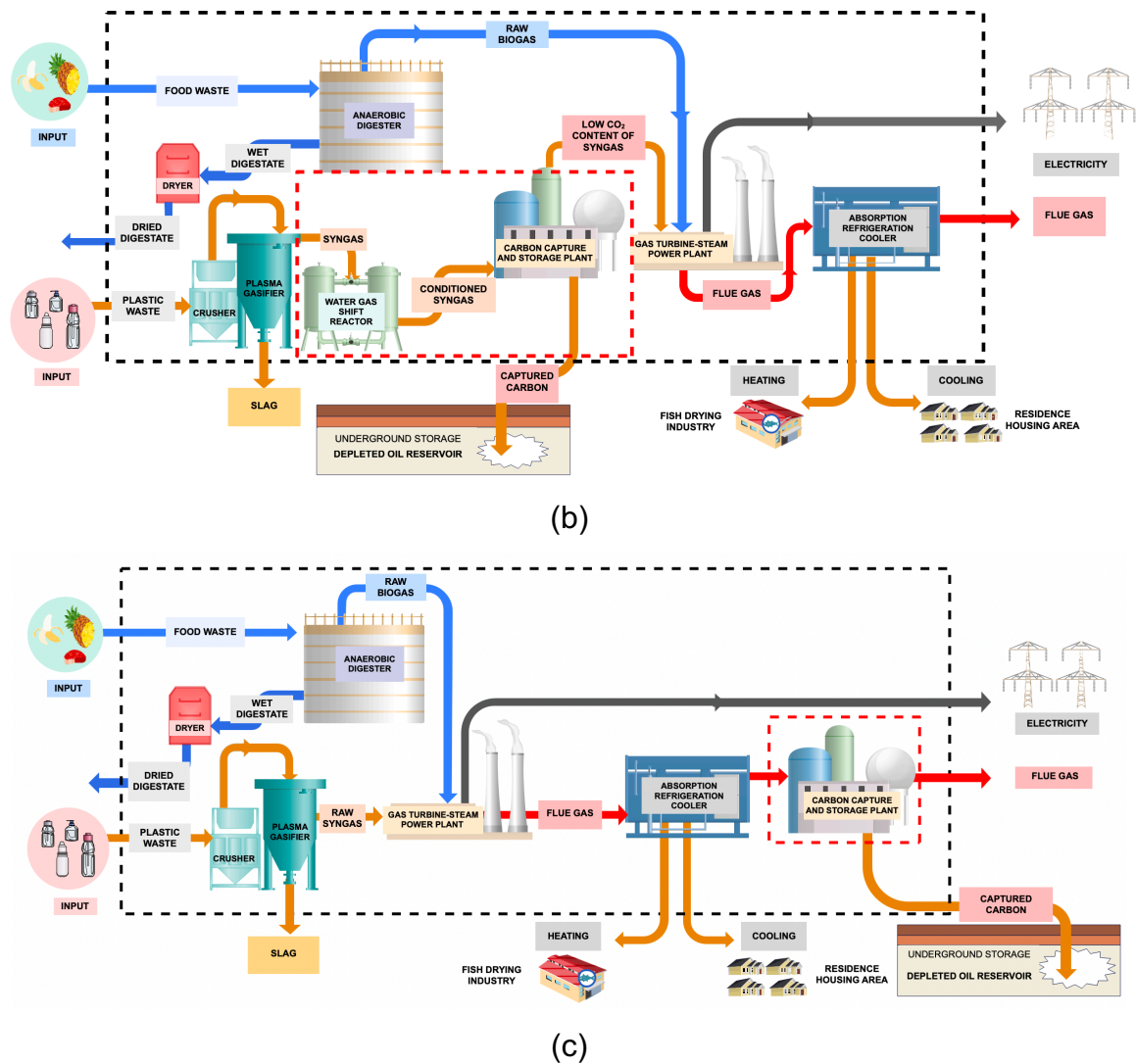
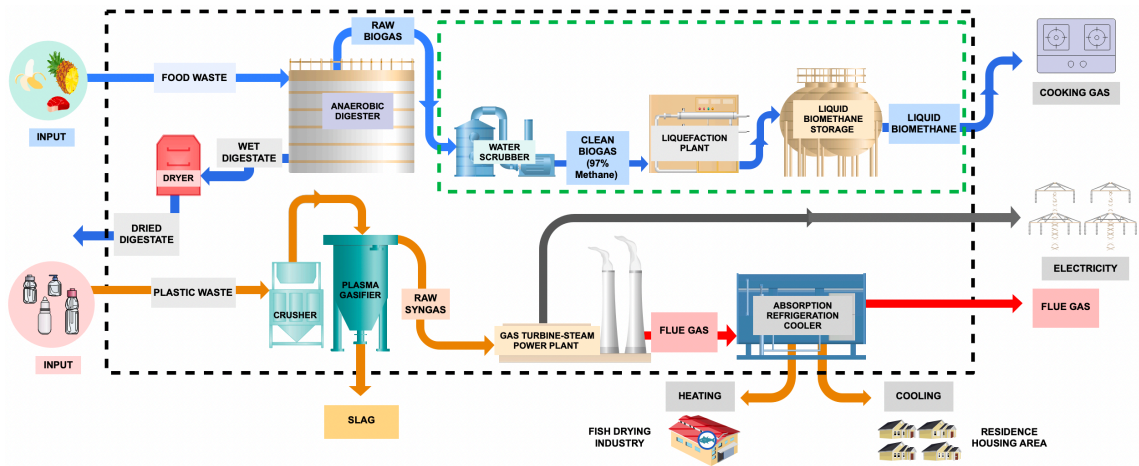


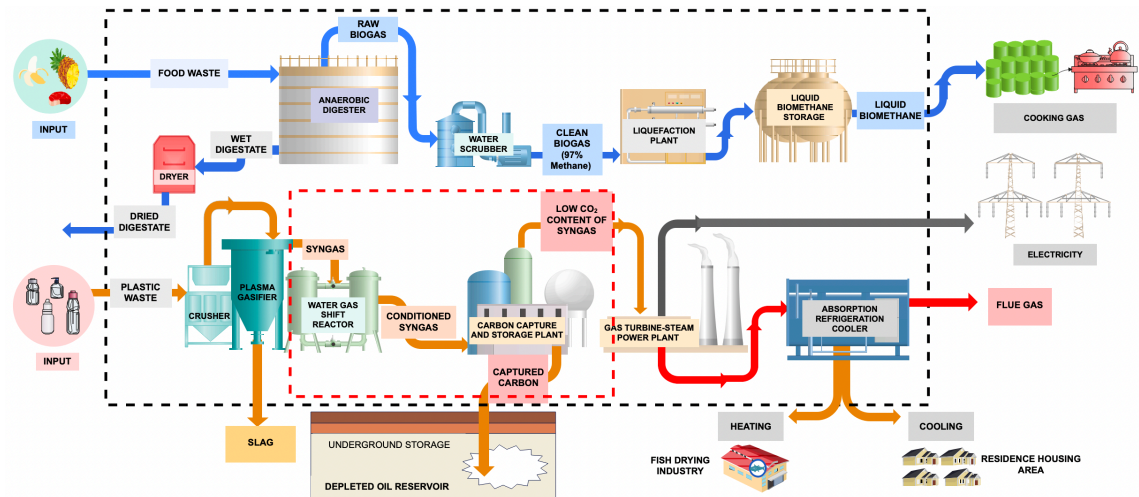
Figure 4-1. Process schematic of the Scenario 1 (products: cooling, heating, and electricity) with three different cases: (a) baseline, (b) integrated with pre-CCS, (c) integrated with post CCS.

4.2.1.2 Scenario 2 (CCHP and fuel production)

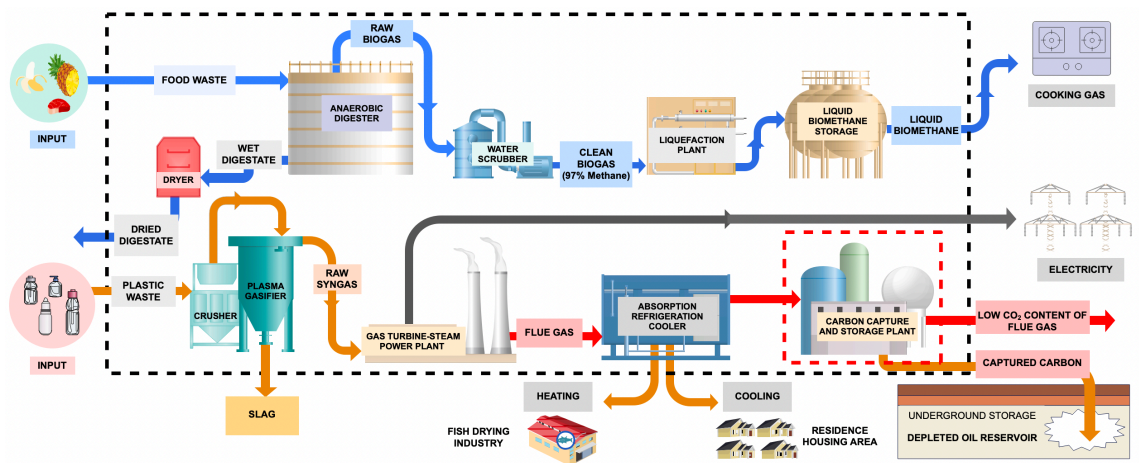
The three configurations in the second scenario are depicted in In this scenario, instead of using the raw biogas produced in the AD unit as a fuel for the CCGT, it is purified in a high-pressure water scrubber (HPWS) to increase the methane concentration. The resulting biomethane is then directed to a methane liquefier (ML) to liquify biomethane, which can be used as cooking gas. In Scenario 2, the CCHP system is fueled solely by syngas to generate electricity. Similar to the Scenario 1, before the flue gas is released from the CCHP, its heat is used by the ARC to generate cooling, and the heat rejected by the ARC is utilized for fish drying.



(a)



(b)



(c)

Figure 4-2. Process schematic of the Scenario 2 (products: cooling, heating, electricity, and liquid fuel) with three different cases: (a) baseline, (b) integrated with pre-CCS, (c) integrated with post CCS.

Figure 4-2a illustrates the baseline configuration, where the flue gas is released directly into the atmosphere. The primary distinction from Scenario 1 is marked by the green dotted box in Figure 4-2a, which highlights the additional process of converting biogas into liquid biomethane. Figure 4-2b shows the configuration integrated with a pre-CCS plant, including the WGS reactor and pre-CCS system, as indicated by the red dotted box. In this configuration, the pre-CCS system only removes carbon from the syngas before it enters the CCHP.

Finally, Figure 4-2c depicts the configuration with the post-combustion carbon capture (post-CCS) plant added to the baseline system, as shown by the red dotted box. The post-CCS system captures carbon from the flue gas produced by syngas combustion before it is released into the atmosphere.

4.2.2 System modelling

The investigated processes were modelled in Aspen Plus V12.1. In order to adapt to standard operational parameters, it was essential to refer to relevant technical literature. Table 4-1 displays the important process design parameters utilised in the modelling.

Table 4-1. Process design parameters used in the Aspen Plus modelling.

Process	Conditions
Pretreatment	Crusher [288]: Model = gyratory crusher, resulting in the particle size reduction to 1 mm. Electricity required for grinding [288]: $\dot{W}_{el,consumption} = 50 \frac{\text{kWh}}{\text{ton of feedstock}}$
PG	Model [289]: Air plasma gasifier combined with the steam injection. Plasma torch temperature [80]: 4000°C. ER and SAMR ratio of plasma gasifier = 0.14 and 0.396, respectively. (ER and SAMR ratio were chosen iteratively until the temperature of HTZ reaching 2500°C). SFR = 0.8. HTZ temperature [80]= 2500°C. LTZ temperature [80]= 1250°C.
AD	Digestion temperature = 38°C

	The dilution ratio [290]: Food waste:water = 2.64:1, kg/kg. Heating demand during the digestion process [259]: $\dot{Q}_{digestion} = 10\% \cdot \dot{E}_{biogas}$
HPWS	Compressor 1 [291]: $P_{output} = 5$ bar Compressor 2 [291]: $P_{output} = 10$ bar Absorber [291]: $P = 10$ bar; $T = 20$ °C Stripper [291]: $P = 1$ bar; $T = 20$ °C Flash [291]: $P = 3$ bar; $T = 20$ °C
GT power cycle	Air compressor [292], [293]: pressure ratio = 19.5; $\eta_{isentropic} = 85\%$; $\eta_{mechanical} = 98.65\%$; $\dot{m}_{air comp}$ is determined until temperature of flue gas prior to gas turbine reach = 1350 °C. Combustion chamber [292]: fuel pressure = 26.83 bar. Gas turbine [292]: $P_{output} = 1.065$ bar; $\eta_{isentropic} = 89.769\%$; $\eta_{mechanical} = 98.65\%$.
ST power cycle	Steam turbine [294]: HP: $P_{input} = 180.2$ bar; $P_{output} = 28.8$ bar; $\eta_{isentropic} = 88.03\%$. IP: $P_{input} = 28.8$ bar; $P_{output} = 3.907$ bar; $\eta_{isentropic} = 92.37\%$. LP: $P_{input} = 3.907$ bar; $P_{output} = 0.0483$ bar; $\eta_{isentropic} = 93.67\%$.
ARC	Working fluid [295]: LiBr/H₂O; $\dot{m}_{LiBr-H_2O} / \dot{m}_{fluegas} = 0.1$; LiBr concentration in a basic solution = 57.4%. Pump [295], [296]: $P_{input} = 0.00672$ bar; $P_{output} = 0.07461$ bar; $\eta_{isentropic} = 75\%$; $\eta_{driver} = 95\%$. Valve [295], [296]: $P_{input} = 0.07461$ bar; $P_{output} = 0.00672$ bar. Designated temperature $T_{air,cooling} = 21$ °C (cooling for residence house) $T_{air,heating} = 52$ °C (heating for fish drying)
ML	Methane compressor [297]: $P_{output} = 200$ bar
CCS unit	Pre-CCS Absorber: $H = 25$ m; $D = 5$ m Stripper: $H = 15$ m; $D = 3.5$ m; $P = 2.4$ bar Post-CCS Absorber: $H = 25$ m; $D = 8.5$ m Stripper: $H = 15$ m; $D = 3.5$ m; $P = 2.6$ bar Carbon sequestration Compressor [298]: multistage; $P_{outlet,1} = 7$ bar; $P_{outlet,2} = 24$ bar; $P_{outlet,3} = 83$ bar Pump [298]: $P_{outlet} = 153$ bar

4.2.3 Process modelling and design in Aspen Plus

4.2.3.1 Plasma Gasifier (PG)

To determine the thermal and physical properties of conventional components and streams involved in the model, the Peng-Robinson equation was applied. For non-conventional components that are not available in the Aspen Plus's database, for example, plastic and food waste, HCOALGEN and DCOALIGT were used to estimate their thermal and physical properties based on the ultimate and proximate analysis of the feedstock. The gasifier was modelled following a thermo-chemical equilibrium approach at atmospheric pressure, with the plastic waste as the feedstock.

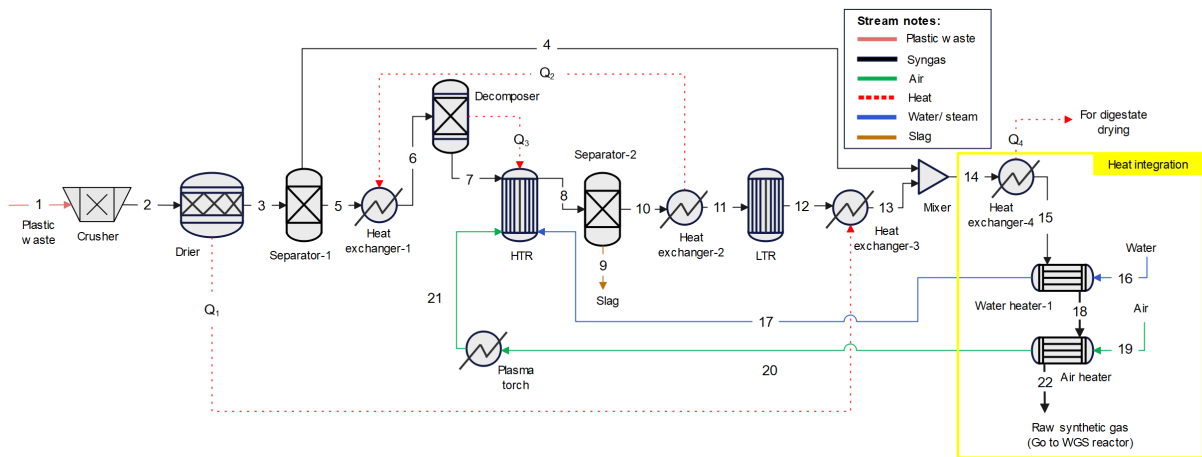


Figure 4-3. Process flow diagram of the PG with the heat integration for the baseline case.

Figure 4-3, Figure 4-4, and Figure 4-5 show the Aspen model of the plasma gasifier together with the heat utilization from the high-temperature syngas for the investigated cases (baseline case, case with pre-CCS, and case with post-CCS, respectively). As shown in all figures, plastic waste is dried in an RStoic reactor by utilising extracting heat from the syngas exiting the gasifier. This is represented by the heat stream Q_1 from heat exchanger-3. The evaporated moisture is separated from the plastic waste using a separator block (separator-1) and mixed with the syngas. The dried plastic waste is further decomposed in an RYield block “Decomposer” using the heat provided by the syngas stream from the high-temperature reactor (HTR) (stream-10). The decomposition process converts the

feedstock into its constituting elements namely carbon, hydrogen, oxygen, sulphur, nitrogen and ash by specifying the yield distribution according to the feedstock's ultimate analysis.

Figure 4-4. Process flow diagram of the heat-integrated PG for the case with pre-CCS.

The heat contained in the raw syngas (stream 14) is extracted through multiple heat exchangers in the three proposed configurations. Heat Exchanger-4 is common in the three configurations where the syngas supplies heat for drying the digestate. Additionally, “Water Heater-1” and “air heater” preheat water and air, respectively, for subsequent use in the PG. Even after passing through the air heater (stream

22), the syngas temperature remains high ($\sim 450^{\circ}\text{C}$), enabling the utilization of this heat in specific configurations.

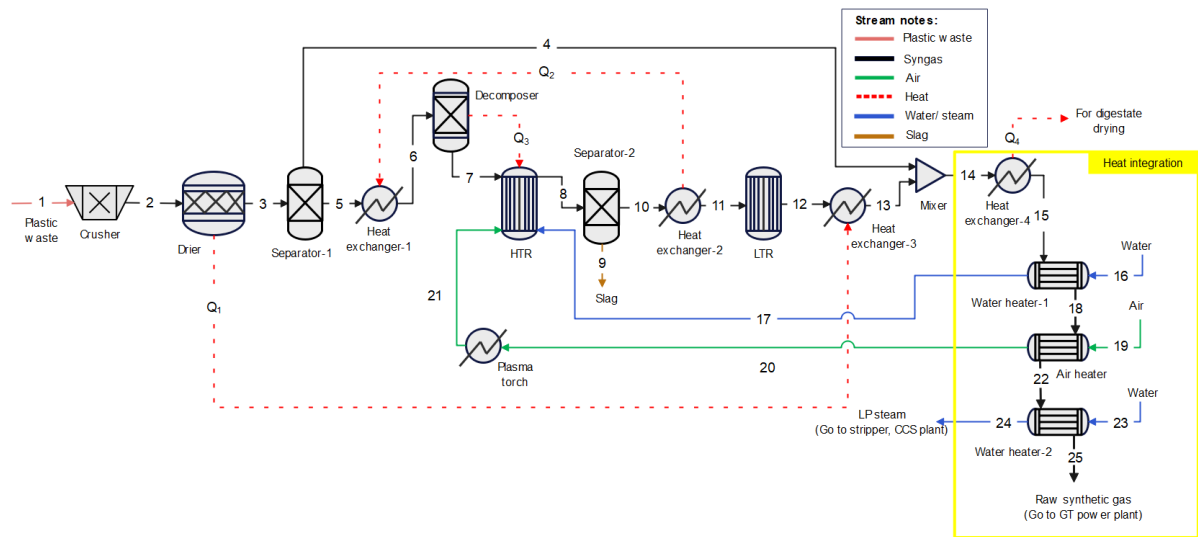


Figure 4-5. Process flow diagram of the heat-integrated PG for the case with post-CCS.

For the baseline case, Figure 4-3 shows two additional heat exchangers which are used to utilise the heat available. “Water Heater-2” is used to produce low-pressure (LP) steam at 284°C and 3.9 bar, which is sent to the steam power plant to boost the electricity generation in the LP steam turbine. The remaining heat is used to maintain the anaerobic digester (AD) temperature at 38°C , which is transferred in the heat exchanger “Water Heater-3”. On the other hand, and as illustrated in Figure 4-4, the syngas exiting the “Air Heater” for the pre-CCS configuration, is directed to the WGS reactor. In the case of the post-CCS configuration (Figure 4-5), the syngas was sent through an additional heat exchanger, “Water Heater-2”, which is used to generate LP steam at 143.5°C and 2.6 bar. This steam serves as the heat source of the stripper in the carbon capture plant, providing around 18% of its heat requirement.

4.2.3.2 Water Gas Shift (WGS) Reactor

A WGS reactor is used to increase the hydrogen content of the produced syngas, by converting CO and water into H_2 and CO_2 . Figure 4-6 describes the process configuration of this section. A fraction of 10% of the raw syngas is combusted to provide heat for steam generation, with the resulting steam subsequently used in

the water-gas shift (WGS) reactor. The amount of water (stream 11) required for steam generation is determined based on the maximum heat that can be absorbed from the flue gas. By using a steam to CO molar ratio (S/CO ratio) of 1.66, the maximum conversion of CO that can be achieved is around ~93%.

Figure 4-6. Process flow diagram of the WGS reactor in case with pre-CCS.

The WGS reaction is carried out at two different temperatures to maximize CO conversion [300]. At high temperatures, the reaction has a fast rate, while at low temperatures, the equilibrium favours the reaction due to its exothermic nature, although the reaction rate is slower [300]. The high temperature water gas shift reactor (HTWGSR) and low temperature water gas shift reactor (LTWGSR) are operated at 450°C using a Fe-Cr catalyst and 200°C using a CuO/ ZnO/ Al₂O₃, respectively [124] [125]. The stream leaving the HTWGSR (stream 4) is cooled down to 200°C. The released heat from this process is utilized in the “water heater-2” to generate LP steam that is used to cover approximately 41% of the heat required by the reboiler of the carbon capture plant. In addition, the stream leaving the LTWGSR (stream 6) is also used to generate LP steam which can cover 13.4% of the overall reboiler duty. For modelling purposes, RPlug block is used to compute the chemical kinetics of the WGSR as mentioned in 2.2.1.

AD processes food waste to generate biogas in a mesophilic environment (38°C). Based on the experiments conducted by Kumar et al. [301] and Blasius et al. [302], it was demonstrated that mesophilic condition yielded higher amount of biogas compared to thermophilic production, making it the preferred option. In addition, the mesophilic condition consumes less energy compared to thermophilic condition

[303].

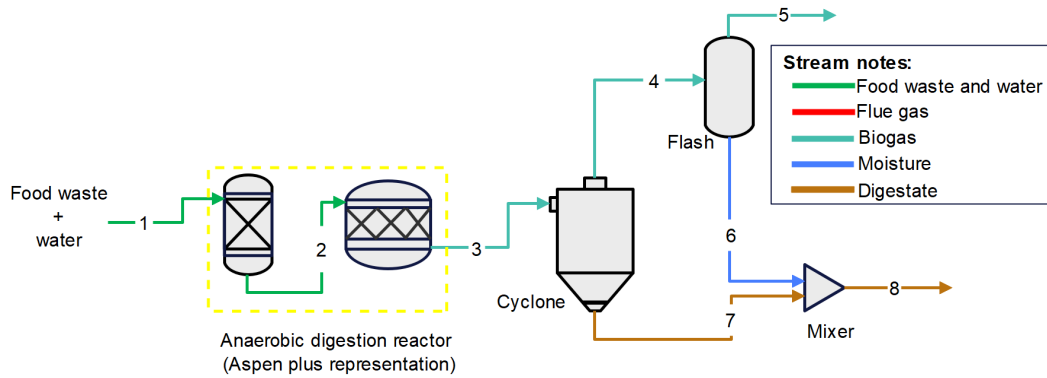


Figure 4-7. Process flow diagram of the AD.

The AD heating demand was assumed to be 10% of the energy content of the biogas [259]. To fulfil this requirement, the AD was heated using the heat recovered from other subsystems. For the baseline case, the required thermal energy was supplied by recovering heat from the raw syngas stream and the flue gas downstream of the waste heat recovery heat exchanger (HX) located after the ARC. For the case with pre-CCS and post-CCS, it was fulfilled by utilising some of the heat available in the vapour stream leaving the stripper prior to the reflux condenser. The AD model developed in this work builds upon the framework established by Naqi et al [304]. As shown in Figure 4-7, two reactors were used to represent the process of anaerobic digestion. The initial reactor, which is represented by a RYield block, models the breakdown of the food waste into its elemental composition. The non-biodegradable portion of the food waste is characterized using theoretical correlations, as described in [305], with detailed calculations provided in [304]. Subsequently, a RStoic reactor is used to represent the production of the components that are typically present in the biogas produced, which can be represented by the following reactions [304]:



All reactions are assumed to occur in the complete conversion of the limiting reactant for each reaction (the details of the limiting reactant can be seen in [304]). The product from the second reactor consists of biogas and digestate. Biogas is

separated from digestate by using a cyclone and a flash block.

4.2.3.4 Combined cycle gas turbine (CCGT)

Figure 4-8 and Figure 4-9 depict the process flow diagram for the CCGT in the baseline, the pre-CCS and post-CCS configurations, respectively. The quantity of air required for the gas turbine (GT) power cycle (stream 1) is calculated in order to achieve an exhaust temperature of 1350°C from the GT combustion chamber (stream 5). Notably, the fuel inputs for Scenarios 1 and 2, of each of the proposed configurations, differ based on the type of fuel used. In Scenario 1, the fuel input is a mixture of syngas and biogas. Specifically, in the baseline and post-CCS cases, the GT is powered by untreated raw syngas and biogas. On the other hand, in the pre-CCS case, the syngas and biogas are preliminary treated in the CCS plant to remove CO₂ before entering the GT. In contrast, Scenario 2 uses syngas only as the sole fuel input for the three cases involved in this study. In the baseline and post-CCS cases, the GT is fuelled by raw syngas, while in the pre-CCS case, the syngas is previously treated in the CCS plant. In the GT, compressed fuel gas (stream 4) is mixed with compressed air in the combustion chamber, and the generated exhaust gas powers the gas turbine. The exhaust gas from the GT (stream 6) is then directed to the steam production system for further energy recovery.

In the steam turbine (ST) power plant, heat from the flue gas is utilized in the heat recovery steam generator (HRSG) to produce low-pressure (LP), intermediate-pressure (IP), and high-pressure (HP) steam. After passing through the HRSG, the flue gas (stream 7) is directed to the absorption refrigeration cooler (ARC) for cooling production. The configuration of the ST power cycle differs between the baseline case, the pre-CCS case, and the post-CCS case. The key difference lies in the utilization of additional LP steam from the waste heat of syngas (stream 13 in Figure 4-8) and the presence of a desuperheater, which converts superheated steam to saturated LP steam (stream 26 in Figure 4-9).

it down to the saturation temperature. The sensible heat generated by the desuperheater is further used to preheat the boiler feed water before entering the HRSG.

4.2.3.5 Absorption Refrigeration Cooler (ARC)

Figure 4-10 shows the process flow diagram of the ARC. The cooling production of the ARC is determined by the quantity of heat absorbed by the working fluid (LiBr-H₂O) from the flue gas after passing through HRSG in the desorber as shown in Figure 4-10. The amount of the necessary working fluid was calculated as 1:10 between the working fluid and flue gas flow rates [296].

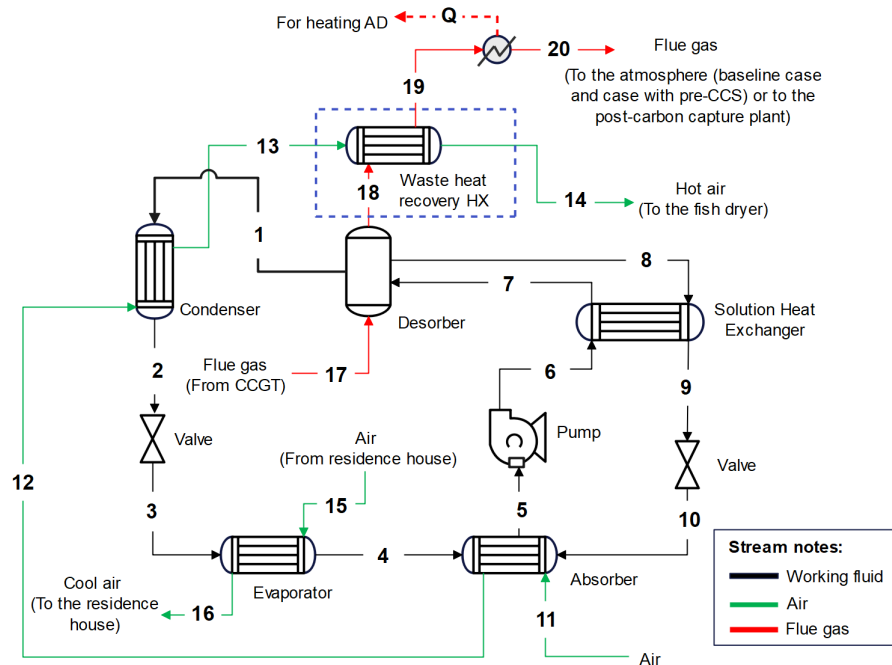


Figure 4-10. Process flow diagram of the ARC.

The working fluid in the ARC (Li-Br and water) absorbs the heat released from the flue gas after the CCGT (stream 17) and converts it into cooling energy that can be distributed to residential buildings (stream 16). During the process, the heat recovered from both absorber and condenser can increase the temperature of air (stream 11) from 25 to 35°C (stream 13). The remaining heat in the flue gas (stream 18) can be recovered to increase the temperature of stream 13 to 52°C. This hot air is then used to reduce the moisture content of fish in a fish-drying industry. In the baseline case, the remaining heat of the flue gas is used for maintaining the operating temperature of the AD. For the pre-CCS case, the flue gas is ejected into the atmosphere, while for the post-CCS plant, the flue gas is

sent to the CO₂ capture plant.

4.2.3.6 Pre and post carbon capture and storage (CCS)

The MEA Steady State Model developed by the DOE Carbon Capture Simulation Initiative (CCSI) is utilised to model the carbon capture section [252]. The operating conditions and design parameters of the carbon capture plant are presented in Table 4-2. A CO₂ capture fraction of 95% in both the pre- and post-combustion capture cases was applied. MEA at 35% w/w was selected as the solvent due to its low cost and its well-understood performance. The absorber packing diameter was determined to satisfy the limitation of 80% flooding condition. More process parameters were based on Sherman FEED Studies [306]. The CO₂ concentration in the gas input affects the rich solvent loading (RSL); in the pre-CCS case, the higher CO₂ concentration creates a stronger driving force for the mass transfer of CO₂ in the absorber, allowing the solvent to capture more CO₂ and therefore achieve a higher RSL.

Table 4-2. Comparisons of several parameters applied in pre- and post-CCS models.

Parameter		Pre-CCS		Post-CCS		Unit
		Scenario 1	Scenario 2	Scenario 1	Scenario 2	
Absorber height	packing	25	25	25	25	m
Absorber diameter	packing	5	4	8.5	7.5	m
Stripper height	packing	15	15	15	15	m
Stripper diameter	packing	3.5	3.5	4	4	m
Stripper pressure		2.6	2.6	2.6	2.6	bar
LSL		0.12	0.12	0.11	0.11	mol CO ₂ /mol MEA
RSL		0.50	0.50	0.48	0.48	mol CO ₂ /mol MEA

To achieve high capture rates of CO₂ (i.e. 95%), lower lean solvent loading (LSL) promotes the mass transfer process of CO₂ by providing a higher driving force between the gaseous CO₂ and the solvent (higher partial pressures difference).

However, this approach may result in a higher energy penalty for solvent regeneration. Due to higher CO₂ concentration in the pre-CCS which promotes better mass transfer, the LSL value of pre-CCS (0.12 molCO₂/molMEA) is set slightly higher than that of post-CCS (0.11 molCO₂/molMEA) to reduce the reboiler duty. The optimum absorber and stripper packing heights were achieved at 25 m and 15 m, respectively while a stripper pressure of 2.6 bar was employed. This pressure is necessary to avoid operation below the inflexion point in the stripper. This point represents a threshold where the energy needed to achieve lower lean loading increases exponentially due to increased boiling of water leading to excess water leaving the stripper [307]. Therefore, a pressure of 2.6 bar enables efficient operation with an LSL value of 0.11 molCO₂/molMEA (post-CCS) and 0.12 molCO₂/molMEA (pre-CCS).

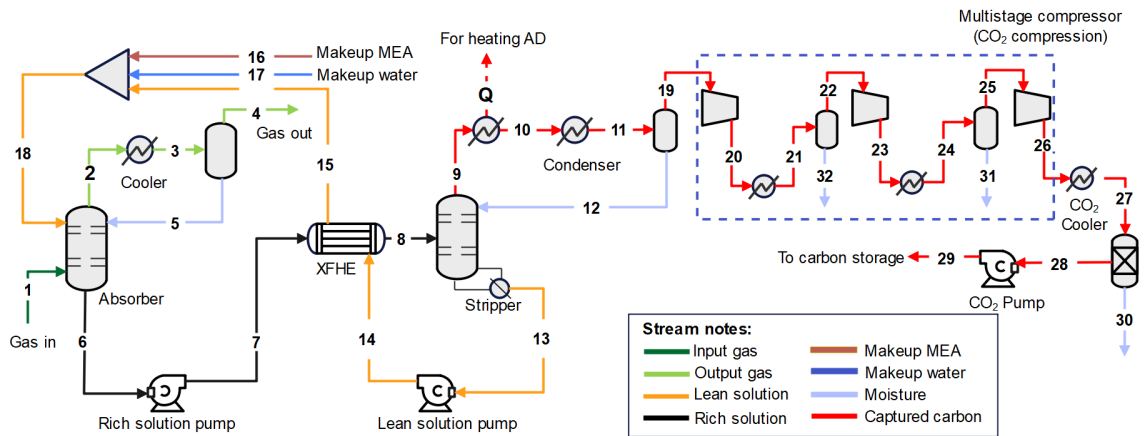


Figure 4-11. Process flow diagram of the CCS plant.

Figure 4-11 illustrates the process flow diagram for the CCS plant. The gas input (stream 1) for the CCS plant varies between Scenario 1 and Scenario 2. In Scenario 1, for the pre-CCS case, the input is a mixture of syngas and biogas, whereas in the post-CCS case, the input is flue gas from the combustion of the syngas-biogas mixture. In Scenario 2, for the pre-CCS case, the gas input is syngas, while in the post-CCS case, the input is flue gas from the combustion of syngas. As for the gas output (stream 4), the same approach applies to both Scenario 1 and Scenario 2: in the pre-CCS case, the output is directed to the gas turbine (GT) power plant, while in the post-CCS case, the output is released to the atmosphere.

As illustrated in Figure 4-11, after the CO₂ is extracted (stream 9), some of the

carried heat is utilized to maintain the temperature of anaerobic digester (AD). Following that, the vapour in stream 10 is condensed to a temperature of 40°C and removed from the CO₂ gas stream. The CO₂ is then compressed through several phases to reach supercritical values and transform into a liquid state. Subsequently, the liquid CO₂ is assumed to be transported and stored deep underground. To achieve the required supercritical conditions for transportation, a series of compression stages are utilized, together with intercooling and flash separators, to compress CO₂. After each compression step, the compressed gas is cooled using cooling water, and the resulting condensate is removed in the flash blocks. At stream 27, the CO₂ stream is maintained at a temperature of 30°C and a pressure of 83 bar, exceeding the critical pressure threshold of 73.8 bar [298]. It indicates that the stream is in the liquid phase. Consequently, a pump is employed to elevate the pressure of the liquid CO₂ to the required level of 153 bar to transport it through pipelines [298].

4.2.3.7 High pressure water scrubber (HPWS)

The process of water scrubbing in HPWS involves using high pressure and low temperature to effectively increase the absorption of CO₂ in water [308]. A typical water gas scrubber is to recover methane from the biogas [309]. As shown in Figure 4-12, raw biogas enters the HPWS (stream 1) to enhance its CH₄ content. The biogas, which has been compressed by compressors 1 and 2, enters the lower section of the absorber (stream 6). Simultaneously, pumped water is introduced into the upper section of the column (stream 9). The subsequent stage involves directing of water enriched with CO₂ from the absorber to the flash column (stream 10). To mitigate the potential release of methane, the pressure is decreased in the flash column (separator). The gas discharged from the flash column (stream 15) is combined with the raw biogas (stream 3) in a mixer before being compressed by the second compressor. The concentrated solution, another output of the flash column (stream 11), enters the stripper column and is exposed to an airflow moving in the opposite direction (stream 16). The water stream exiting this column (stream 13) is recycled to the upper section of the absorber (stream 14).

$$ER(< 1.0)_{gasification} = \frac{(\dot{m}_{air})_{actual}}{(\dot{m}_{air})_{stoichiometric}} \quad \text{Eq. 4-7}$$

where $\dot{m}_{air}/\dot{m}_{PW}$ is the actual ratio of mass flow rate air to plastic waste and $(\dot{m}_{air}/\dot{m}_{PW})_{stoic}$ is the stoichiometric ratio of mass flow rate air to plastic waste. The steam-air mass ratio (SAMR) is a dimensionless parameter that is frequently employed to quantify the rate at which steam is supplied in the air and steam gasification process [289].

$$SAMR = \frac{\dot{m}_{steam}}{\dot{m}_{air}} \quad \text{Eq. 4-8}$$

In the plasma gasification, the energy conversion efficiency ($\eta_{en,pg}$) is calculated as the ratio of the energy content of the cold syngas to the energy content of solid waste and the electricity consumed for plasma torch during the process and it can be written as follows [80]:

$$\eta_{en,pg} = \frac{\dot{E}_{syngas}}{\dot{E}_{pw} + \dot{Q}_{preheat,air} + \dot{Q}_{preheat,water} + \frac{P_{torch}}{\eta_{torch}}} \quad \text{Eq. 4-9}$$

where $\dot{E}_{pw}$ and $\dot{E}_{syngas}$ are the energy content of plastic waste and syngas which can be calculated using Eq. 3-7 and Eq. 3-13, respectively, P_{torch} is the torch power (MW), η_{torch} represents the thermal efficiency of torch, which is set to 90%, $\dot{Q}_{preheat,air}$ and $\dot{Q}_{preheat,water}$ represent the heat utilized to preheat the air (prior to entering the torch) and the water, respectively.

For the exergy performance, the exergy conversion efficiency ($\eta_{ex,pg}$) can be obtained from the following expression:

$$\eta_{ex,pg} = \frac{\dot{Ex}_{syngas}}{\dot{Ex}_{pw} + \dot{Ex}_{Q_{preheat,air}} + \dot{Ex}_{Q_{preheat,water}} + \frac{P_{torch}}{\eta_{torch}}} \quad \text{Eq. 4-10}$$

where $\dot{Ex}_{syngas}$ is the exergy content of syngas (MW) and $\dot{Ex}_{pw}$ is the exergy content of plastic waste (MW), $\dot{Ex}_{Q_{preheat,air}}$ and $\dot{Ex}_{Q_{preheat,water}}$ denote the exergy

associated with heat transfer for preheating air and water, respectively.

4.2.4.2 Anaerobic digestion

In the first scenario, the energy conversion efficiency ($\eta_{en,ad}$) and exergy conversion efficiency ($\eta_{ex,ad}$) of anaerobic digestion can be mathematically represented as follows:

$$\eta_{en,ad} = \frac{\dot{E}_{biogas}}{\dot{E}_{foodwaste} + \dot{Q}_{AD}} \quad \text{Eq. 4-11}$$

$$\eta_{ex,ad} = \frac{\dot{Ex}_{biogas}}{\dot{Ex}_{foodwaste} + \dot{Q}_{AD} \left(1 - \frac{T_0}{T_{heating}}\right)} \quad \text{Eq. 4-12}$$

where $\dot{E}_{biogas}$ is energy content of biogas which can be defined using Eq. 3-13 (MW); $\dot{Ex}_{biogas}$ is the exergy content of biogas which can be determined using Eq. 3-16 (MW); $\dot{E}_{foodwaste}$ is energy content of food waste which can be calculated using Eq. 3-7 (MW), $\dot{Ex}_{foodwaste}$ is exergy content of food waste which can be calculated using Eq. 3-9 (MW), and $\dot{Q}_{AD}$ is the heating requirement to maintain the digester temperature (38°C).

In the context of Scenario 2, it is important to note that the installation of a water scrubber after the AD requires the involvement of two compressors for biogas purification. As a result, the above energy conversion efficiency of the AD ($\eta_{en,ad}$) and exergy conversion efficiency ($\eta_{ex,ad}$) equation are modified as follows:

$$\eta_{en,ad} = \frac{\dot{E}_{biogas}}{\dot{W}_{comp,ws} + \dot{W}_{pump,ws} + \dot{E}_{fw} + \dot{Q}_{AD}} \quad \text{Eq. 4-13}$$

$$\eta_{ex,ad} = \frac{\dot{Ex}_{biogas}}{\dot{W}_{comp,ws} + \dot{W}_{pump,ws} + \dot{Ex}_{fw} + \dot{Q}_{AD} \left(1 - \frac{T_0}{T_{heating}}\right)} \quad \text{Eq. 4-14}$$

where $\dot{W}_{comp,ws}$ is compressor work (MW) in the water scrubber and $\dot{W}_{pump,ws}$ is pump work (MW) in the water scrubber.

4.2.4.3 Gas-turbine power generation

The thermal energy efficiency of the gas-turbine power plant ($\eta_{en,gt}$) and thermal exergy efficiency ($\eta_{ex,gt}$) can be expressed as follows:

For Scenario 1,

$$\eta_{en,gt} = \frac{\dot{W}_{gt} - \dot{W}_{air\ comp,gt} - \dot{W}_{fuel\ comp,gt}}{\dot{E}_{syngas} + \dot{E}_{biogas}} \quad \text{Eq. 4-15}$$

$$\eta_{ex,gt} = \frac{\dot{W}_{gt} - \dot{W}_{air\ comp,gt} - \dot{W}_{fuel\ comp,gt}}{\dot{E}x_{syngas} + \dot{E}x_{biogas}} \quad \text{Eq. 4-16}$$

For Scenario 2,

$$\eta_{en,gt} = \frac{\dot{W}_{gt} - \dot{W}_{air\ comp,gt} - \dot{W}_{fuel\ comp,gt}}{\dot{E}_{syngas}} \quad \text{Eq. 4-17}$$

$$\eta_{ex,gt} = \frac{\dot{W}_{gt} - \dot{W}_{air\ comp,gt} - \dot{W}_{fuel\ comp,gt}}{\dot{E}x_{syngas}} \quad \text{Eq. 4-18}$$

where $\dot{W}_{gt}$ is the gas turbine work (MW), $\dot{W}_{air\ comp,gt}$ is the air compressor work (MW), $\dot{W}_{fuel\ comp,gt}$ is the fuel compressor work (MW).

4.2.4.4 Steam-turbine power generation

The thermal energy efficiency of the steam-turbine power plant ($\eta_{en,st}$) and thermal exergy efficiency ($\eta_{ex,st}$) can be formulated as follows:

$$\eta_{en,st} = \frac{\dot{W}_{hpt} + \dot{W}_{ipt} + \dot{W}_{lpt} - (\dot{W}_{hpp} + \dot{W}_{ipp} + \dot{W}_{lpp})}{\dot{Q}_{fluegas}} \quad \text{Eq. 4-19}$$

$$\eta_{ex,st} = \frac{\dot{W}_{hpt} + \dot{W}_{ipt} + \dot{W}_{lpt} - (\dot{W}_{hpp} + \dot{W}_{ipp} + \dot{W}_{lpp})}{\dot{Q}_{fluegas} \left(1 - \frac{T_0}{T_{fluegas}}\right)} \quad \text{Eq. 4-20}$$

where $\dot{W}_{hpt}$, $\dot{W}_{ipt}$, $\dot{W}_{lpt}$ are the high-pressure, intermediate-pressure, low-pressure turbine work (MW), respectively, and $\dot{W}_{hpp}$, $\dot{W}_{ipp}$, $\dot{W}_{lpp}$ are the high-pressure, intermediate-pressure, low-pressure pump work (MW), respectively.

4.2.4.5 Absorption refrigeration cooling

The energetic coefficient of performance of the ARC ($COP_{en,arc}$) and exergetic coefficient of performance ($COP_{ex,arc}$) can be determined by the following mathematical expression:

$$COP_{en,arc} = \frac{\dot{Q}_{eva}}{\dot{W}_{pump,arc} + \dot{Q}_{gen}} \quad \text{Eq. 4-21}$$

$$COP_{ex,arc} = \frac{\dot{Q}_{eva} \left(\frac{T_0}{T_{eva}} - 1 \right)}{\dot{W}_{pump,arc} + \dot{Q}_{gen} \left(1 - \frac{T_0}{T_{gen}} \right)} \quad \text{Eq. 4-22}$$

where $\dot{W}_{pump,arc}$ is the pump work (MW), $\dot{Q}_{gen}$ and T_{gen} are the heat absorbed by the generator (heat released from flue gas) (MW) and its associated temperature (K), respectively, and $\dot{Q}_{eva}$ and T_{eva} are the heat absorbed by the evaporator (heat released from residence house) (MW) and its associated temperature (K), respectively.

4.2.4.6 Methane liquefaction

The calculation of the energetic coefficient of performance of the ML ($COP_{en,ml}$) and exergetic coefficient of performance ($COP_{ex,ml}$) can be derived using the subsequent equation:

$$COP_{en,ml} = \frac{\dot{Q}_{refg}}{\dot{W}_{comp,ml}} \quad \text{Eq. 4-23}$$

$$COP_{ex,ml} = \frac{\dot{Q}_{refg} \left(\frac{T_0}{T_{refg}} - 1 \right)}{\dot{W}_{comp,ml}} \quad \text{Eq. 4-24}$$

where $\dot{W}_{comp,ml}$ is the compressor work (MW), $\dot{Q}_{refg}$ and T_{refg} are the heat required to liquify CH₄ and its associated temperature (K), respectively.

4.2.4.7 Overall system

In the first scenario, the system output comprises of electricity, cooling, and heating. Therefore, in order to assess the performance of the overall system in Scenario 1, the energy and exergy efficiency of the system as a whole is defined as follows:

$$\eta_{en,overall,sc1,x} = \frac{\dot{W}_{net,x} + \dot{Q}_{cooling} + \dot{Q}_{heating}}{\dot{E}_{pw} + \dot{E}_{fw}} \quad \text{Eq. 4-25}$$

$$\eta_{ex,overall,sc1,x} = \frac{\dot{W}_{net,x} + \dot{E}x_{cooling} + \dot{E}x_{heating}}{\dot{E}x_{pw} + \dot{E}x_{fw}} \quad \text{Eq. 4-26}$$

where:

$$x = \text{baseline, pre} - \text{ccs, post} - \text{ccs} \quad \text{Eq. 4-27}$$

$$\begin{aligned} \dot{W}_{net,baseline} &= \dot{W}_{gross,ccgt,baseline} \\ &\quad - (\dot{W}_{waste\ treatment} + \frac{P_{torch}}{\eta_{torch}})_{pg} \\ &\quad - \dot{W}_{pump,arc,baseline} \end{aligned} \quad \text{Eq. 4-28}$$

$$\begin{aligned} \dot{W}_{net,pre-ccs} &= \dot{W}_{gross,ccgt,pre-ccs} \\ &\quad - (\dot{W}_{waste\ treatment} + \frac{P_{torch}}{\eta_{torch}})_{pg} \\ &\quad - \dot{W}_{pump,arc,pre-ccs} \\ &\quad - (\dot{W}_{pump} + \dot{W}_{comp})_{ccs,pre-ccs} \end{aligned} \quad \text{Eq. 4-29}$$

$$\begin{aligned} \dot{W}_{net,post-ccs} &= \dot{W}_{gross,ccgt,post-ccs} \\ &\quad - (\dot{W}_{waste\ treatment} + \frac{P_{torch}}{\eta_{torch}})_{pg} \\ &\quad - \dot{W}_{pump,arc,post-ccs} \\ &\quad - (\dot{W}_{pump} + \dot{W}_{comp})_{ccs,post-ccs} \end{aligned} \quad \text{Eq. 4-30}$$

where $\dot{W}_{net,baseline}$, $\dot{W}_{net,pre-ccs}$, $\dot{W}_{net,post-ccs}$ are the net work (MW) in baseline, case with pre CCS, case with post CCS, respectively; $\dot{Q}_{cooling}$ and $\dot{E}x_{cooling}$ are the cooling energy and exergy generated from the ARC (MW), respectively; $\dot{Q}_{heating}$ and $\dot{E}x_{heating}$ are the heating energy and exergy released from the system (MW), respectively; $\dot{W}_{gross,ccgt}$ is the gross power generated from the CCGT (MW); $\dot{W}_{waste\ treatment}$ is the power required for crushing the plastic waste (MW); $\dot{W}_{torch}$ is the plasma torch power (MW); $\dot{W}_{pump,arc}$ and $\dot{W}_{pump,ccs}$ are the pump power in the ARC and in the CCS plant (MW), respectively; $\dot{W}_{comp,ccs}$ is the compressor power in the CCS plant (MW).

In Scenario 2, the output consists of power, cooling, heating, and liquid fuel. Hence,

in order to evaluate the performance of the entire system in Scenario 2, the energy and exergy efficiency of the total system can be determined in the following formula:

$$\eta_{en,overall,sc2,x} = \frac{\dot{W}_{net,x} + \dot{E}_{CH_4 liq} + \dot{Q}_{cooling} + \dot{Q}_{heating}}{\dot{E}_{pw} + \dot{E}_{fw}} \quad \text{Eq. 4-31}$$

$$\eta_{ex,overall,sc2,x} = \frac{\dot{W}_{net,x} + \dot{E}x_{CH_4 liq} + \dot{E}x_{cooling} + \dot{E}x_{heating}}{\dot{E}x_{pw} + \dot{E}x_{fw}} \quad \text{Eq. 4-32}$$

where:

$$x = baseline, pre - ccs, post - ccs \quad \text{Eq. 4-33}$$

$$\begin{aligned} \dot{W}_{net,baseline} = & \dot{W}_{gross,ccgt,baseline} \\ & - (\dot{W}_{waste treatment} + \frac{P_{torch}}{\eta_{torch}})_{pg} \\ & - \dot{W}_{pump,arc,baseline} - (\dot{W}_{pump} + \dot{W}_{comp})_{ws} \\ & - \dot{W}_{comp,ml} \end{aligned} \quad \text{Eq. 4-34}$$

$$\begin{aligned} \dot{W}_{net,pre-ccs} = & \dot{W}_{gross,ccgt,pre-ccs} \\ & - (\dot{W}_{waste treatment} + \frac{P_{torch}}{\eta_{torch}})_{pg} \\ & - \dot{W}_{pump,arc,pre-ccs} - (\dot{W}_{pump} + \dot{W}_{comp})_{ws} \\ & - \dot{W}_{comp,ml} - (\dot{W}_{pump} + \dot{W}_{comp})_{ccs,pre-ccs} \end{aligned} \quad \text{Eq. 4-35}$$

$$\begin{aligned} \dot{W}_{net,post-ccs} = & \dot{W}_{gross,ccgt,post-ccs} \\ & - (\dot{W}_{waste treatment} + \frac{P_{torch}}{\eta_{torch}})_{pg} \\ & - \dot{W}_{pump,arc,post-ccs} - (\dot{W}_{pump} + \dot{W}_{comp})_{ws} \\ & - \dot{W}_{comp,ml} - (\dot{W}_{pump} + \dot{W}_{comp})_{ccs,post-ccs} \end{aligned} \quad \text{Eq. 4-36}$$

where $\dot{E}_{CH_4 liq}$ and $\dot{E}x_{CH_4 liq}$ are the energy and exergy content of liquid biomethane (MW), respectively; $\dot{W}_{pump,ws}$ and $\dot{W}_{comp,ws}$ are the pump and compressor power in the WS (MW), respectively; $\dot{W}_{comp,ml}$ is the compressor power in the ML (MW).

4.3 Results and discussion

4.3.1 Model validation

Table 4-3 presents a comparison between the modelling results of PG, AD, GT power cycle, ST power cycle and ML and their related experimental data. Based on these findings, it can be concluded that there is a good agreement between the results derived from the current model and relevant experimental data.

Table 4-3. Comparison between the simulation outcomes of the present study with results of previous studies.

Process	Parameter	Present work	Experimental	Reference
PG	H ₂ (mol %)	62.03	57	[311]
	CO (mol %)	18.44	18.8	
	CO ₂ (mol %)	19.28	15.7	
	Plasma torch power (MW)	34.34	27.2	
AD	CH ₄ (mol %)	38.7	37.4	[290]
	CO ₂ (mol %)	60.5	62.6	
	Specific biogas yield (m ³ /tonnes VS)	605.27	642	
	Specific methane yield (m ³ /tonnes VS)	366.19	402	
	Digestate (kg/s)	1080.89	1113.42	
GT power cycle	Compressor Power (MW)	214.35	215.79	[294]
	Turbine Power (MW)	430.89	430.28	
	Turbine Outlet Temperature (°C)	603.28	604	
ST power cycle	Total steam turbine power (MW)	242.4	232.97	[294]
	Pump power (MW)	3.66	3.37	
	Turbine Outlet Temperature (°C)	34.94	32.25	
ML	Liquefaction temperature (°C)	-152.698	-161.5	[297]
	COP_{actual}	0.26	0.23	

4.3.2 Energy and exergy performance

4.3.2.1 Syngas and biogas generation

In order to generate syngas using a PG, an ER of 0.14, a SAMR of 0.396, and a SFR of 0.8 are employed to attain a temperature of 2500°C in the HTZ. The amount of plasma torch power used in the current PG is around 40.04 MW. Steam is generated at a high temperature of approximately 1157°C by harnessing the heat from the raw syngas. The process flow diagram of the PG for baseline case, the

case with pre-CCS and the case with post CCS can be seen in Figure 4-3, Figure 4-4, and Figure 4-5, respectively.

Table 4-4 presents the mass flow rates of the feedstock and the generated gas, along with their energy and exergy content, fuel gas composition, and the performance parameters of the PG and AD. The mass flow rates of plastic waste (PW) and food waste (FW) are derived from the amounts received at the Benowo landfill [278]. As shown in Table 4-5, the primary components of the raw syngas are predominantly H₂ at 43%, followed by nitrogen N₂ at 25%, and CO at 24%. In comparison with the results obtained by Zhang et al. [312] who used the same type of gasifier and gasification agents (air and steam), the results of the current study demonstrate higher energy and exergy efficiency. Specifically, the energy and exergy efficiency of the PG in the reference study are 50.8% and 44.9%, while the current study achieved 75.28% and 70.75%, respectively. This is a result of the higher carbon and hydrogen content of the feedstock utilized, i.e. waste plastic compared to the solid waste utilized in [312], 79.77% vs 50.5% and 15.47% vs 5.6%, respectively which results in higher LHV for the plastic waste.

Table 4-4. Material and energy balances for the PG and AD systems.

Characteristic	Feed stock		Generated gas			Unit
	Plastic waste	Food waste	Raw syngas	Raw biogas	Syngas after pre-CC	
Mass flow rate	5.00	13.97	18.45	3.31	10.04	kg/s
Energy content	184.11	45.67	191.85	44.26	163.09	MW
Exergy content	193.11	53.37	184.08	48.21	157.67	MW
Composition of fuel (molar fraction)						
H ₂	N/A	N/A	42.56	0.00	67.24	%
CO	N/A	N/A	24.01	0.00	1.73	%
CH ₄	N/A	N/A	0.00	47.50	0.00	%
N ₂	N/A	N/A	25.09	0.00	25.98	%
H ₂ O	N/A	N/A	6.93	5.20	3.92	%
CO ₂	N/A	N/A	1.40	43.40	1.14	%
O ₂	N/A	N/A	0.00	0.00	0.00	%
NH ₃	N/A	N/A	0.00	3.60	0.00	%
H ₂ S	N/A	N/A	0.00	0.40	0.00	%
Performance parameter	Plasma gasifier		Anaerobic digester			Unit
Conversion energy efficiency	75.28		88.34			%
Conversion exergy efficiency	70.75		86.28			%

For the pre-CCS cases, the syngas produced by the PG reacts with steam in the WGS reactor. This reaction results in the generation of CO₂ and hence its composition is further increased in the syngas, which is then captured in the CCS

plant. A 10% portion of the raw syngas has to be combusted to raise heat for steam generation. By optimizing the AFR to a value of 15.69, it is possible to attain the highest temperature of the combustion gas, which is around 633°C. The highest possible conversion of CO into CO₂ and H₂ that can be attained is ~93% by employing a steam to carbon monoxide molar ratio (S/C ratio) of 1.66. Due to the conversion in the WGS reactor, the mass flow rate of CO reduces dramatically from 7.53 to 0.47 kg/s, while there is an increase in the mass flow rate of hydrogen H₂ from 0.961 to 1.318 kg/s. After the syngas passes through the WGS reactor and pre-CCS plant, there is a reduction in the energy and exergy content of the syngas, as shown in the Table 4-4. The process flow diagram of WGS reactor is shown in Figure 4-6.

For biogas generation, in order to maintain the temperature of the AD at 38°C, approximately 10% of the energy content of the biogas, equivalent to 4.42 MW, is required. The heating requirement can be satisfied by extracting 0.83 MW of energy from the flue gas after passing through the ARC, and an additional 3.59 MW from the sensible heat of the raw syngas (for the baseline scenario) or from extracting heat from the top stream of the desorber prior to the condenser (for the case with pre- and post-CCS). The energy conversion efficiency of the AD process can be observed in Table 4-4, with values of around 88.34%. According to [313], the conversion of food waste into biogas has an energy efficiency of approximately 85%, which is in good agreement with the energy efficiency of 88.34% calculated herein.

4.3.2.2 Electricity generation

Table 4-5 contains detailed information about the performance of the CCGT under different scenarios and configurations. Key parameters analyzed include fuel energy input, compressor power consumption, turbine power output, and performance efficiency. It can be seen from this table that the energy input to the GT power cycle remains the same in the baseline case and the case with post-CCS in both scenarios. Nevertheless, the energy content of fuel in the case with pre-CCS is always lower than in the other cases. There are two primary factors contributing to this phenomenon. Firstly, the mass flow rate of CO decreases significantly and the rise of the mass flow rate of H₂ is not enough to offset this.

Secondly, 10% of the raw syngas is extracted to supply heat in the WGSR. The higher energy of fuel on scenario 1 compared to scenario 2 is attributed to the mixture of syngas and biogas used in the combustor of scenario 1, whereas only syngas is utilized in the combustor of scenario 2.

Table 4-5. Thermodynamic results for the CCGT island.

Parameter	SCENARIO 1			SCENARIO 2			UNIT
	BASELINE	W/PRE-CCS	W/POST-CCS	BASELINE	W/PRE-CCS	W/POST-CCS	
Gas turbine performance							
Energy of fuel (to the gas turbine cycle)	236.11	207.32	236.11	191.85	163.09	191.85	MW
Exergy of fuel (to the gas turbine cycle)	231.48	195.30	231.48	184.08	157.67	184.08	MW
Air compressor power in the gas turbine cycle	89.51	81.06	89.51	72.03	62.15	72.03	MW
Fuel compressor power in the gas turbine cycle	15.72	13.33	15.72	14.29	12.55	14.29	MW
Gross gas turbine power	186.03	166.53	186.03	150.61	129.97	150.61	MW
Net power output from GT cycle	80.79	72.14	80.79	64.29	55.27	64.29	MW
Thermal energy efficiency	34.22	34.80	34.22	33.51	33.89	33.51	%
Thermal exergy efficiency	34.90	36.94	34.90	34.92	35.05	34.92	%
Steam turbine performance							
Energy of flue gas (to the steam turbine cycle)	123.48	107.70	115.88	99.42	83.72	96.77	MW
Exergy of flue gas (to the steam turbine cycle)	63.02	54.68	61.20	50.68	42.49	50.06	MW
Pump power in the steam turbine cycle	0.84	0.78	0.89	0.70	0.63	0.74	MW
Steam turbine power	51.11	39.18	37.74	41.61	30.94	32.23	MW
Net power output from ST cycle	50.27	38.40	36.85	40.91	30.31	31.50	MW
Thermal energy efficiency	40.71	35.66	31.80	41.15	36.21	32.55	%
Thermal exergy efficiency	79.77	70.23	60.21	80.72	71.33	62.91	%
Overall combined cycle performance							
Net power output from CCGT	131.06	110.54	117.64	105.20	85.58	95.78	MW
CCGT energy efficiency	55.51	53.32	49.83	54.83	52.47	49.93	%
CCGT exergy efficiency	56.62	56.60	50.82	57.15	54.28	52.03	%

According to Table 4-5, the baseline case and the case with post-CCS in Scenario 1 and 2 generate higher amounts of CCGT fuel, resulting in a bigger amount of air being drawn into the compressor and consequently a higher air compressor power compared to the case with pre-CCS is required. Given that all the cases have the

same turbine inlet temperature, the only variable that impacts the turbine work is the combined mass flow rate of air and fuel. Therefore, the baseline case and the post-CCS case can generate more electricity in the GT compared to the pre-CCS case.

The flue gas from the GT power cycle is utilized to drive a typical steam turbine (ST) cycle. The amount of heat that can be recovered by the boiler feed water (BFW) depends on the temperature at which it enters the HRSG. In the baseline case, the BFW is condensate water that comes out from the LP pump (stream 8, as shown in Figure 4-8) and absorbs most of the available heat. The BFW temperature in the baseline case for Scenarios 1 and 2 is 35°C.

While the baseline case benefits from the lower temperature of the BFW, allowing for greater heat absorption, the situation changes in the pre- and post-CCS cases. The energy absorbed from the flue gas by the BFW (stream 8, as shown in Figure 4-9) is reduced because the BFW has been preheated using sensible heat from the desuperheater. Specifically, the temperature of BFW in the case with pre-CCS for Scenarios 1 and 2 is 90.14°C and 81.62°C, respectively. In the post-CCS case, the BFW temperature rises to 143°C and 119°C for Scenarios 1 and 2, respectively. Additionally, it is noteworthy that the lowest amount of energy delivered to the HRSG through the flue gas in the pre-CCS case is due to the lowest fuel energy.

Finally, the net power output of the combined cycle is maximum in the baseline case because it has a higher flow rate of fuel, and the absence of LP steam necessary for the CCS to meet the reboiler heat duty. This extraction reduces the net power generated by the steam turbine in both cases. Specifically, as shown in Table 4-5, the net power output in the baseline case for Scenarios 1 and 2 is 131.06 MW and 105.20 MW, respectively. In contrast, the post-CCS case, which maintains the same fuel flow rate as the baseline but includes LP steam extraction, shows the second-highest net power output. The net power output for post-CCS in Scenarios 1 and 2 is 117.46 MW and 95.78 MW, respectively, with heat supplies of 58.33 MW and 41.81 MW of thermal energy to the CCS plant. The lowest net power output is observed in the pre-CCS case, with outputs of 110.54 MW and 85.58 MW for Scenarios 1 and 2, respectively, and corresponding heat supplies of 43.63 MW and 35.58 MW for the CCS plant.

4.3.2.3 Cooling and heating generation

The process flow diagram of the ARC is provided in Figure 4-10. Table 4-6 compares various system operating conditions and presents the corresponding energy and exergy analyses for ARC. As shown, the post-CCS case in both scenarios exhibits the highest flue gas temperature due to the limitation in the HRSG (temperature crossover between the flue gas and the incoming BFW).

Additionally, Table 4-6 indicates that the cooling capacity is identical in both the baseline and post-CCS cases under both scenarios, but these are higher than in the pre-CCS case. Since the temperature difference is fixed during the operation of the evaporator, the only remaining factor determining the cooling effect produced by the ARC is the mass flow rate of the ARC working fluid. While the ratio of the mass flow rates of ARC working fluid and flue gas is 1:10 (as listed in Table 4-1), the flow rate of flue gas in each case is the determining factor of the cooling capacity. This explains the difference between the baseline and the post-CCS cases vs the pre-CCS case (4.04 vs 3.51 MW in Scenario 1 and 3.27 vs 2.72 MW in Scenario 2).

Table 4-6. Results for the absorption refrigeration cooler (ARC).

Parameter	SCENARIO 1			SCENARIO 2			UNIT
	Baseline	W/PRE-CCS	W/POST-CCS	Baseline	W/PRE-CCS	W/POST-CCS	
Operating condition							
Flue gas temperature	110.00	110.00	144.24	110.00	110.00	124.78	°C
Pump power	0.12	0.11	0.12	0.10	0.08	0.10	kW
Working fluid flow rate	20.79	18.04	20.79	16.81	13.96	16.81	kg/s
Energy analysis							
Energy of flue gas absorbed by ARC	5.50	4.77	5.50	4.45	3.69	4.45	MW
Cooling energy for residence house	4.04	3.51	4.04	3.27	2.72	3.27	MW
Energy COP	0.74	0.74	0.74	0.74	0.74	0.74	-
Exergy analysis							
Exergy of flue gas absorbed by ARC	1.22	1.06	1.57	0.99	0.82	1.11	MW
Cooling exergy for residence house	0.34	0.29	0.34	0.27	0.23	0.27	MW
Exergy COP	0.28	0.28	0.22	0.28	0.28	0.25	-

	Waste heat from ARC*						
Energy of the waste heat	17.29	27.87	26.84	13.99	25.62	17.12	MW
Exergy of the waste heat	1.62	2.83	3.28	1.49	2.70	3.13	MW
Heating air flow rate	631.10	1016.98	948.52	510.48	934.91	624.86	kg/s

*The heat recovered comes from the absorber, condenser, and the waste heat recovery HX

The amount of heat energy recovered from the ARC is presented in Table 4-6. As shown, the pre-CCS case recovers the highest amount of heat energy due to the highest heating air flow rate. The heating air flow rate is iteratively adjusted until the heating air temperature reaches 52°C. However, in terms of heat exergy recovery, the post-CCS case exhibits the highest value, as the temperature of the recovered heat from the flue gas is higher.

As presented in the Table 4-6, the energy COP of the ARC is consistent across all scenarios, with a value of around 0.74. This result aligns with previous research findings, where the COP was reported as 0.75 in [314] and 0.74 in [315]. The exergy COP in both Scenario 1 and Scenario 2 ranges between 0.22 and 0.28, which is comparable to the COP range of 0.2 to 0.25 identified in [316]. The exergy COP in the case with post-CCS under both scenarios is consistently lower compared to the baseline and pre-CCS cases due to higher heat loss during the heat transfer from the flue gas to the ARC working fluid in the desorber. The higher heat loss is due to the ratio established between the ARC working fluid and flue gas flow rates through the desorber.

4.3.2.4 Liquid biomethane production

Biogas at a flowrate of 3.31 kg/s with CO₂ and CH₄ molar fractions of 43.3% and 47.4%, respectively, is upgraded to 97.7% CH₄ using a HPWS process operating at 10 bar and 25°C. The integration of HPWS with the anaerobic digester results in a slight reduction in both energy and exergy efficiency, from 88.34% to 85.79% and from 86.28% to 84.04%, respectively. This decrease is primarily attributed to the energy consumption required for biogas compression and water pumping within the HPWS.

In the subsequent methane liquefaction (ML) process, 5.24 MW of power is required to compress the upgraded biomethane to 200 bars. The energy COP for

the ML process is 0.29, which is consistent with previous studies, such as the research cited in [297], where a COP of approximately 0.234 was reported.

4.3.2.5 CCS plant

Table 4-7 compares several critical parameters between pre-CCS and post-CCS across two scenarios. Parameters such as liquid-to-gas (L/G) ratio, specific reboiler duty (SRD), and reboiler heat duty are examined to evaluate the performance of the carbon capture processes. Modelling approach and other important parameters such as the dimension of absorber and stripper, lean solvent loading (LSL), rich solvent loading (RSL) as well as the process flow diagram of the CCS plant, are provided in 4.2.3.6.

Based on the Table 4-7, L/G ratio in the case with pre-CCS is higher than that in the case with post-CCS as more CO₂ per cross sectional area has to be captured. The reboiler heat duty in Scenario 1 is consistently higher than in Scenario 2 due to the greater amount of carbon captured in the first scenario. Specifically, in the pre-CCS case, the CCS plant in Scenario 1 captures 18.90 kg/s of carbon, compared to 14.59 kg/s in Scenario 2. Similarly, in the post-CCS case, the CCS plant in Scenario 1 captures 24.69 kg/s, while Scenario 2 captures 17.89 kg/s. Further, the pre-CCS cases exhibit lowers SRDs compared to the post-CCS, i.e. 3.31 and 3.41 GJ/tCO₂ (for Scenario 1 and 2) vs 3.56 and 3.6 GJ/tCO₂ (Scenario 1 and 2), respectively, due to the higher concentrations of CO₂ in the treated gas. Interestingly, despite the differences in captured carbon, the SRD shows minimal variation across the cases, ranging from 3.31 GJ/tCO₂ to 3.6 GJ/tCO₂. These results align with findings from [166], [167], which indicate that with a LSL of 0.11-0.12 mol CO₂/mol MEA and an absorption packing height of 24 m, the SRD is approximately 3.5 GJ/tCO₂.

Table 4-7. Main input and results for the pre- and post-CCS plant.

Parameter	Pre-CCS		Post-CCS		Unit
	Scenario 1	Scenario 2	Scenario 1	Scenario 2	
Gas input flow rate	23.83	20.61	205.12	165.89	kg/s
Absorber height	25	25	25	25	m
Absorber diameter	5	4	8.5	7.5	m
CO ₂ concentration in gas input	53.52	51.10	8.37	7.54	% (mass)
CO ₂ captured flow rate	12.16	10.05	16.33	11.94	kg/s

L/G ratio	5.45	5.28	0.88	0.80	kg/kg
Reboiler heat duty	40.29	34.30	58.18	43.05	MW
SRD	3.31	3.41	3.56	3.60	GJ/tCO ₂

4.3.3 Exergy destruction

Figure 4-14 illustrates the distribution of exergy destruction across various subsystems of the scenarios and configurations explained above. The graphs highlight the differences in inefficiencies among the subsystems, offering insights into where exergy losses occur and where improvements could be made to enhance overall system efficiency. The detailed exergy destruction of each component can be seen in the Table 4-8.

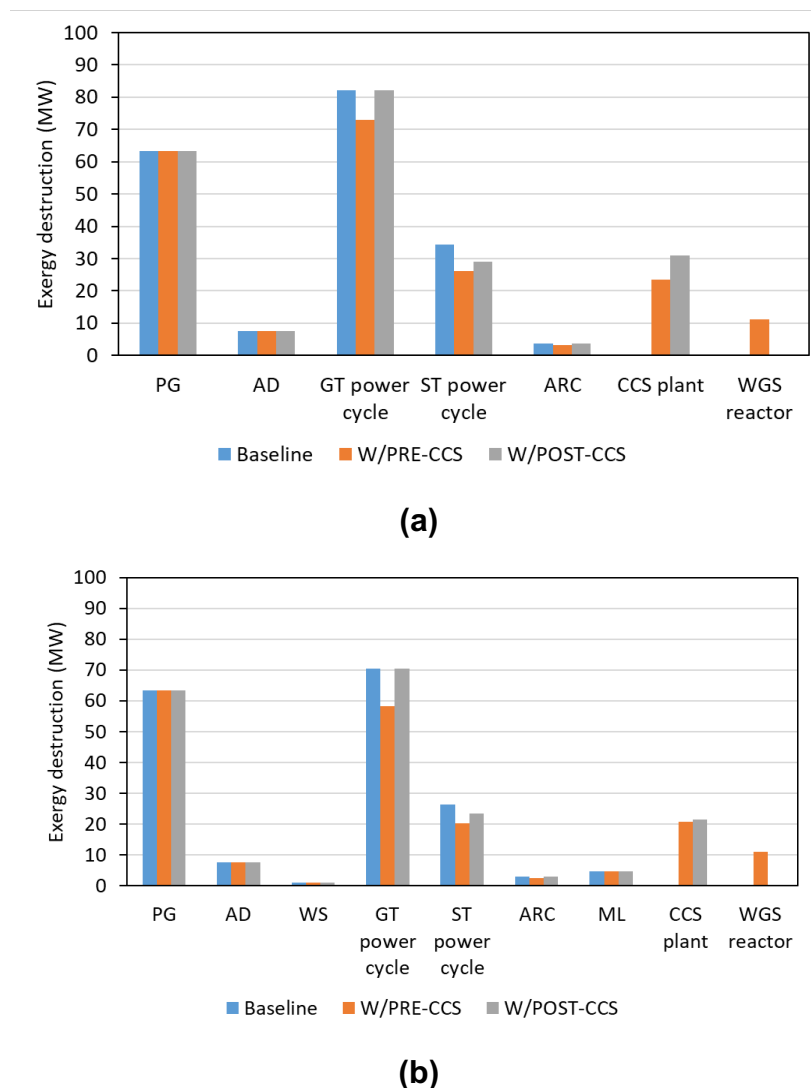


Figure 4-14. Exergy destruction of each energy subsystem in (a) Scenario 1 (b) Scenario 2.

As seen in Figure 4-14, the PG and the GT power cycle have the highest exergy

destruction across all plant scenarios. The PG has an exergy destruction rate of 63.37 MW, primarily due to the irreversibility associated with the high gasification temperature caused by the plasma torch and the chemical reactions taking place for converting the plastic waste into high temperature of syngas and slag [199], [317], [318]. In the GT power cycle, the exergy destruction spans from 58 MW to 83 MW, with the combustion chamber (CC) being the largest contributor. This loss is caused by the irreversibility of the oxidation reactions of the fuel gas to produce thermal energy [319].

Table 4-8. Exergy destruction of each component.

Parameter	Exergy destruction						UNIT
	SCENARIO 1			SCENARIO 2			
	Baseline	W/PRECCS	W/POSTCCS	Baseline	W/PRECCS	W/POSTCCS	
PG	63.37	63.37	63.37	63.37	63.37	63.37	MW
AD	7.56	7.56	7.56	7.56	7.56	7.56	MW
GT power plant							
Fuel							
compressor	5.29	4.52	5.29	4.82	4.27	4.82	MW
Air							
compressor	6.73	6.10	6.73	5.42	4.67	5.42	MW
Gas turbine	9.95	8.97	9.95	8.07	7.01	8.07	MW
Combuster	60.24	53.37	60.24	52.22	42.21	52.22	MW
Sub total	82.22	72.96	82.22	70.53	58.16	70.53	MW
ST power plant							
HRSG	29.50	24.27	27.82	23.69	18.87	22.34	MW
Steam							
Turbine	3.11	1.01	0.66	1.37	0.82	0.65	MW
Pump	0.15	0.00	0.00	0.01	0.00	0.00	MW
Condenser	1.46	0.89	0.58	1.20	0.72	0.58	MW
Sub total	34.22	26.17	29.06	26.26	20.41	23.58	MW
ARC							
Desorber	0.00	0.00	0.00	0.00	0.00	0.00	MW
Condenser	0.06	0.11	0.11	0.04	0.09	0.07	MW
Evaporator	0.32	0.28	0.32	0.30	0.25	0.26	MW
Absorber	0.22	0.19	0.22	0.18	0.15	0.18	MW
Pump	2.43	2.11	2.43	1.97	1.63	1.97	MW
Solution							
heat							
exchanger	0.62	0.54	0.62	0.50	0.42	0.50	MW
Sub total	3.65	3.23	3.70	2.99	2.55	2.97	MW
WS							
Compressor	N/A	N/A	N/A				
1				0.08	0.08	0.08	MW
Cooler 1	N/A	N/A	N/A	0.19	0.19	0.19	MW
Compressor	N/A	N/A	N/A				
2				0.10	0.10	0.10	MW

<i>Cooler 2</i>	N/A	N/A	N/A	0.13	0.13	0.13	MW
<i>Absorber</i>	N/A	N/A	N/A	0.37	0.37	0.37	MW
<i>Flash</i>	N/A	N/A	N/A	0.16	0.16	0.16	MW
<i>Pump</i>	N/A	N/A	N/A	0.00	0.00	0.00	MW
<i>Stripper</i>	N/A	N/A	N/A	0.05	0.05	0.05	MW
Sub total	N/A	N/A	N/A	1.07	1.07	1.07	MW
ML							
<i>Compressor</i>	N/A	N/A	N/A	2.14	2.14	2.14	MW
<i>Heat exchanger</i>	N/A	N/A	N/A	0.52	0.52	0.52	MW
<i>Expander</i>	N/A	N/A	N/A	1.88	1.88	1.88	MW
<i>Flash</i>	N/A	N/A	N/A	0.10	0.10	0.10	MW
Sub total	N/A	N/A	N/A	4.65	4.65	4.65	MW
WGS reactor	N/A	11.15	N/A	N/A	11.15	N/A	MW
CCS plant							
<i>Absorber</i>	N/A	2.13	0.70	N/A	1.66	0.33	MW
<i>Pump 1</i>	N/A	0.06	0.08	N/A	0.03	0.06	MW
<i>Heat exchanger 1</i>	N/A	0.45	0.18	N/A	0.34	0.11	MW
<i>Heat exchanger 2</i>	N/A	0.27	0.71	N/A	0.69	0.57	MW
<i>Flash</i>	N/A	2.41	2.63	N/A	1.96	1.89	MW
<i>Pump 2</i>	N/A	1.30	1.87	N/A	1.09	1.38	MW
<i>Stripper</i>	N/A	12.51	17.15	N/A	10.15	12.61	MW
<i>Condenser</i>	N/A	2.91	5.11	N/A	3.22	3.33	MW
<i>Flash</i>	N/A	0.00	0.00	N/A	0.00	0.00	MW
<i>Compressor prior CCS</i>	N/A	0.11	0.81	N/A	0.10	-0.09	MW
<i>CO₂ Compressor 1</i>	N/A	0.13	0.17	N/A	0.11	0.13	MW
<i>CO₂ Compressor 2</i>	N/A	0.14	0.19	N/A	0.12	0.14	MW
<i>CO₂ Compressor 3</i>	N/A	0.12	0.17	N/A	0.10	0.12	MW
<i>CO₂ Cooler 1</i>	N/A	0.12	0.16	N/A	0.23	0.12	MW
<i>CO₂ Cooler 2</i>	N/A	0.14	0.18	N/A	0.23	0.13	MW
<i>CO₂ Cooler 3</i>	N/A	0.37	0.49	N/A	0.40	0.36	MW
<i>Separator</i>	N/A	0.27	0.36	N/A	0.22	0.26	MW
<i>Pump 3</i>	N/A	0.07	0.09	N/A	0.06	0.07	MW
Sub total	N/A	23.52	31.05	N/A	20.71	21.52	MW

In addition to the PG and GT power cycle, the exergy destruction within the ST power cycle ranges from 20 to 35 MW. The HRSG, ST, and condenser accounted for the majority of exergy destruction in this subsystem. The exergy destruction in

the HRSG is mostly affected by heat transfer irreversibility between the exhaust gases and the steam. Meanwhile, the exergy destruction of the steam turbine is caused by expansions at high pressure and temperature across multiple stages. Finally, the exergy destroyed in the condenser results from heat loss as the expanded LP steam exiting the LP turbine is cooled down by the cooling water, which is at a low temperature of 25°C [320].

Moreover, the carbon capture facility exhibits significant exergy destruction, varying from 20 MW to 31 MW, with the stripper being the primary source. The exergy loss is caused due to the heating of the solvent to release the absorbed CO₂ using LP steam, the chemical reaction that occurs during the desorption process, and the pressure drops experienced by the solutions as they flow through the stripper [321], [322], [323].

Finally, the WGS reactor demonstrates an exergy destruction of approximately 11 MW. The disparity in temperature between the reactants and products leads to significant exergy degradation. The WGS reactor in this study operates at two distinct temperatures: a high-temperature shift of 450°C and a low-temperature shift of 200°C. Other components, such as AD, ARC, ML, and WS, have an exergy destruction below 10 MW. The processes on these sub systems operate under low-temperature conditions, leading to lower exergy destruction compared to high-temperature chemical reactions.

Overall, the exergy destruction analysis shows that the largest inefficiencies are associated with high-temperature reaction and heat transfer process, particularly in the plasma gasifier, combined cycle gas turbine, and carbon capture unit, while the lower temperature subsystems contribute comparatively smaller losses.

4.3.4 Carbon balance

Figure 4-15 and Figure 4-16 illustrate the carbon mole flow in both Scenario 1 and Scenario 2, showing how carbon distribution changes depending on whether the system operates under baseline, pre-CCS, or post-CCS configurations. As shown, CO₂ emissions are higher in all three configurations of Scenario 1 compared to Scenario 2. This is because Scenario 1 releases more flue gas due to the combustion of both syngas and biogas in the gas turbine power plant, whereas in

Scenario 2, only syngas is combusted, and the biogas is converted into liquid biomethane. In Scenario 1, the flue gas contains both biogenic and fossil carbon, while in Scenario 2, the flue gas consists only of fossil carbon.

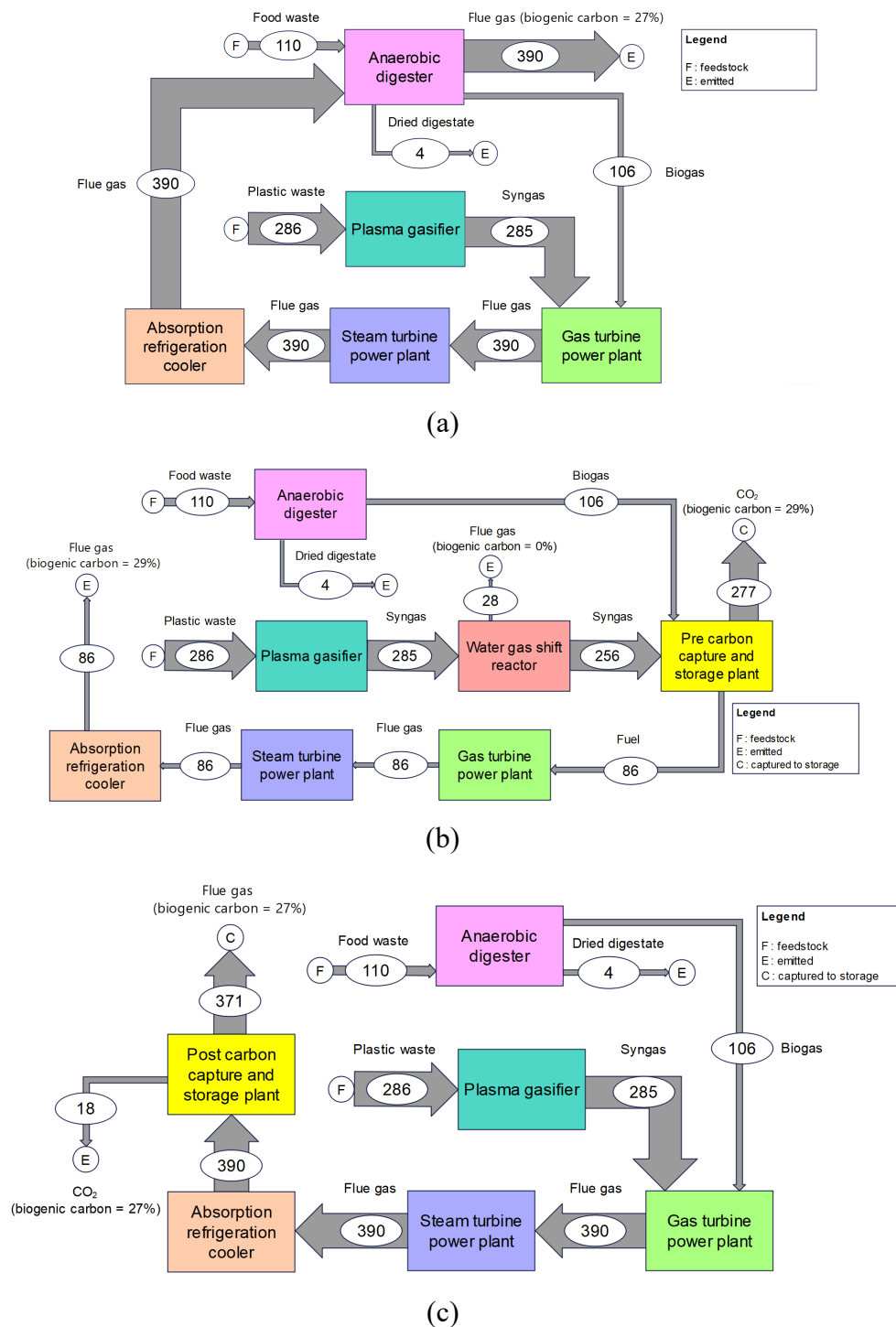
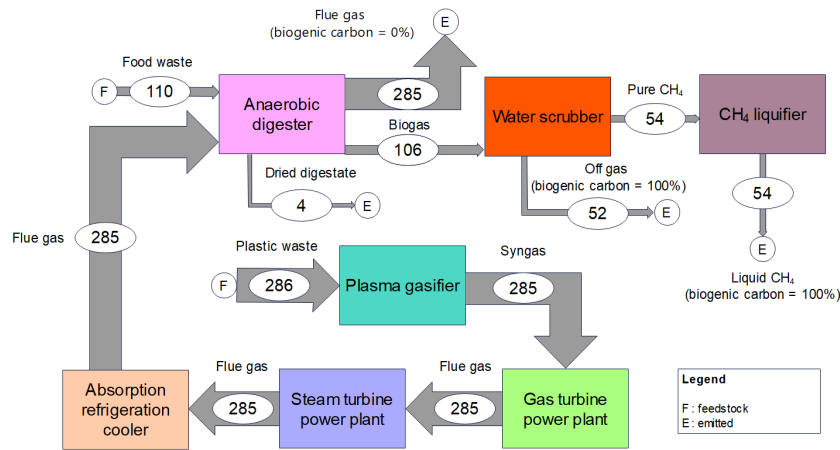


Figure 4-15. Carbon mole flow (mol/s) for Scenario 1: (a) baseline, (b) pre-CCS, (c) post-CCS.

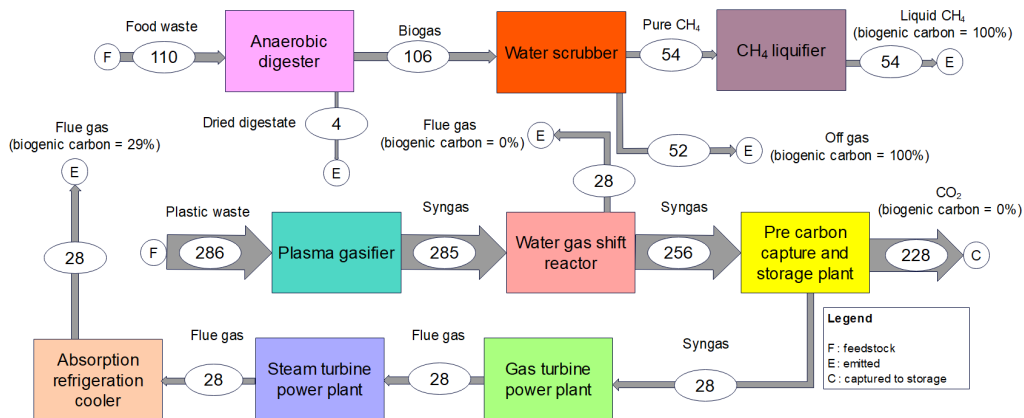
The baseline case in each scenario emits the highest amount of carbon due to the

absence of CCS integration. In Scenario 1, the baseline case emits 0.247 kgCO₂ released per kg of waste processed, and in Scenario 2, the emission factor is 0.213 kgCO₂ per kg of waste processed.

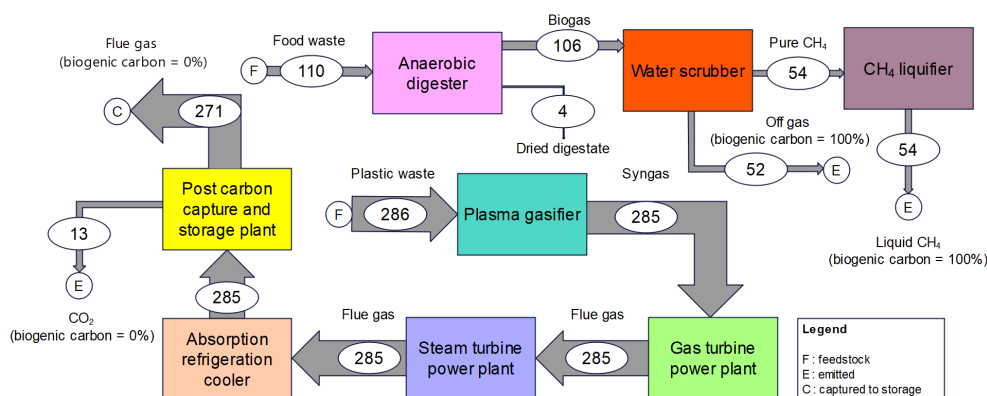
In the pre-CCS configuration, CO₂ emissions are higher than in the post-CCS case. This occurs because, prior to entering the pre-CCS plant, not all CO in the syngas can be fully converted into CO₂ inside the WGS reactor. As a result, the combustion of CO still releases amounts of CO₂ into the atmosphere. As shown in Figure 4-15b & Figure 4-16b, the CO₂ emissions in the pre-CCS case are 0.072 kgCO₂ per kg of waste processed in Scenario 1 and 0.068 kgCO₂ released per kg of waste processed in Scenario 2.



(a)



(b)



(c)

Figure 4-16. Carbon mole flow (mol/s) in Scenario 2: (a) baseline, (b) pre-CCS, (c) post-CCS.

In contrast, in the post-CCS case, the capture process occurs after the conversion of all carbon in the fuel to CO_2 , resulting in a more direct and efficient capture of the generated carbon. The case with post-CCS plant emits the least CO_2 , around 0.011 kgCO_2 released per kg of waste processed and 0.008 kgCO_2 released per kg of waste processed for Scenario 1 and 2, respectively. This finding highlights the significance of post-CCS as the most effective configuration for minimizing direct carbon emissions and improving the overall carbon performance of the system.

4.3.5 Electricity distribution

The electricity generated by the CCGT is used to power the various components within the system. Figure 4-17 illustrates the distribution of electricity consumption across the system for both Scenario 1 and Scenario 2 under three different configurations. The data highlights how electricity demand varies depending on the type of components, energy production method and carbon capture technologies.

By comparing the components' operations in each configuration, it is evident that the plasma gasification unit consistently consumes the most electricity, reaching up to 45.39 MW. This high consumption is primarily due to the need for generating and maintaining plasma at extremely high temperatures, which is essential for efficiently breaking down plastic waste.

The energy vectors generated also affect electricity consumption in each scenario. Scenario 1, which produces electricity, cooling, and heating, has a lower electricity

demand compared to Scenario 2, which generates electricity, cooling, heating, and liquid biomethane. In Scenario 1, the mixture of syngas and biogas is used as fuel for the GT power plant, while Scenario 2 converts biogas into liquid biomethane rather than burning it. The additional steps in Scenario 2, including upgrading biogas into biomethane and the biomethane liquefaction process, lead to higher electricity consumption.

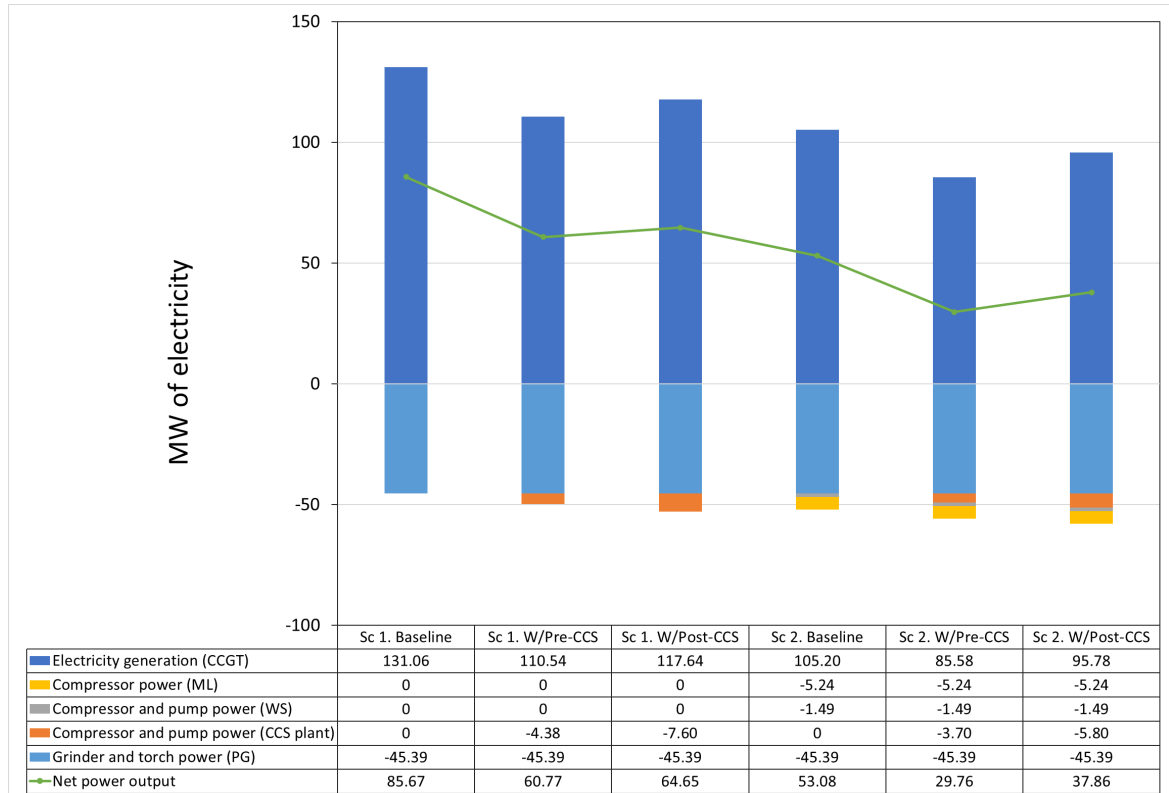


Figure 4-17. Electricity generation and consumption breakdown for Scenario 1 and Scenario 2 across the three configurations.

Additionally, the type of carbon capture technology influences electricity consumption within the system. Pre-CCS requires less electricity than post-CCS due to the lower amount of CO₂ captured and sequestered. The highest electricity consumption in both CCS systems is attributed to the operation of compressors, which are needed to increase the input gas pressure to 1.18 bar and raise the pressure of the captured CO₂ to 83 bar. The type of fuel used in the GT power plant also impacts electricity demand in the CCS systems. In Scenario 1, which uses a mixture of biogas and syngas as fuel, the CCS systems consume more electricity compared to Scenario 2, where only syngas is used. This is because more

electricity is required to remove the higher CO₂ concentration in the fuel mixture.

Finally, as shown in Figure 4-17, the baseline cases in both scenarios yield the highest net power output due to the combination of maximum electricity generation and minimal electricity consumption. Specifically, in Scenario 1, the net power output reaches 85.67 MW, while in Scenario 2, it is 53.08 MW.

4.3.6 Overall system performance

Table 4-9 compares the overall performance of the system in Scenario 1 (CCHP) and Scenario 2 (CCHP, and liquid biomethane) across three configurations: baseline, pre-CCS, and post-CCS. The key parameters examine include total energy generation, total exergy generation, system energy efficiency, and system exergy efficiency.

As shown in Table 4-9, Scenario 2 achieves higher system energy and exergy efficiency compared to Scenario 1, despite generating lower levels of power, cooling, and heating. This increased efficiency is mainly attributed to the high efficiency of biomethane production, which has energy and exergy conversion efficiencies of 85.79% and 84.04%, respectively, as mentioned in 4.3.2.4. In contrast, the CCGT has thermal energy efficiency ranging from 49% to 56% and exergy efficiency between 50% and 58%, as shown in Table 4-9.

Table 4-9. Thermodynamic results for the overall system performance for Scenario 1 and Scenario 2.

Overall system performance	Case			UNIT
	Baseline	W/PRE-CCS	W/POST-CCS	
Scenario 1 (electricity, heating, and cooling)				
Total energy generation	107.01	92.14	95.54	MW
Total exergy generation	87.63	63.89	68.27	MW
System energy efficiency	46.57	40.10	41.58	%
System exergy efficiency	35.55	25.92	27.70	%
Scenario 2 (electricity, heating, cooling, and liquid biomethane)				
Total energy generation	113.59	101.36	101.52	MW
Total exergy generation	101.56	79.41	87.99	MW
System energy efficiency	49.44	44.11	44.18	%
System exergy efficiency	41.20	32.22	35.70	%

The baseline case consistently achieves the highest energy and exergy efficiency in

both scenarios, followed by the post-CCS case, with the pre-CCS case showing the lowest efficiencies. The reason is that the baseline case does not provide heat for the stripper in the carbon capture plant and does not allocate 10% of syngas for WGSR. Although the heat demand in the post-CCS case is higher than in the pre-CCS case, the pre-CCS case has less total energy and exergy generation due to the 10% syngas allocation to the WGSR.

Considering both energy production and carbon emissions, the post-CCS case in Scenario 2 emerges as the most favourable option for waste-to-energy utilization. This configuration balances efficient energy production with significant carbon reduction, making it a key solution to adopt low-carbon technologies and mitigate climate change.

4.4 Conclusion

This study demonstrates the significant potential of integrating waste-to-energy systems with carbon capture and storage (CCS) technologies to address both energy demands and environmental concerns. By utilizing 5 kg/s of plastic waste and 13.97 kg/s of food waste as feedstocks, the research evaluates for the first time the thermodynamic performance of an integrated energy system consisting of a plasma gasifier (PG), anaerobic digester (AD), combined-cycle gas turbine (CCGT), absorption refrigeration cooler (ARC), and biomethane liquefier. Two scenarios, each with three configurations (baseline, pre-CCS, and post-CCS), were explored to assess the system's energy and exergy efficiency, as well as carbon emissions.

The findings of the study indicate that Scenario 1, which is a CCHP system, generated higher amount of electricity, heating and cooling output, but had lower overall efficiency compared to Scenario 2, which includes liquid biomethane production. The baseline case of Scenario 2 achieves the highest overall energy and exergy efficiency, with maximum efficiencies of 49.44% and 41.20%, respectively. On the other hand, the post-CCS configuration of Scenario 2 emits the lowest amount of CO₂, i.e. 0.008 kgCO₂ per kg of waste processed.

These results highlight the effectiveness of integrating advanced waste conversion technologies with CCS in enhancing energy recovery and significantly reducing

greenhouse gas emissions. The findings provide a promising pathway for sustainable waste management and energy production, offering valuable insights for detailed engineering studies, policy development, and academic research. Future research will focus on conducting economic and life cycle assessments for both Scenarios 1 and 2 to further evaluate their feasibility and sustainability.

5 Economic and Life Cycle Assessment of novel hybrid energy and fuel generation systems from municipal waste through plasma gasification and anaerobic digestion coupled with carbon capture and storage

Abstract

Achieving climate goals demands novel system designs that enable the conversion of municipal waste, such as plastic and food waste into energy and fuels with minimal environmental impact. This study proposes an innovative multi-energy generation system that integrates plasma gasification for plastic waste and anaerobic digestion for food waste, coupled with carbon capture and storage (CCS) technologies. This novel conceptual design aims to maximize energy recovery while reducing lifecycle emissions compared to conventional waste-to-energy pathways. Two novel system configurations are assessed: (1) a combined cooling, heating, and power (CCHP) system, and (2) a CCHP system integrated with liquid biomethane production. Each configuration is evaluated under three CCS strategies: no CCS, pre-combustion CCS, and post-combustion CCS. The techno-economic analysis (TEA) and life cycle assessment (LCA) highlight the environmental and economic trade-offs of each design. Specifically, in Scenario 1, the levelized cost of electricity (LCOE) increases from 0.171 USD/kWh (no CCS) to 0.311 and 0.354 USD/kWh while in Scenario 2, the levelized cost of biomethane (LCOBM) rises from 0.176 USD/kWh to 0.314 and 0.374 USD/kWh for pre- and post-combustion CCS, respectively. While CCS raises production costs, they also represent a tangible commitment to reducing emissions and underscore that transitioning to cleaner energy systems often entails higher near-term expenditures. Across both scenarios, the levelized cost of waste treatment (LCOWT) spans 0.081–0.236 USD/kg of waste. Global warming potential (GWP) ranges from -0.191 to 0.662 kgCO₂-eq/kg of feedstock for Scenario 1, and 0.123 to 0.745 kgCO₂-eq/kg for Scenario 2. This work provides the first integrated assessment of such a hybrid waste-to-energy system, offering new insights for sustainable waste valorisation. The proposed novel designs support future detailed engineering studies and inform policymaking for low-carbon waste management.

Keywords: Novel system design; Sustainable waste management; Waste-to-energy; Plasma gasification; Carbon capture and storage; TEA; LCA

5.1 Introduction

Effective waste management is a critical component of sustainable development, as reflected in the United Nations Sustainable Development Goal (SDG) 11, which emphasizes the importance of sustainable cities and communities [324]. Among various waste streams, plastic and food waste are of particular concern. Plastics are resistant to degradation, persisting in the environment and contributing to long-term pollution, while food waste decomposes into methane, a greenhouse gas with a GWP far exceeding that of carbon dioxide. In Indonesia, food and plastic waste account for 28% and 16% of total MSW, respectively, making their management a national priority.

WtE technologies offer a pathway to mitigate these challenges by simultaneously reducing waste volumes and recovering energy. Plasma gasification is particularly effective for plastics, operating at extremely high temperatures to produce a tar-free syngas and vitrified slag suitable for reuse [77], [186]. Anaerobic digestion on the other hand, is suitable for high moisture organic waste such as food waste, producing biogas and digestate that can be used as fertilizer [32], [192]. Integration of plasma gasification, and anaerobic digestion enables the utilization of complementary feedstock, maximizing resource recovery. The addition of CCS can further enhance the environmental performance of such systems by significantly reducing fossil and biogenic CO₂ emissions [217], [218].

Previous studies have primarily focused on single output systems, such as syngas [229], [230], [233], [325], pyrolysis oil [231], [232], electricity [93], [187], hydrogen [234], hydrogen and electricity [185], [221], CCHP [198], [199], and methanol and electricity [228]. Research into dual-feedstock WtE systems producing multiple energy vectors, such as CCHP and liquid biomethane remains limited. Furthermore, comprehensive economic, and life cycle assessment (LCA) of such systems under conditions relevant to developing countries like Indonesia are scarce, and little attention has been given to the role of CCS in these integrated designs.

To fill this gap, this chapter investigates the economic and environmental performance of a novel dual feedstock WtE system that integrates plasma gasification of plastic waste and anaerobic digestion of food waste with CCS. The system is designed to produce CCHP, and liquid biomethane, alongside by products such as vitrified slag and digestate. Three configurations are evaluated: a baseline without CCS, a pre combustion CCS case, and a post combustion CCS case. Economic assessment is applied to determine the LCOE, LCO_{bM}, and LCOWT, while LCA focuses on GWP and WF. Sensitivity and uncertainty analyses are performed to explore investment risks, while policy scenarios assess the influence of financial incentives on system competitiveness.

This chapter presents the research published in the author's second journal paper, which provides the first integrated economic and LCA of a dual-feedstock plasma gasification-anaerobic digestion system with CCS. The study contributes original insights into the trade-offs between cost, carbon performance, and policy design, with direct relevance to the sustainable management of waste and energy transition in Indonesia.

5.2 Methodology

5.2.1 System description

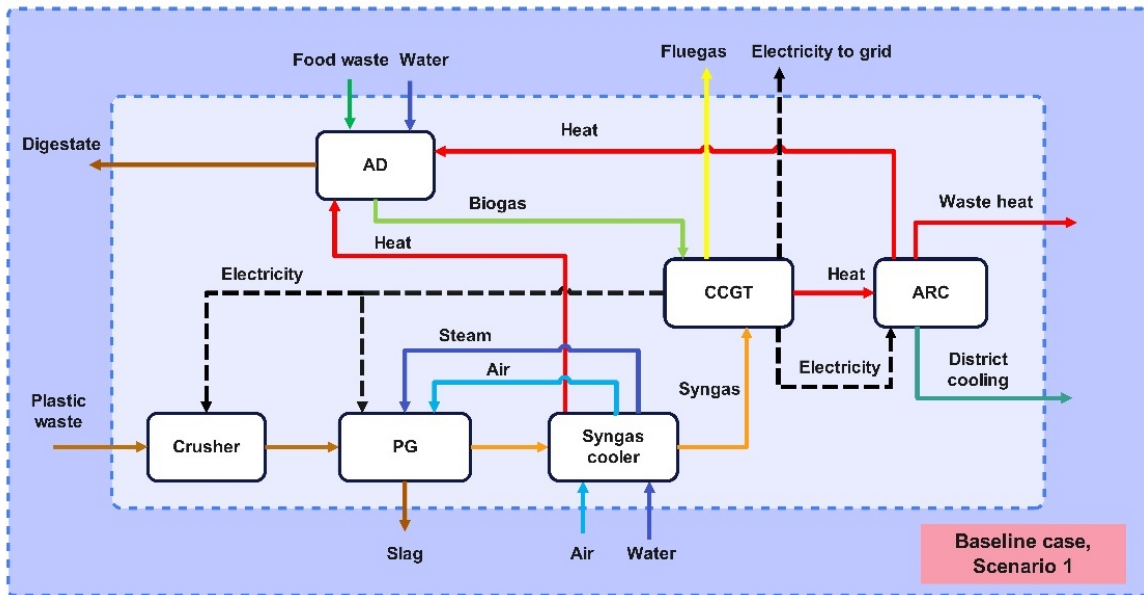
Feedstock is sourced from the 37-hectare Benowo landfill in Surabaya, Indonesia's second-largest city. The system processes mixed waste, approximately 74 wt.% food waste and 26 wt.% plastic, at feed rates of 13 kg/s (46 MW_{th}) for food waste and 5 kg/s (184 MW_{th}) for plastic, respectively. The waste-to-energy system was designed to generate multiple valuable outputs, including electricity, cooling, heating, and liquid fuel, while also meeting the internal heat and electricity demands for system operation. Additionally, the system generates useful by-products, such as slag from plasma gasification, which can be repurposed as construction material, and digestate from anaerobic digestion, which can be used as fertilizer.

Mixed plastic waste, chosen for its low moisture content, is first crushed and fed into a plasma gasifier operating at 2500°C, with air and steam as gasifying agents. A plasma torch heats the incoming air to 4000°C, supplying the thermal energy

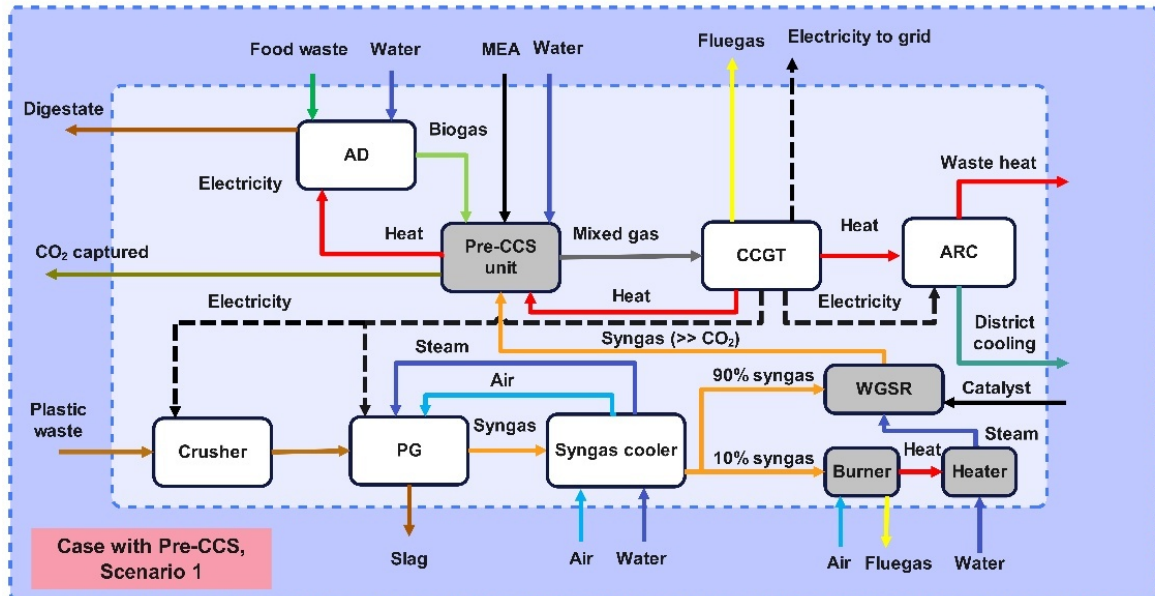
required; this torch is powered by electricity generated on-site. The raw syngas leaving the gasifier has a molar composition of 42.6% H₂, 24% CO, 1.4% CO₂ while slag is produced at a rate of 0.66 kg/s.

Food waste, by contrast, contains 76% moisture and is thus routed to a mesophilic anaerobic digester at 38 °C, producing biogas composed of 47.5% CH₄ and 43.4% CO₂ (molar) and generates digestate “fertilizer” at 15.95 kg/s. In the baseline case, the heating demand required to maintain the operating temperature of the AD process is met by recovering heat from the raw syngas and the flue gas. In contrast, for the scenario incorporating CCS, the heating requirement is fulfilled by utilizing heat recovered from the vapor stream exiting the stripper in the CCS plant, prior to the reflux condenser.

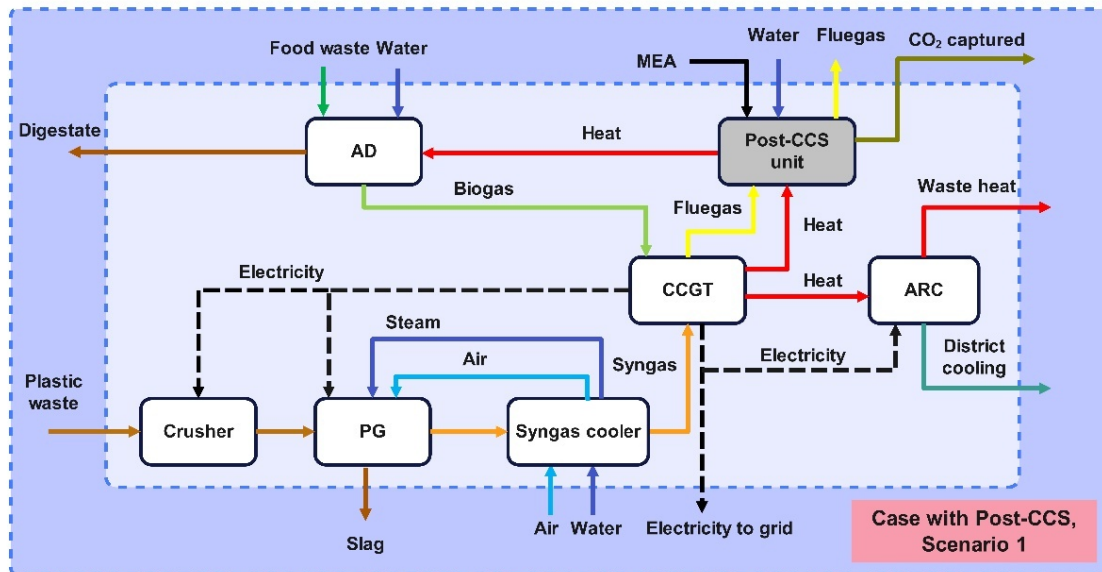
Two scenarios are explored in this study. In Scenario 1, syngas and biogas are utilized as fuels in the combined cycle gas turbine (CCGT) to generate electricity, as illustrated in Figure 5-1. The heat recovered from the flue gas of the CCGT is then used in an absorption refrigeration chiller (ARC) to provide residential cooling, while the remaining waste heat is applied to fish drying industries. Scenario 1 is referred to as a combined cooling, heating, and power (CCHP) production system. Meanwhile, in Scenario 2, only syngas is used as fuel in the CCGT, while the biogas is converted into liquid biomethane through a water scrubber (WS) and methane liquefier (ML), as shown in Figure 5-2. Scenario 2 is referred to as a CCHP and liquid biomethane production system. Both figures outline the system boundaries for the economic analysis and LCA conducted in this study.



(a)

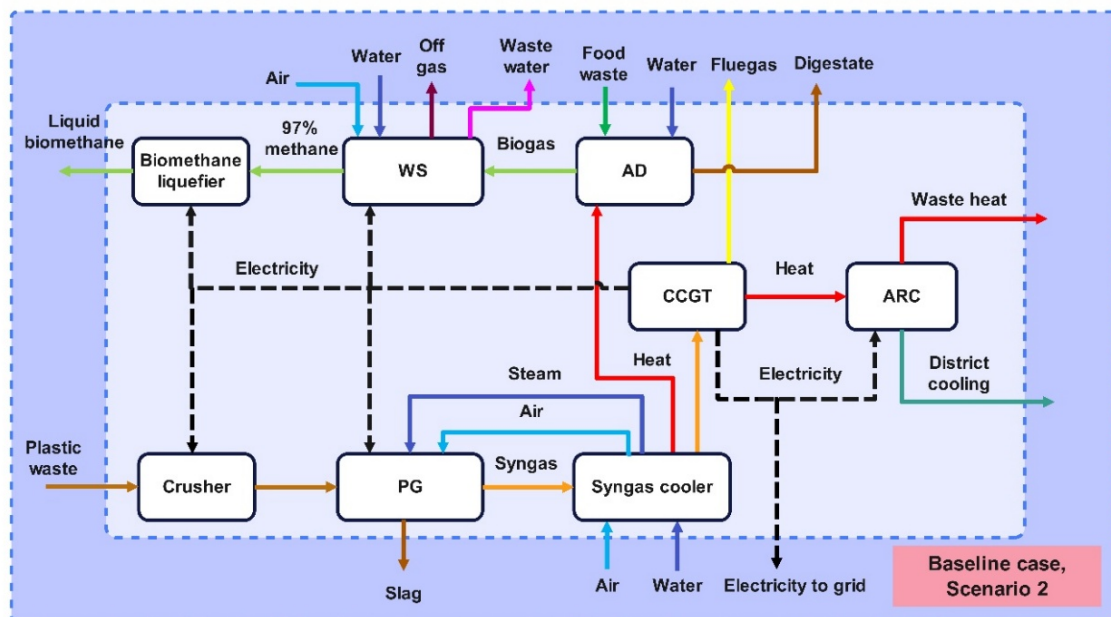


(b)

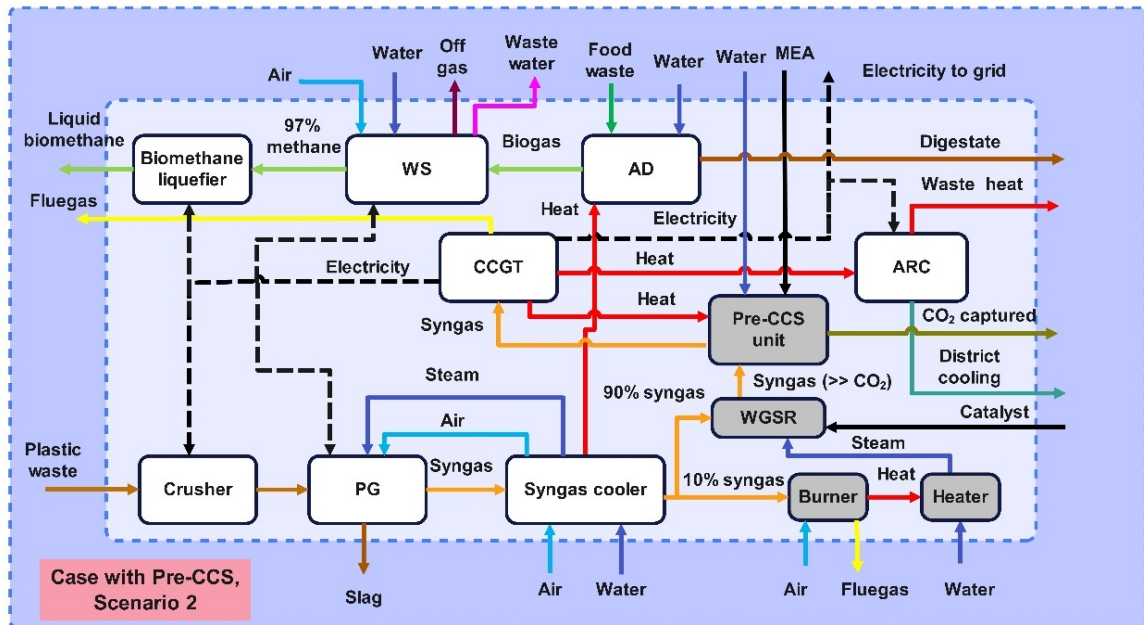


(c)

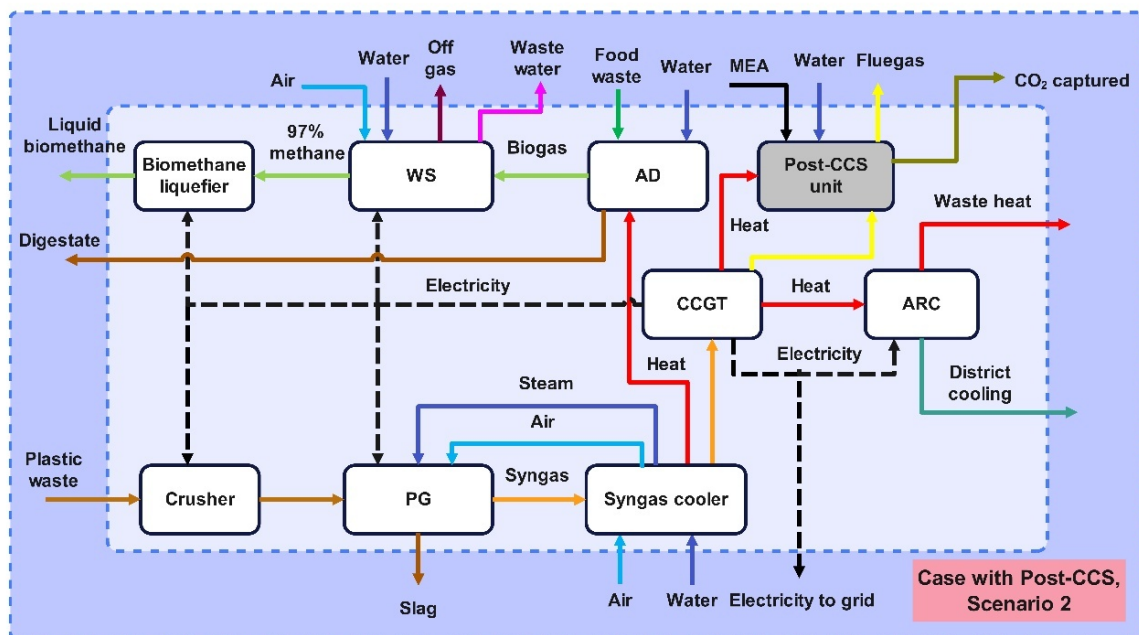
Figure 5-1. System boundaries and configurations for Scenario 1 (products: cooling, heating, and electricity) with three different cases: (a) baseline, (b) w/ pre-CCS, (c) w/ post CCS.



(a)



(b)



(c)

Figure 5-2. System boundaries and configurations for Scenario 2 (products: cooling, heating, electricity, and liquid fuel) with three different cases: (a) baseline, (b) w/ pre-CCS, (c) w/ post CCS.

Each scenario includes three cases: a baseline case without CCS integration, a case integrated with pre-combustion CCS (w/ pre-CCS), and a case integrated with post-combustion CCS (w/ post-CCS). In the w/ pre-CCS case, a WGSR is installed

after the PG, and the CCS plant is positioned between the WGSR and the CCGT. To facilitate the WGSR reaction, 10% of the raw syngas is extracted and burned to generate steam for the process. The water gas shift (WGS) process is carried out in two stages to maximize CO conversion: first, at 450°C in a high-temperature WGSR (HTWGSR) using a Fe-Cr catalyst, and second, at 200°C in a low-temperature WGSR (LTWGSR) employing a CuO/ZnO/Al₂O₃ catalyst. To lower the temperature of the stream between the HTWGSR and LTWGSR, as well as the stream exiting the LTWGSR prior to entering the CCS plant, heat is recovered and used to generate low-pressure (LP) steam, which can subsequently be utilized for heating the stripper in the CCS system.

The syngas exiting the WGSR, now enriched with CO₂, is directed to the pre-CCS plant, where CO₂ is captured at a rate of 95%. In Scenario 1, the syngas is first blended with biogas before entering the CCS plant, whereas in Scenario 2, only syngas is used as the input stream. The chemical solvent employed is MEA at a concentration of 35 wt.%. The CCS plant comprises two main components: an absorber and a stripper. The feed gas enters at the bottom of the absorber column, where it is contacted counter currently with the MEA solution sprayed from the top. The column packing is designed to operate below the 80% flooding limit to ensure stable and efficient mass transfer. Within the absorber, CO₂ is absorbed into the MEA, forming a CO₂-rich solvent, while the cleaned gas exits at the top. The CO₂-rich solvent is then directed to the stripper, where it is heated using LP steam generated within the system to release the absorbed CO₂. Once the CO₂ is removed, the regenerated lean solvent is recycled back to the absorber. The CO₂ stream exiting the stripper is cooled to condense water and separate vapour components. The purified CO₂ is then compressed to supercritical conditions and further cooled into a liquid state for transport and storage in deep geological formations.

Afterwards, the clean gas, containing minimal CO₂, is used as fuel in the CCGT. In the gas turbine (GT) cycle, the required air flow rate for combustion is determined to ensure the turbine exhaust reaches 1350°C. This high-temperature exhaust gas drives the gas turbine to generate electricity. The residual thermal energy in the exhaust is then recovered in the heat recovery steam generator (HRSG) of the

steam turbine (ST) cycle. To maximize power output, the steam cycle operates with three pressure levels: high-pressure (HP), intermediate-pressure (IP), and low-pressure (LP) steam. The LP steam is split into two streams: one directed to the LP turbine, and the other supplied to the CCS plant to meet the reboiler heat duty in the stripper. After passing through the HRSG, the exhaust gas flows into the ARC for further heat recovery.

In the ARC, the working fluid is a mixture of lithium bromide and water (LiBr–H₂O). Lithium bromide serves as the absorbent, enabling water to evaporate at low temperatures by maintaining low system pressure. The primary components of the system include the evaporator, absorber, generator, condenser, throttling devices, and a solution pump. Waste heat from the exhaust gas is utilized in the generator, where it heats the working fluid, causing the separation of water vapor from the LiBr solution. The water vapor then flows to the condenser, where it is cooled and condensed into liquid water. This liquid refrigerant is throttled and directed to the evaporator, where it absorbs heat from the surroundings, producing the cooling effect. The resulting water vapor is drawn into the absorber and absorbed by the concentrated LiBr solution, forming a dilute solution that is pumped back to the generator to complete the cycle. The residual heat in the flue gas exiting the ARC is subsequently used as a thermal source for fish drying in local industries.

In the w/ post-CCS case, the CCS plant is positioned downstream of the ARC to recover waste heat from the CCGT flue gas. The cooled flue gas is then directed to the post-CCS plant, where CO₂ is captured before the remaining gas is released into the atmosphere. The overall process in the post-CCS plant closely resembles that of the w/ pre-CCS case, but with a few notable distinctions. First, the absorber column in the post-CCS configuration features a larger packing diameter to accommodate the higher volumetric flow rate of flue gas, which helps prevent flooding and ensures effective mass transfer. Second, the lean and rich solvent loadings in the post-CCS system are slightly lower than in the pre-CCS system, reflecting the lower CO₂ concentration in the flue gas compared to the gas streams in Scenario 1 (mixed syngas and biogas) and Scenario 2 (syngas only). For clarity, components that are unique to the w/ pre-CCS and w/ post-CCS configurations are highlighted using grey blocks in the system diagrams.

In Scenario 2, where biogas is upgraded to liquid biomethane, a water scrubber and a methane liquefaction system are employed. The water scrubber operates under high pressure and low temperature conditions to enhance the solubility of CO₂ in water, thereby improving CO₂ removal efficiency. The main components of the water scrubbing system include compressors, absorbers, strippers, pumps, and heat exchangers. After treatment in the scrubber, the methane concentration in the biogas can reach up to 97.7%. The upgraded biogas is then directed to a multistage compressor, where its pressure is increased to 200 bar in preparation for liquefaction. In the liquefaction stage, the high-pressure gas is cooled and subsequently expanded to achieve the low temperatures necessary for methane condensation and liquefaction. The flow diagram and operating conditions of each component used in both scenarios are provided in reference [64].

5.2.2 Economic assessment

Gross profit is calculated by monetizing the useful energy outputs of the system, which include electricity, cooling, heating, and liquid biomethane. For electricity, cooling, and heating, the exergy content is used to capture the quality and usefulness of these energy forms, reflecting their economic value more accurately than raw energy content. However, for liquid biomethane, the calculation is based on its energy content rather than exergy. This is because market prices are generally tied to energy content. Using energy content for liquid biomethane aligns the economic assessment with market valuation practices and ensures consistency with fuel pricing.

The gross profit calculation varies by scenario which depends on the main product. For example, in Scenario 1, where the electricity is the main product, the selling price of the co-products (cooling and heating) follows the electricity market price (refer to Table 5-1). In Scenario 2, where the liquid biomethane is the main product, the selling price of the co-products (electricity, cooling and heating) follows the electricity market price. The corresponding gross profit equations are presented in Eq. 5-1 and Eq. 5-2.

$$P_{n,Sc1} = \left(LCOE \times \dot{W}_{elec} + \$_{cooling} \times \frac{\dot{E}x_{cooling}}{COP_{cooler}} + \$_{heating} \times \frac{\dot{E}x_{heating}}{COP_{heater}} \right. \\ \left. + \$_{slag} \times \dot{m}_{slag} + \$_{tipping} \times \dot{m}_{waste} \right. \\ \left. + \$_{carbon\ credit} \times \dot{m}_{captured\ CO_2} \right) \times C_f \times 8000 - OPEX_n \quad \text{Eq. 5-1}$$

$$P_{n,Sc2} = \left(LCOBm \times \dot{m}_{LBM} \times LHV_{LBM} + \$_{elec} \times \dot{W}_{elec} \right. \\ \left. + \$_{cooling} \times \frac{\dot{E}x_{cooling}}{COP_{cooler}} + \$_{heating} \times \frac{\dot{E}x_{heating}}{COP_{heater}} \right. \\ \left. + \$_{slag} \times \dot{m}_{slag} + \$_{tipping} \times \dot{m}_{waste} \right. \\ \left. + \$_{carbon\ credit} \times \dot{m}_{captured\ CO_2} \right) \times C_f \times 8000 - OPEX_n \quad \text{Eq. 5-2}$$

where $\dot{W}_{elec}$ is net electricity (kWh), $\$_{cooling}$ and $\$_{heating}$ are equal to electricity market price ($\$_{elect}$) shown in Table 5-1, $\dot{E}x_{cooling}$ and $\dot{E}x_{heating}$ are the cooling and heating exergy (kWh), respectively, COP_{cooler} was assumed to be 3.45 [326], COP_{heater} was assumed to be 1. $\$_{slag}$ and $\dot{m}_{slag}$ are the price of slag shown in Table 5-1 and mass flow rate of slag (tonne/h), respectively. $\dot{m}_{LBM}$ and LHV_{LBM} are the mass flow rate (kg/s) and lower heating value of liquid biomethane (kJ/kg), respectively. The tipping fee ($\$_{tipping}$) follows the Indonesian waste tipping fee of 14.02 USD/tonne [327], with $\dot{m}_{waste}$ representing the total waste processed. The carbon credit price ($\$_{carbon\ credit}$) is based on the Indonesian market at 8.9 USD/tonne CO_2 [328]. The capacity factor (C_f) and OPEX correspond to a process plant operating 8,000 hours per year.

In addition, to calculate the LCOWT, the main product and co-product was assumed to be same as their market price, thus the gross profit equation for LCOWT calculation ($P_{n,wt}$) for Scenario 1 and 2 can be seen in Eq. 5-3 and Eq. 5-4, respectively.

$$P_{n,wt,sc1} = \left(LCOWT \times \dot{m}_{waste} + \$_{elec} \times \dot{W}_{elec} + \$_{cooling} \times \frac{\dot{E}x_{cooling}}{COP_{cooler}} \right. \\ \left. + \$_{heating} \times \frac{\dot{E}x_{heating}}{COP_{heater}} + \$_{slag} \times \dot{m}_{slag} \right. \\ \left. + \$_{tipping} \times \dot{m}_{waste} + \$_{carbon\ credit} \times \dot{m}_{captured\ CO_2} \right) \\ \times C_f \times 8000 - OPEX_n \quad \text{Eq. 5-3}$$

$$\begin{aligned}
P_{n,wt,sc2} = & \left(LCOWT \times \dot{m}_{waste} + \$_{LNG} \times \dot{m}_{LBM} \times LHV_{LBM} \right. \\
& + \$_{elec} \times \dot{W}_{elec} + \$_{cooling} \times \frac{\dot{E}x_{cooling}}{COP_{cooler}} \\
& + \$_{heating} \times \frac{\dot{E}x_{heating}}{COP_{heater}} + \$_{slag} \times \dot{m}_{slag} \\
& + \$_{tipping} \times \dot{m}_{waste} + \$_{carbon\ credit} \times \dot{m}_{captured\ CO_2} \left. \right) \\
& \times C_f \times 8000 - OPEX_n
\end{aligned}
\tag{Eq. 5-4}$$

where $\$_{LNG}$ is the liquid natural gas market price, as shown in Table 5-1.

Table 5-1. Market price applied in the economic and environmental assessment.

Parameter	Description
Electricity market price [329]	0.105 USD/kWh
LNG market price [330]	0.049 USD/kWh
Vitrified slag price [215]	53.78 USD/tonne
Indonesia's tipping fee [327]	14.02 USD/tonne
Indonesia's carbon credit price [328]	8.9 USD/tonne CO ₂

To calculate the LCOWT, LCOE, and LCOBM, two key metrics, namely capital expenditure (CAPEX) and operating expenditure (OPEX), need to be determined. The calculation of CAPEX involves the utilisation of a methodology referred to as the bottom-up approach. In this approach, the cost of each component is represented as a fraction of the total purchased equipment cost (PEC), as detailed in Table 3-4. The capital cost contribution of each component is calculated by multiplying the total PEC by its corresponding factor. These factors are typically derived from historical cost data of comparable processes to ensure accuracy [331]. The PEC values applied in this study are presented in Table 5-2 and Table 5-3. OPEX includes both fixed and variable costs. The fixed costs are listed in Table 3-5, while the variable component (VC) consists primarily of utilities, which are detailed in Table 5-4.

Table 5-2. PEC at base capacity, PEC formulation and year of reference.

Equipment	Base cost [MM USD]	Base capacity	Unit	Scaling factor	Base year	Ref.	CEPCI
Crusher	2	2000	ton/day (dry feed mass flow)	0.72	2007	[332]	525.4
Drier	19.3	2000	ton/day (dry feed mass flow)	0.72	2007	[332]	525.4
Plasma gasifier	78	39.2	kg/s (feed mass flow)	0.67	2021	[333]	708.8
Substrate heater	0.693	13,149	m ² (heat transfer area)	0.68	2014	[334]	576.1
Anaerobic digester	0.455	727.21	m ³ (feed volume flow)	1	2017	[335]	567.5
Desorber	0.0175	100	m ² (heat transfer area)	0.6	2000	[336]	394.1
Absorber	0.0165	100	m ² (heat transfer area)	0.6	2000	[336]	394.1
Solution heat exchanger	0.012	100	m ² (heat transfer area)	0.6	2000	[336]	394.1
WGSR	3.77	150	kg/s (total gas feed)	0.67	2014	[337]	576.1
Burner ^a	2.14	20	MW (heat duty)	0.83	2014	[337]	576.1
Cooling tower	3.99	4530.3	kg/s (feed mass flow)	0.78	2014	[338]	576.1
Heat exchanger	0.8	8372	m ² (heat transfer area)	0.68	2014	[334]	576.1
$PEC_{AC} = \frac{71.1\dot{m}_{AC}}{0.9 - \eta_{AC}} \left(\frac{P_{AC,out}}{P_{AC,in}} \right) \ln \left(\frac{P_{AC,out}}{P_{AC,in}} \right)$							
Air compressor (AC)	Here, $\dot{m}_{AC}$ denotes the mass flow rate through the air compressor (AC), η_{AC} represents its isentropic efficiency, and $P_{AC,in}$ and $P_{AC,out}$ are the fluid pressures at the air compressor's inlet and outlet, respectively.			0.7 ^b	1994	[339]	368.1
CCS plant	The PEC of all components (compressors, pumps, heat exchangers, columns) in the CCS plant is estimated using the financial calculations provided in the referenced study			0.7 ^c	2022	[340]	816

Table 5-3. PEC from Aspen Plus.

Component	Equipment Cost (USD)					
	Scenario 1			Scenario 2		
	Case 1	Case 2	Case 3	Case 1	Case 2	Case 3
Fuel						
Compressor	\$32,465,200.00	\$26,624,700.00	\$32,465,200.00	\$26,850,200.00	\$25,184,100.00	\$26,850,200.00
Combustion						
Chamber+Gas	\$82,985,000.00	\$76,422,000.00	\$82,985,000.00	\$70,914,000.00	\$63,543,000.00	\$70,914,000.00
turbine						
HRSG	\$3,464,500.00	\$3,775,100.00	\$7,202,200.00	\$2,778,600.00	\$2,886,900.00	\$4,936,500.00
LP-PUMP	\$17,300.00	\$12,700.00	\$10,000.00	\$14,800.00	\$11,300.00	\$9,900.00
IP-PUMP	\$120,500.00	\$117,900.00	\$122,200.00	\$115,500.00	\$95,300.00	\$116,900.00
HP-PUMP	\$1,083,900.00	\$1,039,400.00	\$1,098,900.00	\$1,022,700.00	\$1,005,900.00	\$1,038,800.00
LP-TURBINE	\$5,573,400.00	\$3,821,600.00	\$2,753,400.00	\$4,809,300.00	\$3,263,200.00	\$2,719,100.00
IP-TURBINE	\$3,868,300.00	\$3,647,700.00	\$4,105,300.00	\$3,283,800.00	\$3,009,100.00	\$3,491,400.00
HP-TURBINE	\$2,715,200.00	\$2,560,300.00	\$2,881,500.00	\$2,304,900.00	\$2,112,100.00	\$2,450,600.00
Condenser	\$5,406,300.00	\$3,123,600.00	\$1,998,800.00	\$4,459,100.00	\$2,579,100.00	\$2,186,000.00

Table 5-4. The component of variable cost.

Expenses	Cost
Variable Cost (VC)	VC = $\sum(1 \text{ to } 6)$
Utilities	
(1) Cooling water [341]	0.032 USD/1000 kg
(2) Feed water [342]	1 USD/1000 kg
(3) Waste water treatment [342]	0.53 USD/1000 kg
(4) MEA solvent price ^a [340]	1169.27 USD/1000 kg
(5) WGS catalyst ^{b,c} [343]	10.3 USD/kg
(6) CO ₂ transport and storage [341]	24.27 USD/1000 kg
^a The MEA consumption rate of 2 kg/tCO ₂ is assumed, as estimated by Elliott et al. [165].	
^b The WGS catalysts are to be changed over a period of 3 years [343].	
^c The GHSV for the HTWGS and the LTWGS reactors are 3000 and 4000 hr ⁻¹ respectively [343].	

5.2.3 Sensitivity and uncertainty analysis for the economic and LCA investigation

Table 5-5 provides a summary of the independent variables along with their respective lower and upper bounds. These parameters were chosen due to their significant levels of uncertainty.

Table 5-5. Chosen parameters for the sensitivity and uncertainty analysis.

Parameter	Low value	Nominal	High value	Unit
Economic assessment				
PEC GT	70	100	150	%
PEC CCS	70	100	150	%
Carbon T&S	10.22	24.27	39.60	USD/tonne
Discount rate	8	10	12	%
Tax rate	0	30	40	%
Tipping fee	80	100	120	%
Carbon credit	80	100	120	%

5.2.4 Life cycle assessment

The method for conducting the LCA can be seen in Sub Chapter 3.5 with the corresponding inventory presented in the Appendix 9.2.

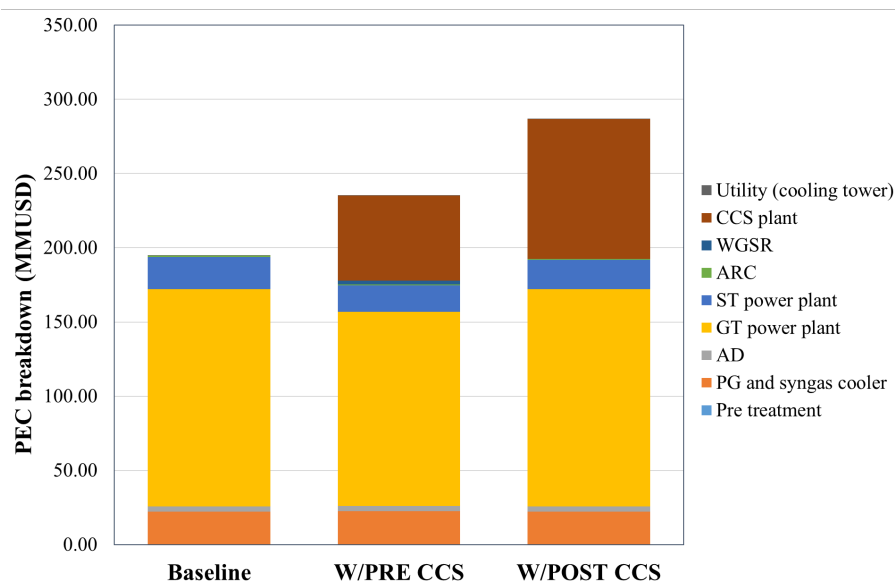
5.3 Result and discussion

5.3.1 Economic performance

5.3.1.1 LCOE, LCO_bM, and LCOWT

Figure 5-3a and Figure 5-3b illustrate the PEC breakdown for Scenarios 1 and 2, respectively. The total PEC for Scenario 1 ranges from 194 to 287 MMUSD, while for Scenario 2, it ranges from 176 to 254 MMUSD depending on the adopted case. From the figures, the GT power plant represents the most significant cost component across all cases, primarily due to the high price of the gas turbine. This is followed by the PG in the baseline cases of both scenarios. These findings align with previous studies by Aich et al. [214] and Tang et al. [213], which reported that the investment cost of a PG is lower than that of the power generation unit.

Meanwhile, in cases involving CCS plants, the CCS plant emerges as the second most significant cost component after the GT power plant. This observation is consistent with prior studies on the integrated gasification combined cycles with CO₂ capture [344], [345]. Furthermore, the cost of the CCS plant in the pre-CCS case is lower than in the post-CCS case, due to the smaller amount of feed gas processed. The PEC directly influences CAPEX, which is calculated based on the methodology outlined in Table 3-4. As observed in Table 5-6, Scenario 1 incurs a higher CAPEX than Scenario 2. Across all cases, the baseline configuration has the lowest CAPEX, followed by the pre-CCS case, while the post-CCS case exhibits the highest CAPEX.



(a)

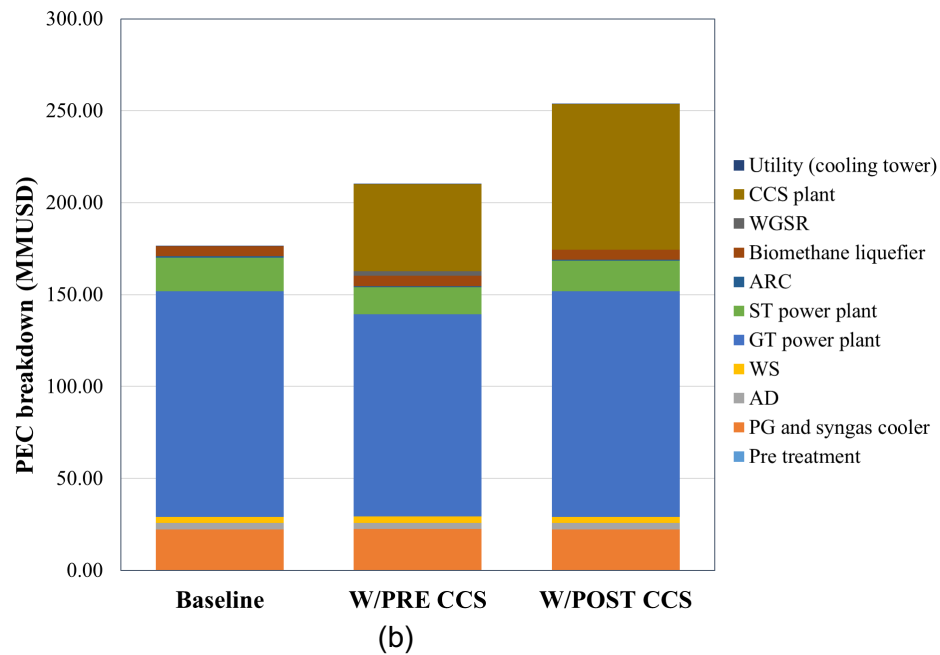


Figure 5-3. Contribution of each component to the total PEC for (a) Scenario 1 and (b) Scenario 2.

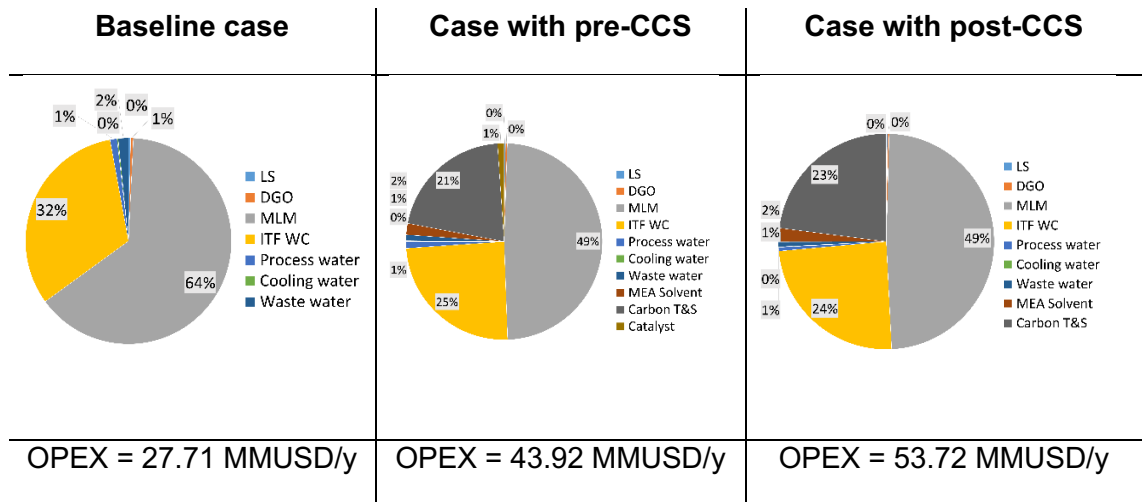
Figure 5-4a and Figure 5-4b present the OPEX breakdown for Scenarios 1 and 2, respectively, along with their estimated total OPEX. For Scenario 1, the total OPEX ranges from 27 to 54 MMUSD/year, whereas for Scenario 2, it ranges from 25 to 46 MMUSD/year depending on the adopted case. In both scenarios, the post-CCS case has the highest OPEX, while the baseline case has the lowest. The higher OPEX in CCS cases compared to the baseline case is primarily due to significant variable costs associated with carbon transport and storage (Carbon T&S).

Table 5-6. CAPEX and Annualised CAPEX (ACAPEX) for Scenario 1 and Scenario 2.

Parameters	Scenario 1			Scenario 2		
	Baseline	W/pre-CCS	W/post-CCS	Baseline	W/pre-CCS	W/post-CCS
CAPEX (MMUSD)	635.52	767.00	935.32	575.67	685.76	827.40
ACAPEX (MMUSD)	74.65	90.09	109.86	67.62	80.55	97.19

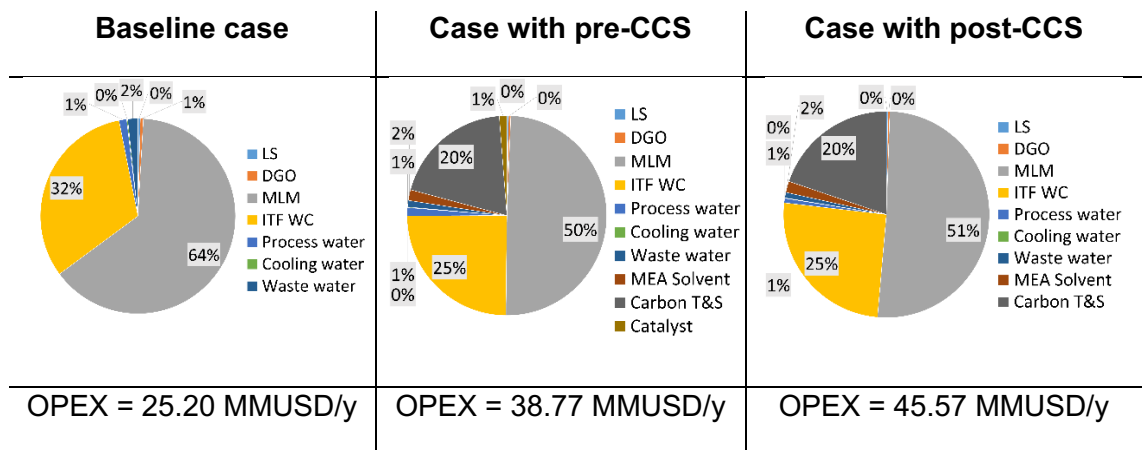
In terms of the OPEX distribution, "maintenance labour and materials" account for the largest percentage in both scenarios. In the baseline cases, the second-largest component is "insurance, tax, and financing WC", while in the CCS cases, the second-

largest cost is attributed to Carbon T&S. It is worth mentioning that OPEX excludes both feedstock costs, which are free, and feedstock transportation costs since the proposed system is located at the landfill site. Comparing the ACAPEX values from Table 5-6 with the OPEX indicates that annual CAPEX exceeds OPEX, which signifies that the financial structure of the current system is more CAPEX-driven.



Note: LS (Labour and supervision), DGO (Direct and general overhead), MLM (Maintenance labour and materials), ITF WC (Insurance, tax and financing WC)

(a)



Note: LS (Labour and supervision), DGO (Direct and general overhead), MLM (Maintenance labour and materials), ITF WC (Insurance, tax and financing WC)

(b)

Figure 5-4. OPEX breakdown for (a) Scenario 1 and (b) Scenario 2.

Figure 5-5 presents a comparison of the LCOE in Scenario 1 with values reported in the literature [266]. The reference study describes an integrated plasma gasification combined cycle (IPGCC) power plant comprising a PG, fabric filter, wet scrubber,

COS-hydrolysis unit, H₂S removal, sour water treatment systems, and a CCGT. The LCOE values used for comparison are from the baseline case in the reference study, where no incentives, such as waste treatment fees, are applied. This approach ensures a consistent basis for evaluation.

The LCOE reported in the literature remains significantly lower than those in all cases of Scenario 1, particularly in configurations involving CCS. This difference should not be interpreted solely as a disadvantage of the present system, but rather as a reflection of its broader process scope and functionality. The proposed system is inherently more capital-intensive because it is designed as a multi output WtE platform, integrating not only electricity generation but also cooling, heating, and waste treatment, while also incorporating carbon capture. In contrast, the reference study considers a more conventional electricity-focused system with fewer integrated process units.

The specific capital cost in the reference study is approximately 298,543 USD per tonne of MSW) processed per day [266], whereas in the current study it ranges from 387,705 to 570,625 USD/TPD MSW. This substantial difference can be attributed to several factors. Firstly, the size of the CCGT differs: the reference case reports a gross electricity output of 99.1 MW, while the current study's baseline, pre-CCS, and post-CCS cases produce 131.06 MW, 110.54 MW, and 117.64 MW, respectively. Figure 5-3a indicates that the CCGT is the largest contributor to PEC, thus its size significantly influences the specific capital cost.

Second, the current study adopts a more integrated process configuration, including additional units such as the ARC, AD, and WGSR, which are not present in the reference case. Third, the inclusion of CCS, particularly in the post-CCS case, adds substantial capital and operating requirements. As illustrated in Figure 5-3a, the CCS plant represents the second-largest PEC contributor after the CCGT. Therefore, the higher LCOE observed here reflects the cost of achieving greater process integration and deepen decarbonization rather than simply indicating poorer economic performance.

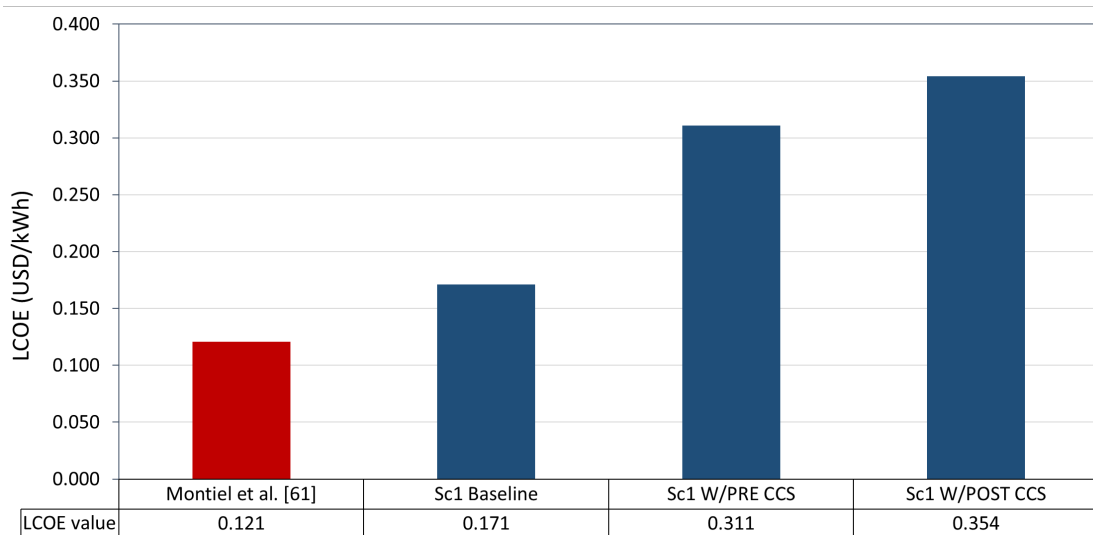


Figure 5-5. LCOE comparison for Scenario 1: baseline, pre-CCS, and post-CCS cases, with the literature value [266].

Figure 5-6 presents the LCOBM results for Scenario 2, comparing the current study's outcomes with the reference case from Capra et al. [40] and the LNG price in Indonesia [330]. The LCOBM values for all cases in Scenario 2 are more than four times higher than both reference benchmarks. This difference is primarily because Capra et al. [40] focus solely on converting biogas into liquid biomethane using a simpler process that includes AD, raw biogas pre-treatment, upgrading and polishing units, and liquefaction. In contrast, the current study integrates multiple complex technologies, such as PG, AD, CCGT, ARC, WS, ML, and CCS resulting in significantly higher CAPEX.

As shown in Figure 5-3b, the largest PEC contribution comes from the CCGT, with costs of 140.91, 124.65, and 139.14 million USD for the baseline, pre-CCS, and post-CCS cases, respectively. This is followed by the CCS plant, with PECs of 47.48 and 79.11 million USD for the pre-CCS and post-CCS cases, respectively. The PG also represents a substantial cost, with a PEC around 22 million USD. These significant equipment costs contribute to the higher LCOBM observed in the current study. However, this higher cost is associated with a system that delivers more than biomethane alone, as it simultaneously handles mixed waste, generates multiple useful outputs, and enables substantial CO₂ mitigation. Thus, the comparison with single output liquefied biomethane systems highlights the additional cost of integration and decarbonization, but also the broader value offered by the proposed configuration.

Furthermore, compared to Indonesian LNG production plants [330], the LCO_bM reported herein is higher than the current LNG market price. This discrepancy arises from the comparatively smaller scale of the system analyzed in this study. Since LNG plants are typically large-scale facilities, they benefit from economies of scale which significantly reduces their production costs. Nevertheless, unlike conventional LNG plants, the present system combines waste treatment, multi output energy recovery, and CCS integration, which provides environmental and functional benefits that are not captured in the direct market price comparison.

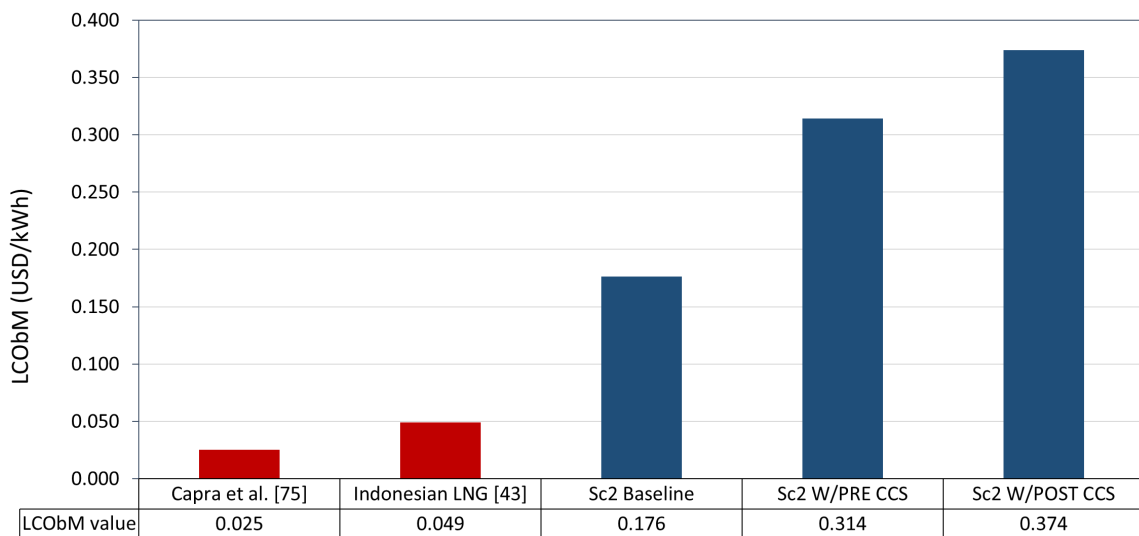


Figure 5-6. LCO_bM comparison for Scenario 2: baseline, pre-CCS, and post-CCS cases, with the literature value: literature 1 [40], [330].

The LCOWT is calculated by assuming the main product, co-products, and by-products sold using market price, as detailed in Eq. 5-3 and Eq. 5-4. Figure 5-7 compares LCOWT values across both scenarios with those from the literature [346], where the reference system utilized a bubble fluidized bed gasifier, syngas cleaner, and combustion engines. LCOWT values in all baseline cases are comparable to the reference. However, CCS cases show LCOWT values approximately double those of the baseline and reference cases, primarily due to the additional costs of integrating a CCS plant. Even so, this increase in cost should be viewed alongside the corresponding benefit of significantly lower direct CO₂ emissions, indicating that the CCS-integrated cases involve a clear economic-environmental trade-off.

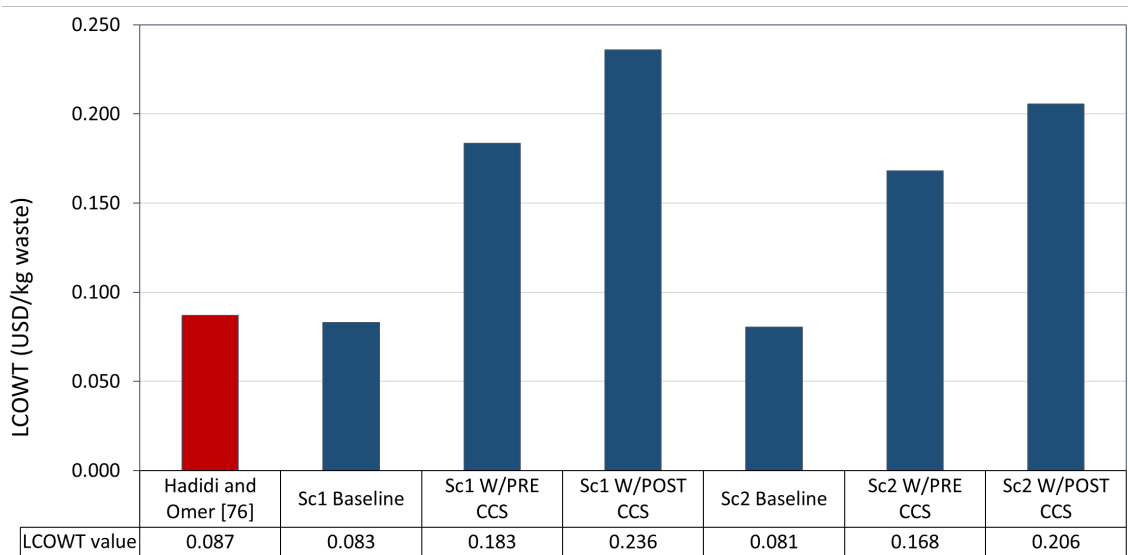


Figure 5-7. Comparison of the LCOWT for Scenarios 1 and 2 under different configurations: baseline, pre-CCS, and post-CCS, along with the literature value from [346].

The LCOWT in Scenario 1 is slightly higher than in Scenario 2, and this is primarily due to the significantly higher CAPEX in Scenario 1. The higher CAPEX observed in Scenario 1 is primarily attributed to the CCGT system's increased size and associated equipment costs. Scenario 1 employs a CCGT fueled by a mixture of syngas from plastic waste plasma gasification and biogas from food waste anaerobic digestion, necessitating a larger gas turbine to handle the greater fuel volume. Conversely, Scenario 2's gas turbine is fueled solely by syngas from plasma gasification, as biogas is converted into liquid biomethane separately, allowing for a smaller, less expensive CCGT system. This difference in fuel strategy directly impacts the size and cost of the CCGT, leading to the higher CAPEX in Scenario 1.

Additionally, Scenario 1 incurs to be slightly higher OPEX compared to Scenario 2. Revenue differences also contribute to the LCOWT variation between the two scenarios. In Scenario 1, the primary product is electricity, with an annual production ranging between 486 GWh and 686 GWh. In contrast, Scenario 2 primarily produces liquid biomethane, with an annual output of 346 GWh. Since the market price of electricity is approximately twice that of liquid biomethane, this pricing disparity further influences the difference in LCOWT between the two scenarios.

5.3.1.2 Comparison with Single-Output WtE Systems

To contextualize the techno-economic performance of the proposed multi-output waste-to-energy system, a comparison with single-output systems documented in the literature is presented in Table 5-7. The present study investigates two scenarios processing mixed plastic and food waste, producing multiple outputs including electricity, cooling, heating, and biomethane, with CAPEX ranging from approximately 576 to 935 million USD, LCOE between 0.172 and 0.354 USD/kWh, and LCOBM between 0.176 and 0.374 USD/kWh.

Anaya-Reza et al. [347] report lower CAPEX per tonne of waste for their baseline biomethane production system compared to the baseline biomethane scenario in this study, supporting that multi-output configurations typically incur higher capital costs relative to single-output systems. Their work further highlights that combined heat and power (CHP) systems, as non-single-output configurations, exhibit increased CAPEX, a trend also observed by Li et al. [348], where biomethane-only systems present lower CAPEX than CHP systems.

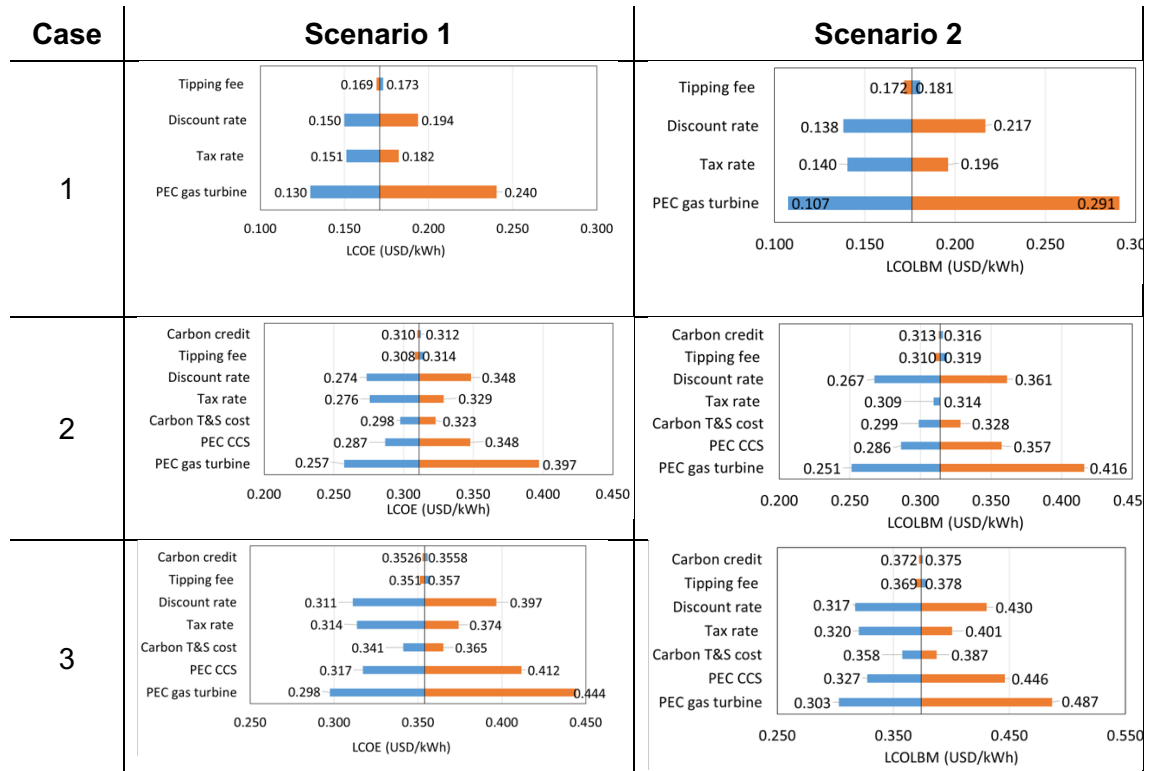
Similarly, single-output electricity generation from plastic waste, as investigated by Javed et al. [349] and Arena et al. [350], demonstrates lower CAPEX than the integrated multi-output system examined herein. This comparative analysis underscores that the greater process complexity in multi-output systems contributes to elevated capital requirements, reflecting the trade-off between enhanced functionality and increased investment. Although multi-output systems require greater capital investment than single output benchmarks, they offer wider technical, economic, and environmental benefits through diversified revenue streams, improved operational flexibility, integrated waste treatment, and the potential for deep decarbonization when combined with CCS.

Table 5-7. Comparison of the current multi-output generation system's performance against single-output generation systems reported in the literature.

Feedstock	Plant Capacity	Economic performance	Main component within the system	Reference
Mixed 5 kg/s of plastic & 13.97 kg/s of food waste (total waste = 2222.62 tonnes/day)	60.77-85.67 MW of electricity, cooling, and heating biomethane	CAPEX: 635.52-935.32 MMUSD LCOE: 0.172-0.3542 USD/kWh	PG, AD, CCGT, ARC, and CCS	Current study
	Electricity, cooling, heating, and 43.26 MW of biomethane	CAPEX: 575.67-827.40 MMUSD LCObM: 0.1762-0.3737 USD/kWh	PG, AD, CCGT, ARC, WS, ML, and CCS	Current study
200 tonnes/ day of OFMSW	30.41 Nm ³ Biomethane/ MT OFMSW	CAPEX: ~44.9 MM USD Biomethane price: 0.554 USD/Nm ³	AD and absorption column	[347]
	126,400 kWh of electricity/day	CAPEX: ~49.5 MMUSD Standard electricity price: 0.071 USD/kWh	AD, H ₂ S removal, steam generator, and steam turbine	[347]
30 tonnes of corn stover, 149 tonnes of tomato residues, 250 tonnes of dairy manure	744 GJ of biomethane	CAPEX: 522,799 USD Biomethane price: 0.54 USD/m ³	AD and HPWS	[348]
	61,995 kWh of electricity, and 314.1 GJ of heat	CAPEX: 602,981 USD Electricity selling price: 0.16 USD/kWh Heat energy selling price: 0.04 USD/kWh	AD and CHP engines	[348]
50 tonnes /day of plastic waste	1850 kWh of electricity /ton of plastic waste	Equipment cost: 1.5 MMUSD Electricity selling price: 0.142 USD/kWh	Gasifier, WGSR, gas purifier, CHP system	[349]
0.95 tonnes /h of end-belt refuse	4 MWe	Total plant cost: 4.79 k€/kWe	Fluidized bed gasifier, cyclone, syngas combustor, heat recovery steam generator, steam turbine	[350]

5.3.1.3 LCOE, LCObM, and LCOWT sensitivity analysis

Figure 5-8 presents the results of the sensitivity analysis on the LCOE and LCObM, while Figure 5-9 illustrates the sensitivity analysis on the LCOWT. In all cases, the PEC of GT emerges as the next most influential parameter. This significant impact is attributed to the fact that GT is the most expensive component in each case across both scenarios.

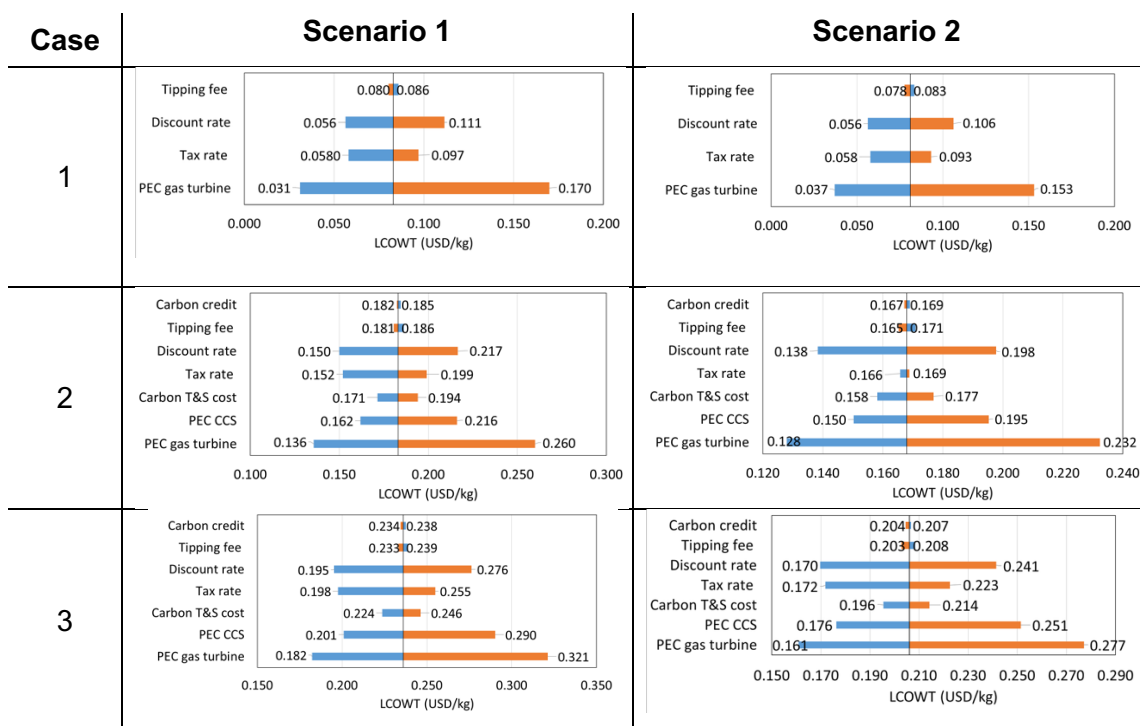


Note: case 1: baseline case; case 2: with pre-CCS case; case 3: with post-CCS case

Figure 5-8. Sensitivity of the LCOE and LCObM to various economic variables.

Since most plastics are derived from fossil fuels, capturing fossil-based carbon emissions during the utilization of syngas from the plasma gasification of plastic waste is essential. To achieve net-zero emissions, biogenic carbon from the biogas also needs to be captured. Consequently, the costs of implementing CCS technology have a significant impact on LCOE, LCObM, and LCOWT. As shown in Figure 5-8 and Figure 5-9, the PEC of CCS gives impact on the LCOE, LCObM and LCOWT. The future cost of CCS equipment could decrease due to technological advancements and economies of scale, as highlighted in [243]. Both Figure 5-8 and Figure 5-9 also demonstrate that the effect of Carbon T&S costs is relatively small. The adoption of T&S facilities depends more on the social factors and public

acceptance rather than economic considerations [341]. Among common financial factors, the discount rate shows a broader fluctuation range compared to the tax rate. Small changes in the discount rate can lead to substantial variations in the LCOE, LCOBM, and LCOWT. While tax allowances can improve the project viability, their impact is limited since even under an optimistic tax-free scenario, the LCOE, LCOBM, and LCOWT would only decrease by less than 17%. Policy-driven incentives, namely the waste tipping fee and carbon credit price, have only a minimal effect on project unit costs.



Note: case 1: baseline case; case 2: with pre-CCS case; case 3: with post-CCS case

Figure 5-9. Sensitivity of the LCOWT to various economic variables.

5.3.1.4 Feedstock composition sensitivity analysis

In municipal refuse, the proportion of plastics and food waste can vary widely; to quantify its impact on system economics, three PW:FW blends (24:76, 50:50 and 90:10) were simulated in Aspen Plus. For each case, all unit operations were resized and mass-/energy-balance outputs compiled (Table 5-8), then subjected to discounted-cash-flow analysis to yield the LCOE in Scenario 1, the LCOBM in Scenario 2, and the LCOWT in both scenarios.

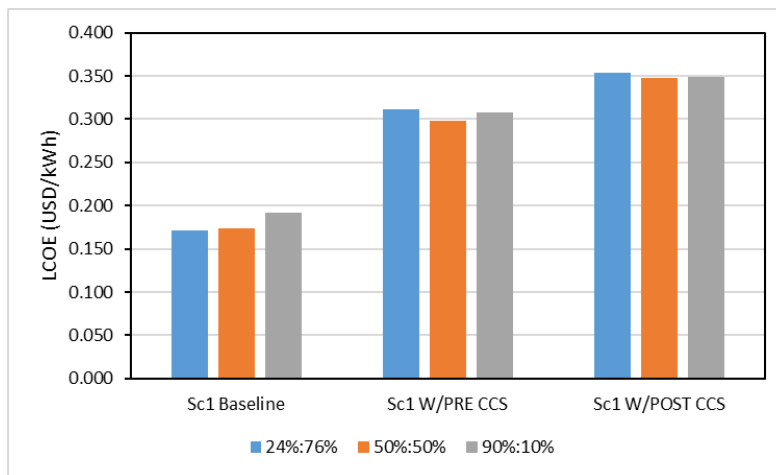
Table 5-8. The output of the system under different proportion.

Output (MW)	Baseline			W/PRE-CCS			W/POST-CCS		
	24:76	50:50	90:10	24:76	50:50	90:10	24:76	50:50	90:10
Scenario 1									
Electricity	85.67	133.79	213.94	60.77	93.76	150.1	64.65	100.47	162.17
Cooling	0.34	0.56	0.95	0.29	0.48	0.81	0.34	0.56	0.95
Heating	1.62	2.69	4.51	2.83	4.65	7.82	3.28	2.72	9.11
Scenario 2									
Electricity	53.08	111.7	209.55	29.76	73.9	145.72	37.86	81.54	154.33
Cooling	0.27	0.52	0.94	0.23	0.44	0.8	0.27	0.52	0.94
Heating	1.49	2.61	4.5	2.7	4.57	7.81	3.13	2.61	9.06
Liquid methane	43.26	29.26	5.81	43.26	29.26	5.81	43.26	29.26	5.81

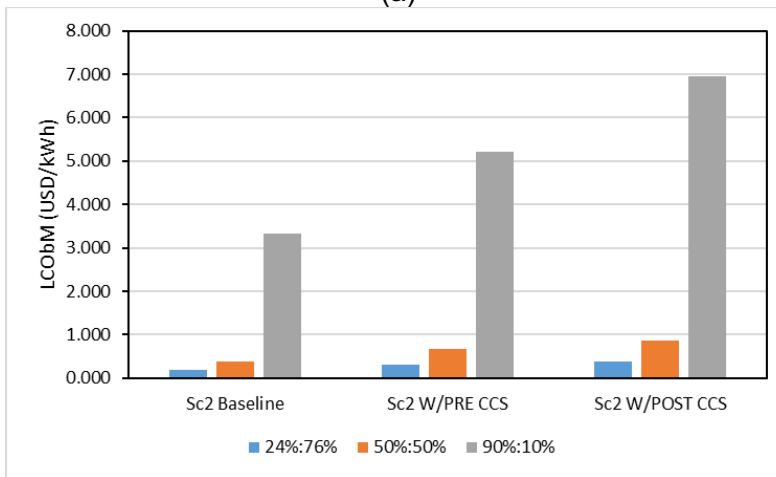
As shown in Table 5-8, increasing the plastic-to-food waste ratio from 24:76 to 90:10 results in a marked shift in the main outputs. Electricity production in both Scenario 1 and Scenario 2 rises with higher plastic content, owing to the greater calorific density of plastics, which yields a more energetic fuel gas stream. Conversely, liquid biomethane output in Scenario 2 declines substantially as the proportion of biodegradable food waste is reduced, diminishing the biogas volume available for upgrading. As a result, the gap between electricity and biomethane outputs widens with more plastic content, highlighting how feedstock composition determines whether the system favours power generation or biomethane production.

Figure 5-10 presents the corresponding levelized costs. Total CAPEX rises with plastic content due to a larger plasma gasifier plus upsized CCGT and CCS units; nevertheless, LCOE in Scenario 1 remains relatively insensitive. This stability occurs because the higher plastic content produces a more energy-dense syngas stream, boosting electricity output and offsetting the added capital costs. In Scenario 2, by contrast, LCO_{BM} climbs sharply between the 50:50 and 90:10 blends. Although CAPEX for biomethane-specific equipment (AD, WS, and liquefaction) falls with reduced biogas flow, the higher CAPEX for processing the plastic-rich feed must still be recovered over a shrinking methane volume. Note that electricity revenue in Scenario 2 is assumed at prevailing market prices rather than

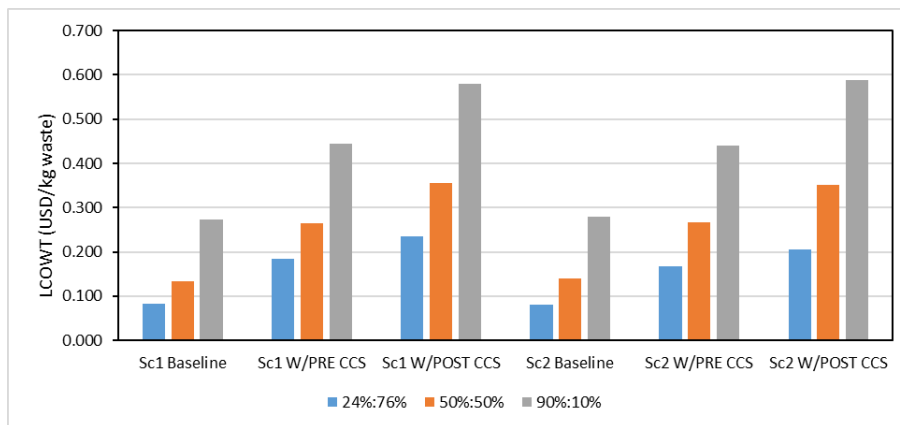
the levelized cost derived from DCFA, as in Scenario 1.



(a)



(b)



(c)

Figure 5-10. Sensitivity of levelized costs to feedstock composition. (a) LCOE in Scenario 1 (CCHP), (b) LCOBM in Scenario 2 (CCHP + biomethane), and (c) LCOWT as the plastic-to-food waste ratio varies (24:76, 50:50, 90:10) under baseline, pre-CCS, and post-CCS configurations.

Figure 5-10c shows how LCOWT varies with feedstock composition. Here, LCOWT is calculated by assuming that all outputs (electricity, cooling, heating, and biomethane) are sold at market prices. As the plastic fraction increases, LCOWT rises because plastic-rich feeds require more capital and energy for gasification (larger plasma gasifier, CCGT, and CCS units), and these extra costs outweigh the modest savings from treating less food waste. Overall, these results indicate that feedstock composition plays an important role in determining the preferred system pathway: plastic-rich municipal waste streams favour Scenario 1 due to their stronger potential for electricity generation, while food waste rich streams favour Scenario 2 because they enhance biomethane yield through anaerobic digestion.

5.3.1.5 LCOE, LCObM, and LCOWT uncertainty analysis

An uncertainty analysis is conducted to evaluate the effect of simultaneous variations in all the uncertainty parameters and to quantitatively determine the range of variability in the economic criteria. A Monte Carlo simulation, developed in the MATLAB environment, applies the economic model to assess the impact of the variables listed in Table 5-5 on the LCOE, LCObM and LCOWT. Each of these three costs is recalculated across 10,000 trials, which produces a probability distribution curve. From these curves, essential statistical metrics such as the mean, median, and standard deviation are determined. Furthermore, the probability range of the LCOE, LCObM and LCOWT are defined using the 95% confidence interval (CI) statistical metric.

Table 5-9 presents the results of the Monte Carlo simulation used to estimate the LCOE, LCObM and LCOWT. The mean values are higher than those estimated in the deterministic economic assessment, reflecting the impact of CAPEX uncertainty. As shown in Table 5-9, the LCOE has less variance than the LCObM. This means that the biomethane price may be more sensitive to economic uncertainties. In addition, Scenario 1 consistently has a higher LCOWT than Scenario 2 across all cases, suggesting that electricity-focused production incurs higher waste treatment costs than biomethane-focused production. This is reasonable, as the PEC of CCGT is the largest cost component, meaning the higher electricity production capacity in Scenario 1 results in greater CAPEX which ultimately raises waste treatment costs compared to Scenario 2.

Table 5-9. Monte Carlo simulation results for estimating the LCOE, LCObM, and LCOWT.

Scenario	Case	Metric	Mean	Median	STD	95% CI	Units
1	1	LCOE	0.1761	0.1749	0.0249	0.127 to 0.225	USD/kWh
	2	LCOE	0.3195	0.3184	0.0366	0.248 to 0.391	of
	3	LCOE	0.3647	0.3635	0.0417	0.283 to 0.446	electricity
2	1	LCObM	0.1839	0.1819	0.0419	0.102 to 0.266	USD/kWh
	2	LCObM	0.3323	0.3306	0.0416	0.251 to 0.414	of liquid
	3	LCObM	0.3864	0.3850	0.0530	0.285 to 0.490	biomethane
1	1	LCOWT	0.0893	0.0878	0.0312	0.028 to 0.150	USD/kg of waste
	2	LCOWT	0.1910	0.1900	0.0325	0.127 to 0.255	
	3	LCOWT	0.2459	0.2448	0.0395	0.168 to 0.323	
2	1	LCOWT	0.0854	0.0841	0.0265	0.033 to 0.137	USD/kg of waste
	2	LCOWT	0.1797	0.1786	0.0263	0.128 to 0.231	
	3	LCOWT	0.2151	0.2143	0.0333	0.150 to 0.280	

Note: case 1: baseline case; case 2: with pre-CCS case; case 3: with post-CCS case

5.3.1.6 Economies of scale

The economies of scale for the proposed plant are assessed by varying the plant size between 100 MW_{th} to 750 MW_{th} of feedstock on the LCOE and LCObM. CAPEX is adjusted using component-specific scaling factors (*f*), as outlined in Table 5-2. Fixed OPEX components are scaled proportionally to CAPEX. Although labor costs are calculated based on the number of system components rather than plant capacity, they are assumed to scale with plant capacity to the power of 0.242 to account for economies of scale, following the labor cost correlation proposed by Peters et al. [257].

Figure 5-11a and Figure 5-11b illustrate the relationship between feedstock consumption and the corresponding LCOE and LCObM, respectively. Both figures demonstrate that as the feedstock consumption increases, the associated LCOE and LCObM decrease. In Scenario 1, for feedstock input levels between 100 MW_{th} and 200 MW_{th}, the LCOE declines significantly. Specifically, the LCOE decreases from 0.24 to 0.18 USD/kWh for Case 1, from 0.42 to 0.33 USD/kWh for Case 2, and from 0.48 to 0.37 USD/kWh for Case 3. Beyond 250 MW_{th}, the rate of decline slows considerably, with values reaching 0.11 USD/kWh for Case 1, 0.21 USD/kWh for Case 2, and 0.24 USD/kWh for Case 3 at a feedstock consumption level of 750 MW_{th}.

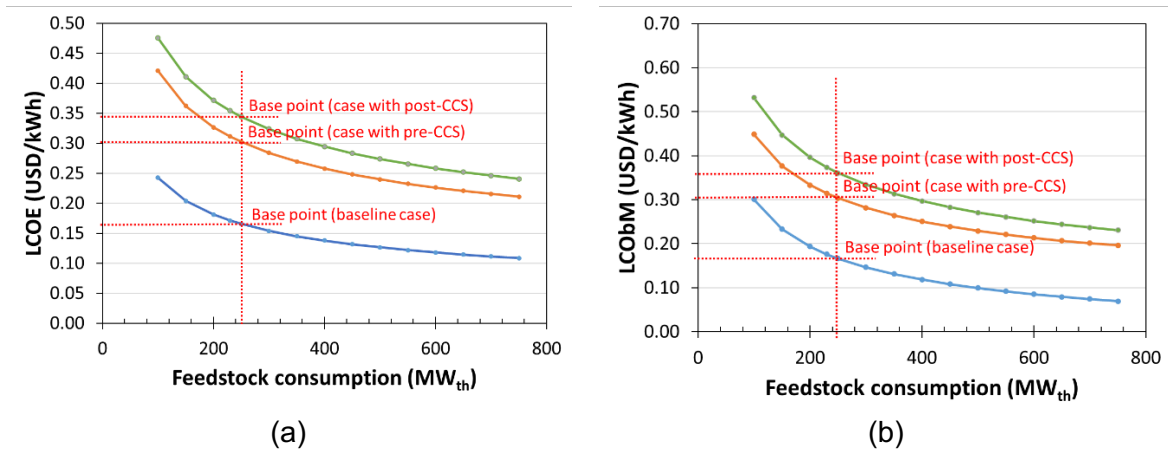


Figure 5-11. Economies of scale for the investigated scenarios: (a) Scenario 1 and (b) Scenario 2.

In Scenario 2, a similar trend is observed. For feedstock input levels between 100 MW_{th} and 200 MW_{th}, the LCOBM decreases sharply, from 0.30 to 0.19 USD/kWh for Case 1, from 0.45 to 0.33 USD/kWh for Case 2, and from 0.53 to 0.40 USD/kWh for Case 3. Beyond 250 MW_{th}, the reduction rate slows significantly, with values stabilizing at 0.07 USD/kWh for Case 1, 0.20 USD/kWh for Case 2, and 0.23 USD/kWh for Case 3 at 750 MW_{th}.

While economies of scale provide initial cost reductions, further scaling beyond 600 MW_{th} of feedstock consumption yields diminishing economic benefits, as indicated by the flattening of the curves.

5.3.2 Environmental performance

The Life Cycle Inventory (LCI) data were processed using the SimaPro software to quantify the contributions to various midpoint impact categories. The LCA is performed by evaluating the input and output flows within the system boundary illustrated in Figure 5-1 and Figure 5-2, without modelling the internal processes. In this section, global warming potential and water footprint are discussed as the key midpoint indicators. Additional impact indicators are available in the Sub Chapter 9.2.2. All impact category results are normalized to one kilogram of waste processed within the system.

5.3.2.1 Global warming potential (GWP)

Figure 5-12a and Figure 5-12b present the GWP impact breakdown for the different

process stages in Scenario 1 and Scenario 2, respectively. As shown in the figures, the baseline case in both scenarios exhibits the highest GWP, with Scenario 1 reaching approximately 0.662 kgCO₂-eq/kg waste and Scenario 2 reaching 0.746 kgCO₂-eq/kg waste. The slightly lower GWP in the baseline case of Scenario 1 compared to Scenario 2 results from the off-gas emitted from the water scrubber, which is used for upgrading biogas by removing CO₂ and H₂S [291]. The stripper of the water scrubber releases off-gas composed of CO₂ (17.9%), CH₄ (0.43%), NH₃ (1%), H₂S (0.16%), O₂ (14.8%), and N₂ (64%), with all carbon emissions being biogenic. While biogenic CO₂ emissions are excluded from the GWP calculation as carbon-neutral, biogenic methane emissions contribute significantly to the GWP. According to the ReCiPe 2016 method [351] and supported by the SimaPro database, biogenic methane is assigned a global warming potential (GWP) of approximately 33.25 kg CO₂-eq/kg CH₄. Consequently, the baseline case in Scenario 2 exhibits a slightly higher GWP compared to Scenario 1 due to the biogenic methane released from the water scrubber.

Compared to the literature, the GWP in the baseline case is higher than the GWP reported by Materazzi et al. [352] for a waste-to-heat-and-power system based on incineration without CCS, where the GWP was approximately 0.425 kg CO₂-eq/kg waste. The difference should be interpreted by considering two separate factors. First, the technology choice differs: the reference study uses a conventional incineration-based system, whereas the present study employs a more integrated gasification-anaerobic digestion-based configuration with combined power generation and additional process units. These different conversion pathways lead to different emission profiles and auxiliary energy demands. Second, the feedstock carbon composition also differs significantly. The reference study used a feedstock containing approximately 64% biogenic carbon, whereas the present study contains only around 28% biogenic carbon. Since biogenic CO₂ is treated as carbon-neutral in the GWP calculation, a higher biogenic fraction naturally lowers the reported GWP. Therefore, the higher GWP observed in the present baseline cases is not driven solely by the core technology, but also by the lower proportion of biogenic carbon in the waste feedstock.

For cases incorporating CCS, the pre-CCS configuration resulted in a higher GWP

than the post-CCS configuration in both scenarios. Specifically, the GWP values for the pre-CCS case are 0.004 kgCO₂-eq/kg waste and 0.222 kgCO₂-eq/kg waste in Scenarios 1 and 2, respectively. In contrast, the post-CCS case achieved lower GWP values of -0.191 kgCO₂-eq/kg waste for Scenario 1 and 0.123 kgCO₂-eq/kg waste for Scenario 2. The higher GWP in the pre-CCS case compared to the post-CCS case is primarily due to the direct release of flue gas emissions from the combustion process in the CCGT, as these emissions are not treated or captured. Conversely, in the post-CCS case, a significant fraction of CO₂ emissions is captured, reducing the overall GWP.

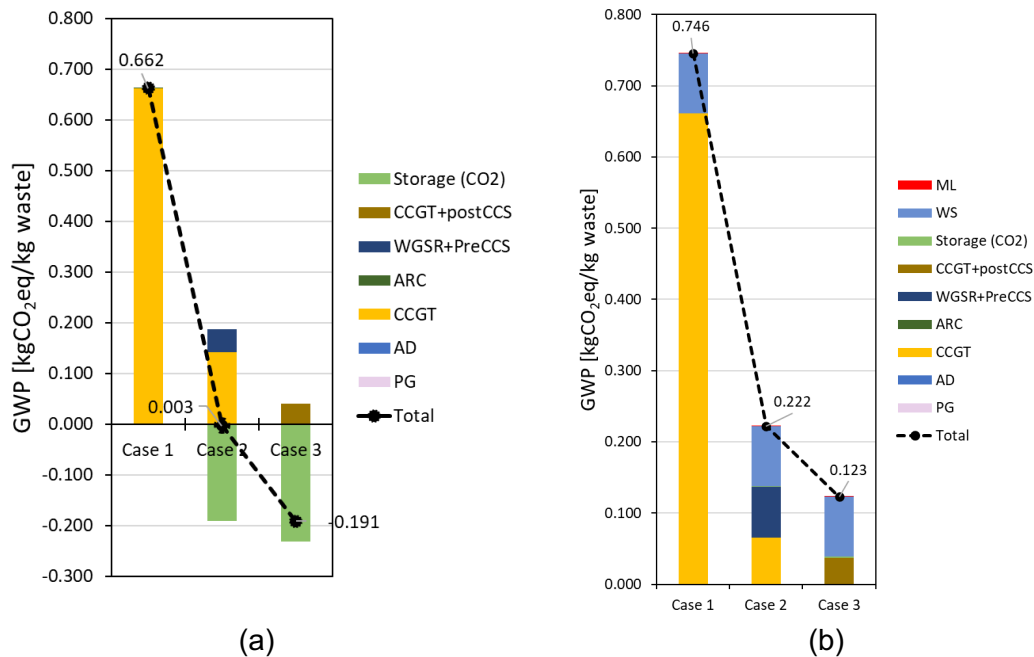


Figure 5-12. The breakdown of the GWP under (a) Scenario 1 and (b) Scenario 2 (**Note:** case 1: baseline case; case 2: with pre-CCS case; case 3: with post-CCS case).

Additionally, the lower GWP in both the pre-CCS and post-CCS cases of Scenario 1 is attributed to the presence of biogenic carbon in the captured CO₂, whereas in Scenario 2, the captured CO₂ is entirely of fossil origin. A previous study [352] reported a lower GWP of -0.416 kgCO₂-eq/kg waste in the incinerator integrated with the CCS, which is lower than the values obtained in this study. This discrepancy is primarily due to a higher proportion of biogenic carbon sequestration in the reference study. Again, this difference can be understood through the combined influence of technology and feedstock composition. From technology

perspective, the reference system is based on incineration with CCS, whereas the present study uses more complex gasification-AD-CCS pathways, which involve different process emissions and energy requirements. From the feedstock perspective, the lower GWP in reference study is strongly influenced by the higher proportion of biogenic carbon available for sequestration. Hence, the more negative GWP reported by Materazzi et al. [352] is mainly associated with greater biogenic carbon capture, rather than reflecting the effects of CCS technology alone.

As shown in Figure 5-12, the CCGT operation is the primary contributor to GWP in all cases due to the release of fossil and biogenic carbon emissions in Scenario 1 and fossil carbon emissions in Scenario 2. The fossil carbon emissions originate from plastic waste, while biogenic carbon emissions stem from food waste. In Figure 5-12a, the implementation of pre-CCS reduces CCGT emissions from 0.661 kgCO₂-eq/kg waste to 0.141 kgCO₂-eq/kg waste, while post-CCS integration further decreases emissions to -0.0392 kgCO₂-eq/kg waste. Similarly, Figure 5-12b, which illustrates the GWP breakdown for Scenario 2, confirms that the operation of CCGT plant remains the dominant emission source. With pre-CCS, CCGT emissions decline from 0.661 kgCO₂-eq/kg waste to 0.066 kgCO₂-eq/kg waste, and with post-CCS, emissions are further reduced to 0.0369 kgCO₂-eq/kg waste.

5.3.2.2 Water footprint (WF)

Figure 5-13a and Figure 5-13b illustrate the WF in Scenario 1 and Scenario 2, respectively. Among all cases, the pre-CCS case exhibits the highest WF, reaching approximately 0.0014 m³/kg waste in Scenario 1 and 0.0013 m³/kg waste in Scenario 2. This is primarily due to the substantial amount of water required for steam generation to support the operation of WGSR. Meanwhile, the post-CCS case exhibits the lowest WF, with values of 0.0010 m³/kg waste for both scenarios. Compared with literature reports on incineration-CCS integration for combined heat and power [352], where WF values of up to 0.0234 m³/kg waste have been observed, our results are markedly lower, likely because on-site wastewater treatment has been incorporated into the present study. Biogas production from energy crops (e.g., maize, sorghum, wheat) typically entails even higher WFs, ranging from 0.025 to 0.217 m³/kg of crop [353]. This disparity is attributed primarily to differences in feedstock moisture content: low-moisture energy crops demand

substantially more water for anaerobic digestion than high-moisture food waste.

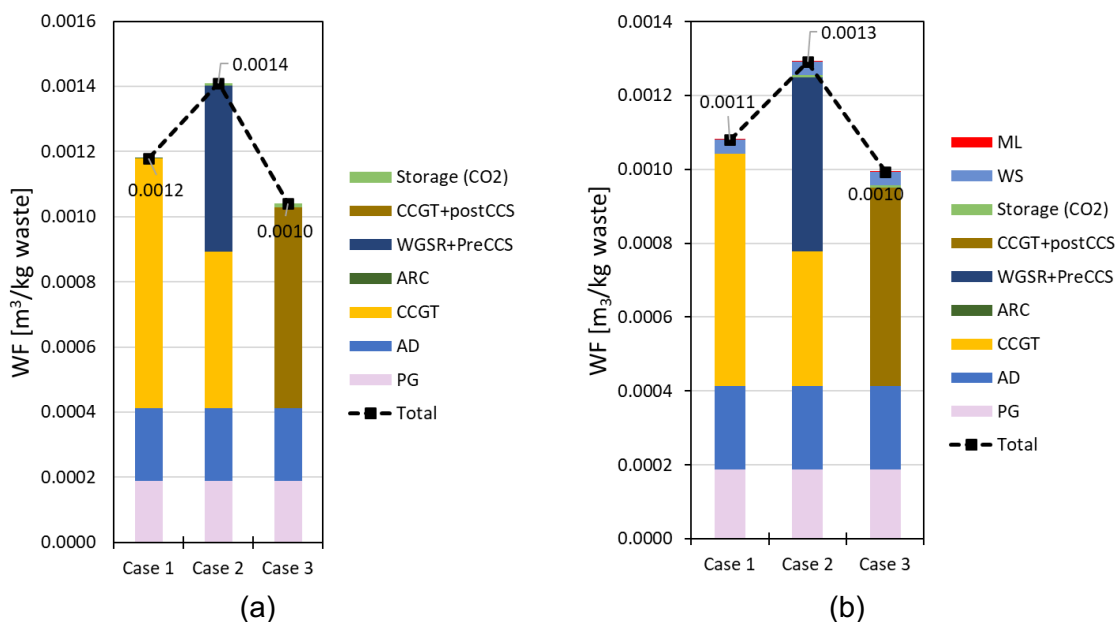


Figure 5-13. The breakdown of the WF under (a) Scenario 1 and (b) Scenario 2 (**Note:** case 1: baseline case; case 2: with pre-CCS case; case 3: with post-CCS case).

In both the baseline and post-CCS configurations, the CCGT constitutes the predominant contributor to the overall WF, accounting for over 50% of the total in each case. The baseline WF is lower than that of the post-CCS configuration owing to the absence of low-pressure (LP) steam diversion for stripper heating; consequently, all LP steam must be condensed, thereby requiring greater cooling-water consumption.

By contrast, in the pre-CCS configuration, the water footprint is dominated by the water required for steam generation in the WGSR and pre-CCS units (denoted “WGSR + Pre-CCS” in the diagram), which together represent approximately 36% of total WF in both scenarios. The CCGT contributes the next largest share, approximately 33 % in Scenario 1 and 28 % in Scenario 2. This reduced share reflects lower cooling-water demand for LP steam condensation, since 10% of syngas is diverted to produce steam for the WGSR rather than for power generation, resulting in reduced fuel use and cooler condenser loads compared to the baseline and post-CCS cases.

5.3.2.3 Policy incentives analysis

This section examines the impact of waste utilization incentives and carbon credit pricing on the economic feasibility of the proposed waste-to-energy system. Figure 5-14 and Figure 5-15 illustrates the effect of employing potential policy schemes on the LCOE and LCObM, respectively. As shown, increasing energy incentives and carbon credit prices leads to a reduction in LCOE and LCObM, though the impact varies between the two configurations. The post-CCS case exhibits a higher LCOE and LCObM range than the pre-CCS due to the higher CAPEX and OPEX associated with the larger size of the CCGT and the additional costs of post-combustion CO₂ capture and processing.

Indonesia has established a carbon credit price of 8.9 USD/tonne CO₂ [328] and provides financial incentives for waste utilization through tipping fees, which vary by municipality. In the city where the proposed system is located, the tipping fee reaches 14.02 USD/tonne of waste [327]. Despite the application of both carbon credits and tipping fees, the reduction in LCOE is only 7.95% for the pre-CCS case and 6.81% for the post-CCS case. Similarly, the LCObM reduction is limited to 8.60% and 7.65% for pre-CCS and post-CCS configurations, respectively. The Indonesian government has set a maximum electricity purchase price for waste-to-energy plants with a capacity exceeding 10 MW at 92.9 USD/MWh for the first 10 years, followed by 59.44 USD/MWh from years 11 to 25 [354]. Given that the LCOE of the current system far exceeds these limits, the project is not financially viable under existing policies.

To assess the potential impact of higher financial incentives, the carbon credit price and tipping fee values from the United States (U.S.) were applied as a benchmark. It is recognized that economic conditions in Indonesia differ markedly from those in the U.S., however, U.S. market data were selected as conservative, high-integrity proxies for two reasons. First, the U.S. carbon-credit market, both compliance and voluntary, is among the most developed globally. Robust regulatory frameworks for greenhouse-gas reductions have been established by state and regional cap-and-trade programs (e.g., California Air Resources Board, Oregon Clean Fuels Program, and the Regional Greenhouse Gas Initiative), and rigorous certification and registry systems for offset projects worldwide are maintained by leading

voluntary standards bodies (Verra, Gold Standard, and the American Carbon Registry) [355]. These combined mechanisms ensure transparent pricing, third-party verification, and publicly accessible transaction data, justifying the adoption of the U.S. carbon credit price of USD 85.00/tonne CO₂ as an upper-bound benchmark [356].

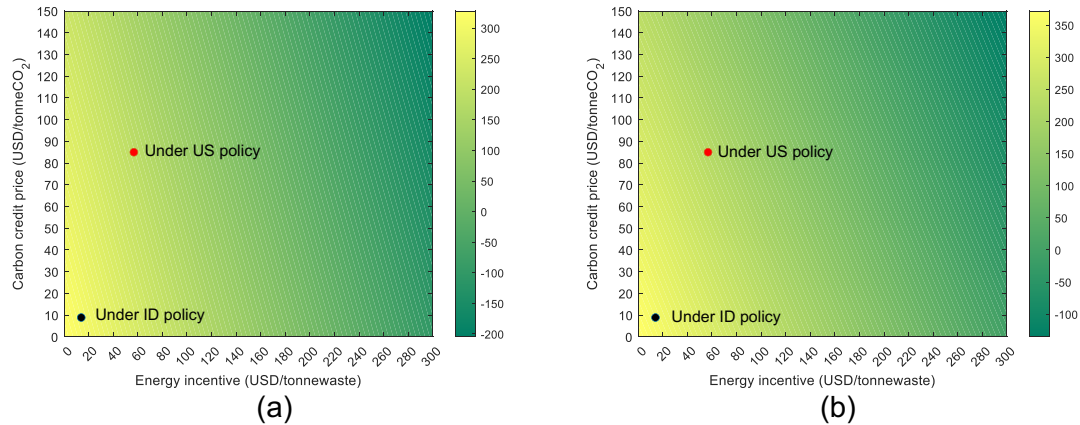


Figure 5-14. Impact of carbon credit price and energy incentive on LCOE for different CCS configurations (a) pre-CCS and (b) post-CCS.

Second, the Environmental Research & Education Foundation’s May 2024 survey of U.S. municipal landfills has been adopted, reporting an unweighted national average tipping fee of USD 56.80 per tonne [357]. This fee is understood to fully internalize capital, operating, and long-term environmental compliance costs, and is therefore presented as a “best-practice” benchmark for cost-recovery assumptions. Under these U.S. benchmark conditions, the reductions in LCOE and LCOBM due to the application of both carbon credits and tipping fees are substantial, ranging from 39% to 47%, demonstrating the potential for enhanced financial viability with stronger policy support.

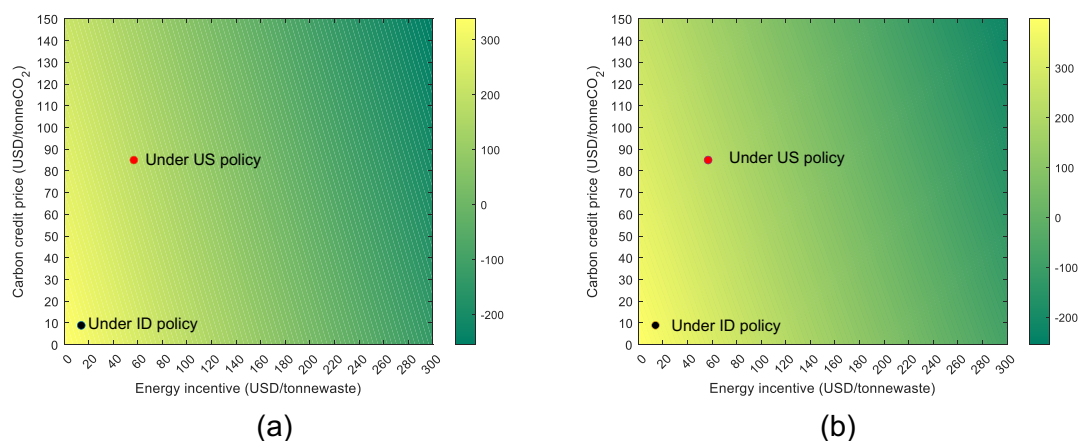


Figure 5-15. Impact of carbon credit price and energy incentive on LCOBm for different CCS configurations (a) pre-CCS and (b) post-CCS.

5.4 Conclusion

This study presents a novel waste-to-energy system integrating plastic waste plasma gasification and food waste anaerobic digestion for the production of CCHP, and liquid biomethane. While plasma gasification has primarily been applied for hydrogen production, its integration with CCS remains underexplored, particularly for plastic waste feedstocks and hybrid waste conversion processes such as anaerobic digestion. This study provides a cradle-to-gate assessment of multiple energy outputs, incorporating heat integration strategies to holistically evaluate economic and environmental performance. Several configurations were evaluated, and the key findings are summarized as follows:

1. In Scenario 1, the LCOE spans from 0.171 to 0.354 USD/kWh, while in Scenario 2, the LCOBm ranges between 0.176 and 0.374 USD/kWh. In both scenarios, the baseline case achieves the lowest levelized cost, whereas the case with post-CCS exhibits the highest levelized cost. This outcome reflects that the integration of a CCS plant increases both LCOE and LCOBm.
2. In both scenarios, LCOWT falls between 0.083 to 0.236 USD/kg waste. The LCOWT in Scenario 1 is slightly higher than in Scenario 2, primarily due to the significantly higher CAPEX in Scenario 1.
3. Based on the sensitivity analysis, PEC of gas turbine power plant has the greatest impact on LCOE, LCOBm, and LCOWT. Lowering this cost yields the largest reductions across all three metrics, highlighting its critical role in

the system's economic feasibility.

4. Monte Carlo simulations reveal greater variability in LCO_bM compared to LCOE, with wider 95% confidence intervals for LCO_bM which indicates that biomethane prices are more sensitive to economic uncertainties.
5. Scenario 1 exhibits GWP values between -0.191 and 0.662 kgCO₂-eq/kg waste, while Scenario 2 ranges from 0.123 to 0.746 kgCO₂-eq/kg waste, with post-CCS configurations yielding the lowest GWP, and the operation of CCGT being the largest GWP contributor across all cases due to the carbon emission.
6. Both scenarios exhibit WF values ranging from 0.0010 to 0.0014 m³/kg waste, with the CCGT as the primary contributor due to its high cooling water demand for LP steam condensation.
7. By applying Indonesia's policy incentives, the LCOE and LCO_bM for the system with a CCS plant are reduced by less than 12% which reflects a limited financial impact on improving the system's economic feasibility.

This research offers new insights into the design and performance of hybrid, low-emission waste-to-energy systems which illustrates the technical feasibility and trade-offs involved in coupling plasma gasification, anaerobic digestion, and CCS. The novel system design contributes to advancing the state-of-the-art in sustainable waste management and supports decision-making for both technology developers and policymakers. Future research can focus on developing a pilot-scale demonstration to validate the system's technical performance, economic feasibility, and environmental impact under real-world conditions.

6 Hydrogen and CCHP Generation from Waste via Combined Plasma Gasification and Anaerobic Digestion with CCS: A Techno-Economic and Life Cycle Assessment

Abstract

Hydrogen has been increasingly recognized as a promising clean energy carrier to support global climate change mitigation efforts. Among various production routes, the utilization of municipal solid waste (MSW), particularly plastic and food waste, offers a dual benefit: reducing the environmental burden of high-volume waste streams and producing a low-carbon energy carrier. Despite growing interest in waste-to-hydrogen technologies, few studies have explored an integrated pathway that simultaneously utilizes both plastic and food waste while incorporating carbon mitigation strategies.

This study proposes and assesses, for the first time, a novel MSW-to-hydrogen and combined cooling, heating and power (CCHP) pathway fitted with carbon capture and storage (CCS). Two CCS configurations are examined: a pre-combustion CCS scheme (Scenario 1) and a post-combustion CCS scheme (Scenario 2). Three process cases are designed and evaluated using techno-economic analysis (TEA) and life cycle assessment (LCA) under conditions representative of Indonesia. In Case 1, plastic and food waste are separately processed via plasma gasification and anaerobic digestion, with syngas and biogas subsequently reformed into hydrogen. Case 2 co-processes the waste streams through plasma gasification. Case 3 builds on Case 1 by adding steam methane reforming (SMR) of biogas prior to mixing.

For a feedstock of 13.97 kg/s of food waste and 5.0 kg/s of plastic waste, the systems deliver 0.27–0.70 kg/s of hydrogen across all cases and scenarios, with energy and exergy efficiencies of 19–42% and 13–34%, respectively. The levelized cost of hydrogen (LCOH) is estimated at 3.52–13.37 USD/kg, while LCA reveals global warming potentials (GWP) ranging from –0.159 to 0.048 kgCO₂-eq/kg of waste treated. These results demonstrate the technical feasibility and environmental promise of integrating plasma gasification, anaerobic digestion, and

CCS for hydrogen production. By valorizing municipal waste while achieving carbon mitigation, this novel multi-integration of plasma gasification, anaerobic digestion, PSA, CCS, CCGT, and ARC represents a compelling and scalable solution for sustainable hydrogen deployment in developing economies.

Keywords: Hydrogen production, Plastic waste, Food waste, Plasma gasification, Anaerobic digestion, CCS, TEA and LCA.

6.1 Introduction

Decarbonising in the energy sector is essential to meet global climate targets, as outlined in the Paris Agreement, while sustainable waste management remains a pressing challenge due to the rapid growth of MSW and its associated environmental impacts [324], [358], [359], [360]. Among different waste streams, plastics and food waste are particularly significant due to their abundance and environmental consequences. Plastics are resistant to degradation and contribute to long-term pollution, whereas food waste decomposes into methane, a greenhouse gas with a significantly higher GWP than CO₂. These challenges highlight the need for integrated solutions that address both waste management and low-carbon energy production.

Hydrogen has emerged as a key energy carrier for supporting deep carbonization, with production pathways ranging from renewable-based electrolysis to fossil-derived routes coupled with CCS [48], [235], [236]. In parallel, WtE technologies provide an opportunity to convert waste into valuable energy products. Plasma gasification is particularly suitable for processing dry, carbon-rich waste such as plastics, producing a clean syngas with low tar content and a vitrified slag byproduct [75], [245], [246]. In contrast, anaerobic digestion is more appropriate for high-moisture organic waste such as food waste, generating biogas and digestate[77]. Integrating these two technologies allows for the effective utilization of heterogeneous MSW, while the inclusion of CCS can significantly reduce carbon emissions associated with energy conversion processes.

Previous studies have demonstrated the technical feasibility of hydrogen production

from waste via plasma gasification, often combined with purification and CCS [185], [214], [249]. However, these studies have largely focused on hydrogen as a single output and on individual waste streams, without considering integrated multi-output systems. Opportunities to enhance system performance through combined hydrogen and CCHP generation remain underexplored. Furthermore, the integration of plasma gasification with anaerobic digestion for dual-feedstock processing has received limited attention, and comprehensive TEA-LCA of such systems, particularly with CCS, are scarce. In addition, existing CCS studies are typically limited to single configurations, making it difficult to assess the comparative performance of pre- and post-combustion capture strategies. To address these gaps, this chapter investigates the techno-economic and environmental performance of a novel integrated WtE system that combines plastic waste plasma gasification with food waste anaerobic digestion for the hydrogen and CCHP production. The system incorporates CCS and evaluates both pre- and post-combustion configurations under consistent assumptions.

This chapter presents the research published in the author's third journal paper, which provides the first integrated TEA-LCA of a dual feedstock plasma gasification-anaerobic digestion system with CCS for hydrogen and CCHP production. The study offers new insights into the trade-offs between cost, efficiency, and carbon mitigation, contributing to the development of sustainable WtE systems.

6.2 Methodology

6.2.1 System description

The proposed system is based on actual waste composition data from one of Indonesia's largest landfills, Benowo Landfill, which spans approximately 37 hectares. Among the various waste types, plastic and food waste represent the highest proportions, accounting for roughly 16% and 28% of the total waste, respectively. In this study, the feed rates of plastic and food waste into the integrated hydrogen and CCHP system were set at 5 kg/s (equivalent to 184 MW_{th}) and 13 kg/s (46 MW_{th}), respectively. The characteristics of these waste streams are summarized in Table 3-1. The system is designed to extend the functionality of the

existing waste-to-energy (WTE) plant at Benowo Landfill, which is currently limited to electricity and heat production. Furthermore, the system produces valuable co-products: slag from plasma gasification, which can be repurposed as a construction material, and digestate from anaerobic digestion, which can be applied as an organic fertilizer.

This study evaluates a waste-to-hydrogen system that also provides cooling and heating under two scenarios: Scenario 1 integrates pre-CCS into the energy system, while Scenario 2 adopts post-CCS. Each scenario is further divided into three cases. Within each scenario, all cases share the same downstream configuration, comprising a water–gas shift reactor (WGSR), a CCS unit, and a PSA system for hydrogen purification. In addition, a combined cycle gas turbine (CCGT) is used for electricity generation, and an ARC is employed to supply cooling and heating.

The following descriptions outline the specific assumptions and configurations for each scenario, with particular focus on how the plastic and food waste streams are handled prior to entering the shared downstream processes. Block flow diagrams of the proposed systems with pre-CCS and post-CCS are presented in Figure 6-1a and Figure 6-1b, respectively.

1. Case 1

Incorporates different treatment technologies for plastic and food waste. Plastic waste is processed in a PG to generate syngas, while food waste undergoes anaerobic digestion to produce biogas. The syngas and biogas are mixed and processed through the downstream system as described above.

2. Case 2

Assumes that both plastic and food waste are co-processed in the plasma gasification unit to generate mixed syngas, which then follows the same to Case 1 hydrogen production and energy recovery steps.

3. Case 3

Similar to Case 1, plastic and food waste are treated separately. However, the biogas is first processed in a SMR to increase its hydrogen content before being mixed with syngas. The combined gas stream then

proceeds through the same WGSR, CCS, PSA, and energy recovery systems as in the previous cases.

Pre-treatment is required prior to the PG and this depends on the waste type. For plastic waste, the material is crushed before being fed into the PG. In contrast, in the case of mixed plastic and food waste, both wastes must first be dried and then crushed. The PG operates at a temperature of 2500°C, achieved using a plasma torch that heats air up to 4000°C. Electricity for the plasma torch is supplied by on-site generated power from a CCGT system. To enhance the lower heating value (LHV) of the syngas, steam is employed as an additional gasifying agent. The raw syngas exiting the PG is at a very high temperature, and its thermal energy is recovered to preheat the incoming air for the plasma torch and to preheat water before it enters the PG. The byproduct from the PG is slag which can be utilized as construction material.

In anaerobic digester (AD), the food waste is processed in a mesophilic environment with the temperature of 38°C. The AD heating demand was assumed to be 10% of the energy content of the biogas [259]. To fulfil this requirement, the AD was heated using the heat recovered from the vapour stream leaving the stripper prior to the reflux condenser. The AD process generates digestate as a byproduct, which can be used as a fertilizer.

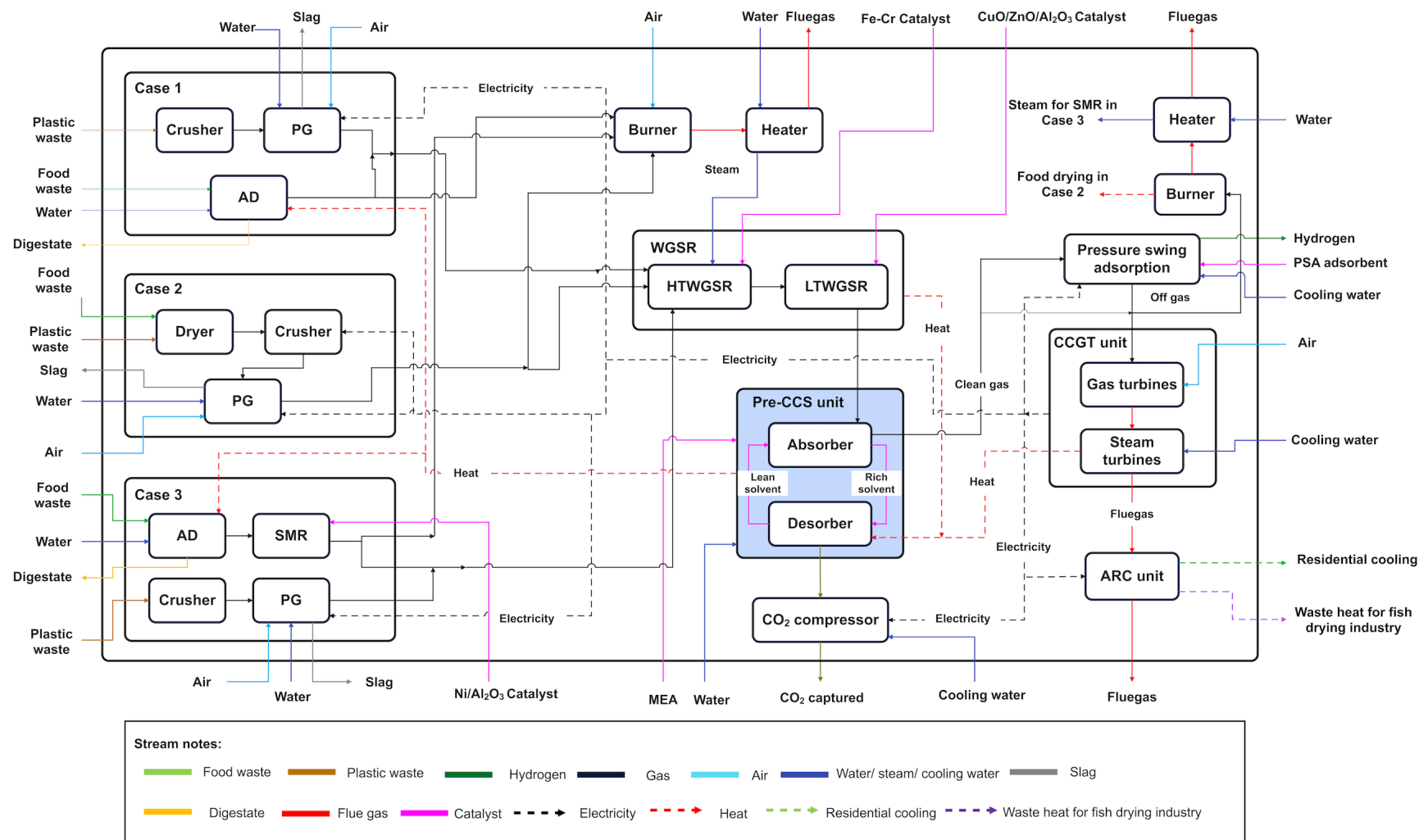
In the WGSR, the feed gas, and steam is subjected to the water-gas shift (WGS) reaction. A fraction of the biogas is combusted to generate heat for steam production, with the resulting steam subsequently used in the (WGSR). This fraction is adjusted to achieve similar CO conversion ($\approx 95\%$) in Cases 1 and 3; Case 2 does not follow this rule because the food waste is not converted to biogas. Biogas is selected as the fuel so that the associated CO₂ emissions in the flue gas are entirely biogenic. The WGS reaction is conducted at two different temperature levels to maximize CO conversion [300]. The high-temperature WGSR (HTWGSR) operates at 450°C with a Fe-Cr catalyst, while the low-temperature WGSR (LTWGSR) operates at 200°C with a CuO/ZnO/Al₂O₃ catalyst [124], [125]. The WGSR is employed in both scenarios to increase the hydrogen content of the gas.

Following the WGSR, the gas enters the pre-CCS unit for Scenario 1, where 95% of

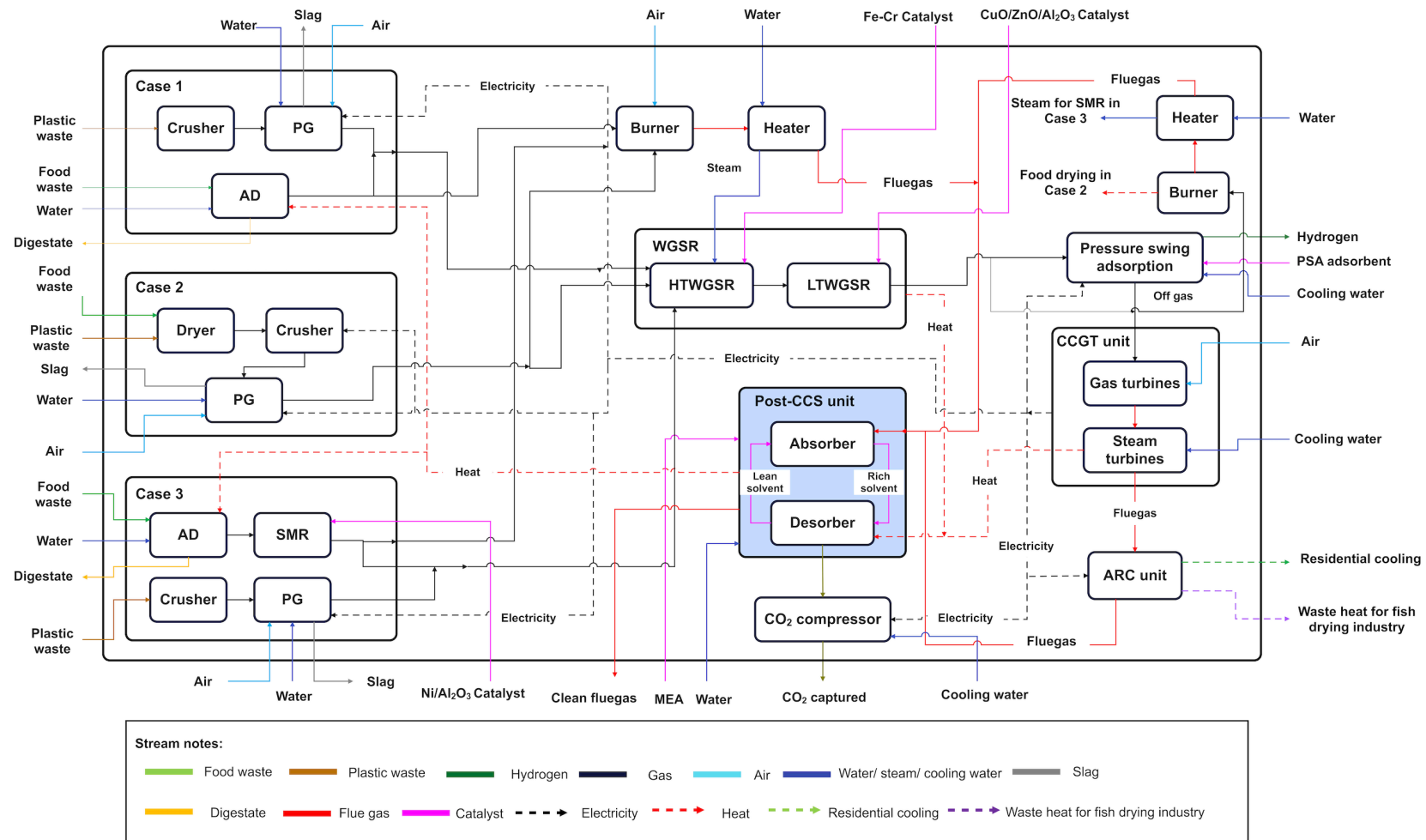
the CO₂ is captured [166]. MEA at a concentration of 35% w/w is chosen as the solvent for its affordability and effective performance and also has extensive performance/ cost data in the literature [166]. The absorber packing diameter is specified to ensure it adheres to the 80% flooding condition limit [298]. The heat required by the desorber of the CCS plant is satisfied from the LP steam produced by utilizing heat in the stream in the WGSR and LP steam split from LP steam in CCGT. The CO₂ is then compressed in multiple stages to achieve supercritical conditions and convert it into a liquid state for transportation and storage underground. In Scenario 2, by contrast, the shifted gas is sent directly to the PSA, with CO₂ capture implemented downstream via post-combustion CCS after flue gas from CCGT passing the generator in the ARC unit.

The gas exiting pre-CCS unit is processed in a PSA system to produce hydrogen with 99% purity [343]. Some portions of PSA feed gas and PSA tail gas are utilized in a CCGT to generate electricity for internal consumption. In the gas turbine (GT), compressed gas is mixed with compressed air in the combustion chamber, and the resulting high-temperature exhaust drives the turbine. This exhaust is then directed to the steam production system for further energy recovery. In the steam turbine (ST) power plant, heat from the flue gas is recovered in a heat recovery steam generator (HRSG) to produce low-pressure (LP), intermediate-pressure (IP), and high-pressure (HP) steam. The flue gas exiting the HRSG is then routed to an ARC for additional thermal utilization.

Waste heat from the CCGT flue gas is recovered in the ARC, which employs a LiBr/H₂O working fluid to produce residential cooling at 21°C. The remaining thermal energy is utilized for fish drying in local industries at 52°C, contributing to efficient multi-output energy recovery. After heat extraction, the flue gas is released into the atmosphere.



(a)



(b)

Figure 6-1. Block flow diagram of the proposed system: (a) Scenario 1 and (b) Scenario 2. The system boundaries cover all components modelled in Aspen Plus and are consistently applied in both the economic and environmental assessments.

Table 6-1. Key assumptions and operating conditions for process simulation.

Process Unit	Parameter	Value / range	Unit	Notes / source
Pretreatment (Model: gyratory crusher [288], resulting in the particle size reduction to 1 mm)	Electricity required for grinding	$\dot{W}_{el,consumption}$ $= \frac{50 \text{ kWh}}{\dot{m}_{feedstock}}$	$\frac{\text{kWh}}{\text{tonne}}$	[288]
Plasma gasifier (Model: air plasma gasifier combined with the steam injection. [289])	Plasma torch temperature	4000	°C	[80]
	Equivalence Ratio (ER)	0.14	-	
	Steam-to-Air Mass Ratio (SAMR)	0.396	-	Iteratively chosen (until the temperature of HTZ reaching 2500°C)
	Steam-to-feedstock Ratio (SFR)	0.8	-	
	HTZ temperature	2500	°C	[80]
	LTZ temperature	1250	°C	[80]
Anaerobic digester	Digestion temperature	38	°C	[290]
	Dilution ratio (food waste: water)	2.64:1	kg/kg	[290]
	Heating demand during the digestion process	$\dot{Q}_{digestion}$ $= 10\% \cdot \dot{E}_{biogas}$	MW	[259]
Steam methane reforming (SMR)	Reactor temperature	650	°C	[136]
	Pressure	1	bar	[136], [361], [362], [363], [364]
	S/CH ₄	3	-	[365]
	L/D	1.3/0.1	m/m	
Pressure swing adsorption	Hydrogen purity	99.9	%	
	Compressor pressure ratio	4	-	[343]
	Hydrogen molar concentration prior PSA	≥70	%	[343]
Gas turbine power plant	Compressor pressure ratio	19.5	-	[292]
	Turbine inlet temperature	1350	°C	[293]
	Fuel pressure	26.83	bar	[292]

Steam turbine power plant	High Pressure steam outlet pressure	28.8	bar	[294]
	Intermediate Pressure steam outlet pressure	3.907	bar	[294]
	Low Pressure steam outlet pressure	0.0483	bar	[294]
Absorption refrigeration cooler	Working fluid	LiBr / H ₂ O	-	[295]
	Ratio of working fluid to flue gas	0.1	-	[295]
	Pump outlet pressure	0.07461	bar	[295], [296]
	Valve outlet pressure	0.00672	bar	[295], [296]
	Cooling temperature	21	°C	Building cooling demand
	Heating temperature	52	°C	Fish drying application
Carbon capture and storage (CCS)	CO ₂ capture rate	95	%	
	Pre CCS			
	Absorber height/diameter	25/5	m/m	Avoid column flooding
	Stripper height/diameter	15/3.5	m/m	
	Stripper pressure	2.4	bar	
	CO₂ sequestration			
	Multistage compressor			
	Outlet pressure 1	7	bar	[298]
	Outlet pressure 2	24	bar	[298]
	Outlet pressure 3	83	bar	[298]
	Pump outlet pressure	153	bar	[298]

6.2.2 Technical details

The key assumptions and operating conditions applied in the process simulation are provided in Table 6-1. These parameters include, among others, the plasma torch temperature, equivalence ratio, steam-to-feedstock ratio, dilution temperature in anaerobic digestion, compressor pressure ratio, turbine inlet temperature, and the dimensions of the absorber and stripper in the CCS unit. Collectively, these assumptions form the foundation for accurately modelling system performance.

To simulate the SMR reactions, the kinetic expressions summarized in Table 2-4 was used. In Aspen Plus, these kinetics are implemented by specifying the driving-force coefficients (Term 1 and Term 2) and the adsorption constants that appear in the denominator of the rate law; the required coefficients are listed in Table 2-5.

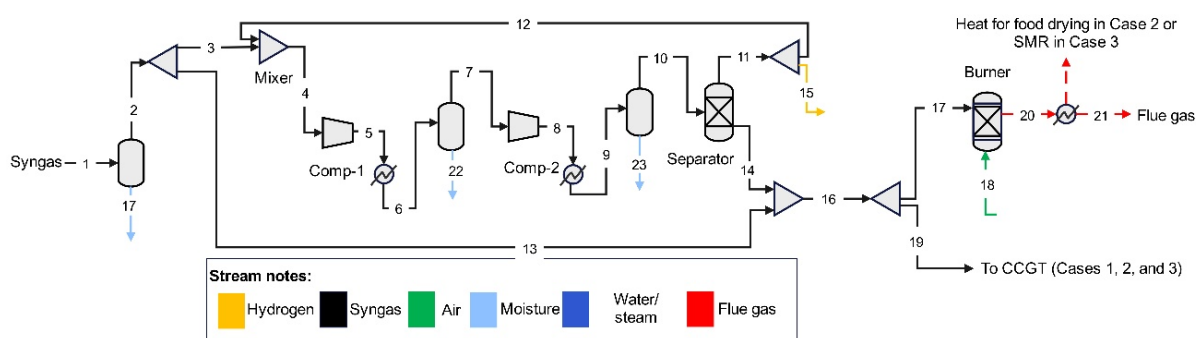


Figure 6-2. Process flow diagram of PSA along with the stream route for fuel supply in the CCGT and food drying (Case 2) or SMR (Case 3).

For the PSA section, the process flow diagram is shown in Figure 6-2. In Scenario 1, the PSA feed is the CO₂-lean gas exiting the pre-CCS unit, whereas in Scenario 2 it is the shifted gas leaving the WGSR. This stream is split: one branch (Stream 3) enters the PSA to produce hydrogen at 99.9% purity, and the other branch (Stream 13) supplies fuel/heat—Case 1: to the CCGT only; Case 2: to the CCGT and the food-drying duty; Case 3: to the CCGT and the SMR heat duty. After the split, the PSA feed is compressed in two stages (pressure ratio 4 per stage) [343]. Consistent with prior work [343], a minimum 70 mol% H₂ at Stream 10 is required for effective purification. Because the feed gas is typically below this threshold, a

controlled recycle of high-purity hydrogen from the PSA product is blended into the feed to meet the 70 mol% specification. The resulting recycle-to-fresh-feed ratios can be seen in Table 6-2.

Table 6-2. Hydrogen recycle-to-fresh-feed ratios (%) at the PSA inlet for Scenarios 1 and 2.

	Scenario 1			Scenario 2		
	Case 1	Case 2	Case 3	Case 1	Case 2	Case 3
Recycle ratio (%)	20	8	1	62	59	52

As shown in Table 6-2, the hydrogen recycle-to-fresh-feed ratios are significantly higher in Scenario 2 (52–62%) than in Scenario 1 (1–20%). This is because, in Scenario 1, the PSA treats a CO₂-lean stream leaving the pre-CCS unit, whereas in Scenario 2 the PSA feed is the shifted gas directly from the WGSR, which contains a higher proportion of CO₂. The lower hydrogen concentration in the PSA feed in Scenario 2 therefore requires a larger recycle of high-purity hydrogen to satisfy the 70 mol% H₂ specification at the PSA inlet. Furthermore, Case 3 exhibits the lowest hydrogen recycle-to-fresh-feed ratios in both scenarios. In this case, the biogas produced from food waste is further processed in the SMR and WGSR, yielding a higher hydrogen content in the PSA feed than in Cases 1 and 2. The higher inlet H₂ concentration correspondingly reduces the amount of high-purity hydrogen that must be recycled to meet the 70 mol% H₂ requirements.

6.2.3 Thermodynamic evaluation

The system outputs consist of hydrogen and CCHP, with all electricity generated allocated to meet the internal power demand. To evaluate the performance of the overall system across all scenarios, the total energy and exergy efficiencies are defined as follows:

$$\eta_{en,overall} = \frac{\dot{E}_{H_2} + \dot{Q}_{cooling} + \dot{Q}_{heating}}{\dot{E}_{pw} + \dot{E}_{fw}} \quad \text{Eq. 6-1}$$

$$\eta_{ex,overall} = \frac{\dot{E}x_{H_2} + \dot{E}x_{cooling} + \dot{E}x_{heating}}{\dot{E}x_{pw} + \dot{E}x_{fw}} \quad \text{Eq. 6-2}$$

where $\dot{E}_{H_2}$ and $\dot{E}x_{H_2}$ represent the energy and exergy content of hydrogen (MW), respectively; $\dot{Q}_{cooling}$ and $\dot{E}x_{cooling}$ denote the cooling energy and exergy generated from the ARC (MW), respectively; $\dot{Q}_{heating}$ and $\dot{E}x_{heating}$ correspond to the heating energy and exergy released from the system (MW), respectively.

6.2.4 Economic evaluation

Gross profit is determined by monetizing the useful energy outputs of the system, namely hydrogen, cooling and heating. For cooling, and heating, the exergy content is applied to account for the quality and utility of these energy forms, thereby providing a more accurate representation of their economic value than raw energy content. In calculating P_n , the co-products, such as cooling and heating, were valued at prevailing market price, as expressed in Eq. 6-3.

$$P_n = \left(LCOH \times \dot{m}_{hydrogen} + \$_{cooling} \times \frac{\dot{E}x_{cooling}}{COP_{cooler}} + \$_{heating} \times \frac{\dot{E}x_{heating}}{COP_{heater}} + \$_{slag} \times \dot{m}_{slag} + \$_{tipping} \times \dot{m}_{waste} + \$_{carbon\ credit} \times \dot{m}_{captured\ CO_2} \right) \times C_f \times 8000 - OPEX_n \quad \text{Eq. 6-3}$$

$$LCOH = \frac{ACAPEX + AOPEX - AREV_{co-products} - AREV_{TIP} - AREV_{CO_2}}{APROD_{H_2}} \quad \text{Eq. 6-4}$$

where $\dot{m}_{hydrogen}$ is the mass flow rate of hydrogen (kg/s), $\$_{cooling}$ and $\$_{heating}$ are valued at the electricity market price ($\$_{elec}$) as listed in

, $\dot{E}x_{cooling}$ and $\dot{E}x_{heating}$ represent the cooling and heating exergy (kWh), respectively. The coefficient of performance (COP) were assumed as $COP_{cooler} = 3.45$ and $COP_{heater} = 1$ [326]. $\$_{slag}$ and $\dot{m}_{slag}$ denote the market price of slag shown in

and its mass flow rate of slag, respectively. The tipping fee ($\$_{tipping}$) is taken as the Indonesian rate of 14.02 USD/tonne [327], with $\dot{m}_{waste}$ representing the waste

processed. The carbon credit price ($\$_{carbon\ credit}$) follows the Indonesian market value of 8.9 USD/tonne CO₂ [328]. The key parameters used in the DCFA are summarized in Table 3-3. C_f is the capacity factor, $OPEX$ represents the annual operating costs of the process plant, assuming 8000 operating hours per year. It includes both fixed and variable costs. The fixed costs are listed in Table 3-5, while the variable component (VC) consists primarily of utilities, which are detailed in Table 6-5.

Table 6-3. Market based assumptions used in the analysis.

Parameter	Description
Electricity market price [329]	0.105 USD/kWh
Vitrified slag price [215]	53.78 USD/tonne
Indonesia's tipping fee [327]	14.02 USD/tonne
Indonesia's carbon credit price [328]	8.9 USD/tonne CO ₂

The estimation of LCOH requires two primary inputs: capital expenditure (CAPEX) and operating expenditure (OPEX). LCOH is calculated as the net annualised cost divided by annual hydrogen production, as expressed in Eq. 6-4. Where $ACAPEX$ is annualised CAPEX, $AOPEX$ is annual OPEX, $AREV_{co-products}$ is annual revenue from co-products, such as heating, cooling, and vitrified slag. The list of PEC is provided in Table 6-4. $AREV_{TIP}$ and $AREV_{CO_2}$ are annual revenues associated with policy incentives (tipping fee and carbon credit price), respectively. $APROD_{H_2}$ denotes annual hydrogen production.

Table 6-4. PEC at base capacity, PEC formulation and year of reference.

Equipment	Base cost [MM USD]	Base capacity	Unit	Scaling factor	Base year	Ref.
Crusher	2	2000	ton/day (dry feed mass flow)	0.72	2007	[332]
Drier	19.3	2000	ton/day (dry feed mass flow)	0.72	2007	[332]
Plasma gasifier	78	39.2	kg/s (feed mass flow)	0.67	2021	[333]
WGSR	3.77	150	kg/s (total gas feed)	0.67	2014	[337]
Substrate heater	0.693	13,149	m ² (heat transfer area)	0.68	2014	[334]
Anaerobic digester	0.455	727.21	m ³ (feed volume flow)	1	2017	[335]
CCGT	142.33	390.99	MW of output power	0.7*	2021	[366]
SMR	25.372	483.01	tonne/hr (H ₂ production)	0.79**	2018	[367]
PSA	7.81	0.294	kmol/s (purge gas)	0.74	2014	[337]
Desorber	0.0175	100	m ² (heat transfer area)	0.6	2000	[336]
Absorber	0.0165	100	m ² (heat transfer area)	0.6	2000	[336]
Solution heat exchanger	0.012	100	m ² (heat transfer area)	0.6	2000	[336]
Burner***	2.14	20	MW (heat duty)	0.83	2014	[337]
Cooling tower	3.99	4530.3	kg/s (feed mass flow)	0.78	2014	[338]
Heat exchanger	0.8	8372	m ² (heat transfer area)	0.68	2014	[334]
CCS plant	The PEC of all components (compressors, pumps, heat exchangers, columns) in the CCS plant is estimated using the financial calculations provided in the referenced study					[340]

(*) The scale factor used from the reference [337]

(**) The scale factor used from the reference [340]

(***) The burner combusts a 10% of raw syngas and generates heat to supply steam required for the WGSR process.

Table 6-5. The component of variable cost [259], [368].

Expenses	Cost
Variable Cost (VC)	VC = $\sum(1 \text{ to } 6)$
(1) Cooling water [341]	0.032 USD/1000 kg
(2) Feed water [342]	1 USD/1000 kg
(3) Waste water treatment [342]	0.53 USD/1000 kg
(4) MEA solvent price ^b [340]	1169.27 USD/1000 kg
(5) WGS catalyst ^{c,d} [343]	10.3 USD/kg
(6) SMR catalyst ^e [369]	33.07 USD/kg
(7) PSA adsorbent [370]	1.4 USD/kg
(8) CO ₂ transport and storage [341]	24.27 USD/1000 kg

^a Operating labour work is assumed to be 2000 hours per year, with a wage rate of 2.0415 USD/h based on Indonesian standards [260].

^b The MEA consumption rate of 2 kg/tCO₂ is assumed, as estimated by Elliott et al. [165].

^c The WGS catalysts are to be changed over a period of 3 years [343].

^d The GHSV for the HTWGS and the LTWGS reactors are 3000 and 4000 hr⁻¹ respectively [343].

^e The SMR catalysts are to be changed over a period of 2 years [371].

6.2.5 Sensitivity and uncertainty analysis for the economic investigation

To improve the credibility and robustness of the economic outcomes, both sensitivity and uncertainty analyses were undertaken. These analyses help reveal which input variables exert the greatest influence on the calculated economic indicators. The sensitivity analysis adopts a one-at-a-time approach, in which each parameter is individually varied while holding others constant, allowing the isolation of its specific impact. Table 6-6 outlines the key variables examined, along with their respective upper and lower bounds, chosen due to their inherent uncertainty.

Table 6-6. Economic variables and ranges used in sensitivity and uncertainty analysis.

Economic input	Lower bound	Nominal	Upper bound	Unit
PEC GT	70	100	150	%
PEC CCS	70	100	150	%
PEC PG	70	100	150	%
Tipping fee	80	100	120	%
Carbon credit	80	100	120	%

Tax rate	8	10	12	%
Discount rate	0	30	40	%

6.2.6 Life cycle assessment

The method for conducting the LCA can be seen in Sub Chapter 3.5 with the corresponding inventory presented in the Appendix 9.3.

6.3 Result and discussion

The results are presented in three parts: technical, economic, and environmental performance and are discussed in detail in separate subsections. Case 1 demonstrates the most balanced technical performance across both scenarios, combining high hydrogen yield with strong energy/exergy efficiencies, while also achieving the lowest LCOH. Case 2 exhibits the highest internal energy penalty because co-processing food and plastic waste increases the plasma load, along with drying and handling requirements. This results in higher auxiliary electricity consumption and lower net efficiency. Case 3 improves hydrogen recovery by upgrading AD biogas via SMR, however this advantage is accompanied by additional heat demand. The CCS configuration also has a strong influence on system performance. Post-CCS (Scenario 2) requires a higher heat duty than pre-CCS (Scenario 1). From an environmental perspective, the key drivers are the amount of net CO₂ captured and emissions associated with utilities, both of which vary depending on the waste routing strategy and CCS configuration. Overall, these findings show that upstream waste routing and the location of the CCS are the dominant design choices shaping the main cost and environmental trade-offs.

6.3.1 Technical performance

Table 6-7 summarizes the key process model outputs, including energy generation, CO₂ capture, and overall system efficiency. These results provide the essential baseline data for the subsequent TEA and LCA discussed in the following sections.

Table 6-7. The performance of the two scenarios.

Parameters	Scenario 1			Scenario 2			Unit
	Case 1	Case 2	Case 3	Case 1	Case 2	Case 3	
Process							
Heat demand within the system	Value	Value	Value	Value	Value	Value	Unit
Heat for the desorber in the CCS plant	42.47	48.98	47.23	56.67	57.44	57.38	MW
Heat for food waste dryer prior to PG	N/A	27.08	N/A	N/A	27.08	N/A	MW
Heat for WGSR	20.81	26.38	25.23	20.81	26.38	25.23	MW
Heat for anaerobic digester	4.43	N/A	4.43	4.43	N/A	4.43	MW
Heat for SMR and pre heating the feed gas	N/A	N/A	24.97	0.00	N/A	24.97	MW
Total heat demand	67.70	102.43	101.86	81.90	110.90	112.01	MW
Power demand within the system	Value	Value	Value	Value	Value	Value	Unit
Power for crusher	0.90	0.90	0.90	0.90	0.90	0.90	MW
Power for torch	42.72	54.08	42.72	42.72	54.08	42.72	MW
Power for CCS compressors and pumps	4.80	5.58	5.38	4.67	4.46	4.60	MW
Power for PSA compressors	8.12	3.62	5.72	13.31	6.23	8.96	MW
Total power demand	56.54	64.18	54.72	61.59	65.67	57.18	MW
CO ₂ captured in CCS plant	0.66	0.76	0.73	0.86	0.87	0.86	kg/kg waste
Production							
H ₂	0.70	0.35	0.62	0.53	0.27	0.45	kg/s
Cooling (energy based)	2.02	2.23	1.86	2.26	2.31	2.12	MW

Heating (energy based)	8.44	9.53	7.94	9.65	9.89	9.05	MW
Cooling (exergy based)	0.17	0.19	0.16	0.19	0.19	0.18	MW
Heating (exergy based)	0.78	0.89	0.74	0.90	0.92	0.85	MW
Slag	0.66	1.14	0.66	0.66	1.14	0.66	kg/s
Process efficiency							
Overall energy efficiency	41.30	23.45	36.49	32.96	19.34	28.84	%
Overall exergy efficiency	33.85	17.11	29.71	25.74	13.21	22.26	%

6.3.1.1 Heat demand

Table 6-7 presents the heat demand of the system for Cases 1, 2, and 3 in both Scenarios 1 and 2. In all cases, the dominant heat load arises from the desorber in the CCS plant. In Scenario 1, the CCS heat duty varies more substantially, from approximately 42 to 49 MW, whereas in Scenario 2 it lies in a narrower range of about 56-58 MW. The difference reflects the way each configuration affects the CO₂ load to the capture unit: in Scenario 1, the pre-combustion CCS is located downstream of the WGSR, so changes in the upstream conversion routes in Cases 1-3 (i.e. whether food and plastic waste are processed separately or as a mixed feed, and the presence or absence of SMR, which affects the extent of CO formation and subsequent CO shift) directly alter the CO₂ flow rate and composition entering CCS, leading to larger variations in desorber duty. In Scenario 2, by contrast, the post combustion CCS treats the combined flue gases from all relevant combustion units, namely: fuel combustion to provide heat for the WGSR (Cases 1-3), the CCGT exhaust (Cases 1-3), fuel combustion for food waste drying in Case 2, and fuel combustion for providing heat to the SMR in Case 3. As these streams are all ultimately derived from the same overall waste feed, the total amount of carbon oxidised to CO₂, and hence the CO₂ flow to the capture unit, remains very similar across Cases 1-3.

While energy consumption in the desorber is often considered the primary performance metric in solvent selection, it represents only part of the picture. Solvent management costs, including degradation, reclamation, and handling of by-products such as nitrosamines, can be substantial for blended amines, whereas MEA exhibits relatively low management costs [164]. In addition, MEA is well-characterized, non-proprietary, low-cost, and biodegradable, with proven reclamation methods and minimal environmental or health impacts, making it a reliable benchmark for post-combustion CO₂ capture [165]. Its extensive documentation provides a stable baseline for modelling, techno-economic comparison, and policy-relevant analysis, not because it is technically superior in every respect, but because it is the most widely studied and adopted reference solvent in the field.

In both scenarios, Case 2 has a major additional heat requirement associated with drying. Because Case 2 co-processes food and plastic waste, with a combined moisture content of 56% (each moisture content can be seen in Table 3-1), drying is required to reduce the moisture to 10%. The required drying duty is supplied by combusting approximately 15-17% of the PSA off-gas, and the model predicts a duty of 27.08 MW. This value aligns with the findings of Li et al. [372], who reported a thermal demand of approximately 2.7–2.8 GJ per tonne of water evaporated using a flash dryer.

In Cases 1 and 3, the second-largest heat demand is associated with the WGSR. WGSR heat duty differs across all cases because the CO and H₂O contents of the incoming syngas are not identical, which leads to different amounts of steam that must be generated and added. Case 2 has the highest CO flow rates and therefore requires the largest steam addition to achieve the targeted CO conversion ($\approx 95\%$). In Case 3, an additional significant heat demand arises from the SMR unit, where biogas and water are preheated to 330°C and the reformer outlet temperature reaches 650°C. This duty is supplied by combusting part of the PSA feed gas together with PSA off-gas.

By contrast, the lowest heat demand is associated with anaerobic digestion. The AD were assumed to require heat equivalent to 10% of the biogas energy content [259], supplied using heat recovered from the vapor stream leaving the CCS desorber prior to the reflux condenser. In addition to biogas, the AD process generates digestate as a co-product, which can be used as fertilizer.

6.3.1.2 Electricity demand

The electricity generated in the system is entirely used for internal operation, including powering components such as the waste crusher, plasma torch, compressors and pumps in the CCS unit, and compressors in the PSA system. As shown in Table 6-7, among all components, the plasma torch accounts for the highest electricity demand. This is linked to the high temperature assumed for the plasma gasifier (2500°C), which supports deep conversion and low tar formation but requires substantial electrical input. As a result, there is a clear trade-off: higher operating temperatures improves syngas quality, but it raises electricity

consumption and subsequently reduces overall system energy efficiency. An alternative approach is to operate at a lower gasification temperature to reduce electricity demand. However, this typically requires additional process steps, including catalytic tar reforming/ cracking and optimization of gasifying-agent ratios to maintain syngas quality suitable for downstream use [91], [92]. Such process steps introduce additional challenges, including increased tar management requirements, enhanced control strategies, catalyst handling/ recovery, and potentially additional gas cleaning units.

Among three cases, case 2 exhibits the highest overall electricity consumption because it processes the largest waste flow through the plasma gasifier (food and plastic waste are treated together as a mixed feedstock). This increases the plasma load and, where applicable, increases upstream conditioning requirements (e.g., drying and handling of high moisture fractions), which further raises internal energy and demand. In contrast, Cases 1 and 3 show lower plasma torch electricity demand because food waste is routed through anaerobic digestion rather than being co-processed in the plasma unit, thereby reducing the amount of material requiring high-temperature conversion.

The electricity demand also varies with how much gas is routed to hydrogen purification. Case 1 shows slightly higher PSA compressor power than Case 3 because a larger share of the processed gas is directed to the PSA unit for hydrogen separation, whereas in Cases 2 and 3, a larger portion of the gas is diverted to satisfy higher heat demands, reducing the PSA compressor load.

6.3.1.3 Hydrogen production

As shown in Table 6-7, hydrogen production is highest in Case 1, primarily because hydrogen output is directly influenced by the amount of feed gas entering the PSA unit. In Case 1, a larger fraction of the processed gas is available for hydrogen purification, as no additional fuel is required for food waste drying or SMR. Consequently, Case 1 allocates the largest share of fuel to the PSA unit, with approximately 58% of the gas stream directed to hydrogen purification in Scenario 1 and 53% in Scenario 2; the remaining fraction is supplied to the CCGT for electricity generation. The slightly higher PSA feed fraction in Scenario 1 reflects its

lower overall electricity demand, which reduces the amount of fuel that must be diverted to the CCGT.

Case 3 achieves the second-highest hydrogen production, as it incorporates two hydrogen-enhancing mechanisms: the water-gas shift reaction and SMR. However, the substantial thermal energy required to operate the SMR reduces the fraction of gas available for hydrogen purification, with only about 42% of the gas stream sent to the PSA in Scenario 1 and 36% in Scenario 2. Despite benefiting from both conversion mechanisms, the higher internal energy demand in Case 3 ultimately leads to a lower net hydrogen yield than in Case 1.

Case 2 has the lowest hydrogen yield. This is attributed to the significant thermal energy required for drying food waste prior to PG and the highest plasma torch power demand among the cases. In Case 2, plastic and food waste are co-gasified, which increases the plasma energy requirement. Consequently, only around 25% of the gas stream is directed to the PSA unit in Scenario 1 and 23% in Scenario 2.

6.3.1.4 Cooling and heating production

As shown in Table 6-7, Case 2 exhibits the highest levels of heating and cooling generation in both scenarios. This is due to the largest volume of fuel combusted in the CCGT system to meet its substantial electricity demand, resulting in the highest flue gas output. The increased flue gas flow enhances both steam production and the heat recovery available to drive the absorption refrigeration cooler (ARC). Cases 1 and 3 show comparable levels of heating and cooling production, with Case 1 being slightly higher due to a greater volume of flue gas generated from increased off-gas combustion in the CCGT, driven by its higher electricity demand relative to Case 3.

6.3.1.5 Performance efficiency

The system's performance is assessed using both energy and exergy efficiencies as calculated by the equation provided above. As summarized in Table 6-7, Case 1 achieves the highest energy and exergy efficiencies in both scenarios, primarily because it produces the largest amount of hydrogen while requiring less internal heating. This allows a greater portion of the fuel to be directed toward hydrogen

production rather than being consumed for process heat.

Case 3 ranks second in terms of energy and exergy efficiency. Its strong performance is attributed to two hydrogen-enhancing mechanisms: SMR of biogas and the application of WGSR to the syngas stream. However, its higher internal heating demand compared to Case 1 slightly reduces its overall efficiency. In contrast, Case 2 demonstrates the lowest energy and exergy efficiencies, primarily due to its lower hydrogen output. This case requires substantial thermal energy to dry the food waste prior to gasification and has the highest plasma torch electricity demand, as it processes a mixture of food and plastic waste. As a result, a smaller portion of the fuel reaches the PSA unit for hydrogen recovery. These combined factors lead to a notable reduction in the system's energy and exergy performance.

In the comparison between scenarios, Scenario 1 exhibits higher overall energy and exergy efficiencies than Scenario 2. This is primarily because Scenario 1 produces more hydrogen, since a smaller fraction of the conditioned syngas needs to be diverted to meet internal heat and power demands. In both scenarios, the PSA feed must be split between hydrogen production and fuel to the CCGT and auxiliary combustion units; therefore, higher internal energy requirements directly reduce the gas available for hydrogen purification. In Scenario 1, the total heat demand is lower, mainly due to the reduced desorber duty in the pre-combustion CCS unit, which treats a CO₂-rich fuel gas stream with a lower volumetric flow rate.

In Scenario 2, the CCS unit captures CO₂ from the combined flue gas of the CCGT and other combustion sources rather than from the shifted with a higher fuel gas. As a result, CO₂ is removed from a more diluted stream at higher total flow, which increases the CCS desorber heat duty and marginally raises overall electricity demand. Consequently, a larger fraction of the PSA feed gas must be burned to cover these internal energy demands, leaving less gas for hydrogen production and leading to lower process efficiencies in Scenario 2.

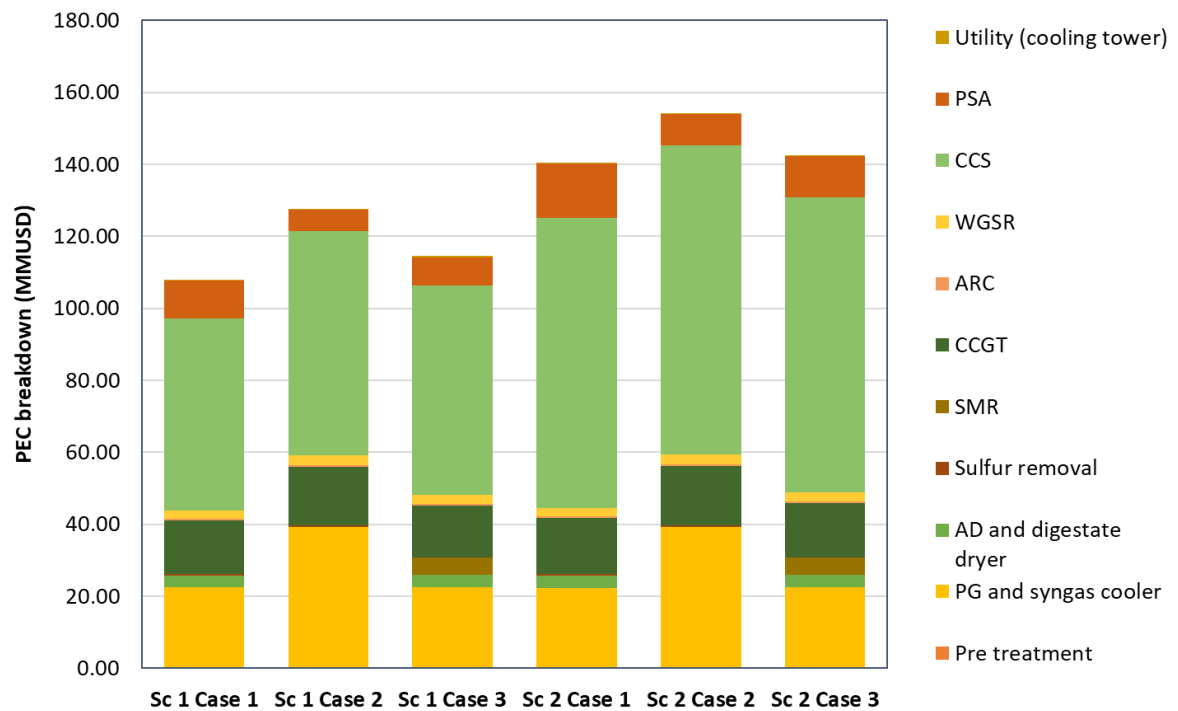
6.3.2 Economic performance

6.3.2.1 CAPEX and OPEX

Figure 6-3a and Figure 6-3b show the PEC and OPEX, respectively. As shown in

Figure 6-3a, Case 2 incurs the highest PEC in both scenarios, estimated at approximately 128 million USD in Scenario 1 and 154 million USD in Scenario 2, due to several factors. First, the CCS unit is significantly larger because plasma gasification in Case 2 treats the combined plastic and food waste streams, generating a higher syngas flow and elevated CO concentrations; consequently, the WGSR produces a greater CO₂ flow rate to the CCS unit than in other cases. By contrast, in Case 1, the food waste is converted to CH₄ and CO₂ in the AD unit and bypasses the WGSR pathway, while in Case 3, the conversion of CH₄ to CO in the SMR unit is constrained by the available heat duty, leading to incomplete reforming and a lower overall CO₂ load to the CCS plant.

Second, the plasma gasifier must be scaled up to process the larger volume and heterogeneous composition of the mixed feedstock. Third, Case 2 requires a substantial investment in the CCGT, as it is designed to meet the highest electricity demand among the three cases, largely due to the energy-intensive plasma torch operation associated with treating mixed waste.



(a)

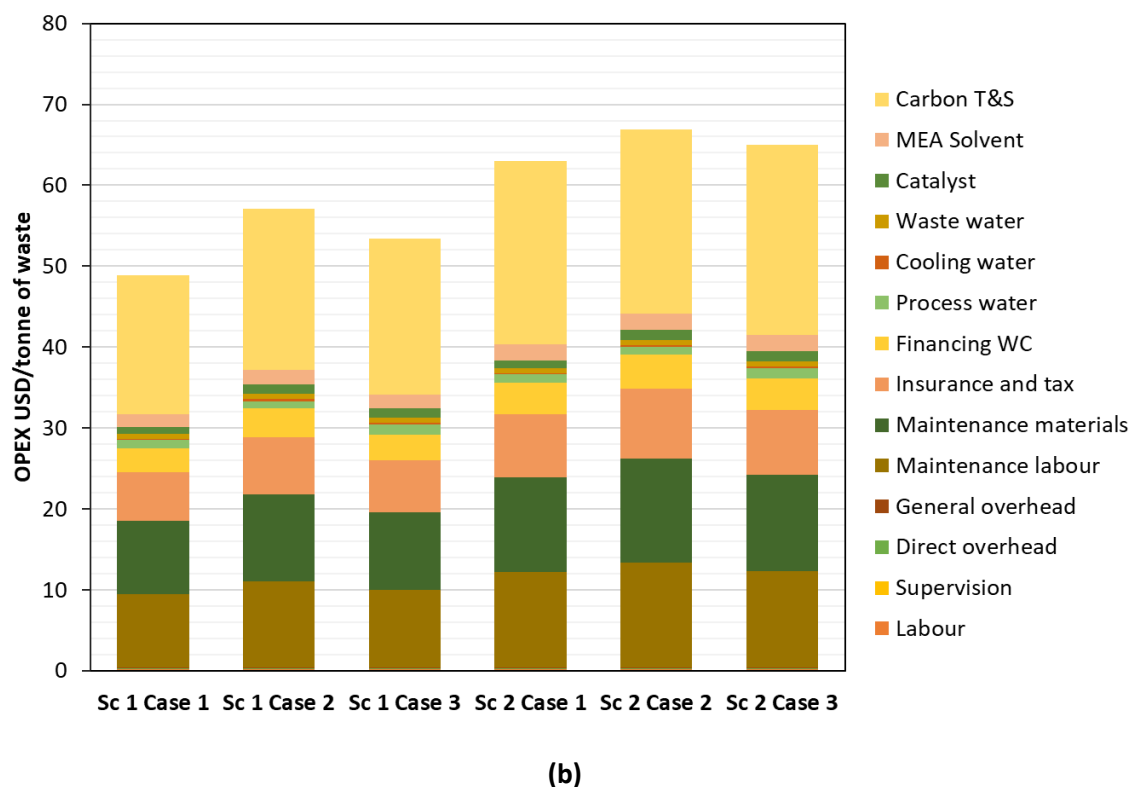


Figure 6-3. (a) PEC & (b) OPEX of Case 1, 2, and 3 under Scenarios 1 and 2.

In contrast, Case 1 has the lowest PEC, at 108 million USD in Scenario 1 and 140 million USD in Scenario 2, mainly due to the smaller plasma gasification unit and the absence of a SMR. Further, Case 1 compared to the other cases has the largest share of PEC attributed to the PSA unit, as a greater portion of the fuel gas stream is directed to PSA. This is because of the lower heat demand relative to the other two cases, which requires a smaller split fraction of the PSA feed gas for heating, thereby allowing more syngas to be processed for hydrogen recovery.

Case 3 shows a moderate PEC of 114 million USD in Scenario 1 and 142 million USD in Scenario 2, mainly resulting from the inclusion of the SMR unit and the corresponding increase in CCS capacity. The additional CO₂ to be captured arises from the conversion of biogenic methane into CO in the SMR, which is subsequently shifted to CO₂ in the WGSR. Importantly, a smaller portion of the syngas is routed to the PSA unit in Case 3, since a significant fraction of the gas is diverted and combusted to supply the thermal energy required for the SMR process. This reduces the PSA processing load and lowers its capital cost relative to Case 1.

Figure 6-3b compares the OPEX of three cases for both scenarios. Case 2 again exhibits the highest OPEX, at 57 USD per tonne of waste in Scenario 1 and 67 USD per tonne of waste in Scenario 2, driven by several contributing factors. The most significant is the FOM cost, which is directly influenced by the high CAPEX in this case. In addition, Case 2 incurs the highest cost for carbon transport and storage (T&S) because it captures the largest amount of CO₂, driven by the higher CO flow rate to the WGSR when mixed plastic and food waste are gasified. The consumption of MEA solvent is also highest, reflecting the greater CO₂ concentration in the process gas entering the carbon capture unit. Moreover, catalyst usage in the WGSR is elevated due to the higher flowrate of feed gas requiring treatment.

In contrast, Case 1 presents the lowest OPEX at 49 USD per tonne of waste in Scenario 1 and 63 USD per tonne of waste in Scenario 2, primarily because of its lower FOM, which correlates with the lowest CAPEX, and its minimal CO₂ capture requirement, given the simpler system configuration and absence of biogas reforming. Case 3 results in an intermediate OPEX of 53 USD per tonne of waste in Scenario 1 and 65 USD per tonne of waste in Scenario 2, reflecting a balance between system complexity and performance. Its FOM is lower than Case 2 due to moderate capital investment, and its CO₂ capture-related costs are reduced relative to Case 2 because of less CO₂ production and capture volume, despite the inclusion of the SMR. Across all cases, the annualized CAPEX (ACAPEX) is lower than OPEX (see Table 6-8), indicating that the system's financial structure is predominantly OPEX-driven.

Table 6-8. CAPEX, Annualized CAPEX (ACAPEX), and OPEX for Case 1, 2, and 3.

Parameters	Scenario 1			Scenario 2		
	Case 1	Case 2	Case 3	Case 1	Case 2	Case 3
CAPEX (MMUSD)	107.93	127.69	114.55	140.43	154.27	142.53
ACAPEX (MMUSD)	12.68	15.00	13.46	16.50	18.12	16.74
OPEX (MMUSD)	26.72	31.20	29.17	34.40	36.56	35.03

6.3.2.2 Levelized cost of hydrogen (LCOH)

The LCOH for the three evaluated cases in Scenario 1 and 2 is summarized in Table 6-9, expressed in USD per kilogram of hydrogen produced. In this study, the LCOH ranges from 3.52 to 13.36 USD/kg H₂, depending on the case configuration. Scenario 1 consistently yields lower LCOH values than Scenario 2 for all cases, reflecting the higher hydrogen production and lower internal energy demands in the pre-CCS configuration, as well as the reduced CCS equipment cost, which constitutes the largest cost component and has the strongest influence on the overall CAPEX.

Table 6-9. LCOH across three scenarios.

Parameter	Scenario 1			Scenario 2		
	Case 1	Case 2	Case 3	Case 1	Case 2	Case 3
LCOH (USD/kg)	3.52	8.34	4.33	6.16	13.37	7.44

Case 1 in both scenarios achieves the lowest LCOH, making it the most cost-effective configuration. This is largely due to its simpler design, which combines plasma gasification of plastic waste with anaerobic digestion of food waste, thereby avoiding the energy-intensive drying process required in Case 2. Furthermore, Case 1 incurs the lowest capital and operating costs due to reduced CO₂ capture requirements, lower overall energy demand, and the absence of SMR integration. When compared with literature [203], this LCOH is higher than the reported 2.11 USD/kg. The discrepancy arises mainly from differences in system scope: while the literature system considered only plasma gasification, the current study employs both plasma gasification and AD, as well as an internal CCGT unit to supply electricity, which increases system complexity and cost.

In contrast, Case 2 results in the highest LCOH. This is attributed to the large cost of the plasma gasifier. Such an approach requires a larger gasifier, greater solvent and catalyst consumption, an expanded CCS unit, and substantially more electricity, particularly for the plasma torch. These factors combine to produce the highest capital and operating expenditures among the three cases.

Case 3 yields an intermediate LCOH by integrating SMR of biogas with plasma gasification. Although the addition of SMR and heat integration increases capital cost and complexity, the reduced feedstock input to plasma gasification allows for a smaller gasifier, lower PEC, and reduced plasma torch electricity consumption. Consequently, less PSA feed gas is diverted for electricity generation, which lowers overall electricity demand and reduces the size and cost of the CCGT unit. These factors contribute to a markedly lower LCOH relative to Case 2, though it remains above that of Case 1. Overall, the results suggest that system simplicity, effective waste separation, and minimized energy consumption are critical drivers of cost-competitive hydrogen production.

Compared with blue hydrogen, produced via natural gas SMR with CCS at 1.4–3.2 USD/kg, the LCOH obtained in this study for the configuration including SMR (Case 3) is higher. This difference is primarily due to the greater process efficiency of natural gas reforming, which achieves around 76–77% hydrogen conversion efficiency [373] and the relative simplicity of its conversion pathway. By contrast, the present system attains overall waste-to-hydrogen efficiencies of only 36% and 28% in Case 3 for Scenario 1 and Scenario 2, respectively (Table 6-7).

Beyond fossil-based pathways such as blue hydrogen, electrolysis-based hydrogen production is also included for reference, with an LCOH range of 3.20–7.70 USD/kg H₂ [236]. While this range partially overlaps with the lowest cost achieved in the present study (Case 1), it remains more competitive than Case 2 and broadly comparable to Case 3. It should be emphasized, however, that electrolysis is highly sensitive to both electricity price and carbon intensity of the power supply, and generally provides limited waste valorization potential compared with plasma gasification-based pathways.

In summary, although the LCOH reported here is higher than some benchmark technologies, the integrated, multi-feedstock configuration analyzed in this study offers the added benefits of waste reduction and carbon mitigation. These co-benefits may justify the higher cost under sustainability-oriented policy frameworks.

6.3.2.3 Sensitivity analysis on LCOH

Figure 6-4 illustrates the sensitivity of the LCOH to variations in capital costs, financial parameters, and incentive-related revenues across the three cases in Scenarios 1 and 2. Comparing the two scenarios, Scenario 2 consistently exhibits longer bars for the key parameters than Scenario 1, indicating that LCOH in the post-CCS configuration is more sensitive to parameter uncertainty. This is because Scenario 2 has higher CAPEX and OPEX and lower hydrogen outputs. Case 2 in both scenarios show the greatest sensitivity because, unlike Cases 1 and 3 where food waste is treated by AD and only plastic waste enters the plasma gasifier, both plastic and food waste are directly processed in the plasma gasifier. This configuration requires a larger plasma gasifier to accommodate the higher input load, a larger CCGT to meet the increased electricity demand, and a larger CCS unit to capture the greater volume of CO₂, thereby amplifying cost sensitivity.

In all cases, the PEC of CCS shows the strongest influence, followed by the PEC of the PG and CCGT. In Scenario 1, varying PEC of CCS shifts LCOH from 3.00 to 4.38 USD/kg (Case 1), 7.15 to 10.35 USD/kg (Case 2), and 3.69 to 5.39 USD/kg (Case 3). In Scenario 2, the same parameter produces an even larger effect, increasing LCOH from 5.14 to 7.86 USD/kg (Case 1), 11.20 to 16.97 USD/kg (Case 2), and 6.21 to 9.49 USD/kg (Case 3). The next most influential technical contributors are typically PEC of PG, and PEC of CCGT.

Financial assumptions also play a significant role. Variations in the discount rate produce a much broader fluctuation in LCOH than changes in the tax rate; for example, in Scenario 1 Case 1, varying the discount rate changes LCOH from 3.12 to 3.95 USD/kg, compared with 3.14 to 3.73 USD/kg for the tax rate. This trend indicates that project viability is highly sensitive to financing conditions. By contrast, incentive factors such as carbon credits and tipping fees reduce LCOH by less than 0.16 USD/kg and 0.20 USD/kg in Scenario 1 and 2, respectively, highlighting their limited impact on overall economics. It should be noted that the baseline values for carbon credits and tipping fees are derived from current Indonesian market prices, and their influence on LCOH could become more significant if higher incentive levels were applied.

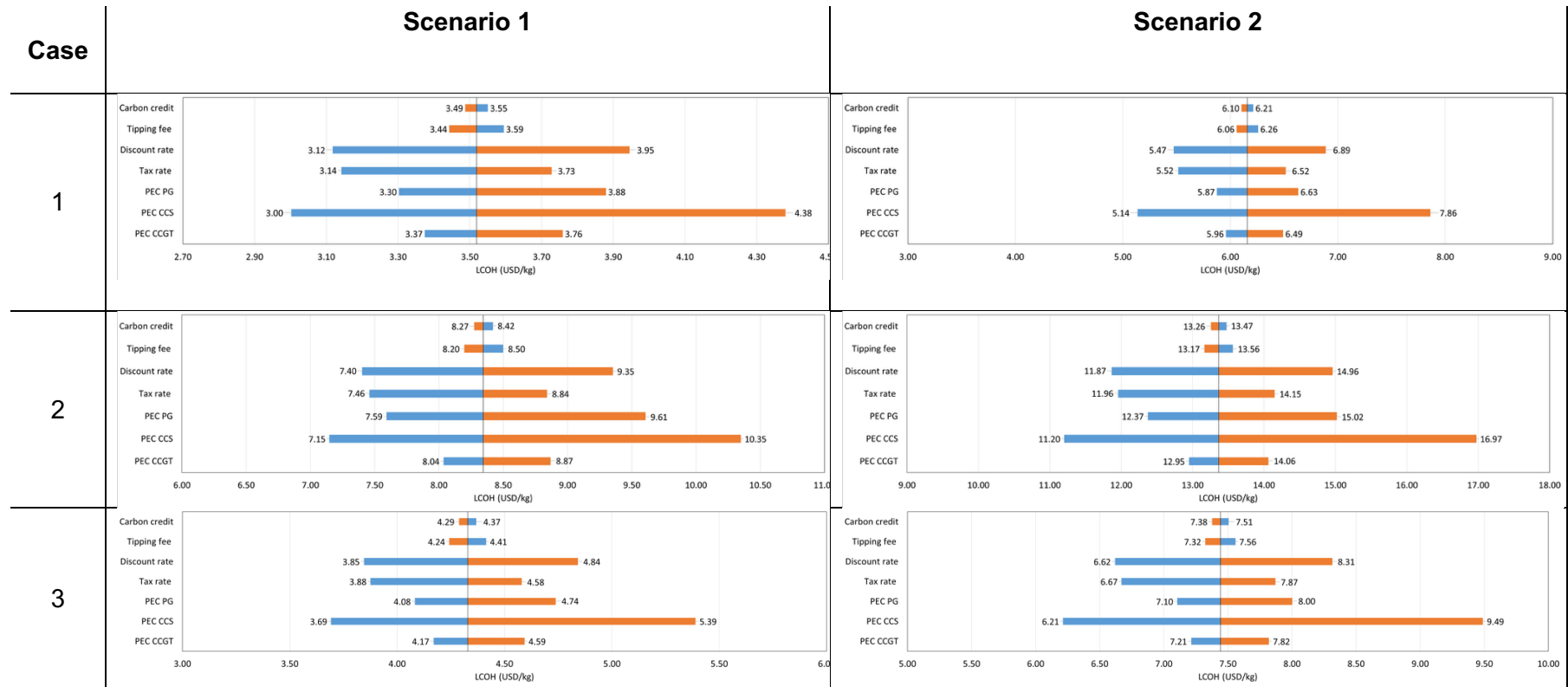
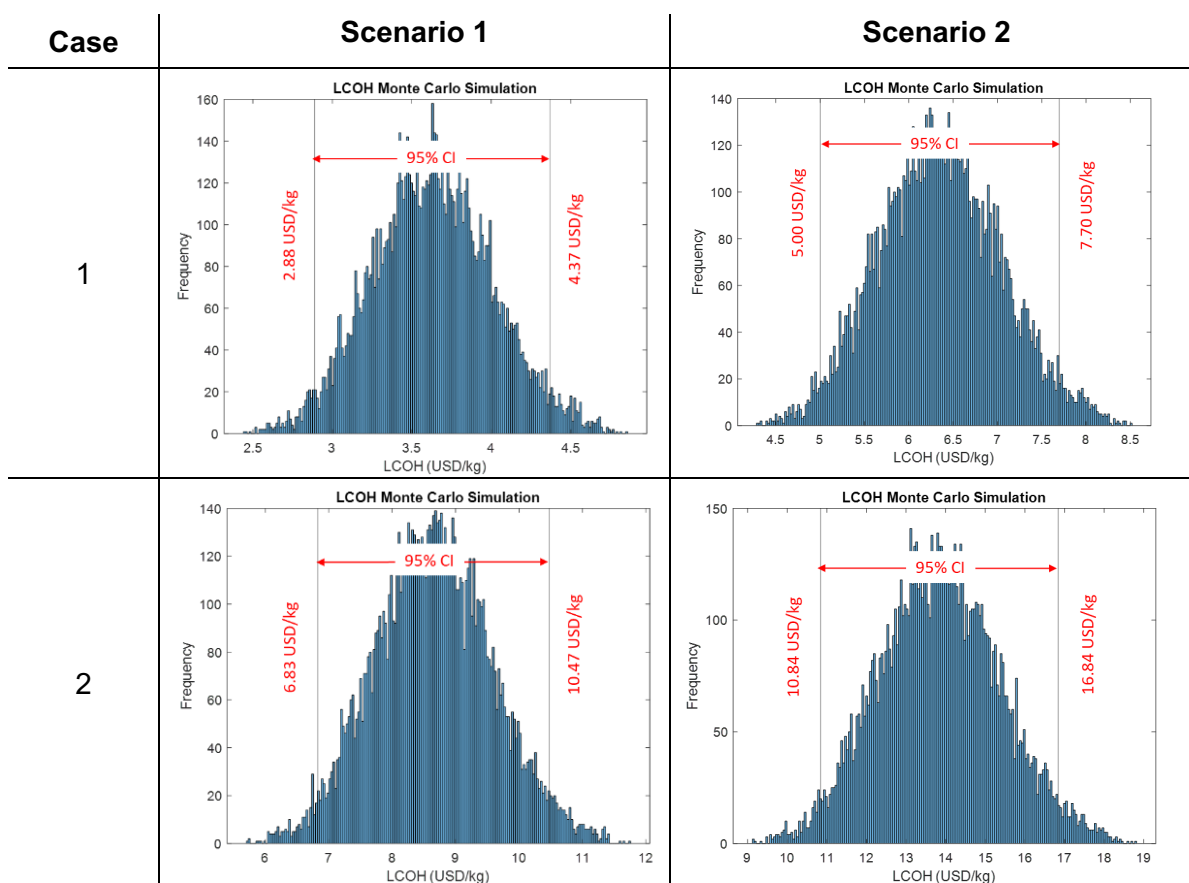


Figure 6-4. Sensitivity of the LCOH to various economic variables across three cases in Scenario 1 and Scenario 2.

6.3.2.4 Uncertainty analysis on LCOH

Uncertainty analysis was undertaken to quantify the effect of joint variation in all uncertain inputs on the economic criterion (LCOH). A Monte Carlo simulation implemented in MATLAB repeatedly evaluates the economic model while sampling the parameters listed in Table 6-6 from their specified probability distributions. For each scenario, 10,000 iterations are executed, yielding an empirical distribution of LCOH from which summary statistics (mean, median, standard deviation) are computed. The uncertainty range is reported as the central 95% interval of the simulated LCOH distribution.



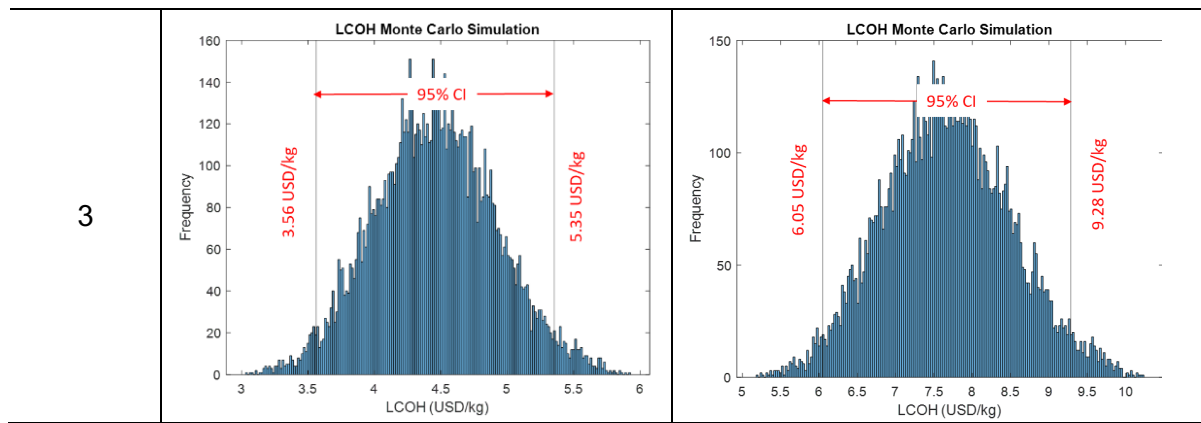


Figure 6-5. Monte Carlo simulation results for the LCOH across three cases in Scenario 1 and Scenario 2.

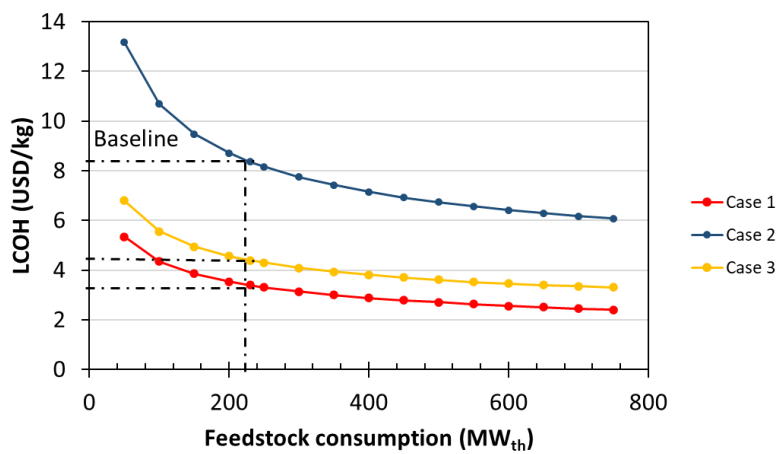
Figure 6-5 presents the Monte Carlo simulation results for the LCOH across the three cases in Scenarios 1 and 2. The figures indicate that Scenario 1 exhibits a narrower LCOH distribution (smaller standard deviation) than Scenario 2 for all cases. This arises from both lower cost intensity (lower CAPEX and OPEX) and higher hydrogen output in Scenario 1, which make the LCOH less sensitive to parameter uncertainty.

In all cases, the mean LCOH values obtained from the Monte Carlo analysis are higher than those estimated in the deterministic economic assessment, reflecting the impact of input uncertainty. Case 1 demonstrates the lowest and most stable cost outcome; its relatively narrow distribution indicates low uncertainty, reflecting the benefits of a simpler system configuration with plasma gasification of plastic waste combined with anaerobic digestion of food waste.

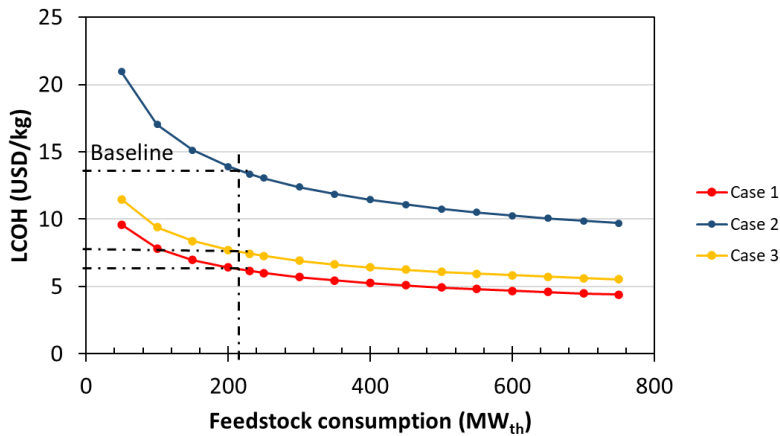
In contrast, Case 2 shows the highest and most uncertain LCOH. The broader spread arises from the need for larger plasma gasification, CCGT, and CCS units due to the absence of waste separation, which drives up both cost and variability. Case 3 yields an intermediate result. While SMR integration enhances hydrogen yield, it also increases cost variability compared to Case 1. Overall, Case 1 is the most economically favorable and least uncertainty-sensitive pathway, Case 2 is the least competitive, and Case 3 is intermediate on both cost and uncertainty.

6.3.2.5 Economies of scale

The economies of scale were evaluated by varying the feedstock thermal input from 50 to 750 MW_{th} and computing the resulting LCOH. CAPEX was rescaled using component-specific scaling factors (f), as outlined in Table 6-4. Fixed OPEX components were assumed proportional to CAPEX. Labor costs, while initially estimated from the number of equipment items (Eq. 3-20), were further adjusted to scale with plant capacity to the power of 0.242 to reflect economies of scale, following the correlation proposed by Peters et al. [257]. Variable OPEX components were scaled linearly with feedstock consumption.



(a)



(b)

Figure 6-6. Economies of scale across the three scenarios.

Figure 6-6 illustrates the relationship between feedstock consumption and the corresponding LCOH for each case in Scenarios 1 and 2. In all cases in both scenarios, LCOH declines with increasing feedstock consumption, reflecting

economies of scale. The magnitude of the scale effect differs across cases. Case 2, processing mixed waste entirely via plasma gasification, exhibits the steepest initial decline due to its high capital and power intensity (larger PG, CCGT, and CCS), yet remains the most expensive across the range. Case 1 is consistently the least costly and displays a gentler slope, indicating the lowest capital and energy intensity enabled by waste separation and the absence of SMR. Case 3 is intermediate in both level and slope. From the graph, even at the feedstock consumption of 700-750 MW_{th}, Case 2 remains the most expensive because processing mixed waste via plasma gasification requires larger, more energy intensive PG, CCGT, and CCS units, thus scaling cannot close the gap with the separated-waste configuration. While economies of scale provide substantial early reductions in LCOH, the cost decreases more slowly beyond ~650 MW_{th}, as evidenced by reduced slope of the curves at high feedstock composition. This means that increasing plant capacity further still lowers LCOH, but the benefit becomes smaller at larger capacities.

6.3.3 Environmental performance

The Life Cycle Inventory (LCI) data were processed using SimaPro software to quantify the contributions to various midpoint impact categories. The developed inventory based on the mass and energy balance which has been normalized on a basis of 1 kg of waste can be seen in the Sub Chapter 9.3.1. The LCA is performed by evaluating the input and output flows within the system boundary illustrated in Figure 6-1, without modelling the internal processes.

In this section, GWP and WF are discussed as the key midpoint indicators. Additional impact indicators are available in the Sub Chapter 9.3.2. All impact category results are normalized to one kilogram of waste processed within the system.

6.3.3.1 Global warming potential (GWP)

Table 6-10 summarizes total GWP across the two scenarios. To aid interpretation, Figure 6-7 and Figure 6-8 illustrate the carbon inflows and outflows within the defined system boundary for each scenario. According to the LCA results, Scenario

1 has lower GWP than Scenario 2. Although both pre and post CCS configurations assume the same CO₂ capture fraction (95%) in the CCS unit, their overall carbon removal performance differs. In Scenario 1, the CCS unit is placed downstream of the WGSR and treats the shifted gas. The CO₂-lean gas leaving the pre-CCS section still contains carbon and does not entirely proceed to the PSA. Instead, it is split, with one fraction sent to the PSA for hydrogen purification and the remainder mixed with the PSA off-gas and combusted in the CCGT and auxiliary burners supplying heat for food waste drying (Case 2), and SMR (Case 3). The resulting flue gases are released directly to the atmosphere without further CO₂ capture.

In Scenario 2, by contrast, no pre-CCS is applied to the shifted gas, but the combined flue gases from the CCGT and the other combustion units are routed through a post-CCS unit before discharge. Consequently, almost all of the carbon oxidised during power and heat generation passes through the CCS unit, so the fraction of total system carbon that is actually captured is higher than in Scenario 1, even though the capture efficiency of each individual CCS unit is the same (95%). This leads to lower GWP values for Scenario 2 and also results in smaller variation between cases, with GWP values ranging from approximately -0.159 to -0.155 kgCO₂-eq/ kg of waste treated. In Scenario 1, by contrast, the overall GWP is more sensitive to the upstream conversion routes prior to the WGSR, and thus varies more widely, from -0.038 to 0.048 kgCO₂-eq/ kg of waste treated. The corresponding amounts of CO₂ captured in the CCS plant per kg of waste treated are reported in Table 6-7.

In both scenarios, Case 1 demonstrates the highest GWP, whereas Cases 2 and 3 both achieve net-negative GWP values, with Case 2 exhibiting the most favourable performance. The elevated GWP in Case 1 is primarily attributed to the absence of SMR for biogas. As a result, a significant fraction of methane (CH₄) remains unconverted, and thus does not contribute to additional CO or CO₂ formation via the WGSR. While all cases rely on combustion to supply heat and electricity, Case 1 results in significantly higher direct CO₂ emissions due to the greater amount of carbon-containing fuel combusted. The most notable source is the combustion process within the CCGT system. This is confirmed by the carbon balance results (see Figure 6-7 and Figure 6-8), which show substantially higher carbon flow in the

CCGT flue gas for Case 1 than for Cases 2 & 3 in both scenarios. In addition, the unreformed CH₄ in Case 1 reduces the amount of CO converted and subsequently captured, contributing the lowest CO₂ capture among the three cases.

Table 6-10. Total GWP from SimaPro for Cases 1-3 under Scenario 1 (pre-CCS) and Scenario 2 (post CCS), expressed per 1 kg of waste treated.

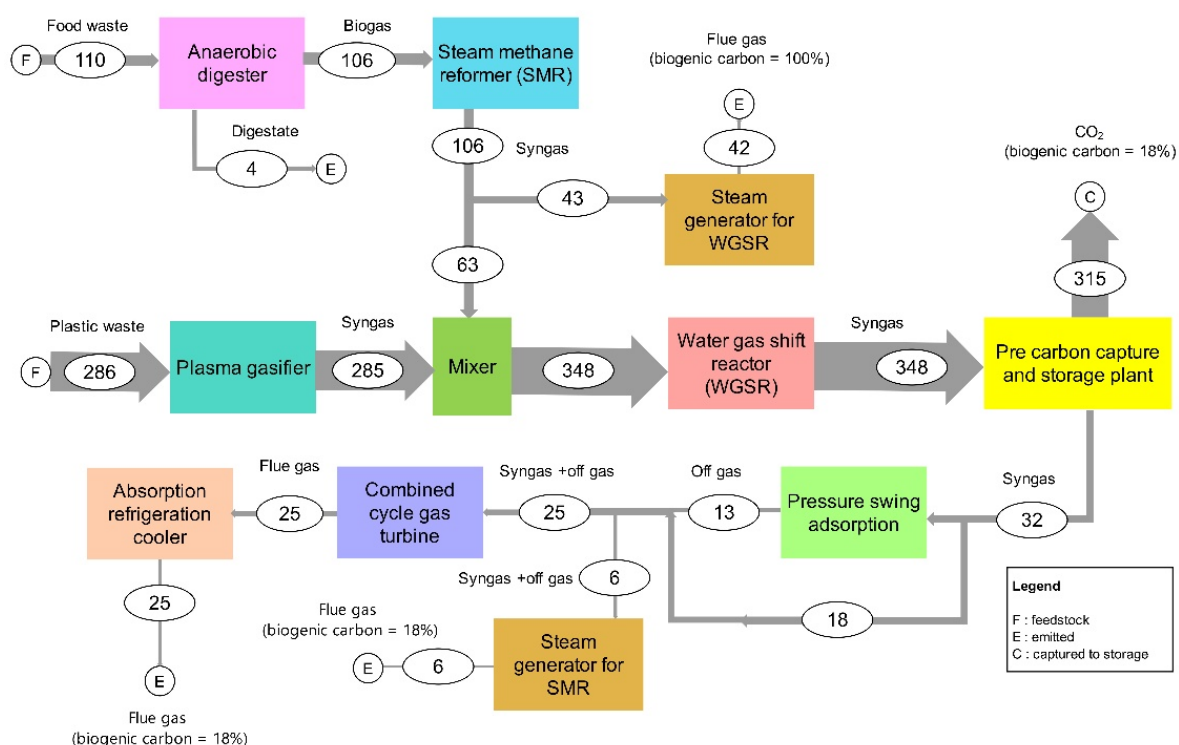
Parameter	Scenario 1			Scenario 2		
	Case 1	Case 2	Case 3	Case 1	Case 2	Case 3
Total GWP (kgCO ₂ -eq/ kg of waste treatment)	0.048	-0.038	-0.021	-0.155	-0.159	-0.155

Case 2 in both scenarios achieves the lowest GWP, primarily due to the high CO content in the raw syngas produced from the plasma gasification of mixed food and plastic waste. This CO is effectively converted to CO₂ via the WGSR and subsequently captured. In Case 3, biogas is processed through SMR, but the CH₄ conversion efficiency is limited to 89% to avoid excessive thermal input requirements. Improving conversion would require more combustion of feed gas and PSA off-gas, reducing hydrogen recovery and increasing emissions. As a consequence, residual CH₄ is carried into the fuel stream for the CCGT, where its combustion results in higher CO₂ emissions. Furthermore, the incomplete CH₄ conversion in the SMR in Case 3 limits CO formation, which in turn reduces the amount of CO₂ generated via the WGSR, thereby lowering the CO₂ yield entering the carbon capture stage. This effect is also evident in Figure 6-7 and Figure 6-8, where the CO₂ captured flow rate in Case 3 is 315 mol/s, slightly lower than the 327 mol/s captured in Case 2.

Compared to the literature, the study by S. Chari et al. [221] investigated hydrogen production from mixed plastic waste through the integration of plasma gasification, WGSR, carbon capture using 30% MEA, a gas turbine cycle, and PSA. Their LCA results reported a GWP of 1.453 kg CO₂-eq per kilogram of waste processed. In contrast, the present study demonstrates a lower GWP, particularly in Cases 2 and 3. The primary factor contributing to this difference is the absence of biogenic carbon in the feedstock used by Chari et al. [221], which consisted entirely of plastic waste. In the current work, the feedstock is a mixture of plastic and food waste, and the biogenic carbon content originating from the food waste allows a portion of the

Another relevant study by Z. Zhang et al. [250] examined hydrogen production from a combination of biomass and coal and evaluated its environmental impact. Their LCA reported a GWP of 11 kg CO₂-eq per kg of biomass processed. The system configuration included plasma gasification, a water-gas shift reaction, acid gas removal using propylene carbonate, and PSA for hydrogen purification.





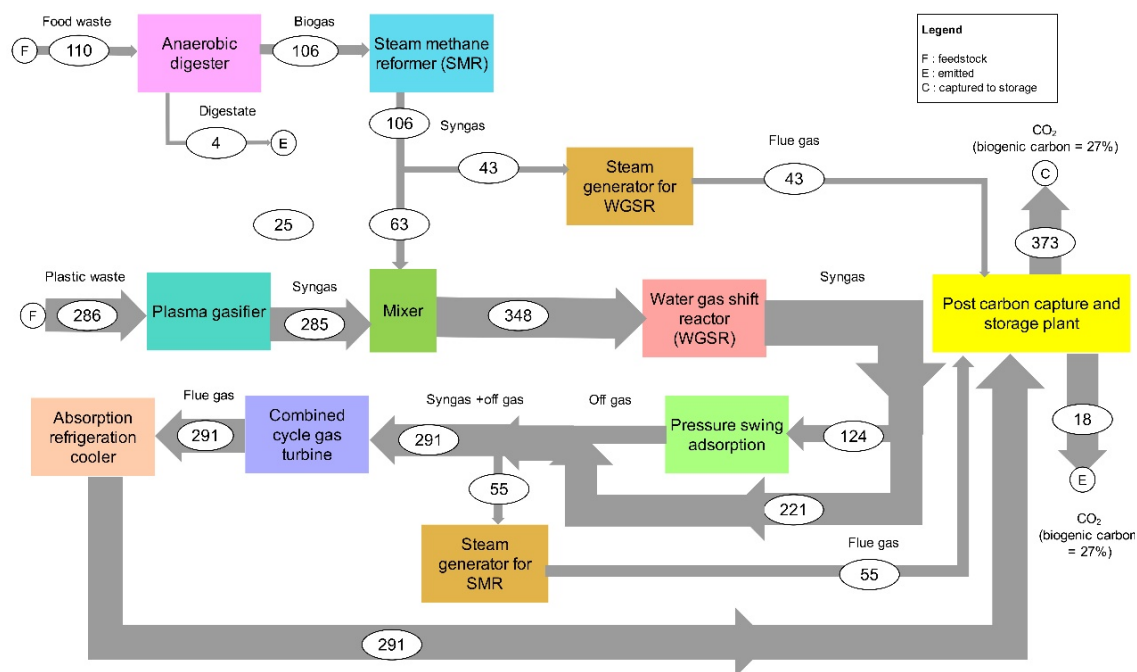
(c)

Figure 6-7. Carbon balance for the proposed H₂ production cases in Scenario 1, accounting for all carbon containing streams: (a) Case 1, (b) Case 2; (c) Case 3.

The substantial difference between their GWP result and the findings in the present study is mainly explained by differences in the system boundaries. In Zhang et al.'s work [250], the system boundary extends from raw material production and transportation to hydrogen transportation. By contrast, in the present study, the plant is assumed to be located on or near a landfill site, eliminating the need for dedicated raw material transport, and hydrogen transportation is not included within the boundary. Feedstock composition may also contribute to the difference: Zhang et al. [250] used coal (fossil-based) mixed with biomass, though the proportion was not explicitly stated, while the present study uses plastic (fossil-derived) and food waste.

Additionally, Zhang et al. [250] did not account for the environmental credit from biogenic CO₂ capture, whereas the current study does. This credit, associated with the permanent storage of CO₂ originating from food waste, contributes significantly to the net reduction in GWP across the three cases evaluated in this study.





(c)

Figure 6-8. Carbon balance for the proposed H₂ production cases in Scenario 2, accounting for all carbon containing streams: (a) Case 1, (b) Case 2; (c) Case 3.

6.3.3.2 Water footprint (WF)

Table 6-11 reports the water footprint per kilogram of waste treated. In both scenarios, Case 1 exhibits the lowest WF, whereas Cases 2 and 3 are higher. The primary driver is cooling water demand required to condition the syngas leaving the WGSR to the CCS absorber inlet temperature. After the WGSR, the hot wet gas is first heat-recovered to generate LP steam for the desorber, yet, even after this recovery, the gas remains at ~150 °C. For amine absorption, the CCS feed must be ~40 °C to limit solvent degradation. In Case 1, only the plasma gasifier syngas passes through the WGSR, while the biogas mixed with it after WGSR. This mixing dilutes the hot stream, lowering the cooling water requirement. In Case 2, syngas from mixed plastic and food waste gasification passes through WGSR, so the entire pre-CCS feed is hotter, yielding the highest cooling water demand. In Case 3, the addition of SMR steam ($S/C = 3$) increases moisture content of the gas before the pre-CCS, raising the cooling load relative to Case 1, but remaining below Case 2.

Table 6-11. Total WF from SimaPro for Cases 1-3 under Scenario 1 (pre-CCS) and Scenario 2 (post CCS), expressed per 1 kg of waste treated.

Parameter	Scenario 1			Scenario 2		
	Case 1	Case 2	Case 3	Case 1	Case 2	Case 3
Total WF (m ³ /kg of waste treatment)	0.005430	0.007420	0.006620	0.005200	0.007880	0.006420

Water use in the present study is lower than values reported in the literature. For example, Materazzi et al. [352] report a water footprint of 0.0717 m³/kg MSW for a configuration integrating a fluidized bed gasifier, a WGSR, a pre-CCS plant, a PSA, and a tail gas engine. The discrepancy likely reflects differences in treatment and recycling of condensate and cooling-system blowdown, and the assumed cooling-tower cycles of concentration.

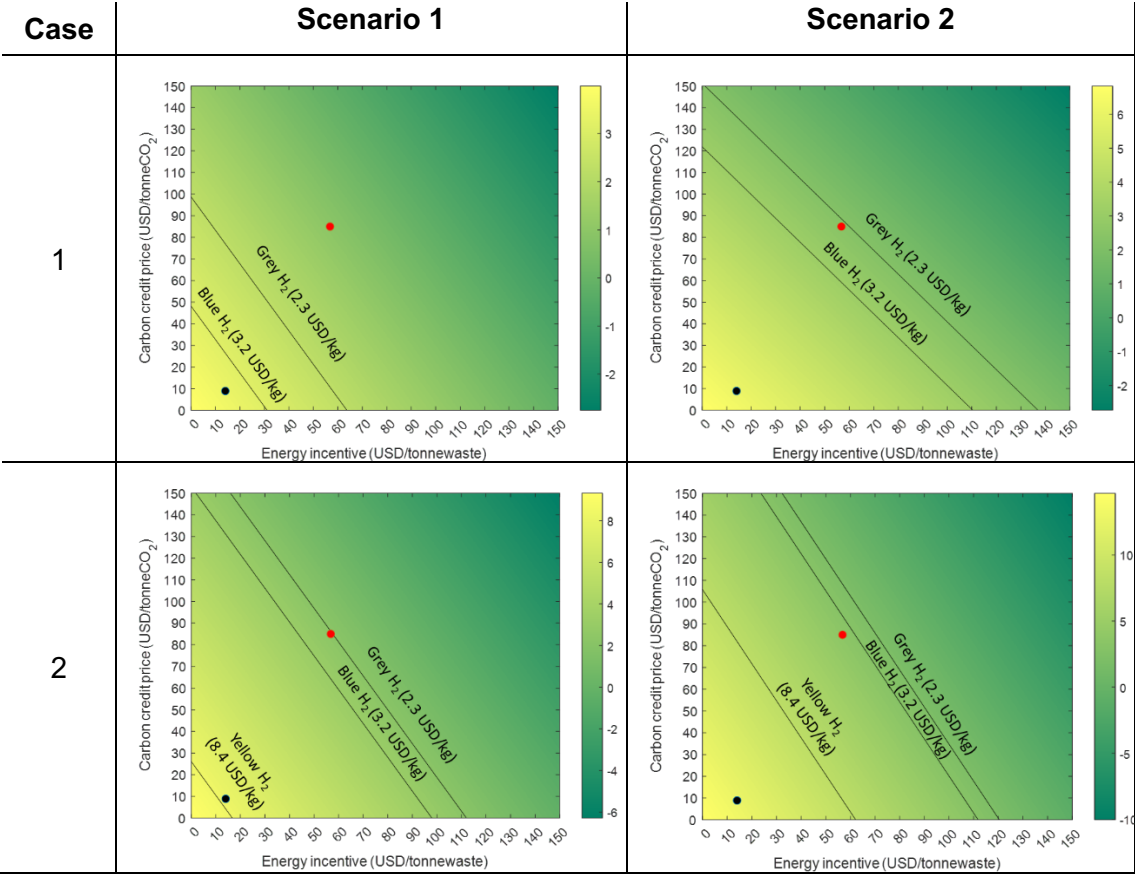
6.3.4 Policy incentives analysis

Figure 6-9 maps the LCOH as a function of two policy variables: the carbon credit price (USD/tonne CO₂) and energy incentive (tipping fee) (USD/tonne waste) for the three cases in Scenarios 1 and 2. As shown, increasing support leads to a reduction in LCOH. The lines mark reference cost levels for alternative hydrogen pathways (grey, blue, yellow): grey and blue correspond to natural gas SMR without and with CCS, respectively, while yellow refers to electrolysis. The black dot denotes current Indonesian levels (carbon credit price of 8.9 USD/tonne CO₂ [328]; the tipping fee reaches 14.02 USD/tonne of waste [327]). At this level, none of the cases reaches cost equivalence with the grey/blue benchmarks: Cases 1 and 3 lie closest to the blue hydrogen level, whereas Case 2 sits nearer the yellow hydrogen. This implies that to become cost competitive, Indonesia would require higher carbon credits and/or higher tipping fees. If support is constrained, Case 1 is the most promising candidate because it requires the least additional support.

The red dot marks a USA benchmark case (85.0 USD/tonne of CO₂ [356]; 56.8 USD/tonne of waste [357]), included because the USA hosts a mature carbon-credit market, robust GHG policy, and widely cited best practices in municipal waste management [355], [357]. At this benchmark, Cases 1 and 3 fall below the grey hydrogen cost benchmark, whereas Case 2 remains between the grey and blue hydrogen. Thus, USA benchmark incentives would be sufficient to enable Cases 1

and 3 cost competitive with grey hydrogen, but insufficient for Case 2, which would still require additional support or internal cost reductions.

Comparing the two scenarios, the contour maps for Scenario 2 are systematically shifted towards higher LCOH values relative to Scenario 2. For a given combination of carbon credit and energy incentive (e.g. the Indonesian or USA benchmarks), Scenario 2 always yields a higher LCOH than Scenario 1, indicating that the post-CCS configuration requires stronger policy support to achieve the same competitiveness level. Consequently, cost equivalence with grey or blue hydrogen is achieved at lower incentive levels in Scenario 1 than in Scenario 2.



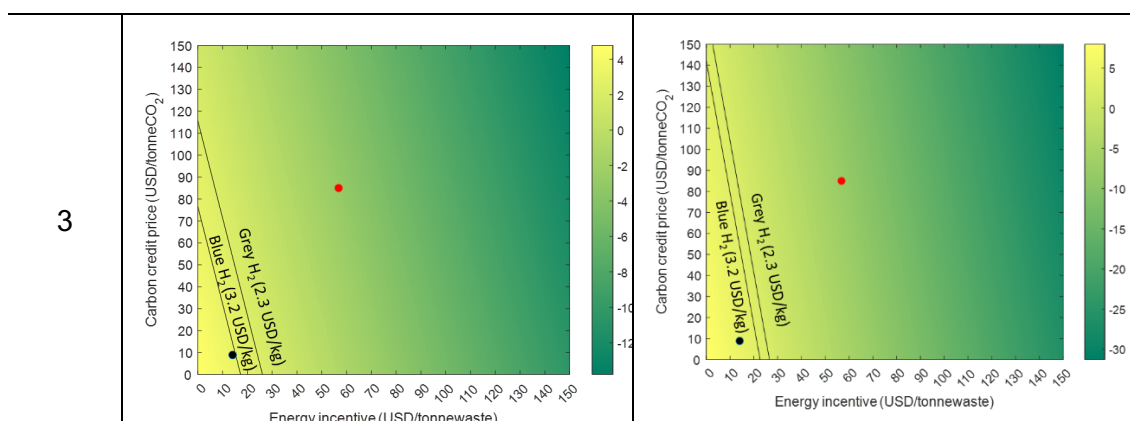


Figure 6-9. Effect of policy incentives on LCOH: heat maps show LCOH (USD/kg) versus carbon credit price (USD/tonne CO₂) and energy incentive (USD/tonne waste) for (a) Case 1, (b) Case 2, (c) Case 3 with cost benchmarks for grey/blue/yellow hydrogen and markers for the Indonesia baseline (black dot) and USA benchmark (red dot).

6.3.5 Practical deployment

The proposed system is designed as a retrofit-expansion rather than a fully greenfield development. It is based on the existing WtE facility at Benowo landfill, Surabaya (Indonesia's second-largest city), where current operations are limited to electricity and heat generation from a gasification-based WtE process and a landfill-gas power plant. The gasification unit treats approximately 1000 tonnes of mixed waste per day and generates around 9 MW of electricity, while the landfill-gas plant treats around 500 tonnes of waste per day and generates approximately 2 MW of electricity [374], [375]. Building on the existing infrastructure (waste handling, grid connection, utilities, and operational experience), the present work proposes an upgrade pathway that expands product outputs beyond electricity/heat to include hydrogen and CCHP.

Compared with existing commercial WtE technologies, the proposed configurations are more complex and capital-intensive because they integrate multiple subsystems, including plasma gasification, anaerobic digestion, CCGT, product upgrading, and CCS. Therefore, practical deployment is more likely to occur through phased expansion of an existing facility than the immediate implementation

of the full integrated system. This staged approach is more consistent with current industrial practice and improves the scalability of the concept under real operating conditions.

A key additional challenge is that the current Benowo WtE facility does not include CO₂ capture and more broadly, Indonesia still has limited mature CO₂ transport and storage infrastructure. Nevertheless, Indonesia has been reported to have substantial CO₂ storage potential (approximately 577.6 Gt, dominated by saline aquifers (~572.8 Gt) with additional capacity in depleted oil and gas reservoirs (~4.8 Gt) [376]. A candidate storage location relatively close to Surabaya is the Bawean Basin (~200 km), which has been identified as a deep saline aquifer storage option (reported capacity ~1.16 Gt), although it is not yet in operation [377]. Given that CO₂ transport and storage infrastructure is still emerging in Indonesia, CCS could be implemented as a retrofit step once a viable transport and storage pathway becomes available. Potential enablers include collaboration with national energy/industrial stakeholders to develop shared CO₂ infrastructure and the use of international partnerships and climate-finance mechanism to support detailed engineering design, monitoring frameworks, and early CCS projects in Indonesia.

It is also important to note that the economically preferred configuration does not necessarily coincide with the environmentally preferred one. The baseline cases are generally more attractive in terms of cost because they avoid the additional CAPEX and OPEX associated with CCS integration, whereas the CCS-integrated cases, particularly post-CCS, achieve superior environmental performance through lower direct CO₂ emissions and GWP. This indicates that real world deployment will depend not only technical feasibility, but also on whether policy support, carbon pricing, or climate finance mechanisms are available to bridge the gap between economic and environmental optima.

6.3.6 Integrated comparison of TEA-LCA performance

Table 6-12 shows the comparative summary of key performance indicators (KPIs) for Cases 1, 2, and 3 in Scenarios 1 and 2, highlighting technical, economic, and environmental trade-offs. The reported TEA/LCA results are specific to the assumed waste composition and operating conditions; however, the modelling

framework can be readily adapted to other MSW contexts by updating feedstock properties and key operating parameters. Across both scenarios, Case 1 demonstrates the most balanced performance, achieving the highest hydrogen yield and the highest overall energy and exergy efficiencies. It also delivers the lowest LCOH, driven by lower CAPEX and OPEX requirements. However, its environmental benefit is limited, since methane in the biogas remains unreformed into CO and H₂.

Table 6-12. KPIs of three cases in Scenarios 1 and 2.

KPI	Scenario 1			Scenario 2		
	Case 1	Case 2	Case 3	Case 1	Case 2	Case 3
Hydrogen yield (kg/s)	0.70	0.35	0.62	0.53	0.27	0.45
Energy efficiency (%)	41.30	23.45	36.49	32.96	19.34	28.84
Exergy efficiency (%)	33.85	17.11	29.71	25.74	13.21	22.26
CO ₂ captured (kg/kg waste)	0.66	0.76	0.73	0.86	0.87	0.86
CAPEX (MMUSD)	107.93	127.69	114.55	140.43	154.27	142.53
OPEX (MMUSD)	26.72	31.20	29.17	34.40	36.56	35.03
LCOH (USD/kgH ₂)	3.52	8.34	4.33	6.16	13.37	7.44
GWP (kgCO ₂ -eq/kg waste)	0.048	-0.038	-0.021	-0.155	-0.159	-0.155

Case 2 is technically the simplest but performs the worst across most KPIs. Due to co-processing mixed plastic and food waste, it requires the largest gasifier and CCS unit, leading to the highest CAPEX and OPEX. As a result, it has the lowest efficiencies and the highest LCOH. On the other hand, it achieves the best climate performance, owing to the high CO content in the raw syngas produced from the plasma gasification of mixed food and plastic waste, which is effectively converted to CO₂ via the WGSR and then captured.

Case 3 provides an intermediate option. It achieves better efficiencies compared to Case 2, due to lower electricity and heat demand. The reduced internal energy requirement allows a larger fraction of the PSA feed gas to be directed to hydrogen purification rather than being combusted to meet process energy needs. Its environmental performance is favourable, and CO₂ capture is relatively high.

Economically, it delivers a moderate LCOH, with CAPEX and OPEX values between Cases 1 and 2.

Comparing the two scenarios, Scenario 1 (pre-CCS) consistently shows higher hydrogen yields and higher energy and exergy efficiencies than Scenario 2 for the corresponding cases, together with lower CAPEX, OPEX, and LCOH. Scenario 2 (post-CCS), by contrast achieves greater climate benefits: the specific CO₂ captured per kilogram of waste is higher, and the GWP values are more negative in all cases. Thus, Scenario 1 is more favourable from a techno-economic perspective, whereas Scenario 2 provides stronger environmental performance at the expense of higher cost.

Overall, Case 1 is the most economically viable under limited policy support. Case 2 offers the strongest environmental benefit but at prohibitive cost, and Case 3 represents a balanced compromise, offering carbon mitigation benefits at a more moderate cost than Case 2

6.4 Conclusion

This study evaluated an integrated waste to hydrogen and CCHP system with CCS under Indonesia-representative conditions, comparing three cases, i.e., separate treatment of plastics and food waste; mixed co-processing; and biogas upgrading via SMR under two CCS configurations (pre- and post-CCS). Case 1 (plastics to plasma gasification and food waste to anaerobic digestion) demonstrated the most balanced performance across both CCS scenarios, achieving the highest hydrogen output and the highest energy/exergy efficiencies, while also delivering the lowest LCOH. This improved performance is primarily attributed to lower internal energy requirements and reduced equipment sizing.

From a techno-economic perspective, Scenario 1 (pre-CCS) consistently outperformed Scenario 2, delivering higher hydrogen production and lower CAPEX and OPEX; as a result, the LCOH ranged from 3.52-8.34 USD/kgH₂ across Cases 1-3, compared with 6.16-13.37 USD/kgH₂ in Scenario 2. From a carbon mitigation perspective, Scenario 2 (post-CCS) achieved higher CO₂ capture and more negative GWP values. Specifically, the amount of CO₂ captured increased from

0.66-0.76 kg/kg waste in Scenario 1 to 0.86-0.87 kg/kg waste in Scenario 2. Consequently, the GWP shifted from -0.038 to 0.048 kgCO₂-eq/kg waste in Scenario 1 to approximately -0.159 to -0.155 kgCO₂-eq/kg waste in Scenario 2. Overall, these results highlight a clear trade-off: pre-CCS offers advantages in terms of cost and process efficiency, whereas post-CCS provides stronger climate benefits at higher cost.

Sensitivity and uncertainty analyses showed that the purchased equipment cost for CCS is the dominant contributor to LCOH across all configurations, highlighting CCS cost reduction and improved heat integration as key levers for improving competitiveness. Under current Indonesian carbon credit and tipping-fee levels, none of the configurations reaches cost competitiveness with grey or blue hydrogen; higher incentives and/or technology cost reductions are required, particularly for the mixed-feed case.

This work demonstrates that integrating plasma gasification for plastics with anaerobic digestion for food waste, coupled with CCS, represents a feasible pathway for converting MSW into hydrogen while reducing climate impacts. Future work should focus on pilot-scale demonstration at landfill/WTE sites, improved feedstock separation strategies, and CCS cost optimization to support practical deployment in Indonesia.

7 CONCLUSION AND FUTURE WORK

7.1 Importance and contribution of research

The three studies address the technical, economic, and environmental dimensions of hybrid WTE pathways, with specific relevance to Indonesia but also transferable to global contexts.

1. Novel system integration and technical contribution

- Demonstrates, for the first time, the thermodynamic feasibility of coupling PG and AD with CCS for the co-production of CCHP, liquid biomethane, and hydrogen.
- Highlights the role of waste separation strategies in enhancing efficiency and hydrogen yield.
- Establishes benchmarks for system performance, achieving energy efficiencies up to ~49% and identifying scenarios where CCS integration reduces CO₂ emissions to near zero.

2. Economic and environmental insights

- Provides comprehensive TEA and LCA, quantifying LCOE, LCO_bM, LCOH, and LCOWT under different configurations.
- Shows that while CCS integration raises costs, it achieves substantial GHG mitigation, in some cases reaching negative net emissions.
- Demonstrates the influence of key cost drivers (e.g. gas turbine, plasma gasifier, CCS unit) through sensitivity and Monte Carlo analysis, offering guidance for cost reduction priorities.
- Establishes the environmental trade-offs of cooling water demand, carbon emissions from CCGT, and methane management in AD-SMR pathways.

3. Policy and practical relevance

- Provides evidence-based insights for the policy frameworks such as carbon credits, and tipping fees, showing their limited but important role in improving economic feasibility.
- Offers a pathway for Indonesia to address municipal waste challenges while progressing towards net-zero targets and SDG commitments (especially SDG 11 on sustainable cities).

- Suggest that international policy environments (e.g., USA benchmarks) could bring hybrid WTE-hydrogen systems closer to cost-competitiveness with grey and blue hydrogen.

4. Academic contribution

- Advances the state of the art in hybrid WTE-CCS systems, broadening the research frontier beyond single-technology studies.
- Creates a foundation for pilot-scale demonstration projects, providing a roadmap to validate technical performance, economic feasibility, and real-world environmental benefits.
- Contributes to the academic literature by integrating thermodynamic modelling, economic, and life cycle assessment in a unified framework for WTE research.

In summary this thesis contributes to knowledge by (i) establishing the technical viability of hybrid WtE-CCS systems for producing clean energy and fuels; (ii) quantifying their economic and environmental trades offs through detailed TEA and LCA; (iii) identifying policy mechanisms and cost drivers that determine system feasibility; and (iv) providing a foundation for both future academic research and industrial-scale deployment. These contributions collectively advance the vision of sustainable waste management and a low carbon energy future.

7.2 Case-specific and general conclusions

7.2.1 Case specific conclusions

7.2.1.1 Research case study 1

The thermodynamic evaluation of the integrated system revealed distinct performance characteristics between the two scenarios: Scenario 1 (for producing CCHP), Scenario 2 (CCHP and liquid biomethane) by integrating a PG, AD, CCGT, ARC and biomethane liquefier, using 5 kg/s of plastic waste and 13.97 kg/s of food waste as feedstocks. The comparative performance of the two scenarios and CCS configurations can be summarized as follows:

- In terms of system efficiency, Scenario 1 achieved energy and exergy efficiencies in the range of 40.10–46.57% and 25.92–35.55%, respectively, while Scenario 2 performed better with efficiencies of 44.11–49.44% and

32.22–41.20%.

- Exergy destruction analysis showed that the plasma gasifier and gas turbine combustion chamber were the dominant sources of irreversibility, contributing 63 MW and 58–83 MW, respectively. Additional losses occurred in the steam cycle (20–35 MW), CCS plant (20–31 MW, mainly in the stripper), and the WGS reactor (~11 MW), while subsystems such as AD, ARC, and methane liquefaction exhibited minimal exergy destruction (<10 MW).
- CCS performance indicated that pre-combustion CCS reduced emissions effectively but lowered efficiency due to syngas diversion and WGS heat demand, whereas post-combustion CCS achieved deeper CO₂ reductions but required greater steam for solvent regeneration, with SRDs of 3.31–3.60 GJ/tonne of CO₂.
- Carbon balance and CO₂ emissions analysis revealed that Scenario 1 released more CO₂ overall because both fossil (plastic) and biogenic (food waste) carbon were combusted, while Scenario 2 achieved lower emissions by converting biogas into liquid biomethane instead of burning it. Specifically, emissions in Scenario 1 were 0.247 (baseline), 0.072 (pre-CCS), and 0.011 (post-CCS) kg CO₂/kg waste, compared to 0.213, 0.068, and 0.008 kg CO₂/kg waste in Scenario 2.
- Energy outputs further reflected these trade-offs: Scenario 1 delivered higher net power (up to 85.67 MW) and larger heating/cooling supply, while Scenario 2 produced lower net power (up to 53.08 MW) but gained efficiency through high-yield biomethane production (energy efficiency 85.79%, exergy efficiency 84.04%).
- Electricity consumption was dominated by the plasma torch (~45 MW). CCS configurations, especially post-CCS, increased auxiliary demand through compression and solvent regeneration, while biomethane upgrading and liquefaction further raised consumption in Scenario 2.
- Baseline systems provided the highest efficiency but lacked carbon mitigation. Pre-CCS improved carbon capture moderately but imposed higher efficiency penalties, whereas post-CCS best balanced emissions reduction with acceptable efficiency losses. Scenario 2 with post-CCS

emerged as the most favourable configuration, offering efficient multi-vector energy production alongside deep CO₂ mitigation.

- The dual feedstock, multi-output system demonstrates technical feasibility for sustainable WtE applications. The outputs can displace conventional energy sources across multiple sectors: electricity can substitute fossil-based grid supply, waste heat can provide renewable heating for industrial or community use, cooling reduces electricity demand from air conditioners in hot humid regions, and liquid biomethane offers a sustainable alternative to LPG. Beyond energy substitution, the system also supports circular economy practices (slag reuse and digestate fertilizer) and enhances energy security.

7.2.1.2 Research case study 2

This study assessed the economic and environmental performance of a novel hybrid WtE system that integrates plasma gasification of plastic waste and anaerobic digestion of food waste, with optional integration of CCS. Two scenarios were explored: Scenario 1, focused on CCHP, and Scenario 2, which coupled CCHP with liquid biomethane production. Each scenario was evaluated under baseline, pre-CCS, and post-CCS configurations. The system was analyzed using economic and life cycle assessment, capturing key indicators such as LCOE, LCO_bM, and LCOWT, GWP, and WF. The comparative findings are summarized below:

- The LCOE in Scenario 1 ranged from 0.171-0.354 USD/kWh, while the LCO_bM in Scenario 2 ranged from 0.176-0.374 USD/kWh, with baseline cases consistently achieving the lowest values and post-CCS cases the highest.
- The LCOWT fell between 0.083-0.236 USD/kg waste, with Scenario 1 slightly higher due to its larger gas turbine capacity and associated capital costs.
- CAPEX values spanned 576-935 MMUSD, dominated by the CCGT and CCS plants, while annual OPEX reached 25-54 MMUSD, with carbon transport and storage emerging as a major cost in CCS case.
- Sensitivity analysis confirmed that the PEC of GT demonstrated the greatest influence on costs, followed by CCS plant costs, while Monte Carlo simulations revealed wider uncertainty ranges for LCO_bM than LCOE, indicating that biomethane prices are more exposed to market and investment risks.

- Feedstock composition strongly influenced system economics: higher plastic fractions increased electricity outputs and stabilized LCOE, while reduced food waste lowered biomethane volumes and sharply increased LCOBM, highlighting the importance of balanced feedstock supply.
- Economies of scale demonstrated clear cost reductions with increased plant size, though diminishing returns occurred beyond ~600 MWth, suggesting an optimal deployment scale for financial feasibility.
- GWP ranged from -0.191 to 0.662 kgCO₂-eq/kg waste in Scenario 1 and 0.123 to 0.746 kgCO₂-eq/kg waste in Scenario 2.
- Post CCS cases achieved the lowest GWP values, even reaching net negative emissions in Scenario 1 (due to biogenic CO₂ capture), while Scenario 2 remained positive as its captured carbon was fossil-derived.
- The CCGT was identified as the dominant contributor to GWP in all cases, reinforcing its role as a critical target for decarbonization.
- WF was modest across all cases (0.0010-0.0014 m³/kg waste), with the pre-CCS case highest due to steam demand in the WGS reactor and the post CCS case lowest due to reduced cooling water requirements.
- Under current Indonesian incentives (tipping fees of 14.02 USD/tonne and carbon credits of 8.9 USD/tonne CO₂), cost reductions were modest (<9%), insufficient to achieve competitiveness with electricity and LNG benchmarks.
- Under stronger policy conditions modelled on U.S. benchmarks (tipping fee ~56.8 USD/tonne; carbon credits ~85 USD/tonne CO₂), costs declined by up to 47% demonstrating the pivotal role of supporting policy frameworks in enabling deployment.
- The integration of CCS, while increasing costs, delivers substantial carbon reductions, with post-CCS in Scenario 1 achieving net-negative GWP.
- Scenario 2 with post CCS offers a strong compromise, combining efficient biomethane production with deep fossil carbon mitigation.
- Multi-output systems face higher CAPEX compared to single single-output designs but benefit from diversified revenue streams and enhanced resilience against market fluctuations.

7.2.1.3 Research case study 3

This study investigated three integrated configurations under two CCS scenarios for converting MSW, specifically plastic and food waste into hydrogen and CCHP under Indonesia-representative conditions. The systems were evaluated through TEA and LCA, considering not only hydrogen yield, and efficiency but also levelized costs, environmental burdens, and policy impacts. The comparative findings are summarized below:

- For both Scenario 1 and Scenario 2, Case 1 demonstrated the best overall techno-economic performance, achieving the highest hydrogen yield as well as the highest energy and exergy efficiencies.
- In contrast, Case 2 showed the lowest hydrogen production and the poorest efficiencies in both scenarios, mainly due to the high drying duty and electricity demand associated with co-processing food and plastic waste in plasma gasifier.
- Case 3 exhibited intermediate performance in both scenarios, as the inclusion of SMR improved hydrogen production but increased the overall system heat demand.
- Across the three cases in Scenario 1, hydrogen production ranged from 0.35 to 0.70 kg/s, whereas in the three cases in Scenario 2 it ranged from 0.27 to 0.53 kg/s.
- The LCOH for the three cases in Scenario 1 ranged from 3.52 to 8.34 USD/kgH₂, while for the three cases in Scenario 2 it ranged from 6.16 to 13.37 USD/kgH₂, indicating that Scenario 1 was more economically favourable than Scenario 2.
- In both scenarios, Case 1 recorded the lowest CAPEX and OPEX, whereas Case 2 recorded the highest, indicating that waste separation can reduce system complexity, equipment size, and overall cost.
- Sensitivity analysis revealed that the cost of the CCS unit was the most influential parameter affecting hydrogen production cost across all three cases in both scenarios.
- From an environmental perspective, Scenario 2 outperformed Scenario 1 across all three cases, as post-combustion CCS enabled a greater proportion of total system carbon to be captured, resulting in more negative GWP values.

- For the three cases in Scenario 1, GWP ranged from -0.038 to 0.048 kgCO₂-eq/kg waste, whereas for the three cases in Scenario 2, it ranged from -0.159 to -0.155 to kgCO₂-eq/kg waste.
- Overall, among the three cases in each scenario, Case 1 was identified as the most economically viable option, Case 2 provided the greatest environmental benefit but at the highest cost, and Case 3 represented a balanced compromise between hydrogen production, economic performance, and carbon mitigation.

7.2.2 General conclusions

This thesis has explored the integration of plasma gasification, anaerobic digestion, and CCS technologies as a novel approach to sustainable WtE systems. Across three interconnected studies, the research demonstrates that municipal solid waste, when strategically separated and processed, can be transformed into multiple high-value energy products, including CCHP, liquid biomethane, and hydrogen, while simultaneously contributing to significant reductions in greenhouse gas emissions.

From a technical perspective, the modelling results show that hybrid WtE systems can achieve high energy and exergy efficiencies, with values approaching 49% and 41% in certain configurations. The findings also emphasize the critical role of waste separation, system integration, and CCS deployment in shaping overall performance. In particular, routing plastics to plasma gasification and food waste to anaerobic digestion enhances hydrogen yield and energy efficiency, while scenarios with extended reforming and capture achieve the greatest reductions in carbon emissions.

From an economic and environmental standpoint, the thesis provides a comprehensive TEA and LCA of hybrid systems under Indonesia-representative conditions. Although CCS integration increases the levelized cost of energy and fuels, it also delivers substantial climate benefits, with some configurations achieving net-negative greenhouse gas emissions. Sensitivity and Monte Carlo analyses identify the capital cost of major equipment, particularly the CCS unit, as the dominant factor influencing economic outcomes, thereby informing priority areas for future cost reductions.

The results also highlight the role of supportive policies and incentives. While current Indonesia policy mechanisms such as tipping fee and carbon credit offer only modest improvements to system feasibility, international benchmarks indicate that with stronger incentives, hybrid WtE hydrogen systems can approach cost competitiveness with grey and blue hydrogen. This underscores the importance of aligning technological innovation with enabling policy frameworks to accelerate deployment.

Overall, this thesis demonstrates that the integration of plasma gasification, anaerobic digestion, and CCS presents a technically feasible and environmentally beneficial pathway for the valorisation of municipal waste into clean energy carries. By addressing the technical, economic, and policy aspect in an integrated framework, the research advances the state of knowledge in sustainable waste management and provides a foundation for pilot-scale demonstration and industrial scale adoption. In doing so, it contributes to the broader objectives of achieving low-carbon energy transitions and supporting progress towards global sustainability goals.

7.3 Limitations of the study and future work

7.3.1 Limitations

While this thesis provides novel insights into integration of plasma gasification, anaerobic digestion, and carbon capture and storage (CCS) for sustainable WtE systems, several limitations remain that warrant further investigation.

- All three studies were based on steady state Aspen Plus simulations, which do not capture dynamic behaviour, start-up/shutdown conditions, or system flexibility under variable operating loads. This simplification omits operational realities, such as transient responses to feedstock fluctuations or equipment disturbances.
- The modelling framework assumed fixed waste composition (plastic and food waste fractions, proximate/ ultimate analyses, and moisture contents). In reality, municipal waste is heterogeneous and seasonal, which may significantly influence hydrogen yields, efficiencies, and CO₂ capture performance.
- Subsystems such as plasma torch, waste dryer, crusher design were

represented with simplified performance assumptions. More detailed cycle modelling and experimental validation would increase reliability.

- The cost analysis relied on AACE Class 4-5 early-stage cost estimates, using scaling factors, and generic databases. This introduces large uncertainty ($\pm 30\text{--}50\%$), limiting accuracy for investment decisions.
- These estimates may not fully reflect Indonesia-specific conditions, including:
 - Supply chain cost: additional expenses from importing specialized equipment (e.g. plasma torches, CCS absorbers, PSA units), custom duties, limited domestic suppliers, and infrastructure/ logistics challenges.
 - Financing conditions: higher interest rates, limited access to low cost green finance, and increased investor risk premiums compared with developed economies. Together, these factors could substantially raise actual project costs relative to modelled values.
- Cost estimations were derived primarily from literature and assumptions, which may differ from actual market conditions. Future work should incorporate updated cost databases.
- Although larger plants may offer potential economic benefits through economies of scale, scale up also introduces non-linear challenges related to securing sufficient feedstock supply, managing more complex process heat integration, and controlling the higher capital and operational risks associated with highly integrated systems. Future work should therefore assess the optimum scale at which economic performance can be improved without compromising technical operability and financial viability.
- The LCA applied a cradle to gate boundary, excluding downstream distribution, storage, and end use stages of hydrogen and CCHP. This means potential substitution benefits (e.g., displacement of LPG, grid electricity, or synthetic fertilisers) were not fully quantified.
- Although the environmental inventory included infrastructure construction, its contribution was minor when allocated per functional unit. However, the analysis did not account for end-of-life impacts associated with plant decommissioning, dismantling, and material disposal, which may add to the overall footprint in a full cradle to grave perspective.
- Broader social and institutional barriers, such as public acceptance of CCS, land

availability for infrastructure, and governance capacity, were not assessed, but will be critical for implementation in developing countries.

7.3.2 Future research directions

Building on the above limitations, future studies should:

- Develop dynamic models to capture operational transients, and system flexibility under variable feedstock supply and operating conditions. This would enable a more realistic assessment of operational stability, start-up, and shutdown dynamics, and integration with intermittent renewable energy sources.
- Conduct pilot or demonstration scale projects to validate the technical integration of plasma gasification, anaerobic digestion, CCS under real world conditions. These projects would help identify unforeseen operational challenges, such as solvent degradation in CCS unit, catalyst deactivation in reforming and WGS reactors, feedstock pretreatment constraints. The latter may include the mechanical limitations of crushing and shredding plastics at scale, the cost and complexity of separating mixed municipal waste streams). Addressing these practical challenges is essential for ensuring continuous and reliable operation.
- Future design could explore carbon utilisation pathways as complementary options to storage. Converting captured CO₂ into synthetic fuels (e.g., methanol, aviation fuels), chemicals, or building materials could generate additional revenue streams, improve carbon circularity, and enhance the economic competitiveness of the systems.
- Collect empirical cost data from demonstration projects and industrial scale operations to reduce uncertainty in TEA. Reliance on literature based scaling factors introduces significant variability; real world cost and performance data are crucial for improving investment reliability.
- Expand LCA beyond a cradle to gate boundary to a full cradle to grave perspective, explicitly including plant decommissioning, and downstream distribution and use phase. In addition, future LCAs should give particular attention to social and equity impacts in developing country contexts, where waste management and energy access issues are highly interlinked.

In summary, while thesis establishes the technical feasibility and strategic potential

of hybrid waste to energy CCS systems, future research is essential to translate these findings into practical, scalable solutions that can contribute meaningfully to national and global decarbonisation pathways.

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9 APPENDIX

9.1 Aspen Plus model implementation details

Relevant Aspen Plus model configurations for representative cases are provided in table below to improve transparency and reproducibility of the simulation methodology.

Table 9-1. Aspen Plus blocks used for process simulation.

Equipment	Aspen block type	Property method	Justification of property method choice
Plasma gasifier	RStoic, Mixer, Splitter, Heater, Separator, RYield, RGibbs, Heatx	IDEAL	IDEAL was used because the streams are predominantly gas phase syngas at high temperature and moderate pressure; under these conditions, ideal-gas behaviour is a reasonable approximation.
Anaerobic digester	Preheater: Heater Digester: RYield, RStoic Separator: Cyclone, Flash Mixer: Mixer	The Soave–Redlich–Kwon (SRK)	SRK was selected because the AD model focuses on biogas output (mainly CH ₄ , and CO ₂), for which SRK provides adequate gas-phase prediction, computational simplicity, and stable integration within the overall Aspen Plus model.
High pressure water scrubber	Absorber: RadFrac Stripper: RadFrac Compressor: Compressor Cooler: HeatX Pump: Pump	Non-Random Two-Liquid (NRTL)	NRTL was selected because the water scrubber involves a highly non-ideal liquid mixture and gas-liquid absorption, for which NRTL is suitable for representing liquid-

	Separator: Flash, FSplit Mixer: Mixer		phase non-ideality and phase equilibrium between water, CO ₂ , and other absorbed gases.
WGSR	RGibbs, Splitter, Mixer, RPlug, Heatx	IDEAL	IDEAL was used because the streams are predominantly gas phase syngas at high temperature and moderate pressure; under these conditions, ideal-gas behaviour is a reasonable approximation.
CCS	Absorber: Radfrac, Heater, Flash Desorber: Radfrac, Heater Pump: Pump Lean/rich heat exchanger: Heatx Compressor: Compressor Cooler: Heatx	Electrolyte Non-Random Two-Liquid (ELECNRTL)	ELECNRTL was used because CO ₂ capture with MEA occurs in a liquid solution, and the solvent is non-ideal.
PSA	Flash, Compressor, Heatx, Separator, Mixer, Splitter	IDEAL	IDEAL was used because the PSA unit is represented as a simplified gas separation (using split/purity and recovery assumptions) and all PSA streams are in the gas phase.
SMR	Compressor, Pump, Separator, Mixer, Heater, RPlug	IDEAL	IDEAL was used because the streams are predominantly gas phase syngas at high temperature and moderate pressure; under these conditions, ideal-gas behaviour is a reasonable approximation.
Methane liquefaction	Compressor: MCompr Heat exchanger: Heatx Expansion valve: Valve Separator: Flash Mixer: Mixer	Peng-Robinson (PENG-ROB)	PENG-ROB was selected because it is well suited for hydrocarbon-rich systems and provides reliable vapour-liquid equilibrium prediction for methane at elevated pressure and low-temperature liquefaction

			conditions.
CCGT	Gas turbine components Compressor: Compressor Combustor: RGibbs Turbine: Compressor Steam turbine components HRSG: Mheatx Turbine: Compressor Condenser: Heatx Pump: Pump Additional: Mixer, Stripper, Flash, Heater	Peng–Robinson Boston–Mathias (PRBM)	PRBM provides reliable thermodynamic properties for combustion/flue gas mixtures at elevated temperatures and pressures and is widely applied in gas turbine and syngas combustion simulations.
ARC	Evaporator: Heatx Absorber: Heatx Pump: Pump Condenser: Heatx Expansion valve: Valve Desorber: Flash, Heater Solution heat exchanger: Heater	Electrolyte Non-Random Two-Liquid (ELECNRTL)	ELECNRTL was applied because ARC performance depends on liquid-solution behaviour (absorption/desorption), which cannot be represented well using ideal-gas assumptions.

9.2 Life cycle assessment of WtE in Research Case 2

Research Case 2 applies TEA and LCA to the first process model, which converts plastic and food waste into CCHP and liquid biomethane through two scenarios. Scenario 1 produces CCHP, while Scenario 2 produces CCHP and liquid biomethane. Both scenarios are assessed without CCS, with pre-combustion CCS, and with post-combustion CCS. The results of Research Case 2 have been previously published [65].

9.2.1 Life cycle inventory of Research Case 2

9.2.1.1 The entire process

The selected inventory data in SimaPro for the process materials, energy, and processes for Scenarios 1 and 2 are in Table 9-2 and Table 9-3, respectively. The functional unit for this assessment is 1 kg of waste, composed of 74% food waste and 26% plastic waste.

Table 9-2. Life cycle inventory of Scenario 1.

Scenario 1 (production of electricity, heating, and cooling)				
	BASELINE	W/ PRE-CCS	W/ POST-CCS	
Products				
Outputs to technosphere: Products and co-products	Amount			Unit
Electricity	1.25E-03	8.90E-04	9.47E-04	MWh/kg feedstock
Cooling	4.96E-06	4.30E-06	4.96E-06	MWh/kg feedstock
Heating	2.37E-05	4.15E-05	4.80E-05	MWh/kg feedstock
Inputs				
Inputs from nature	Amount			Unit
Water (from AD, PG, WGSR, CCS)	5.14E-04	8.45E-04	5.14E-04	m ³ /kg feedstock
Air (from PG and air comp GT)	1.04E+01	1.10E+01	1.04E+01	kg/ kg feedstock
Cooling water (gas conditioner before CCS and CCS)	2.59E-03	2.28E-03	1.90E-03	m ³ / kg feedstock
Inputs from technosphere: materials/fuels	Amount			Unit
Synthetic gas factory {RoW} synthetic gas factory construction Cut-off, U	1.65E-10	1.65E-10	1.65E-10	p
Anaerobic digestion plant, agricultural {RoW} anaerobic digestion plant construction, agricultural Cut-off, U	2.27E-10	2.27E-10	2.27E-10	p

Heat and power co-generation unit, 1MW electrical, common components for heat+electricity {RoW} heat and power co-generation unit construction, 1MW electrical, common components for heat+electricity Cut-off, U	3.71E-09	3.35E-09	3.47E-09	p
Absorption chiller, 100kW {RoW} absorption chiller production, 100kW Cut-off, U	6.49E-11	5.96E-11	6.49E-11	p
Chemical factory, organics {RoW} chemical factory construction, organics Cut-off, U (wgsr+CCS)		1.22E-20	3.75E-20	kg/kg feedstock
Monoethanolamine {RoW} ethanolamine production Cut-off, U		1.28E-03	1.72E-03	kg/kg feedstock
Ferrochromium, high-carbon, 55% Cr {RoW} ferrochromium production, high-carbon, 55% Cr Cut-off, U		2.73E-05		kg/kg feedstock
Copper oxide {RoW} copper oxide production Cut-off, U		1.02E-05		kg/kg feedstock
Zinc oxide {RoW} zinc oxide production Cut-off, U		8.18E-06		kg/kg feedstock
Aluminium oxide, metallurgical {IAI Area, Asia, without China and GCC} aluminium oxide production Cut-off, U		2.04E-06		kg/kg feedstock
Outputs				
Emissions to air	Amount			Unit
CH ₄ (flue gas from GT: baseline, pre and post CCS and flue gas from WGSr: pre CCS)	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
N ₂ (flue gas from GT: baseline, pre and post CCS and flue gas from WGSr: pre CCS)	8.00E+00	8.45E+00	8.00E+00	kg/kg feedstock
O ₂ (flue gas from GT: baseline, pre and post CCS and flue gas from WGSr: pre CCS)	1.48E+00	1.62E+00	1.48E+00	kg/kg feedstock
CO ₂ (flue gas from GT: baseline, pre and post CCS and flue gas from WGSr: pre CCS)	9.06E-01	2.64E-01	4.22E-02	kg/kg feedstock
CO₂ fossil	6.60E-01	1.87E-01	3.08E-02	kg/kg feedstock
CO₂ biogenic	2.46E-01	7.73E-02	1.15E-02	kg/kg feedstock
H ₂ O (flue gas from GT: baseline, pre and post CCS and flue gas from WGSr: pre CCS)	5.68E-01	7.88E-01	2.45E-01	kg/kg feedstock
H ₂ (flue gas from GT: baseline, pre and post CCS and flue gas from WGSr: pre CCS)	3.60E-07	4.70E-07	3.60E-07	kg/kg feedstock
CO (flue gas from GT: baseline, pre and post CCS and flue	9.97E-06	2.19E-06	9.97E-06	kg/kg feedstock

gas from WGSR: pre CCS)				
CO fossil	7.27E-06	1.55E-06	7.27E-06	kg/kg feedstock
CO biogenic	2.71E-06	6.40E-07	2.70E-06	kg/kg feedstock
H ₂ S (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
NH ₃ (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
COS (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	2.73E-03	4.08E-03	2.73E-03	kg/kg feedstock
SO ₂ (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
Emissions to water	Amount			Unit
H₂O	0.00E+00	0.00E+00	0.00E+00	m ³ /kg feedstock
N ₂	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
O ₂	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
NO ₂	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
NO	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
S	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
SO ₂	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
SO ₃	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
H ₂	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
Cl ₂	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
HCl	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
C	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
CO	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
CO ₂	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
COS	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock

CH₄	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
H₂S	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
NH₃	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
Outputs to technosphere: Waste treatment				
Wastewater, average {RoW} treatment of wastewater, average, wastewater treatment Cut-off, U	6.08E-04	5.63E-04	4.63E-04	m ³ /kg feedstock

Table 9-3. Life cycle inventory of Scenario 2.

Scenario 2 (production of electricity, heating, cooling and liquid fuel)				
	BASILINE	W/ PRE CCS	W/ POST CCS	
Products				
Outputs to technosphere: Products and co-products	Amount			Unit
Liquid Fuel	6.84E-04	6.84E-04	6.84E-04	MWh/kg feedstock
Electricity	7.77E-04	4.36E-04	5.54E-04	MWh/kg feedstock
Cooling	4.01E-06	3.33E-06	4.01E-06	MWh/kg feedstock
Heating	2.18E-05	3.96E-05	4.58E-05	MWh/kg feedstock
Inputs				
Inputs from nature	Amount			Unit
Water (AD, WS, PG)	5.40E-04	8.71E-04	5.40E-04	m ³ /kg feedstock
Air (WS, PG, GT)	8.83E+00	9.10E+00	8.49E+00	kg/ kg feedstock
Cooling water (WS, ST, CC)	2.18E-03	1.90E-03	1.69E-03	m ³ / kg feedstock
Inputs from technosphere: materials/fuels	Amount			Unit
Synthetic gas factory {RoW} synthetic gas factory construction Cut-off, U	1.65E-10	1.65E-10	1.65E-10	p
Anaerobic digestion plant, agricultural {RoW} anaerobic digestion plant construction, agricultural Cut-off, U	2.27E-10	2.27E-10	2.27E-10	p
Heat and power co-generation unit, 1MW electrical, common components for heat+electricity {RoW} heat and power co-generation unit construction, 1MW electrical, common components	2.39E-09	2.11E-09	2.26E-09	p

for heat+electricity Cut-off, U				
Absorption chiller, 100kW {RoW} absorption chiller production, 100kW Cut-off, U	2.51E-11	2.25E-11	2.51E-11	p
Chemical factory, organics {RoW} chemical factory construction, organics Cut-off, U (water scrubber)	4.02E-22	4.02E-22	4.02E-22	kg
Natural gas processing plant {GLO} natural gas processing plant production Cut-off, U	7.63E-23	7.63E-23	7.63E-23	p
Liquid storage tank, chemicals, organics {RoW} liquid storage tank production, chemicals, organics Cut-off, U	2.52E-12	2.52E-12	2.52E-12	p
Chemical factory, organics {RoW} chemical factory construction, organics Cut-off, U (case 2: WGSR and CCS, case 3: CCS)		8.78E-21	2.44E-20	kg/kg feedstock
Monoethanolamine {RoW} ethanolamine production Cut-off, U		1.06E-03	1.26E-03	kg/kg feedstock
Ferrochromium, high-carbon, 55% Cr {RoW} ferrochromium production, high-carbon, 55% Cr Cut-off, U		2.73E-05		kg/kg feedstock
Copper oxide {RoW} copper oxide production Cut-off, U		1.02E-05		kg/kg feedstock
Zinc oxide {RoW} zinc oxide production Cut-off, U		8.18E-06		kg/kg feedstock
Aluminium oxide, metallurgical {IAI Area, Asia, without China and GCC} aluminium oxide production Cut-off, U		2.04E-06		kg/kg feedstock
Outputs				
Emissions to air	Amount			Unit
CH4 biogenic (WS)	1.02E-03	1.02E-03	1.02E-03	kg/kg feedstock
CH4 fossil (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
N2 (WS, flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	6.78E+00	7.12E+00	6.78E+00	kg/kg feedstock
O2 (WS, flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	1.29E+00	1.40E+00	1.29E+00	kg/kg feedstock
CO2 biogenic (WS)	1.17E-01	1.17E-01	1.17E-01	kg/kg feedstock
CO2 fossil (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	6.60E-01	1.31E-01	3.08E-02	kg/kg feedstock
H2O (WS, flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	4.61E-01	6.82E-01	2.05E-01	kg/kg feedstock
H2 (WS, flue gas from GT: baseline, pre and post CCS and flue gas	2.90E-07	4.00E-07	2.90E-07	kg/kg feedstock

from WGSR: pre CCS)				
CO fossil (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	7.20E-06	7.20E-07	7.20E-06	kg/kg feedstock
CO biogenic (WS)	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
H ₂ S (WS, flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	8.29E-04	8.29E-04	8.29E-04	kg/kg feedstock
NH ₃ (WS, flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	2.53E-03	2.53E-03	2.53E-03	kg/kg feedstock
COS (WS, flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
SO ₂ (WS, flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
Emissions to water	Amount			Unit
H ₂ O (WS+PG)	0.00E+00	0.00E+00	0.00E+00	m ³ /kg feedstock
N ₂ (WS+PG)	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
O ₂ (WS+PG)	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
NO ₂ (PG)	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
NO	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
S	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
SO ₂	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
SO ₃	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
H ₂ (WS+PG)	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
Cl ₂ (PG)	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
HCl	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
C	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
CO	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
CO₂ (WS+PG)	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
COS	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
CH₄ (WS+PG)	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
H₂S (WS+PG)	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
NH₃ (WS+PG)	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
Outputs to technosphere: Waste treatment	Amount			Unit

Wastewater, average {RoW} treatment of wastewater, average, wastewater treatment Cut-off, U	5.17E-04	4.78E-04	4.12E-04	m ³ /kg feedstock
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In this study, the LCA is conducted not only for the entire process but also for individual process breakdowns within the system. The life cycle inventory (LCI) for each process is detailed in the following tables. The inventory for plasma gasification and anaerobic digestion is identical for Scenarios 1 and 2 and is presented in **Table 9-4** and **Table 9-5**, respectively. However, other processes, such as CCGT and ARC, have different inventories between Scenarios 1 and 2, and therefore, they are presented separately in different tables.

9.2.1.2 The plasma gasification process

Table 9-4. Life cycle inventory of plasma gasification process.

Plasma gasification (FU = 1 kg of plastic waste)		
PRODUCTS		
Outputs to technosphere: Products and co-products	Amount	Unit
Syngas	2.70E-03	MWh/kg feedstock
Slag	3.46E-02	kg/kg feedstock
INPUT		
Inputs from nature	Amount	Unit
Water	2.21E-04	m ³ /kg feedstock
Air	5.33E-01	
Inputs from technosphere: materials/fuels	Amount	Unit
Synthetic gas factory {RoW} synthetic gas factory construction Cut-off, U	1.65E-10	p
OUTPUT		
Emissions to water	Amount	Unit
H ₂ O	0	m ³ /kg feedstock

N ₂	0	kg/kg feedstock
O ₂	0	kg/kg feedstock
NO ₂	0	kg/kg feedstock
NO	0	kg/kg feedstock
S	0	kg/kg feedstock
SO ₂	0	kg/kg feedstock
SO ₃	0	kg/kg feedstock
H ₂	0	kg/kg feedstock
Cl ₂	0	kg/kg feedstock
HCl	0	kg/kg feedstock
C	0	kg/kg feedstock
CO	0	kg/kg feedstock
CO ₂	0	kg/kg feedstock
COS	0	kg/kg feedstock
CH ₄	0	kg/kg feedstock
H ₂ S	0	kg/kg feedstock
NH ₃	0	kg/kg feedstock
Outputs to technosphere: Waste treatment		
Wastewater, average {RoW} treatment of wastewater, average, wastewater treatment Cut-off, U	1.05E-05	m ³ /kg feedstock

9.2.1.3 The anaerobic digestion process

Table 9-5. Life cycle inventory of anaerobic digestion process.

Anaerobic digestion (FU = 1 kg of food waste)		
PRODUCTS		
Outputs to technosphere: Products and co-products	Amount	Unit
Biogas	7.06E-04	MWh/kg feedstock

INPUT		
Inputs from nature	Amount	Unit
Water	2.93E-04	m ³ /kg feedstock
Inputs from technosphere: materials/fuels	Amount	Unit
Anaerobic digestion plant, agricultural {RoW} anaerobic digestion plant construction, agricultural Cut-off, U	2.27E-10	kg/ kg feedstock
OUTPUT		
Emissions to water	Amount	Unit
H ₂ O	0	m ³ /kg feedstock
N ₂	0	kg/kg feedstock
O ₂	0	kg/kg feedstock
NO ₂	0	kg/kg feedstock
NO	0	kg/kg feedstock
S	0	kg/kg feedstock
SO ₂	0	kg/kg feedstock
SO ₃	0	kg/kg feedstock
H ₂	0	kg/kg feedstock
Cl ₂	0	kg/kg feedstock
HCL	0	kg/kg feedstock
C	0	kg/kg feedstock
CO	0	kg/kg feedstock
CO ₂	0	kg/kg feedstock
COS	0	kg/kg feedstock
CH ₄	0	kg/kg feedstock
H ₂ S	0	kg/kg feedstock
NH ₃	0	kg/kg feedstock
Outputs to technosphere: Waste treatment		
Wastewater, average {RoW} treatment of wastewater, average, wastewater treatment Cut-off, U	1.39E-05	m ³ /kg feedstock

9.2.1.4 The CCGT process

Table 9-6. Life cycle inventory of CCGT process (Scenario 1).

CCGT (FU = 1 kg of waste)				
PRODUCTS				
Outputs to technosphere: Products and co-products	Amount (case 1)	Amount (case 2)	Amount (case 3: CCGT+POST CCS)	Unit
Electricity	1.25E-03	8.90E-04	9.47E-04	MWh/kg feedstock
Heating				
Outputs to technosphere: Products and co-products	Amount (case 1)	Amount (case 2)	Amount (case 3: CCGT+POST CCS)	Unit
INPUT				
Inputs from nature	Amount (case 1)	Amount (case 2)	Amount (case 3)	Unit
Water (for case 3: water is for make up CC)			2.64E-13	m ³ /kg feedstock
Air	9.89E+00	8.95E+00	9.89E+00	kg/kg feedstock
Cooling water	2.59E-03	1.57E-03	1.90E-03	m ³ /kg feedstock
Inputs from technosphere: materials/fuels	Amount (case 1)	Amount (case 2)	Amount (case 3)	Unit
Heat and power co-generation unit, 1MW electrical, common components for heat+electricity {RoW} heat and power co-generation unit construction, 1MW electrical, common components for heat+electricity Cut-off, U	3.71E-09	3.35E-09	3.47E-09	p
Onshore petroleum field infrastructure {GLO}			0.00E+00	p
Chemical factory, organics {RoW}			1.08E-10	kg/kg feedstock
Monoethanolamine {RoW}			1.72E-03	kg/kg feedstock
OUTPUT				
Emissions to air	Amount (case 1)	Amount (case 2)	Amount (case 3)	Unit
CH ₄ (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
N ₂ (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	8.00E+00	7.24E+00	8.00E+00	kg/kg feedstock
O ₂ (flue gas from GT: baseline, pre and post CCS)	1.48E+00	1.33E+00	1.48E+00	kg/kg feedstock

and flue gas from WGSR: pre CCS)				
CO ₂ (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	9.06E-01	1.98E-01	4.22E-02	kg/kg feedstock
CO₂ fossil	6.60E-01	1.40E-01	3.08E-02	kg/kg feedstock
CO₂ biogenic	2.46E-01	5.80E-02	1.15E-02	kg/kg feedstock
H ₂ O (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	5.68E-01	7.35E-01	2.45E-01	kg/kg feedstock
H ₂ (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	3.60E-07	4.70E-07	0.00E+00	kg/kg feedstock
CO (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	9.97E-06	2.19E-06	9.97E-06	kg/kg feedstock
CO fossil	7.27E-06	1.55E-06	7.27E-06	kg/kg feedstock
CO biogenic	2.71E-06	6.40E-07	2.70E-06	kg/kg feedstock
H ₂ S (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
NH ₃ (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
COS (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	2.73E-03	4.08E-03	2.73E-03	kg/kg feedstock
SO ₂ (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
Emissions to water	Amount (case 1)	Amount (case 2)	Amount (case 3)	Unit
Outputs to technosphere: Waste treatment	Amount (case 1)	Amount (case 2)	Amount (case 3)	Unit
Wastewater, average {RoW} treatment of wastewater, average, wastewater treatment Cut-off, U	5.83E-04	3.54E-04	4.38E-04	m ³ /kg feedstock

Table 9-7. Life cycle inventory of CCGT process (Scenario 2).

CCGT (FU = 1 kg of waste)				
PRODUCTS				
Outputs to technosphere: Products and co-products	Amount (case 1)	Amount (case 2)	Amount (case 3: CCGT+POST CCS)	Unit
Electricity	7.77E-04	4.36E-04	5.54E-04	MWh/kg feedstock
Heating				
Outputs to technosphere: Avoided products	Amount (case 1)	Amount (case 2)	Amount (case 3)	Unit
INPUT				
Inputs from nature	Amount (case 1)	Amount (case 2)	Amount (case 3)	Unit
Water (for case 3: water is for make up CC)			2.64E-13	m ³ /kg feedstock
Air	7.96E+00	6.86E+00	7.96E+00	kg/kg feedstock
Cooling water	2.13E-03	1.28E-03	1.64E-03	m ³ /kg feedstock
Inputs from technosphere: materials/fuels	Amount (case 1)	Amount (case 2)	Amount (case 3)	Unit
Heat and power co-generation unit, 1MW electrical, common components for heat+electricity {RoW} heat and power co-generation unit construction, 1MW electrical, common components for heat+electricity Cut-off, U	2.39E-09	2.11E-09	2.26E-09	p
Chemical factory, organics {RoW} chemical factory construction, organics Cut-off, U (case 2: WGSR and CCS, case 3: CCS)			2.44E-20	kg/kg feedstock
Monoethanolamine {RoW} ethanolamine production Cut-off, U			1.26E-03	kg/kg feedstock
OUTPUT				
Emissions to air	Amount (case 1)	Amount (case 2)	Amount (case 3)	Unit
CH ₄ (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
N ₂ (flue gas from GT: baseline, pre and post CCS)	6.52E+00	5.64E+00	6.52E+00	kg/kg feedstock

and flue gas from WGSR: pre CCS)				
O ₂ (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	1.22E+00	1.03E+00	1.22E+00	kg/kg feedstock
CO ₂ (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)				kg/kg feedstock
CO₂ fossil	6.60E-01	6.47E-02	3.08E-02	kg/kg feedstock
CO₂ biogenic	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
H ₂ O (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	4.57E-01	6.25E-01	2.00E-01	kg/kg feedstock
H ₂ (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	2.90E-07	4.00E-07	2.90E-07	kg/kg feedstock
CO (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)				kg/kg feedstock
CO fossil	7.20E-06	7.20E-07	7.20E-06	kg/kg feedstock
CO biogenic	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
H ₂ S (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
NH ₃ (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
COS (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
SO ₂ (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	0.00E+00	0.00E+00	0.00E+00	kg/kg feedstock
Emissions to water	Amount (case 1)	Amount (case 2)	Amount (case 3)	Unit
Outputs to technosphere: Waste treatment	Amount (case 1)	Amount (case 2)	Amount (case 3)	Unit
Wastewater, average {RoW} treatment of wastewater, average, wastewater treatment Cut-off, U	4.80E-04	2.87E-04	3.75E-04	m ³ /kg feedstock

9.2.1.5 The absorption refrigeration cooler (ARC) process

Table 9-8. Life cycle inventory of ARC process (Scenario 1).

Absorption Refrigeration Cooler (FU = 1 kg of waste)				
PRODUCTS				
Outputs to technosphere: Products and co-products	Amount (case 1)	Amount (case 2)	Amount (case 3)	Unit
Cooling	4.96E-06	4.30E-06	4.96E-06	MWh/kg feedstock
Heating	2.37E-05	4.15E-05	4.80E-05	MWh/kg feedstock
Allocation cooling	1.73E+01	9.40E+00	9.36E+00	%
Allocation heating	8.27E+01	9.06E+01	9.06E+01	%
INPUT				
Inputs from technosphere: materials/fuels	Amount (case 1)	Amount (case 2)	Amount (case 3)	Unit
Absorption chiller, 100kW {RoW} absorption chiller production, 100kW Cut-off, U	8.42E-10	7.74E-10	8.42E-10	p
OUTPUT				
Emissions to air	Amount (case 1)	Amount (case 2)	Amount (case 3)	Unit
Emissions to water	Amount (case 1)	Amount (case 2)	Amount (case 3)	Unit
Outputs to technosphere: Waste treatment	Amount (case 1)	Amount (case 2)	Amount (case 3)	Unit

Table 9-9. Life cycle inventory of ARC process (Scenario 2).

Absorption Refrigeration Cooler (FU = 1 kg of waste)				
PRODUCTS				

Outputs to technosphere: Products and co-products	Amount (case 1)	Amount (case 2)	Amount (case 3)	Unit
Cooling	4.01E-06	3.33E-06	4.01E-06	MWh/kg feedstock
Heating	2.18E-05	3.96E-05	4.58E-05	MWh/kg feedstock
Allocation cooling	1.55E+01	7.76E+00	8.05E+00	%
Allocation heating	8.45E+01	9.22E+01	9.19E+01	%
Outputs to technosphere: Avoided products	Amount (case 1)	Amount (case 2)	Amount (case 3)	Unit
INPUT				
Inputs from nature	Amount (case 1)	Amount (case 2)	Amount (case 3)	Unit
	0			m ³ /kg feedstock
Inputs from technosphere: materials/fuels	Amount (case 1)	Amount (case 2)	Amount (case 3)	Unit
Absorption chiller, 100kW {RoW} absorption chiller production, 100kW Cut-off, U	2.51E-11	2.25E-11	2.51E-11	p
OUTPUT				
Emissions to air	Amount (case 1)	Amount (case 2)	Amount (case 3)	Unit
Emissions to water	Amount (case 1)	Amount (case 2)	Amount (case 3)	Unit
Outputs to technosphere: Waste treatment	Amount (case 1)	Amount (case 2)	Amount (case 3)	Unit

9.2.1.6 The water gas shift reactor (WGSR) and carbon capture and storage (CCS) process

Table 9-10. Life cycle inventory of WGSR and CCS process (Scenario 1).

WGSR & CCS (FU = 1 kg of waste)

PRODUCTS		
Outputs to technosphere: Products and co-products	Amount (case 2)	Unit
Clean gas	3.04E-03	MWh/kg feedstock
INPUT		
Inputs from nature	Amount (case 2)	Unit
Water	3.31E-04	m ³ /kg feedstock
Air	1.53E+00	kg/kg feedstock
Cooling water	7.03E-04	m ³ /kg feedstock
Inputs from technosphere: materials/fuels	Amount (case 2)	Unit
Chemical factory, organics {RoW}	1.22E-20	kg/kg feedstock
Monoethanolamine {RoW}	1.28E-03	kg/kg feedstock
Ferrochromium, high-carbon, 55% Cr {RoW}	2.73E-05	kg/kg feedstock
Copper oxide {RoW}	1.02E-05	kg/kg feedstock
Zinc oxide {RoW}	8.18E-06	kg/kg feedstock
Aluminium oxide, metallurgical {IAI Area, Asia}	2.04E-06	kg/kg feedstock
OUTPUT		
Emissions to air	Amount (case 2)	Unit
CH ₄ (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	0.00E+00	kg/kg feedstock
N ₂ (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	1.21E+00	kg/kg feedstock
O ₂ (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	2.93E-01	kg/kg feedstock
CO ₂ (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	6.60E-02	kg/kg feedstock
CO₂ fossil	4.81E-02	kg/kg feedstock
CO₂ biogenic	1.79E-02	kg/kg feedstock
H ₂ O (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	5.26E-02	kg/kg feedstock
H ₂ (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	0.00E+00	kg/kg feedstock
CO (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	0.00E+00	kg/kg feedstock
CO fossil	0.00E+00	kg/kg feedstock
CO biogenic	0.00E+00	kg/kg feedstock
H ₂ S (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	0.00E+00	kg/kg feedstock
NH ₃ (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	0.00E+00	kg/kg feedstock
COS (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	0.00E+00	kg/kg feedstock
SO ₂ (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	0.00E+00	kg/kg feedstock
Emissions to water	Amount (case 2)	Unit

Outputs to technosphere: Waste treatment	Amount (case 2)	Unit
Wastewater, average {RoW} treatment of wastewater, average, wastewater treatment Cut-off, U	1.85E-04	m ³ /kg feedstock

Table 9-11. Life cycle inventory of WGSR and CCS process (Scenario 2).

WGSR & CCS (FU = 1 kg of food waste)		
PRODUCTS		
Outputs to technosphere: Products and co-products	Amount (case 2)	Unit
Clean gas	2.39E-03	MWh/kg feedstock
Outputs to technosphere: Avoided products	Amount (case 2)	Unit
INPUT		
Inputs from nature	Amount (case 2)	Unit
Water	3.31E-04	m ³ /kg feedstock
Air	1.53E+00	kg/kg feedstock
Cooling water	5.74E-04	m ³ /kg feedstock
Inputs from technosphere: materials/fuels	Amount (case 2)	Unit
Chemical factory, organics {RoW} chemical factory construction, organics Cut-off, U (case 2: WGSR and CCS)	8.78E-21	kg/kg feedstock
Monoethanolamine {RoW} ethanolamine production Cut-off, U	0.001059	kg/kg feedstock
Ferrochromium, high-carbon, 55% Cr {RoW} ferrochromium production, high-carbon, 55% Cr Cut-off, U	2.73E-05	kg/kg feedstock
Copper oxide {RoW} copper oxide production Cut-off, U	1.02E-05	kg/kg feedstock
Zinc oxide {RoW} zinc oxide production Cut-off, U	8.18E-06	kg/kg feedstock
Aluminium oxide, metallurgical {IAI Area, Asia, without China and GCC} aluminium oxide production Cut-off, U	2.04E-06	kg/kg feedstock
OUTPUT		
Emissions to air	Amount (case 2)	Unit
CH ₄ (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	0.00E+00	kg/kg feedstock
N ₂ (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	1.21E+00	kg/kg feedstock
O ₂ (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	2.93E-01	kg/kg feedstock
CO ₂ (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)		kg/kg feedstock

CO₂ fossil	6.60E-02	kg/kg feedstock
CO₂ biogenic	0.00E+00	kg/kg feedstock
H ₂ O (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	5.26E-02	kg/kg feedstock
H ₂ (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	0.00E+00	kg/kg feedstock
CO (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)		kg/kg feedstock
CO fossil	0.00E+00	kg/kg feedstock
CO biogenic	0.00E+00	kg/kg feedstock
H ₂ S (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	0.00E+00	kg/kg feedstock
NH ₃ (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	0.00E+00	kg/kg feedstock
COS (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	0.00E+00	kg/kg feedstock
SO ₂ (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	0.00E+00	kg/kg feedstock
Emissions to water	Amount (case 2)	Unit
Outputs to technosphere: Waste treatment	Amount (case 2)	Unit
Wastewater, average {RoW} treatment of wastewater, average, wastewater treatment Cut-off, U	1.53E-04	m ³ /kg feedstock

9.2.1.7 Water scrubber process

Table 9-12. Life cycle inventory of water scrubber process.

Water scrubber (FU = 1 kg of waste)		
PRODUCTS		
Outputs to technosphere: Products and co-products	Amount	Unit
Biomethane	6.48E-04	MWh/kg feedstock
Outputs to technosphere: Avoided products	Amount	Unit
INPUT		
Inputs from nature	Amount	Unit
Water	2.60E-05	m ³ /kg feedstock
Air	3.38E-01	kg/kg feedstock
Cooling water	5.13E-05	m ³ /kg feedstock
Inputs from technosphere: materials/fuels	Amount	Unit

Chemical factory, organics {RoW} chemical factory construction, organics Cut-off, U (water scrubber)	4.02E-22	kg/ kg feedstock
OUTPUT		
Emissions to air	Amount	Unit
CH4 biogenic (WS)	1.02E-03	kg/kg feedstock
CH4 fossil (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)		kg/kg feedstock
N ₂ (WS, flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	2.66E-01	kg/kg feedstock
O ₂ (WS, flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	7.05E-02	kg/kg feedstock
CO ₂ biogenic (WS)	1.17E-01	kg/kg feedstock
CO ₂ fossil (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	0.00E+00	kg/kg feedstock
H ₂ O (WS, flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	4.68E-03	kg/kg feedstock
H ₂ (WS, flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	0.00E+00	kg/kg feedstock
CO fossil (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	0.00E+00	kg/kg feedstock
CO biogenic (WS)	0.00E+00	kg/kg feedstock
H ₂ S (WS, flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	8.29E-04	kg/kg feedstock
NH ₃ (WS, flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	2.53E-03	kg/kg feedstock
COS (WS, flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	0.00E+00	kg/kg feedstock
SO ₂ (WS, flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	0.00E+00	kg/kg feedstock
Emissions to water	Amount	Unit
H₂O	0	m ³ /kg feedstock
N ₂	0	kg/kg feedstock
O ₂	0	kg/kg feedstock
NO ₂	0	kg/kg feedstock
NO	0	kg/kg feedstock
S	0	kg/kg feedstock
SO ₂	0	kg/kg feedstock

SO ₃	0	kg/kg feedstock
H ₂	0	kg/kg feedstock
Cl ₂	0	kg/kg feedstock
HCl	0	kg/kg feedstock
C	0	kg/kg feedstock
CO	0	kg/kg feedstock
CO ₂	0	kg/kg feedstock
COS	0	kg/kg feedstock
CH ₄	0	kg/kg feedstock
H ₂ S	0	kg/kg feedstock
NH ₃	0	kg/kg feedstock
Outputs to technosphere: Waste treatment	Amount	Unit
Wastewater, average {RoW} treatment of wastewater, average, wastewater treatment Cut-off, U	1.27E-05	m ³ /kg feedstock

9.2.1.8 Methane liquefier process

Table 9-13. Life cycle inventory of methane liquefier.

Methane liquifier (FU = 1 kg of waste)		
PRODUCTS		
Outputs to technosphere: Products and co-products	Amount	Unit
Liquid biomethane	6.33E-04	MWh/kg feedstock
Outputs to technosphere: Avoided products	Amount	Unit
INPUT		
Inputs from nature	Amount	Unit
Water	0	m ³ /kg feedstock
Air	0	kg/kg feedstock
Cooling water	2.10E-08	m ³ /kg feedstock
Inputs from technosphere: materials/fuels	Amount	Unit

Natural gas processing plant {GLO} natural gas processing plant production Cut-off, U	7.63E-23	m ³ /kg feedstock
Liquid storage tank, chemicals, organics {RoW} liquid storage tank production, chemicals, organics Cut-off, U	2.52E-12	p
OUTPUT		
Emissions to air	Amount	Unit
CH ₄ biogenic (WS)	0	kg/kg feedstock
CH ₄ fossil (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	0	kg/kg feedstock
N ₂ (WS, flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	0	kg/kg feedstock
O ₂ (WS, flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	0	kg/kg feedstock
CO ₂ biogenic (WS)	0	kg/kg feedstock
CO ₂ fossil (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	0	kg/kg feedstock
H ₂ O (WS, flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	0	kg/kg feedstock
H ₂ (WS, flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	0	kg/kg feedstock
CO fossil (flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	0	kg/kg feedstock
CO biogenic (WS)	0	kg/kg feedstock
H ₂ S (WS, flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	0	kg/kg feedstock
NH ₃ (WS, flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	0	kg/kg feedstock
COS (WS, flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	0	kg/kg feedstock
SO ₂ (WS, flue gas from GT: baseline, pre and post CCS and flue gas from WGSR: pre CCS)	0	kg/kg feedstock
Emissions to water	Amount	Unit
H₂O	0	m ³ /kg feedstock
N ₂	0	kg/kg feedstock
O ₂	0	kg/kg feedstock
NO ₂	0	kg/kg feedstock
NO	0	kg/kg feedstock
S	0	kg/kg feedstock
SO ₂	0	kg/kg feedstock
SO ₃	0	kg/kg feedstock
H ₂	0	kg/kg feedstock
Cl ₂	0	kg/kg feedstock
HCl	0	kg/kg feedstock
C	0	kg/kg feedstock
CO	0	kg/kg feedstock
CO ₂	0	kg/kg feedstock
COS	0	kg/kg feedstock

CH ₄	0	kg/kg feedstock
H ₂ S	0	kg/kg feedstock
NH ₃	0	kg/kg feedstock
Outputs to technosphere: Waste treatment		Amount
Wastewater, average {RoW} treatment of wastewater, average, wastewater treatment Cut-off, U	5.19E-09	m ³ /kg feedstock

9.2.2 Life cycle assessment: midpoint environmental impacts of Research Case 2

Table 9-14. Midpoint environmental impacts of Scenarios 1 and 2 (FU: 1 kg of waste).

Impact category	Unit	Scenario 1			Scenario 2		
		Baseline case	Case with pre-CCS	Case with post-CCS	Baseline case	Case with pre-CCS	Case with post-CCS
Global warming	kg CO ₂ eq	6.62E-01	3.47E-03	-1.91E-01	7.45E-01	2.22E-01	1.23E-01
Stratospheric ozone depletion	kg CFC11 eq	9.21E-10	1.56E-09	1.60E-09	7.57E-10	1.29E-09	1.26E-09
Ionizing radiation	kBq Co-60 eq	6.17E-05	2.08E-04	2.43E-04	4.31E-05	1.66E-04	1.76E-04
Ozone formation, Human health	kg NO _x eq	3.33E-06	1.75E-05	2.14E-05	2.47E-06	1.43E-05	1.57E-05
Fine particulate matter formation	kg PM2.5 eq	9.14E-07	9.74E-05	1.25E-05	6.64E-07	8.01E-06	9.14E-06
Ozone formation, Terrestrial ecosystems	kg NO _x eq	3.45E-06	1.83E-05	2.24E-05	2.55E-06	1.50E-05	1.64E-05
Terrestrial acidification	kg SO ₂ eq	5.19E-06	2.51E-05	2.70E-05	4.95E-03	4.97E-03	4.97E-03
Freshwater eutrophication	kg P eq	1.82E-06	4.07E-06	4.10E-06	1.48E-06	3.41E-06	3.15E-06
Marine eutrophication	kg N eq	3.36E-06	6.78E-06	7.47E-06	2.85E-06	5.67E-06	5.86E-06
Terrestrial ecotoxicity	kg 1,4-DCB	2.19E-03	1.53E-02	1.06E-02	1.69E-03	2.17E-02	7.86E-03
Freshwater ecotoxicity	kg 1,4-DCB	6.85E-05	2.01E-05	3.15E-04	5.23E-05	4.85E-04	2.33E-04
Marine ecotoxicity	kg 1,4-DCB	2.11E-05	2.75E-04	9.79E-05	1.60E-05	1.38E-04	7.23E-05

Human carcinogenic toxicity	kg 1,4-DCB	6.29E-05	2.01E-05	2.60E-05	4.56E-07	1.67E-05	1.90E-05
Human non-carcinogenic toxicity	kg 1,4-DCB	3.54E-03	2.75E-04	1.95E-04	2.65E-05	2.45E-04	1.43E-04
Land use	m ² a crop eq	7.27E-05	2.20E-04	2.57E-04	6.08E-05	1.85E-04	1.96E-04
Mineral resource scarcity	kg Cu eq	7.99E-06	4.79E-05	4.17E-05	5.83E-06	4.15E-05	3.06E-05
Fossil resource scarcity	kg oil eq	2.81E-04	2.50E-03	3.20E-03	2.01E-04	2.04E-03	2.34E-03
Water consumption	m ³	1.18E-03	1.41E-03	1.04E-03	1.08E-03	1.31E-03	9.94E-04

9.3 Life cycle assessment of Research Case 3

Research Case 3 produces CCHP and hydrogen from food and plastic waste and is assessed using TEA and LCA. Two scenarios are considered based on the location of carbon capture: Scenario 1 applies pre-CCS, while Scenario 2 applies post-CCS. Each scenario consists of three cases. In Case 1, food and plastic waste are separated, with food waste processed through anaerobic digestion and plastic waste through plasma gasification. In Case 2, the two waste streams are not separated and are treated together using plasma gasification. Case 3 build upon Case 1 by integrating a Steam Methane Reformer (SMR) to enhance hydrogen production from the biogas generated during anaerobic digestion. The results of Research Case 3 have been previously published [67].

9.3.1 Life cycle inventory of Research Case 3

9.3.1.1 Life cycle inventory of Scenario 1

The selected inventory data in SimaPro for the process materials, energy, and processes for Scenarios 1, and 2 are in Table 9-15. The functional unit for this assessment is 1 kg of waste, composed of 74% food waste and 26% plastic waste.

Table 9-15. Life cycle inventory of Scenario 1.

	Case 1	Case 2	Case 3	
Products				
Outputs to technosphere: Products and co-products	Amount			Unit
Hydrogen	1.20E-03	6.02E-04	1.06E-03	MWh/kg feedstock
Electricity	0.00E+00	0.00E+00	0.00E+00	MWh/kg feedstock
Cooling	2.48E-06	2.73E-06	2.28E-06	MWh/kg feedstock
Heating	1.15E-05	1.30E-05	1.09E-05	MWh/kg feedstock
Outputs to technosphere: Avoided products	Amount			Unit
Inputs				
Inputs from nature	Amount			Unit
Water	9.21E-04	8.08E-04	1.16E-03	m ³ /kg feedstock
Air	6.08E+00	7.70E+00	6.44E+00	kg/ kg feedstock
Cooling water	4.96E-03	7.40E-03	6.25E-03	m ³ / kg feedstock
Inputs from technosphere: materials/fuels	Amount			Unit
Plastic waste	2.64E-01	2.64E-01	2.64E-01	kg/kg feedstock
Food waste	7.36E-01	7.36E-01	7.36E-01	kg/kg feedstock
Synthetic gas factory {RoW} synthetic gas factory construction Cut-off, U (PG)	1.65E-10	1.65E-10	1.65E-10	p
Anaerobic digestion plant, agricultural {RoW} anaerobic digestion plant construction, agricultural Cut-off, U (AD)	2.27E-10	2.27E-10	2.27E-10	p
Heat and power co-generation unit, 1MW electrical, common components for heat+electricity {RoW} heat and power co-generation unit construction, 1MW electrical, common components for heat+electricity Cut-off, U (CCHP)	1.66E-09	1.78E-09	1.62E-09	p
Absorption chiller, 100kW {RoW} absorption chiller	1.88E-11	2.00E-11	1.79E-11	p

production, 100kW Cut-off, U (ARC)				
Hydrogen, gaseous {RoW} hydrogen production, steam reforming Cut-off, U (SMR)	0.00E+00	0.00E+00	8.26E-06	p
Chemical factory, organics {RoW} chemical factory construction, organics Cut-off, U (WGSR)	3.09E-15	3.47E-15	3.41E-15	p
Chemical factory, organics {RoW} chemical factory construction, organics Cut-off, U (PSA)	6.22E-16	4.11E-16	5.77E-16	p
Chemical factory, organics {RoW} chemical factory construction, organics Cut-off, U (CCS)	1.71E-15	1.84E-15	1.80E-15	p
Monoethanolamine {RoW} ethanolamine production Cut-off, U (MEA CCS)	1.31E-03	1.52E-03	1.46E-03	kg/kg feedstock
Zeolite, powder {RoW} zeolite production, powder Cut-off, U (PSA catalyst)	2.82E-05	1.16E-05	1.62E-05	kg/kg feedstock
Ferrochromium, high-carbon, 55% Cr {RoW} ferrochromium production, high-carbon, 55% Cr Cut-off, U (HTWGSR)	2.87E-05	2.15E-05	9.42E-06	kg/kg feedstock
Copper oxide {RoW} copper oxide production Cut-off, U (LTWGSR)	1.08E-05	1.43E-05	2.70E-06	kg/kg feedstock
Zinc oxide {RoW} zinc oxide production Cut-off, U (LTWGSR)	8.61E-06	1.14E-05	2.16E-06	kg/kg feedstock
Aluminium oxide, metallurgical {IAI Area, Asia, without China and GCC} aluminium oxide production Cut-off, U (LTWGSR)	2.15E-06	2.86E-06	5.40E-07	kg/kg feedstock
Nickel, class 1 {GLO} smelting and refining of nickel concentrate, 16% Ni Cut-off, U (SMR-Ni Al₂O₃)	0.00E+00	0.00E+00	2.78E-05	kg/kg feedstock
Inputs from technosphere: electricity/heat	Amount			Unit
Outputs				
Emissions to air	Amount			Unit
Combuster for steam generation in the WGSR				
CH ₄	2.63E-27	1.79E-27	5.60E-27	kg/kg feedstock
CH₄ biogenic	2.63E-27	5.46E-28	5.60E-27	kg/kg feedstock

CH₄ fossil	0.00E+00	1.24E-27	0.00E+00	kg/kg feedstock
N ₂	4.61E-01	5.82E-01	4.79E-01	kg/kg feedstock
O ₂	5.10E-02	8.09E-02	6.69E-02	kg/kg feedstock
CO ₂	1.12E-01	9.13E-02	9.94E-02	kg/kg feedstock
CO₂ biogenic	1.12E-01	2.78E-02	9.94E-02	kg/kg feedstock
CO₂ fossil	0.00E+00	6.34E-02	0.00E+00	kg/kg feedstock
H ₂ O	5.35E-02	7.76E-02	1.09E-01	kg/kg feedstock
H ₂	1.92E-07	2.49E-07	3.69E-07	kg/kg feedstock
CO	7.19E-06	5.22E-06	6.02E-06	kg/kg feedstock
CO biogenic	7.19E-06	1.59E-06	6.02E-06	kg/kg feedstock
CO fossil	0.00E+00	3.63E-06	0.00E+00	kg/kg feedstock
H ₂ S	7.24E-16	1.30E-16	0.00E+00	kg/kg feedstock
NH ₃	1.24E-13	1.62E-13	2.87E-13	kg/kg feedstock
COS	1.22E-16	1.23E-17	0.00E+00	kg/kg feedstock
SO ₂	7.14E-04	1.57E-04	0.00E+00	kg/kg feedstock
Combuster in the PSA unit for providing heat for food waste (Case 2) and for SMR (Case 3)				
CH ₄		8.60E-29	1.69E-27	kg/kg feedstock
CH₄ biogenic		2.62E-29	3.07E-28	kg/kg feedstock
CH₄ fossil		5.98E-29	1.38E-27	kg/kg feedstock
N ₂		6.92E-01	6.03E-01	kg/kg feedstock
O ₂		8.64E-02	6.84E-02	kg/kg feedstock
CO ₂		1.13E-02	1.38E-02	kg/kg feedstock
CO₂ biogenic		3.46E-03	2.49E-03	kg/kg feedstock
CO₂ fossil		7.87E-03	1.13E-02	kg/kg feedstock

H ₂ O		1.13E-01	1.04E-01	kg/kg feedstock
H ₂		2.30E-07	4.82E-07	kg/kg feedstock
CO		3.88E-07	1.18E-06	kg/kg feedstock
CO biogenic		1.19E-07	2.14E-07	kg/kg feedstock
CO fossil		2.70E-07	9.65E-07	kg/kg feedstock
H ₂ S		9.76E-17	0.00E+00	kg/kg feedstock
NH ₃		1.54E-13	4.17E-13	kg/kg feedstock
COS		7.36E-19	0.00E+00	kg/kg feedstock
SO ₂		2.31E-04	0.00E+00	kg/kg feedstock
Combuster in the CCGT				
CH ₄	1.10E-27	4.92E-28	5.24E-28	kg/kg feedstock
CH₄ biogenic	1.84E-28	1.50E-28	9.50E-29	kg/kg feedstock
CH₄ fossil	9.12E-28	3.42E-28	4.29E-28	kg/kg feedstock
N ₂	4.21E+00	4.64E+00	3.86E+00	kg/kg feedstock
O ₂	7.11E-01	8.17E-01	6.80E-01	kg/kg feedstock
CO ₂	1.36E-01	5.45E-02	5.94E-02	kg/kg feedstock
CO₂ biogenic	2.29E-02	1.66E-02	1.08E-02	kg/kg feedstock
CO₂ fossil	1.13E-01	3.79E-02	4.86E-02	kg/kg feedstock
H ₂ O	4.27E-01	5.19E-01	4.28E-01	kg/kg feedstock
H ₂	2.81E-07	3.36E-07	2.76E-07	kg/kg feedstock
CO	1.56E-06	6.14E-07	6.69E-07	kg/kg feedstock
CO biogenic	2.62E-07	1.87E-07	1.21E-07	kg/kg feedstock
CO fossil	1.30E-06	4.27E-07	5.48E-07	kg/kg feedstock
H ₂ S	8.95E-17	1.23E-16	0.00E+00	kg/kg feedstock
NH ₃	1.52E-12	1.88E-12	1.54E-12	kg/kg feedstock
COS	2.51E-18	1.14E-18	0.00E+00	kg/kg feedstock

SO ₂	8.47E-04	1.12E-03	0.00E+00	kg/kg feedstock
Total emission to air				
CH ₄	3.73E-27	2.37E-27	7.81E-27	kg/kg feedstock
CH₄ biogenic	2.81E-27	7.22E-28	6.00E-27	kg/kg feedstock
CH₄ fossil	9.12E-28	1.64E-27	1.81E-27	kg/kg feedstock
N ₂	4.67E+00	5.92E+00	4.95E+00	kg/kg feedstock
O ₂	7.62E-01	9.84E-01	8.15E-01	kg/kg feedstock
CO ₂	2.48E-01	1.57E-01	1.73E-01	kg/kg feedstock
CO₂ biogenic	1.35E-01	4.79E-02	1.13E-01	kg/kg feedstock
CO₂ fossil	1.13E-01	1.09E-01	5.99E-02	kg/kg feedstock
H ₂ O	4.81E-01	7.10E-01	6.40E-01	kg/kg feedstock
H ₂	4.73E-07	8.15E-07	1.13E-06	kg/kg feedstock
CO	8.75E-06	6.22E-06	7.86E-06	kg/kg feedstock
CO biogenic	7.46E-06	1.90E-06	6.35E-06	kg/kg feedstock
CO fossil	1.30E-06	4.32E-06	1.51E-06	kg/kg feedstock
H ₂ S	8.14E-16	3.51E-16	0.00E+00	kg/kg feedstock
NH ₃	1.65E-12	2.20E-12	2.25E-12	kg/kg feedstock
COS	1.25E-16	1.42E-17	0.00E+00	kg/kg feedstock
SO ₂	1.56E-03	1.51E-03	0.00E+00	kg/kg feedstock
Emissions to water				
Outputs to technosphere: Waste treatment				
Wastewater, average {RoW} treatment of wastewater, average, wastewater treatment Cut-off, U	1.64E-04	2.77E-04	2.74E-04	m ³ /kg feedstock

9.3.1.2 Life cycle inventory of Scenario 2

Table 9-16. Life cycle inventory of Scenario 2.

	Case 1	Case 2	Case 3	
Products				
Outputs to technosphere: Products and co-products	Amount			Unit
Hydrogen	9.14E-04	4.59E-04	7.73E-04	MWh/kg feedstock
Electricity	0.00E+00	0.00E+00	0.00E+00	MWh/kg feedstock
Cooling	2.77E-06	2.84E-06	2.60E-06	MWh/kg feedstock
Heating	1.32E-05	1.35E-05	1.24E-05	MWh/kg feedstock
Outputs to technosphere: Avoided products	Amount			Unit
Inputs				
Inputs from nature	Amount			Unit
Water	9.21E-04	8.08E-04	1.16E-03	m ³ /kg feedstock
Air	6.04E+00	7.06E+00	6.40E+00	kg/ kg feedstock
Cooling water	4.66E-03	7.86E-03	6.02E-03	m ³ / kg feedstock
Inputs from technosphere: materials/fuels	Amount			Unit
Plastic waste	2.64E-01	2.64E-01	2.64E-01	kg/kg feedstock
Food waste	7.36E-01	7.36E-01	7.36E-01	kg/kg feedstock
Synthetic gas factory {RoW} synthetic gas factory construction Cut-off, U (PG)	1.65E-10	1.65E-10	1.65E-10	p
Anaerobic digestion plant, agricultural {RoW} anaerobic digestion plant construction, agricultural Cut-off, U (AD)	2.27E-10	2.27E-10	2.27E-10	p
Heat and power co-generation unit, 1MW electrical, common components for heat+electricity {RoW} heat and power co-generation unit construction, 1MW electrical, common components for	1.73E-09	1.80E-09	1.68E-09	p

heat+electricity Cut-off, U (CCHP)				
Absorption chiller, 100kW {RoW} absorption chiller production, 100kW Cut-off, U (ARC)	2.01E-11	2.04E-11	1.94E-11	p
Hydrogen, gaseous {RoW} hydrogen production, steam reforming Cut-off, U (SMR)	0.00E+00	0.00E+00	8.26E-06	p
Chemical factory, organics {RoW} chemical factory construction, organics Cut-off, U (WGSR)	3.09E-15	3.47E-15	3.41E-15	p
Chemical factory, organics {RoW} chemical factory construction, organics Cut-off, U (PSA)	5.28E-16	3.49E-16	4.78E-16	p
Chemical factory, organics {RoW} chemical factory construction, organics Cut-off, U (CCS)	3.54E-15	3.89E-15	3.66E-15	p
Monoethanolamine {RoW} ethanolamine production Cut-off, U (MEA CCS)	1.72E-03	1.73E-03	1.72E-03	kg/kg feedstock
Zeolite, powder {RoW} zeolite production, powder Cut-off, U (PSA catalyst)	1.06E-04	4.96E-05	7.12E-05	kg/kg feedstock
Ferrochromium, high-carbon, 55% Cr {RoW} ferrochromium production, high-carbon, 55% Cr Cut-off, U (HTWGSR)	3.11E-05	2.33E-05	3.52E-05	kg/kg feedstock
Copper oxide {RoW} copper oxide production Cut-off, U (LTWGSR)	1.17E-05	1.43E-05	1.19E-05	kg/kg feedstock
Zinc oxide {RoW} zinc oxide production Cut-off, U (LTWGSR)	9.33E-06	1.14E-05	9.49E-06	kg/kg feedstock
Aluminium oxide, metallurgical {IAI Area, Asia, without China and GCC} aluminium oxide production Cut-off, U (LTWGSR)	2.33E-06	2.86E-06	2.37E-06	kg/kg feedstock
Nickel, class 1 {GLO} smelting and refining of nickel concentrate, 16% Ni Cut-off, U (SMR-Ni Al2O3)	0.00E+00	0.00E+00	2.78E-05	kg/kg feedstock
Inputs from technosphere: electricity/heat	Amount			Unit
Outputs				
Emissions to air	Amount			Unit

Combuster for steam generation in the WGSR				
CH ₄	2.63E-27	1.79E-27	5.60E-27	
CH₄ biogenic	2.63E-27	5.46E-28	5.60E-27	kg/kg feedstock
CH₄ fossil	0.00E+00	1.24E-27	0.00E+00	kg/kg feedstock
N ₂	4.61E-01	5.82E-01	4.79E-01	kg/kg feedstock
O ₂	5.10E-02	8.09E-02	6.69E-02	kg/kg feedstock
CO ₂	1.12E-01	9.13E-02	9.94E-02	
CO₂ biogenic	1.12E-01	2.78E-02	9.94E-02	kg/kg feedstock
CO₂ fossil	0.00E+00	6.34E-02	0.00E+00	kg/kg feedstock
H ₂ O	5.35E-02	7.76E-02	1.09E-01	kg/kg feedstock
H ₂	1.92E-07	2.49E-07	3.69E-07	kg/kg feedstock
CO	7.19E-06	5.22E-06	6.02E-06	
CO biogenic	7.19E-06	1.59E-06	6.02E-06	kg/kg feedstock
CO fossil	0.00E+00	3.63E-06	0.00E+00	kg/kg feedstock
H ₂ S	7.24E-16	1.30E-16	0.00E+00	kg/kg feedstock
NH ₃	1.24E-13	1.62E-13	2.87E-13	kg/kg feedstock
COS	1.22E-16	1.23E-17	0.00E+00	kg/kg feedstock
SO ₂	7.14E-04	1.57E-04	0.00E+00	kg/kg feedstock
Combuster in the PSA unit for providing heat for food waste (Case 2) and for SMR (Case 3)				
CH ₄		9.83E-27	1.13E-26	kg/kg feedstock
CH₄ biogenic		3.00E-27	2.04E-27	kg/kg feedstock
CH₄ fossil		6.83E-27	9.21E-27	kg/kg feedstock
N ₂		5.19E-01	5.07E-01	kg/kg feedstock
O ₂		4.53E-02	4.48E-02	kg/kg feedstock
CO ₂		1.23E-01	1.29E-01	kg/kg feedstock

CO₂ biogenic		3.75E-02	2.33E-02	kg/kg feedstock
CO₂ fossil		8.54E-02	1.05E-01	kg/kg feedstock
H ₂ O		1.10E-01	1.07E-01	kg/kg feedstock
H ₂		4.04E-07	4.06E-07	kg/kg feedstock
CO		7.91E-06	8.63E-06	kg/kg feedstock
CO biogenic		2.42E-06	1.56E-06	kg/kg feedstock
CO fossil		5.50E-06	7.06E-06	kg/kg feedstock
H ₂ S		4.24E-16	0.00E+00	kg/kg feedstock
NH ₃		3.37E-13	3.39E-13	kg/kg feedstock
COS		3.75E-17	0.00E+00	kg/kg feedstock
SO ₂		2.06E-04	0.00E+00	kg/kg feedstock
Combuster in the CCGT				
CH ₄	1.13E-26	1.08E-26	1.02E-26	kg/kg feedstock
CH₄ biogenic	1.90E-27	3.28E-27	1.85E-27	kg/kg feedstock
CH₄ fossil	9.41E-27	7.47E-27	8.37E-27	kg/kg feedstock
N ₂	4.18E+00	4.32E+00	3.93E+00	kg/kg feedstock
O ₂	6.34E-01	6.78E-01	6.21E-01	kg/kg feedstock
CO ₂	7.91E-01	6.96E-01	6.75E-01	kg/kg feedstock
CO₂ biogenic	1.33E-01	2.12E-01	1.22E-01	kg/kg feedstock
CO₂ fossil	6.58E-01	4.84E-01	5.53E-01	kg/kg feedstock
H ₂ O	5.07E-01	5.64E-01	5.12E-01	kg/kg feedstock
H ₂	3.67E-07	4.01E-07	3.64E-07	kg/kg feedstock
CO	9.98E-06	8.63E-06	8.35E-06	kg/kg feedstock
CO biogenic	1.68E-06	2.63E-06	1.51E-06	kg/kg feedstock
CO fossil	8.30E-06	6.00E-06	6.84E-06	kg/kg feedstock
H ₂ S	1.31E-16	1.86E-16	0.00E+00	kg/kg feedstock

NH ₃	2.10E-12	2.36E-12	2.13E-12	kg/kg feedstock
COS	1.80E-17	2.02E-17	0.00E+00	kg/kg feedstock
SO ₂	8.47E-04	1.18E-03	0.00E+00	kg/kg feedstock
Total emission to air				
CH ₄	1.39E-26	2.24E-26	2.71E-26	kg/kg feedstock
CH₄ biogenic	4.53E-27	6.83E-27	9.49E-27	kg/kg feedstock
CH₄ fossil	9.41E-27	1.55E-26	1.76E-26	kg/kg feedstock
N ₂	4.64E+00	5.43E+00	4.91E+00	kg/kg feedstock
O ₂	6.84E-01	8.04E-01	7.33E-01	kg/kg feedstock
CO ₂	4.21E-02	4.25E-02	4.21E-02	kg/kg feedstock
CO₂ biogenic	7.08E-03	7.15E-03	7.08E-03	kg/kg feedstock
CO₂ fossil	3.50E-02	3.53E-02	3.50E-02	kg/kg feedstock
H ₂ O	1.39E-01	1.62E-01	1.47E-01	kg/kg feedstock
H ₂	5.60E-07	1.05E-06	1.14E-06	kg/kg feedstock
CO	1.72E-05	2.18E-05	2.30E-05	kg/kg feedstock
CO biogenic	2.89E-06	3.66E-06	3.87E-06	kg/kg feedstock
CO fossil	1.43E-05	1.81E-05	1.91E-05	kg/kg feedstock
H ₂ S	8.55E-16	7.41E-16	0.00E+00	kg/kg feedstock
NH ₃	2.23E-12	2.86E-12	2.75E-12	kg/kg feedstock
COS	1.40E-16	7.00E-17	0.00E+00	kg/kg feedstock
SO ₂	1.56E-03	1.54E-03	0.00E+00	kg/kg feedstock
Emissions to water	Amount			Unit
Outputs to technosphere: Waste treatment	Amount			Unit
Wastewater, average {RoW} treatment of wastewater, average, wastewater treatment Cut-off, U	1.53E-04	2.80E-04	2.77E-04	m ³ /kg feedstock

9.3.2 Life cycle assessment: midpoint environmental impacts of Research Case 3

Table 9-17. Midpoint environmental impacts of Scenario 1 (FU: 1 kg of waste).

Impact category	Unit	Case		
		1	2	3
Global warming	kg CO ₂ eq	4.84E-02	-3.80E-02	-2.09E-02
Stratospheric ozone depletion	kg CFC11 eq	1.08E-09	1.30E-09	1.34E-09
Ionizing radiation	kBq Co-60 eq	1.89E-04	2.11E-04	2.06E-04
Ozone formation, Human health	kg NO _x eq	1.68E-05	1.91E-05	1.88E-05
Fine particulate matter formation	kg PM2.5 eq	9.66E-06	1.11E-05	1.06E-05
Ozone formation, Terrestrial ecosystems	kg NO _x eq	1.76E-05	2.00E-05	1.97E-05
Terrestrial acidification	kg SO ₂ eq	1.58E-03	1.54E-03	6.06E-05
Freshwater eutrophication	kg P eq	3.09E-06	3.76E-06	3.55E-06
Marine eutrophication	kg N eq	4.65E-06	5.87E-06	5.70E-06
Terrestrial ecotoxicity	kg 1,4-DCB	2.34E-02	2.78E-02	1.38E-02
Freshwater ecotoxicity	kg 1,4-DCB	5.24E-04	6.47E-04	5.34E-04
Marine ecotoxicity	kg 1,4-DCB	1.50E-04	1.84E-04	1.56E-04
Human carcinogenic toxicity	kg 1,4-DCB	2.03E-05	2.35E-05	2.56E-05
Human non-	kg 1,4-DCB	2.73E-04	3.30E-04	2.21E-04

carcinogenic toxicity				
Land use	m2a crop eq	2.10E-04	2.35E-04	2.33E-04
Mineral resource scarcity	kg Cu eq	4.79E-05	5.51E-05	1.22E-04
Fossil resource scarcity	kg oil eq	2.45E-03	2.80E-03	2.77E-03
Water consumption	m ³	5.43E-02	7.42E-03	6.62E-03

Table 9-18. Midpoint environmental impacts of Scenario 2 (FU: 1 kg of waste).

Impact category	Unit	Case		
		1	2	3
Global warming	kg CO ₂ eq	-1.55E-01	-1.59E-01	-1.55E-01
Stratospheric ozone depletion	kg CFC11 eq	1.46E-09	1.51E-09	1.69E-09
Ionizing radiation	kBq Co-60 eq	2.56E-04	2.46E-04	2.66E-04
Ozone formation, Human health	kg NO _x eq	2.24E-05	2.20E-05	2.36E-05
Fine particulate matter formation	kg PM2.5 eq	1.27E-05	1.27E-05	1.29E-05
Ozone formation, Terrestrial ecosystems	kg NO _x eq	2.35E-05	2.31E-05	2.47E-05
Terrestrial acidification	kg SO ₂ eq	1.59E-03	1.57E-03	6.93E-05
Freshwater eutrophication	kg P eq	3.89E-06	4.19E-06	4.52E-06
Marine eutrophication	kg N eq	5.79E-06	6.52E-06	6.67E-06
Terrestrial ecotoxicity	kg 1,4-DCB	2.77E-02	2.95E-02	2.95E-02

Freshwater ecotoxicity	kg 1,4-DCB	6.36E-04	6.92E-04	8.50E-04
Marine ecotoxicity	kg 1,4-DCB	1.84E-04	1.98E-04	2.43E-04
Human carcinogenic toxicity	kg 1,4-DCB	2.67E-05	2.69E-05	3.11E-05
Human non-carcinogenic toxicity	kg 1,4-DCB	3.41E-04	3.60E-04	3.71E-04
Land use	m ² a crop eq	2.65E-04	2.64E-04	2.82E-04
Mineral resource scarcity	kg Cu eq	6.37E-05	6.26E-05	1.46E-04
Fossil resource scarcity	kg oil eq	3.27E-03	3.23E-03	3.43E-03
Water consumption	m ³	5.20E-03	7.88E-03	6.42E-03

