

**INTEGRATION OF METHANE REFORMING AND CHEMICAL
LOOPING TECHNOLOGIES FOR POWER GENERATION FROM
WASTE PLASTIC: TECHNICAL AND ECONOMIC ASSESSMENT**

BY

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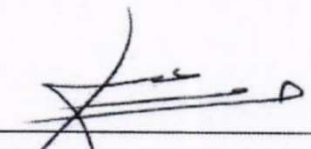
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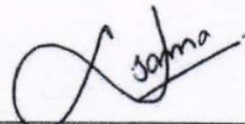
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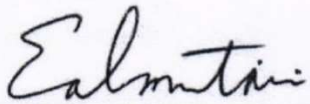
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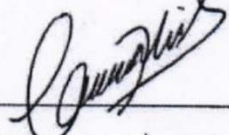
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In honor of all those who supported and inspired me along this academic voyage.

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LIST OF ABBREVIATIONS

ASU	:	Air Separation Unit
CAPEX	:	Capital Expenditure
CCPP	:	Combined Cycle Power Plant
CCS	:	Carbon Capture and Storage
CLC	:	Chemical Looping Combustion
DMR	:	Dry Methane Reforming
GHG	:	Green-House Gas
GT	:	Gas Turbine
HCR	:	Hydrogen to Carbon Ratio
HEN	:	Heat Exchanger Network
HHV	:	Higher Heating Value
HRSG	:	Heat Recovery Steam Generator
IGCC	:	Integrated Gasification and Combined Cycle
KSA	:	Kingdom of Saudia Arabia
LCOE	:	Levelized Cost of Electricity
LHV	:	Lower Heating Value
MSW	:	Municipal Solid Waste

NGCC	:	Natural Gas Combined Cycle
NPV	:	Net Present Value
OPEX	:	Operating Expenditure
PE	:	Poly-Ethylene
POX	:	Partial Oxidation
PP	:	Poly-Propylene
PSA	:	Pressure Swing Adsorption
SMR	:	Steam Methane Reforming
USPC	:	Ultra-Supercritical Pulverized Coal
VPSA	:	Vacuum Pressure Swing Adsorption
VSA	:	Vacuum Swing Adsorption

ABSTRACT

Full Name : Faisal M. Alqarzaee

Thesis Title : Integration of Methane Reforming and Chemical Looping
Technologies for Power Generation from Waste Plastic: Technical and
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The improper disposal of plastic waste disrupts ecosystems, endangers the environment. Techno-economic obstacles and a lack of government assistance are two obstacles that hinder efforts to promote plastic waste recycling. In this study, four recycling pathways - (1) gasification+steam methane reforming (SMR) with air separation unit (ASU), (2) gasification+SMR with chemical looping combustion (CLC), (3) gasification+partial oxidation (POX) with CLC, and (4) gasification+POX+dry methane reforming (DMR) with CLC- were created and examined in order to address this concern and produce power from plastic waste. Detailed simulations using Aspen Plus V12's produced material and energy balances that allowed for the sizing of equipment and the techno-economic analysis of these waste recycling pathways. Techno-economic analysis combines economic analysis, such as cash flow analysis to estimate levelized costs, with technical analysis, such as sensitivity analysis and energy/efficiency estimates. According to simulation results, cases 1 and 2, using similar SMR approaches, produced syngas with an HCR of 3.5 at a rate of 850 kg/hr. However, case 3, which used oxygen in the POX process, had the highest HHV because of its 99.2% combustible content even though it produced the least amount of syngas. Numerous factors affected both power production and efficiency, with cases 2 and 4 exhibiting higher efficiencies. For cases 2-4, heat integration in CLC cases made it feasible to produce more steam, which increased power output and overall process efficiency. According to sensitivity analysis,

power production in SMR increased linearly with CH₄ input. Significant disparities in CAPEX were found between the cases, with case 3 showing to be the most economical at \$94.6/MW compared to case 1's \$211.34/MW LCOE. |

ملخص الرسالة

الاسم الكامل: فيصل محمد القرزعي

عنوان الرسالة: تكامل تقنيات إصلاح الميثان والحلقات الكيميائية لتوليد الطاقة من نفايات البلاستيك: التقييم الفني والاقتصادي

التخصص: هندسة كيميائية

تاريخ الدرجة العلمية: ديسمبر 2024

إن التخلّص غير المناسب من مخلفات البلاستيك يحدث الخلل في النظم البيئية، ممّا يهدد البيئة. وتعتبر العقبات التقنية الاقتصادية ونقص الدعم الحكومي من العوائق الرئيسية التي تعترض جهود تعزيز عملية إعادة التدوير نفايات البلاستيك. في هذه الدراسة، تمّ إنشاء وفحص أربع طرق لإعادة التدوير وهي: (1) إصلاح غاز الميثان بالبخر بوحدة فصل الأكسجين من الهواء (ASU)، (2) إصلاح غاز الميثان بالبخر (SMR) وتوريد الأكسجين عن طريق الاحتراق بالحلقة الكيميائية (CLC)، (3) الأكسدة الجزئية بالأكسجين (POX) وتوريد الأكسجين عن طريق الاحتراق بالحلقة الكيميائية، و(4) إصلاح غاز الميثان الجاف (DMR) وتوريد الأكسجين عن طريق الاحتراق بالحلقة الكيميائية (CLC). وقد أنتجت المحاكاة التفصيلية باستخدام برنامج أسبن بلس في إصداره الـ 12 موازين المواد والطاقة. تم بعد ذلك تحليل الجدوى الاقتصادية التقنية التي تجمع بين التحليل الاقتصادي مثل تحليل التدفق النقدي لتقدير تكلفة التشغيل، والتحليل التقني مثل تحليل الحساسية وتقديرات الطاقة والكفاءة. وفقاً لنتائج المحاكاة، أظهرت الحالات 1 و 2، باستخدام أساليب SMR مماثلة، إنتاج سنجاز (غاز الهيدروجين واحادي اكسيد الكربون) بنسبة HCR تبلغ 3.5 بمعدل 850 كجم/ساعة. ومع ذلك، أحدثت الحالة 3، التي استخدمت الأكسجين في عملية POX، أعلى قيمة HHV بسبب احتوائها على 99.2% من المواد القابلة للاحتراق على الرغم من إنتاجها لأقل كمية من السنجاز. تأثرت العديد من العوامل كل من إنتاج الطاقة والكفاءة، حيث أظهرت الحالات 2 و 4 كفاءات أعلى. بالنسبة للحالات 2-4، جعل تكامل الحرارة في حالات CLC من الممكن إنتاج المزيد من البخار، مما زاد من إنتاج الطاقة وكفاءة العملية بشكل عام. ووفقاً لتحليل الحساسية، زاد إنتاج الطاقة في SMR بشكل خطي مع إدخال CH_4 . ووجدت اختلافات كبيرة في CAPEX بين الحالات، حيث أظهرت الحالة 3 أنها الأكثر اقتصادية بقيمة 94.6 دولار/ميغواط مقارنة بـ 211.34 دولار/ميغواط من LCOE في الحالة 1. |

CHAPTER 1

INTRODUCTION

1.1 Background

Because of their many uses and affordable production, polymers and plastics are utilized extensively around the world. Their useful qualities, which are superior to those of wood, metals, and ceramics in terms of price, chemical resistance, durability, and lower density, make them perfect for a range of applications. Every year, increasing amounts of plastic are produced; in 2015, there were 322 million tons of plastic produced worldwide [1]. For the past ten years, plastics have changed how people live. Traditional and innovative composite plastics, which are primarily employed in the packaging, automotive, electrical, construction, and electronics industries, are being produced in greater quantities [1].

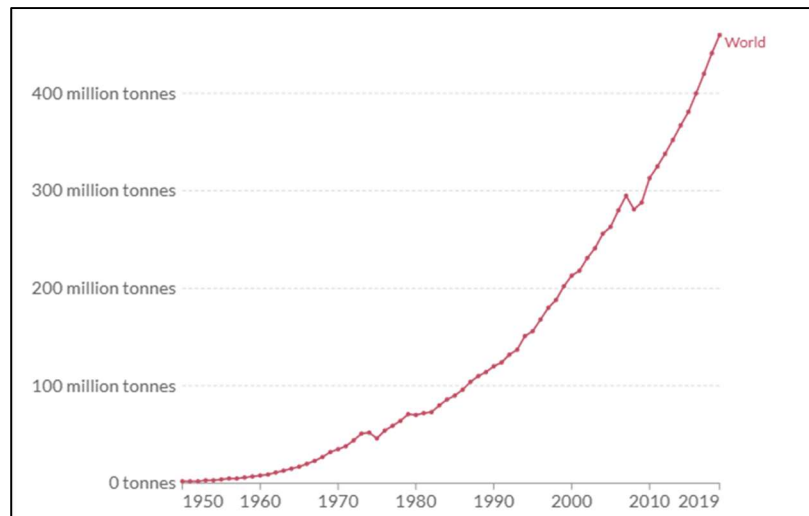


Figure 1 Global plastic production [2]

Plastic production has grown since 1950, and at the start of the twenty-first century, 200 million tons were produced worldwide, as Figure 1 illustrates. The production of plastic

was predicted to surpass 400 million tons annually by the end of 2019 as a result of industrial activity and societal lifestyle preferences [2].

Waste plastic that has not been recycled poses a risk to the environment, and if it is not controlled, the Kingdom of Saudi Arabia (KSA) may suffer serious consequences. Recycling used plastic garbage is expected to bring in 32 billion dollars annually for KSA by 2035 [3]. KSA is adopting policies pertaining to the reuse, recycling, and reconstitution of waste plastic in order to prioritize the nation's sustainable growth. The Saudi Arabian government has partnered with SIRC and SABIC to promote sustainable development.

According to a study, the US produces 2 billion tons of municipal solid waste (MSW) annually, and by 2050, the country is predicted to produce 3.4 billion tons of plastic waste [4]. Since the majority of solid and plastic garbage is disposed of in landfills or dumped into the ocean, this has grown to be a global problem. As a result, both the marine ecology and life have suffered. Polyethylene and polypropylene comprise 29.6% and 18.9%, respectively, of the most often used plastic wastes, while the remaining polymers make up 51.5% [4]. Incinerating plastic waste has become a popular substitute for landfilling; however, it releases harmful chemicals into the atmosphere, including CO₂ [2]. Most of these gases are greenhouse gases, which have the ability to alter the earth's climate on a worldwide scale. MSW includes waste from homes and businesses, as well as waste from agriculture, slaughterhouses, commercial solid waste, building and construction industry waste, and organic waste. By 2025, it is anticipated that the yearly production of MSW will total 2.2 billion tons [5].

Due to industrialization, MSW generation is higher in industrialized countries than in underdeveloped ones. It is anticipated that MSW generation in emerging countries will increase in this period of industrialization [4]. The MSW generation per capita in wealthy countries ranges from 350 to 1200 kg per year, whilst ranging from 250 to 550 kg per year in underdeveloped countries [4], [5]. Figure 2 illustrates the various types of MSW;



Figure 2 Types of waste [5]

The rise in municipal plastic waste was a result of the growth in plastic manufacture. As previously indicated, the used plastic is either disposed of in landfills or dumped into the ocean. In 2012, it was predicted that 25 million tons of waste plastic will be disposed of throughout the year. With the growth in global population, it is anticipated to increase [5].

Only 9% of the 6.3 billion tons of waste plastic produced worldwide between 1950 and 2018 was recycled. The incineration of waste plastic is extremely common as only 12% of it gets burned and the other 79% is released into the environment. Furans, mercury, and dioxins have all been made more prevalent by the released waste plastic. Due to their deadly nature, these hazardous substances pose a threat to the ecosystem and marine life. As the plastic manufacture sector took control, marine life has already been seriously disrupted. It is projected that 4-12 million tons of polymer garbage has been dumped annually into the water thereby disrupting the marine life [8].

1.2 Recycling Routes of Waste Plastics

Plastic waste reduction and reprocessing methods are illustrated below in Figure 3:

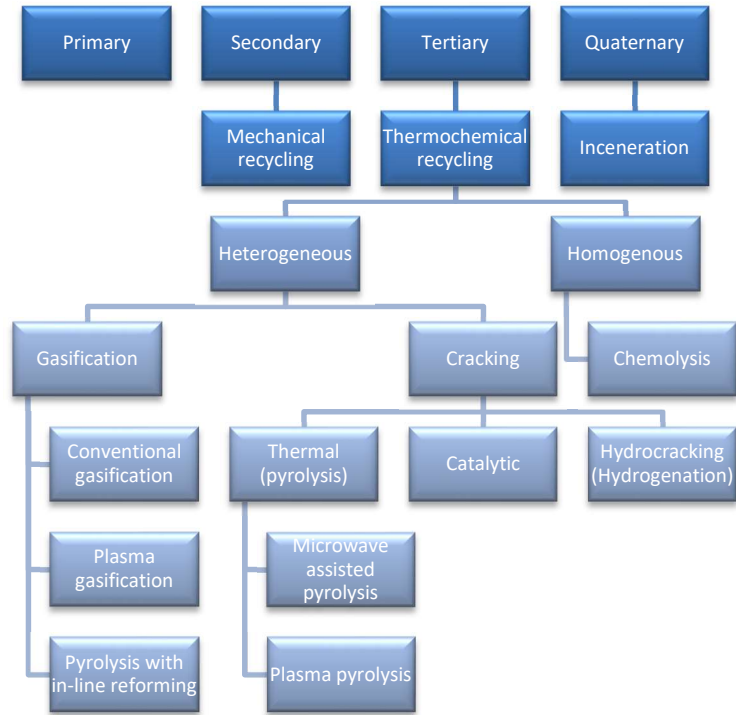


Figure 3 Plastic waste recycling routes [5]

Closed loop recycling is another moniker for primary recycling of waste plastic which involves mechanical reprocessing to create components of a high-quality product. Secondary recycling, also known as mechanical recycling, is the process of downgrading discarded plastic in order to create new plastic products. Tertiary recycling, which is essentially a chemical recycling process, is a commonly utilized technique. While recovering the recyclable materials, the waste plastic is broken down to monomers in this process. Then, the monomers are polymerized once more to create useful products. Plastic garbage is burned down (incinerated) as part of the quaternary recycling process to recover energy [5].

However, certain recycling methods have limitations. Contaminated plastics cannot undergo primary recycling, making municipal solid waste recycling unsuitable for this technique. Due to significant losses of high-quality goods, secondary recycling lacks

viability as well. Tertiary recycling has the capacity to recover high grade materials that can ultimately lead to the manufacture of raw materials. It is extremely beneficial since it encourages recycling and the effective recovery of usable items without harming the environment. The method of pyrolysis turns plastic waste into fuel, which could be a significant asset for sustainable growth. Gasification methods can also be used commercially to convert waste plastic into fuel or syngas. As a stopgap measure for trash recycling, the quaternary technique shortens the lifetime of the material. Given that 299 million tons of plastic trash were created globally in 2013, with Europe producing 19% of that total [3], tertiary recycling is highly recommended. Tertiary recycling may be advantageous in this specific case, even if waste plastic can be recycled into fuel or syngas for sustainable development, which makes about 26% of Europe's plastic waste [7].

Implementing trash disposal regulations and encouraging the recycling of used plastics have been the subject of substantial discussion. The destiny of future generations will be shaped by the existing policies on environmental sustainability if this issue is not addressed, since it could have serious negative effects on the environment. There has been extensive study on solutions for reducing plastic waste, recycling plastic, and turning waste plastic into energy. The viability of this initiative has improved, and there is a lot of promise for turning plastic trash into valuable fuels. The recycling process consists of four key processes i.e., gathering of plastic wastes, separation, manufacture and marketing of the new product.

1.3 Justification for the Syngas-Based Waste Plastic Conversion Process: A Sustainable Alternative to Direct Incineration

The reason for adopting the process of converting plastic to syngas, utilizing the syngas for electricity generation, and capturing CO₂ is based on key principles of efficiency, sustainability, and resource valorization. This approach is strategically designed to maximize energy recovery, minimize environmental impact, and create additional value from waste plastics. Compared to direct incineration, this process offers significant advantages that make it a more comprehensive and sustainable solution. The following arguments outline the rationale behind this approach.

Energy Efficiency and Electricity Generation

The proposed process transforms plastic into syngas (a mixture of hydrogen and carbon monoxide), which serves as a cleaner and more energy-dense intermediate fuel. Unlike direct incineration, where the plastic is combusted inefficiently and may produce unregulated toxic byproducts such as dioxins and furans, the gasification process enables precise control over the chemical reactions. This improves thermal efficiency, resulting in higher electricity generation per unit of plastic waste compared to simple incineration. Syngas combustion also allows better integration with power generation systems, such as combined cycle gas turbines, further enhancing energy recovery.

CO₂ Capture Efficiency

The proposed process captures CO₂ from a controlled and concentrated source after burning the syngas, making it more efficient and economical. In contrast, incineration releases CO₂ into a flue gas with a lower concentration of CO₂ and higher levels of nitrogen

and other impurities, complicating the capture process. By first converting plastic to syngas, the combustion step generates a more concentrated stream of CO₂, which is easier to separate and capture. This reduces the overall energy and cost requirements of CO₂ capture, supporting a more economically and environmentally viable carbon management strategy.

Syngas Utilization and Flexibility

The process creates syngas, which is a versatile intermediate product. Syngas can not only be burned for electricity generation but also used as a feedstock for producing high-value chemicals such as methanol, ammonia, or synthetic fuels. This provides flexibility for integrating the process with various industrial applications, transforming waste plastic into a resource rather than just waste to be disposed of. In contrast, incineration limits the potential to extract additional value from the plastic beyond heat generation.

Reduced Environmental Impact

The proposed process is designed to minimize the release of harmful pollutants. Incineration often leads to the formation of toxic byproducts like heavy metal emissions, particulate matter, and unburned hydrocarbons, especially if the process is not well-controlled. The gasification step operates under controlled conditions, breaking down plastics into syngas without producing these hazardous byproducts. By avoiding the environmental challenges associated with incineration, the process aligns with stricter regulatory standards and contributes to cleaner energy generation.

Circular Economy and Resource Valorization

This process supports a circular economy approach by converting plastic waste into valuable energy and resources. Incineration destroys the material completely, offering no pathway for the recovery of chemical value beyond CO₂ capture. The gasification-based route treats waste plastic as a resource, not merely waste, contributing to sustainable resource utilization and reducing dependency on fossil fuels for chemical and energy production.

1.4 Waste plastic gasification and Saudi Arabia's Vision 2030

The proposed process aligns with Saudi Arabia's Vision 2030 by addressing multiple sustainability goals, including waste management, carbon recycling, and the generation of cleaner energy. Vision 2030 emphasizes the transition to a circular carbon economy, where carbon emissions are minimized, reused, and recycled. This process contributes to these objectives by transforming plastic waste, which would otherwise be landfilled or incinerated, into valuable syngas for clean energy generation while capturing CO₂ for further utilization.

The integration of waste polyethylene/polypropylene (PE/PP) into the process directly tackles the issue of plastic waste, which represents a growing environmental challenge. Unlike incineration, which releases CO₂ and other pollutants without significant material recovery, the proposed system provides a dual benefit. First, it generates syngas, a versatile feedstock for power generation or chemical production, ensuring higher material and energy efficiency. Second, the CO₂ produced during the process is captured and either recycled within the system (e.g., for dry methane reforming) or utilized for other

applications, such as enhanced oil recovery or carbon-based products, thus contributing to the circular carbon economy framework.

While direct incineration of plastic waste may appear simpler, it is inherently less efficient in terms of material recovery and long-term environmental impact. Incineration converts all the waste's carbon content into CO₂, which, even when captured, does not yield additional value beyond storage. In contrast, the proposed process leverages the carbon content in plastic waste to produce high-value syngas, offering flexibility for downstream applications such as hydrogen production or synthetic fuels, aligning with Vision 2030's goals of diversifying the economy and fostering sustainable energy systems.

Furthermore, this approach underscores the importance of advancing clean technologies and innovations, which are key pillars of Vision 2030. By incorporating dry methane reforming (DMR) and chemical looping combustion (CLC) with CO₂ recycling, the process demonstrates a cutting-edge solution that addresses carbon management challenges while enhancing energy security. This aligns with Saudi Arabia's commitment to leading global efforts in sustainable development and climate action.

CHAPTER 2

LITERATURE REVIEW

2.1 Hydrogen Production Techniques

The technology for producing hydrogen that is most practical and effective is steam-methane reforming (SMR). It accounts for around 50% of the world's hydrogen generation, however there are certain drawbacks, including coke deposition and pressure drops [14]. However, one of its main benefits, which drew investors' attention, is that SMR has an H_2/CH_4 ratio of 3:1, making it the most effective method for producing hydrogen [6]. SMR is an endothermic reaction (205.8 kJ/mol) that needs a temperature of 800-1000°C to function requiring 14.7 GJ of energy for 1000 m³ of hydrogen [7]. To stop the unwanted generation of CO₂, the necessary ratios of H/C must be adjusted [6]. The chosen syngas composition will determine the steam pressure, which can range from 3 to 25 bar [6]. The process is facilitated by increased H/C ratios since they improve output quality and efficiency. The levelized cost of hydrogen produced using SMR, the most advanced method for producing hydrogen, is roughly \$1,300–1,500 per tonne [7]. Whereas the inclusion of carbon capture options into the process increases the cost up to \$1,854 per tonne [7]. The inclusion of carbon capture reduces the CO₂ emissions to 0.98 kg CO₂/kg H₂ at a carbon avoidance cost of 75 USD/tonne CO₂. In the USA, the cost of producing hydrogen utilizing SMR with CO₂ capture equipment is \$2.27 per kg, compared to \$2.08 per kg for production without CO₂ capturing technology. In the SMR-only example, the energy cost of hydrogen production is 1.99 \$/GJ [8].

To get higher returns and high purity hydrogen, the gasifier is introduced in tandem with SMR. Because it can run at higher temperatures and handle a variety of feedstocks, the

entrained flow gasifier is well suited to be combined with SMR. The H/C ratio and hydrogen production both significantly improve as a result of entrained flow gasifier integration with SMR.

In the case of integrated SMR technology, the yield is greater than 70%, and the price of producing hydrogen as a result of integrated SMR is \$2.01 per kg. In this instance, the cost of producing hydrogen is less than the cost in SMR alone. The energy required for hydrogen generation is reduced when compared to SMR technology alone since integrated SMR technology has an efficient heat recovery system [23]. In this scenario, the cost of energy used to produce hydrogen is 1.44\$/GJ [9].

Carbon dioxide greenhouse gas, thereby addressing environmental concerns associated with their emissions [10], [11]. However, the endothermic nature of the reaction and propensity for carbon formation pose significant challenges in dry reforming processes. Energy demand for the DMR process lies on the higher side among the reforming processes, requiring 16.3 kWh/kg H₂. Whereas it produces syngas with lowest H₂/CO of 1 which is not suitable for hydrogen production but can be utilized for liquid fuels production through Fischer Tropsch synthesis and power generation [12], [13].

Minh et al. studied the thermodynamics of the SMR and DMR processes and concluded that the later should be operated at 850°C where the production of solid carbon is minimum and yield of H₂ is on the higher side [12]. Whereas the selection of CO₂/CH₄ is a tradeoff between higher conversion, selectivity and energy consumption.

Whereas the probability for carbon formation poses significant challenges in dry reforming processes. To do this, it is possible to stop the catalyst from degrading due to carbonaceous

deposition and sintering by using suitable supports or promoters with a mesoporous structure and basic or redox characteristics [12].

Partial oxidation represents another viable pathway for syngas production, entailing the reaction of methane with a limited amount of oxygen or air. Operating at lower temperatures compared to steam reforming, partial oxidation is well-suited for small-scale applications and distributed energy systems. POX processes work at high temperatures and high pressures using non-stoichiometric fuel-to-air combinations [14]. A common H_2/CO ratio is between 1.7 and 1.8, and high temperatures are necessary to enable complete conversion and prevent soot production [15]. The usual temperature range for large-scale reactors is 1150–1500°C, with pressures ranging from 25–80 bar. Compared to a steam reformer, the POX reactor is smaller, but because of its higher operating temperatures, it is less efficient, requiring more fuel to be consumed and heat recycling.

Catalyst selection and control of reaction conditions play pivotal roles in achieving desired product distribution and minimizing carbon formation, thereby optimizing the efficiency and sustainability of partial oxidation processes.

2.2 Chemical Looping

Chemical looping is a novel idea that makes it possible to run the process with contaminants while still making it simple to separate and purify the finished product [16], [17]. It relies on metal oxides' natural reactivity to help separate hydrogen from water without letting the gas come into direct contact with air [18].

This innovative approach saves money and the environment because it doesn't require the removal of CO_2 upstream nor downstream [19].

2.3 Potential Feedstocks for Syngas Production Via Gasification

Pretreatment of biomass and residual wastes may be necessary to enable thermochemical processes such as pyrolysis and gasification. While most treatment methods involve some kind of size reduction, dehydration, or removal of undesired elements, raw wastes can be handled without any pre-treatment at all using technologies like melting and plasma gasification. Mechanical or mechanical and biological combination pretreatments are also possible. Heavy metals, non-ferrous metals, and ferrous metals are often divided into three different stages during the pre-treatment step. The secondary fuel made of the recovered metals is a byproduct of recycling MSW. However, extra steps can be necessary depending on the type of gasifier. It will be necessary to perform further grinding and shredding for entrained flow gasifiers, for instance. Another solution for mechanical treatment is the addition of biological treatments like bio-stabilization and composting. The final application determines whether the biodegradation is aerobic or anaerobic. According to several study teams, stabilization pre-treatment can increase syngas production while decreasing coke and char formation. To change the inorganic and organic composition of the feedstock, chemical treatment can be utilized in solid waste pre-treatment [2].

Biomass or waste may have significantly varying moisture contents depending on the material. The spectrum is broad, ranging from about 15% in wheat straw to over 90% in algal biomass. When employing biomass or waste in thermochemical conversion processes, pre-drying the feedstock prior to feeding it to the reactor is one technique to get past the moisture content's deleterious impact on the heating value and efficiency. Furthermore, because of its increased deterioration tendency, which reduces the energy and cost balancing by producing energy loss during storage, high moisture content poses

logistical challenges. The effectiveness of the process is directly correlated with the moisture content. Pre-conversion dewatering and steam are frequently utilized to obtain high conversion efficiency. There are numerous methods for lowering the feedstock's moisture content. However, the type of reactor used, the operating circumstances, and the makeup of the feedstock all affect the process' overall efficiency [2]. According to Table 1, this won't be a problem in our instance because the moisture content of PP and PE is quite low, at 0% in PP and 0.02% in PE [20].

When gasification is applied, organic resources like coal, petroleum coke, asphalt, biomass, and municipal and industrial wastes can all be converted into syngas [2].

Gasification is a technique used to create energy, fuels, and chemicals from a range of feedstocks. Differentiating characteristics of feedstocks include their energy content, volume, structure, chemical makeup, bulk density, ash concentration, and moisture content, to name just a few. Below are a few of the most prevalent sources of inputs [21]:

- Waste Wood
- Industrial Wastes
- Municipal Solid Waste (MSW)
- Agricultural Wastes
- Coal
- Petroleum Coke
- Feed Treatment

- Aluminum Wastes
- Plastics
- Sewage Sludge
- Crop Residues
- Mill Waste
- Grasses
- Auto-Shredder Residue (ASR)
- Black Liquor
- Corn Stover
- Wood Pellets and Chips
- Refuse-Derived Fuel (RDF)

Palletization was typically a successful first step in improving fuel homogeneity and boosting feeding device reliability. In using biofuel or plastic granules in a catalyst bed, good outcomes in terms of concentration of hydrogen ions (up to 32% vol.) were obtained despite the rise in tar production and the rate of fine/carbon elutriation as compared to wood pellets [2][13].

The chemical makeup of the feedstock is an important consideration in thermochemical conversion processes. Element behavior during thermochemical transitions and how elements are distributed among the gas, liquid, and solid phases once the conversion

process is complete are influenced by a number of factors. The temperature, pressure, heating rate, equivalent ratio, oxidizing agent (air, oxygen, steam), fuel moisture, particle size, fuel composition, fuel residence time, and other factors are a few examples of these. The most important phase of an element's action is the release of volatiles. The devolatilization conditions have a significant impact on the relative concentrations of each element in the gas, solid, and liquid phases [2].

Table 1 Polyethylene and polypropylene proximate analysis (wt%) [50]

	PE	PP
Moisture	0.02	0
Ash	0.15	0.7
Volatile Matter	99.83	99.3
Total	100	100
LHV (MJ/kg)	38.04	44.7

2.4 Lessons from Existing Waste Plastic to Electricity Plants

The only plastic gasification facility in Thailand that uses waste plastic to make syngas is located in Samutsakorn, run by Thermal Tech Ltd. The down-craft type reactor is utilized for gasification, and the resulting syngas is used to produce electricity [22].

Technical and non-technical barriers can be distinguished from one another. This is the reason why this technology does not qualify as proven technology and is currently experiencing a number of technoeconomic problems on a commercial basis [20].

One of the main problems with this gasification method is the moisture content, which could raise the gasification's running costs. Moisture content alters the makeup of gas and lowers its heating value. Unfortunately, many businesses that had begun operating this plant had to stop since they were unable to meet the necessary gas composition and heating value. A technical hurdle in the gasification of waste plastic is the high tar concentration. Scrubbing as a unit has failed in terms of reliability and sustainability over the long run. Due to the increased production of condensate, there are environmental risks [22].

The successful commercialization of this technology faces numerous non-technical obstacles. Government regulations, a heavy reliance on fossil fuels, a dearth of viable technologies that have been proven effective, energy policies, and the cost of energy production are the main non-technical obstacles. The commissioning of this plant has been attempted by numerous enterprises in an effort to benefit from higher adder rates on power sales. Investors that are only interested in obtaining bigger profits in tariff increases for the purchase and sale of energy have been drawn to the waste plastic-fueled facilities. The management of the plant involves several technical difficulties, and the majority of the nations who commission it import it from a less developed nation. The inability to manage the plant technically constitutes one of the main obstacles as a result. Because there is a dearth of competent labor that can run the plant successfully, many plants are compelled to shut down and wind up their operations. The advantages of selecting waste plastic over biomass are;

Solid waste has more advantages over biomass in terms of government pricing.

Most industry owners pick landfills for the disposal of their waste plastic since they are far less expensive than gasification technology.

Using biomass is not cost-effective, and getting enough biomass for large-scale operations might be challenging. |

CHAPTER 3

AVAILABLE TECHNOLOGIES FOR WASTE TO ENERGY

3.1 Waste Plastic Conversion Methods

3.1.1 Pyrolysis

The thermochemical process of pyrolysis converts biomass into liquid fuels like bio-oil, char, and producer gas while operating in an oxygen-deprived environment. The temperature, heating gradient, retention time, and reactor type all have a significant impact on how the pyrolysis product is distributed. The physical and chemical properties of the feedstock, such as moisture content, elemental composition, and particle size, have an impact on the yields of bio-oil, char, and gases [23]. Gases like hydrogen, carbon dioxide, carbon monoxide, and methane typically cannot condense and escape as producing gas. Due to increased cracking reactions at high temperatures, char production is suppressed. Pyrolysis side reactions are favored by the longer vapor holding period in the reactor, which causes slower rates of condensation of volatile components. Secondary reactions provide chemicals with large molecular weights, such as tars and char. Tar and char are the end products of secondary polymerization processes. Unlike oils derived from biomass, oil produced from the thermal cracking of plastics frequently contains heavy crude, naphtha, mid-distillates, light oil, and gases. Before pyrolysis oil can be utilized as fuel for automobiles and power plants or as a feedstock to produce fine chemicals with additional value, it must be improved. In addition to having a high calorific value, the pyrolysis oil made from plastics also has other fuel properties that are comparable to those of commercial fossil fuels [24].

3.1.2 Gasification

Syngas is a product of the gasification process, which involves organic materials reacting with a gasifier agent at a high temperature. The components of syngas gas are CO and H₂, which are created as a result of breakdown of larger organic molecules via heating along with water gas-shift process which will be elaborated later on [9]. It is a type of process in which wasted plastic or other organic materials are transformed into H₂ and CO which can then be utilized as fuel or to create valuable products [10]. Given that this technique produces relatively little greenhouse gas emissions, it is a very environmentally friendly method [11]. The decision to choose gasification over other technologies was made because of its advantages in the economy and environment [23]. In Table 2, the syngas manufacturing techniques for plastic gasification are contrasted. In comparison to pyrolysis, the gasification route also offers the highest revenue per ton of waste by three-folds [23].

3.1.3 Liquefaction

Another thermochemical biomass-to-liquid process is liquefaction, which converts waste organic material into energy-dense bio-oil under high pressure and typically with the help of catalysts. Pressures between 1 and 25 MPa and temperatures between 250 and 450 °C are typically needed for liquefaction [25]. Compared to the chemically reactive bio-oil obtained from the pyrolysis of lignocellulose and other biogenic wastes, the oil product obtained from the liquefaction of plastics has a lower water content. The oils made from the liquefaction of plastic goods also have a low oxygen concentration, which makes them valuable as a fuel and heat source. Catalytic fuel upgrading procedures including hydro

processing, hydrodeoxygenation, and hydrotreating typically have much lower severity since plastic-derived oils have low moisture content.

3.1.4 Hydrothermal Gasification

By degrading biowaste in hot, compressed water, hydrothermal gasification operates as a thermochemical conversion pathway that generates syngas under sub - critical or supercritical conditions [9]. The method is advantageous for biomass with a substantial moisture content since it avoids the drying and dehydration phases, which lowers operational costs. Several gas-to-liquid processes, such as Fischer-Tropsch syntheses or syngas fermentation with microorganisms, can further transform the generated syngas into liquid fuels [23].

3.1.5 Fermentation

One of the best biochemical conversion methods for producing biofuels is syngas fermentation. Using this method, microorganisms are grown on syngas, which is created thermochemically from biomasses. Carbon-containing lignocellulosic biomass is frequently converted by carbon-fixing bacteria into gases including carbon dioxide, carbon monoxide, and methane, which are subsequently processed further to create biofuels. Forest leftovers, coconut shells, empty palm oil fruit bunches, municipal solid waste, etc. are a few examples of these resources [26].

3.1.6 Anaerobic Digestion

By breaking down organic material without the use of oxygen, a biological process known as anaerobic digestion (AD) produces biogas, which is mostly composed of carbon dioxide and methane. There could be three common steps in anaerobic digestion processes. First, complex biological species break down waste to produce simple soluble molecules like

fatty acids, monosaccharides, and amino acids. This process is carried out by bacteria. The production of materials such as hydrogen, volatile fatty acids, and carbon dioxide is the second phase. The third phase of anaerobic digestion is methanogenesis, which turns organic acid into methane.

Anaerobic digestion produces biogas, which can be used to create chemicals, fuel cells made of solid oxide, gas turbines, and gas motors. Cryogenic methods can be used to improve the biogas. Microorganisms like bacteria and enzymes are utilized in biochemical waste processing methods to break down biomass. The conversion of biochemical waste is an excellent and sustainable way to generate energy from trash. Anaerobic treatment and ferment are the two most often used biochemical processes that use microorganisms to transform waste into energy [27].

Table 2 Comparison of waste plastic gasification technologies [23]

Technology	Products	Reaction Temperature (°C)	Typical Processing Time	Total Capital Expenditure (\$/kW)
Liquefaction	Tar and bio-oil	200-400	0.5-2 h	300
Pyrolysis	Bio-oil, syngas and biochar	300-600	1-5sec for flash pyrolysis 10-100 min for slow pyrolysis	800
Thermochemical Gasification	Biochar, syngas and tar	700-1000	30-60 min	3600-4500

Hydrothermal Gasification	Biochar, syngas and tar	350-650	Batch reactors: 15-60 min Continuous reactors: 30-90 sec	9900
Anaerobic Digestion	Digestate and biogas	37-55	14-60 days	1400-2200
Fermentation (EtOH, BuOH, and syngas)	Butyric acid, acetic acid, acetone, ethanol, and butanol	32-38	EtOH fermentation: 6-48 h BuOH fermentation: 6-72 h Syngas fermentation: <15 days	1400-2000

3.2 Selected Process

The most techno-economical and mature way for recycling waste plastic is thermochemical recycling. Several significant fuels and chemicals are produced using this syngas as a raw source. Nowadays, the gasification of waste plastics has gained popularity as a means of recycling waste plastic [28], [29]. Gasification holds great promise for addressing energy-related concerns in the future as it can be applied to various processes such as producing chemicals through syngas synthesis, producing biofuels like bioethanol, producing hydrogen following RWGS, and producing electricity [30]. Furthermore, compared to combustion, gasification is a more developed technique that produces fewer emissions into the environment [29]. Gasification is therefore chosen as the MSW conversion method in this study due to its broad range of applications and environmental advantages; all of these options will be further discussed in this paper.

3.3 Choice of Gasifier Agent

The most widely utilized gasifying agents are air, steam, a combination of the two, oxygen, or CO₂. The type of gasifier agent utilized determines how the gasification process is categorized. Because the composition of the syngas generated during gasification reactions is determined by the gasifier compounds chosen, this choice is very important [31]. The use of air as a gasification agent appears to be quite advantageous on paper, however there is a significant amount of nitrogen present, which can lower the heating value of the resulting syngas [32]. Oxygen as a gasification agent is employed to address this issue, however the expense of separating pure oxygen from air is very high. This could result in greater syngas production operating costs and affect the process's commercial viability. Moreover, applying an oxygen gasification agent could cause the oxidation reactions to produce CO₂ and water [32].

Due to its lower cost compared to oxygen or air gasification agents, steam gasification has a distinct advantage over the other gasifier agents. Moreover, the use of steam greatly promotes the steam-char and steam-methane reformation agents, which results in the formation of hydrogen [1]. The use of steam with the proper H₂/CO ratio results in a larger hydrogen content as well. For this reason, high temperature steam has been employed as the gasifier agent for the process in this paper [31].

In order to create syngas, researchers have examined the use of steam with air. It is utilized to circumvent the issue of nitrogen dilution of syngas, which occurs frequently when solely air is used as a gasification agent [33]. This is because many endothermic reactions will be influenced significantly if only oxygen is added. Due to the compressing and cooling cycles necessary for the isolation of oxygen from nitrogen, oxygen-aided gasification is typically

expensive. However, by combining steam and air, the gasification process becomes more efficient and cost-effective without having to deal with the issue of nitrogen dilution of syngas. Nevertheless, steam and air co-gasification yields less amount of hydrogen compared to steam gasification [33].

The ability to achieve a variety of H_2/CO ratios is the main advantage of employing CO_2 as a gasification agent. As a result, as CO_2 concentration rises, syngas's lower heating value (LHV) can also increase accordingly [33]. However, when utilized as a gasification agent at elevated temperature (800–1000°C), CO_2 has a significant impact on the endothermic reactions [33].

Table 3 Different types of gasification agents and their influence on endothermic reactions [33]

Parameter	Order of Effectiveness
%Hydrogen (% H_2)	Steam > H_2O_2 > O_2 > Air > CO_2
%Carbon monoxide (%CO)	CO_2 > O_2 > H_2O_2 > Steam > Air
%Carbon dioxide (% CO_2)	O_2 > H_2O_2 > Steam > Air > CO_2 (MER = 0.22-0.26) O_2 > H_2O_2 > Steam > CO_2 > Air (MER = 0.28-0.36)
%Methane (% CH_4)	CO_2 > Steam > H_2O_2 > Oxygen > Air
LHV (MJ/Nm ³)	CO_2 > Steam > H_2O_2 > Oxygen > Air
H_2/CO molar ratio	Steam > H_2O_2 > Air > O_2 > CO_2
Hydrogen yield (g/kg biomass)	Steam > H_2O_2 > Air > O_2 > CO_2
CO yield (g/kg biomass)	CO_2 > Air > O_2 > H_2O_2 > Steam
Gas production rate (Nm ³ /hr)	Air > Steam > H_2O_2 > CO_2 > O_2

3.4 Process Reactions

Gasification as a thermochemical technique that has a limit on how much polyethylene (PE) and polypropylene (PP) waste plastic can be gasified [8]. The process is initially endothermic in nature and results in water vapor, which is present in moisture form [8]. The initial stage of gasification, known as pyrolysis, produces pyrolytic char and combustible materials. The volatile components could be steam, pyrolytic tar, pyrocarbons, CO, or CO₂. The reaction temperature rises to 800-1000°C as a result of the pyrolytic tar's reaction with oxygen in the interim. The leftover material is gasified using this energy as activation energy. As illustrated in table 4 below, the reactions affect the gasification's yield.

Table 4 Comparison of various thermal treatment processes for solid fuels

Process	Temperature (°C)	Residence Time	Particle Size	Product Yield		
				%Liquid	%Solid	%Gaseous
Torrefaction	220-320	Short	-	0-5	80	20
Slow Pyrolysis	<450	Long	Large	30	35	35
Fast Pyrolysis	450-650	Short	Small	75	12	13
Gasification	>750	-	-	5	10	85

Torrefaction is a process that uses thermochemical reactions to lower the amount of water and volatiles in biomass or other organic sources. The resulting composition in this process is 5% liquid, 80% solid, and 20% gas, with an operating temperature range of 220-320°C [33]. Torrefaction has a shorter residence period, which accounts for the larger solid

product content. It has a considerable potential to enhance the fuel's qualities, such as by improving energy density, enhancing hydrophobic behavior, making it easier to grind, and having a homogeneous composition [33].

The three types of pyrolysis are slow, flash, and intermediate pyrolysis. These classifications are determined by the temperature increase and duration of vapor retention. The characteristics that distinguish rapid and flash pyrolysis include high temperatures, rapid heating rates, and brief vapor residence durations. These characteristics result in the production of more bio-oil than char and persistent fumes. On the other hand, moderate temperatures, modest heating rates, and extended vapor residence times are characteristics of slow pyrolysis. These characteristics raise char yields as opposed to bio-oil yields. Fast pyrolysis is commonly defined as a temperature range of 400–500°C, a greater heating rate of 10–200°C/s, and a vapor retention period of 30–1500s [33].

Table 5 Pyrolysis method comparison in terms of operating characteristics

	Slow pyrolysis	Flash pyrolysis	Intermediate pyrolysis
Temperature range (°C)	300-700	800-1000	500-600
Heating rate (°C/s)	0.1-1	1000+	2-10
Vapor residence time (min)	10-100	0.5	0.15-0.33

below depicts the processes that take place throughout the gasification process.

Table 6 Thermochemical reaction involved in the gasification process with corresponding standard enthalpies at 298K

Reaction	ΔH°_R (kJ/mol)	Reaction #
$C_a H_b O_c \leftrightarrow \left(a - \frac{b}{4} - \frac{c}{2}\right) C + \frac{b}{4} CH_4 + \frac{c}{2} CO_2$	218.9	1
$C + CO_2 \leftrightarrow 2CO$	-172.5	2
$C + H_2O \leftrightarrow CO + H_2$	131.3	3
$C_n H_m + nH_2O \leftrightarrow nCO + \left(n + \frac{m}{2}\right) H_2$	>0	4
$C_n H_m + 2nH_2O \leftrightarrow nCO_2 + \left(2n + \frac{m}{2}\right) H_2$	>0	5
$CH_4 + H_2O \leftrightarrow CO + 3H_2$	205.8	6
$CH_4 + 2H_2O \leftrightarrow CO_2 + 4H_2$	164.6	7
$CO + H_2O \leftrightarrow CO_2 + H_2$	-50.8	8
$C_n H_m O_z \leftrightarrow zCO + (n - z)C + \frac{m}{2} H_2$	<0	9
$CH_4 + CO_2 \leftrightarrow 2CO + 2H_2$	247.2	10
$CH_4 \leftrightarrow C + 2H_2$	75	11

3.5 Types of Gasifiers

Depending on the compatibility of the feedstock and desired result, traditional gasifiers come in many different forms.

3.5.1 Traditional Gasifier

The arrangement around the reactor with the feedstock funnel is connected to the fundamental architecture of the conventional gasifier. With each design of the conventional gasifier, the heating mechanism may differ, as well as where the gasifier agent injection and syngas removal nozzles are placed [22]. Figure 4 below depicts the classic gasifiers that are frequently used;

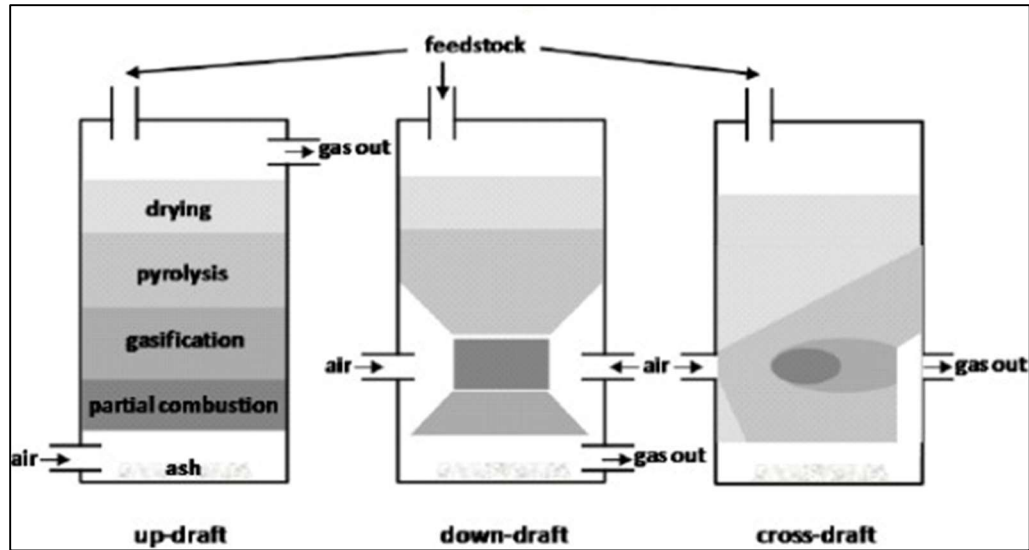


Figure 4 Traditional gasifiers types [20]

In order to pyrolyze, gasify, and dry the feed entering, the stack zones of an up-draft gasifier are precisely defined. As the air/gasifying agent enters the reactor chamber from the bottom, it serves as the gasifier agent inside this system. In the conversion zone, there is a counter-current action with the trash made of plastic and biomass. The gases generated inside the reactors are directed upward, which makes it easier to remove it from the top. The performance and efficiency of the gasification reaction in the up-draft gasifier was enhanced by the upward direction of the air/gasifying agent and gas. The hot gases are subsequently put to use for other purposes, such as heating, drying, and the production of hydrogen [22].

Another kind of updraft gasifier is the fluidized bed gasifier. These gasifiers successfully produce a fluid-like flow of the feedstock and gas inside the chamber by suspending the feedstock in an oxygen-rich environment. The suspension speeds up the process of heat transmission from the gas to the feedstock, allowing the ash to leave the suspensions rather

than ascending with the syngas. Feedstocks that have the potential to produce corrosive ash can be gasified in fluidized bed systems without endangering the chamber [54].

3.5.2 Downdraft Gasifier

In a downdraft gasifier, air is injected from the top or center, acting as the gasifier's agent. The syngas is taken out of the downdraft gasifier's bottom. When the gas flows downward, the temperature of the gas rises due to the energy or heat generation entering the gasifier from the top. The heat from the gas leaving the gasifier is used to heat the upper part of the reactor because of its extremely high temperature. The gas must pass through ash in order to reduce the amount of tar created in the syngas [22].

3.5.3 Cross-Draft Gasifier

Cross-draft gasifiers receive their feed and gasification agents from the opposite ends of the chamber. The feedstock enters the chamber from the top, while the oxidant enters from the bottom. Heat and mass are transferred counter-currently inside the chamber [20]. Due to the generation of extreme velocity and temperature syngas, with an ineffective CO₂ reduction approach when comparing to the other gasifiers, as such, this type of gasification is not typically employed for gasification. This gasifier's feedstock compatibility is likewise quite constrained; only coal, oil, and wood are acceptable [20].

3.5.4 Entrained Flow Gasifier

Due to its compatibility and viability at commercial scale, entrained flow gasifiers remain fairly common. It is a particular kind of downdraft gasifier where the feedstock and gasifier agent are fed at greater temperatures from the gasifier's top. While creating syngas, this mixture descends, and the entrained flow gasifier's efficiency increases. This is why it is applicable for the gasification of waste plastic, methane, and coal. For that, entrained flow

gasifier was opted in this work. Co-current interface between the oxidant and feedstock is the main flow regime in the process. Due to the reaction's significant pace, the residence period is typically quite short. To enhance mass transfer in between feed and oxidant in this type of gasifier, the feed size must be less than 200 microns. The operating temperature needs to be maintained higher because of the reaction's short residence duration. The ideal operating conditions for this kind of gasifier are between 1200° and 1500°C and between 20 and 80 bars of pressure [8]. The working temperature of the gasifier must be kept above the melting point of the ash. In figure 5 below, the entrained flow gasifier is depicted;

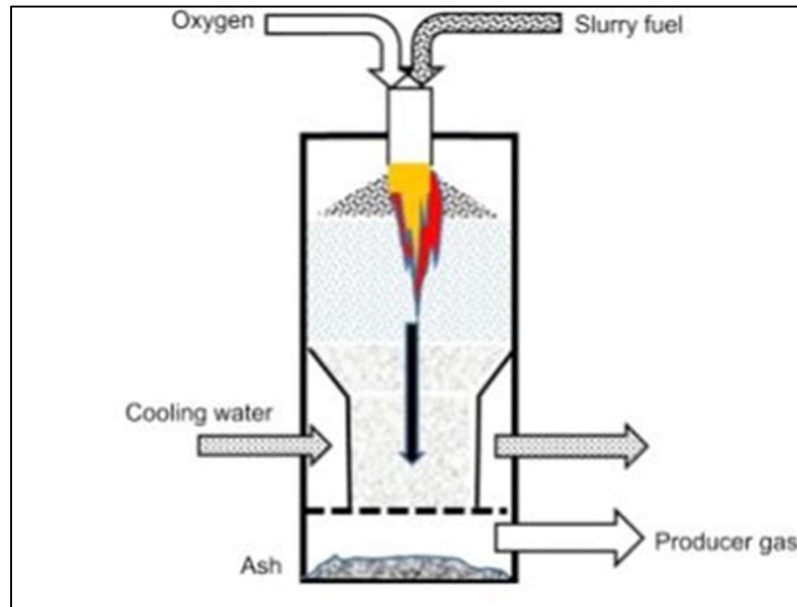


Figure 5 Entrained flow gasifier [20]

3.5.5 Plasma Gasifier

Plasma gasifiers are commonly used to dispose of hazardous material at high pressures and temperatures. The temperature in the plasma gasifier typically runs at about 3000°C and has very high operating expenses [60]. Figure 6 below depicts the plasma gasifier;

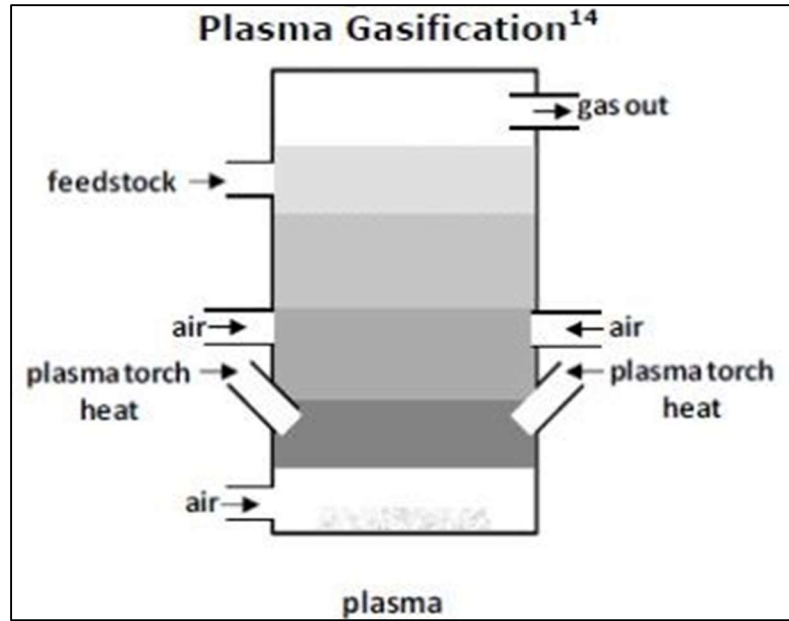


Figure 6 Plasma gasifier [20]

Comparisons of several types of gasifier reactors are made and contrasted below:

Table 7 Comparison of gasification options [34]

Gasifier Type	Tar Content	Cold Gas Efficiency	Energy Requirement	Permissible Particle Size	Maximum Moisture Content	Dust Content
Entrained Flow	Very low	> 80%	Low	< 1/25 in	~ 10%	High
Updraft	Very high	> 80%	Low	< 2 in	~ 50%	Low
Downdraft	Low	> 80%	Low	< 4 in	~ 40%	Medium
Fluidized Bed	High	> 90%	Moderate	< 1/4 in	~ 10%	High
Plasma	Very low	> 90%	High	N/A	> 50%	Low

Table 8 Composition of syngas for using different types of gasifiers and gasifier agents [15]

Gasifier Type	Gasifying Agent	Syngas Composition (%mol)				
		H ₂	CO	CO ₂	CH ₄	N ₂
Circulating Fluidized Bed	Air	10-12	16-19	14-18	6-8	48-52
	Steam	35-55	15-30	15-25	8-12	0
	O ₂	24-28	20-22	40-44	-	-
Updraft Fixed Bed	Air	19	23	12	5	41
Downdraft Fixed Bed	Air	15-18	15-21	13-15	1-2	44-56
Entrained Flow Gasifier	O ₂	28-33	20-25	46	2	-

3.6 Why Syngas is Important?

The International Energy Agency (IEA) has published multiple articles arguing that the daily increase in greenhouse gas emissions (GHG) is caused by increased energy use [14]. Energy consumption inevitably rises as a result of population development, and the main energy source is fossil fuels. Fossil fuels have the disadvantage that their supplies are finite and that when they are burned, they release carbon dioxide, a significant greenhouse gas that causes climate changes. According to the IPCC 2014 report, fossil fuels are accountable for 35% of the world GHG emissions, while carbon dioxide comprised 75% of those emissions. In 2010, coal accounted for 40% of global energy production, or electricity. By 2035, compared to 2009, the amount of coal used to generate electricity is expected to increase by 33%. The demand on coal and fossil fuels can be significantly decreased by heating used plastic to higher temperatures for the transformation into useable

energy [6]. Furthermore, the calorific value of hydrogen per unit mass, that is in the syngas, is higher by three folds compared to natural gas, hence the importance of hydrogen [58].

3.7 Comparison of Syngas Technologies from Natural Gas

Syngas can be derived from natural gas using various technologies as shown below table:

Table 9 Comparison of syngas technologies from natural gas [55].

Technology	Pros	Cons	Licensors
Partial Oxidation (POX)	Desulfurization of feedstock is not necessary.	Often demands for an oxygen plant. Process operating temperature that is quite high.	Texaco Inc. Royal Dutch/Shell
Steam Methane Reforming (SMR)	Lowest process operating temperature, no need for oxygen. The greatest depth of industrial experience. For the manufacture of liquid fuels, the best H ₂ /CO ratio.	Maximum air emissions. More expensive than ATR and POX. Uses membranes to recycle CO and surplus hydrogen is removed.	Foster Wheeler Corp Kinetics Technology Uhde GmbH Lurgi AG International BV Haldor Topsoe AS

Autothermal Reforming (ATR)	By changing the reformer outlet temperature, the amount of methane in syngas may be customized. Process temperature requirement is lower than for POX.	Often demands for an oxygen plant. little Industrial experiences.	Haldor Topsoe Lurgi
Dry Methane Reforming (DMR)	About 100% of CO₂ can be consumed rather than releasing it into the atmosphere.	Since the reaction occurs at 873 K, additional heat is needed. Coke formation on a catalyst.	Carbon Sciences
Combined Dry Steam Reforming (CSDR)	Deposition of coke can be greatly decreased. Optimum H₂/CO ratio for producing liquid fuels.	High cost of project installation. Requires a separation unit for the unreacted methane from SMR reaction.	Midrex Process
Tri-Methane Reforming (TMR)	Utilizes flue gases directly rather than first separating and purifying the CO₂ from the flue gases.	Low H₂/CO ratio ratios prevent its widespread use in the synthesis of F-T and MeOH.	Topsoe AS Haldor

	It is possible to achieve an 80% CO ₂ and 95% methane conversions.	Often demands an oxygen plant.	
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3.8 Syngas Utilization

As demonstrated below, syngas can serve as a precursor for the synthesis of numerous beneficial products:

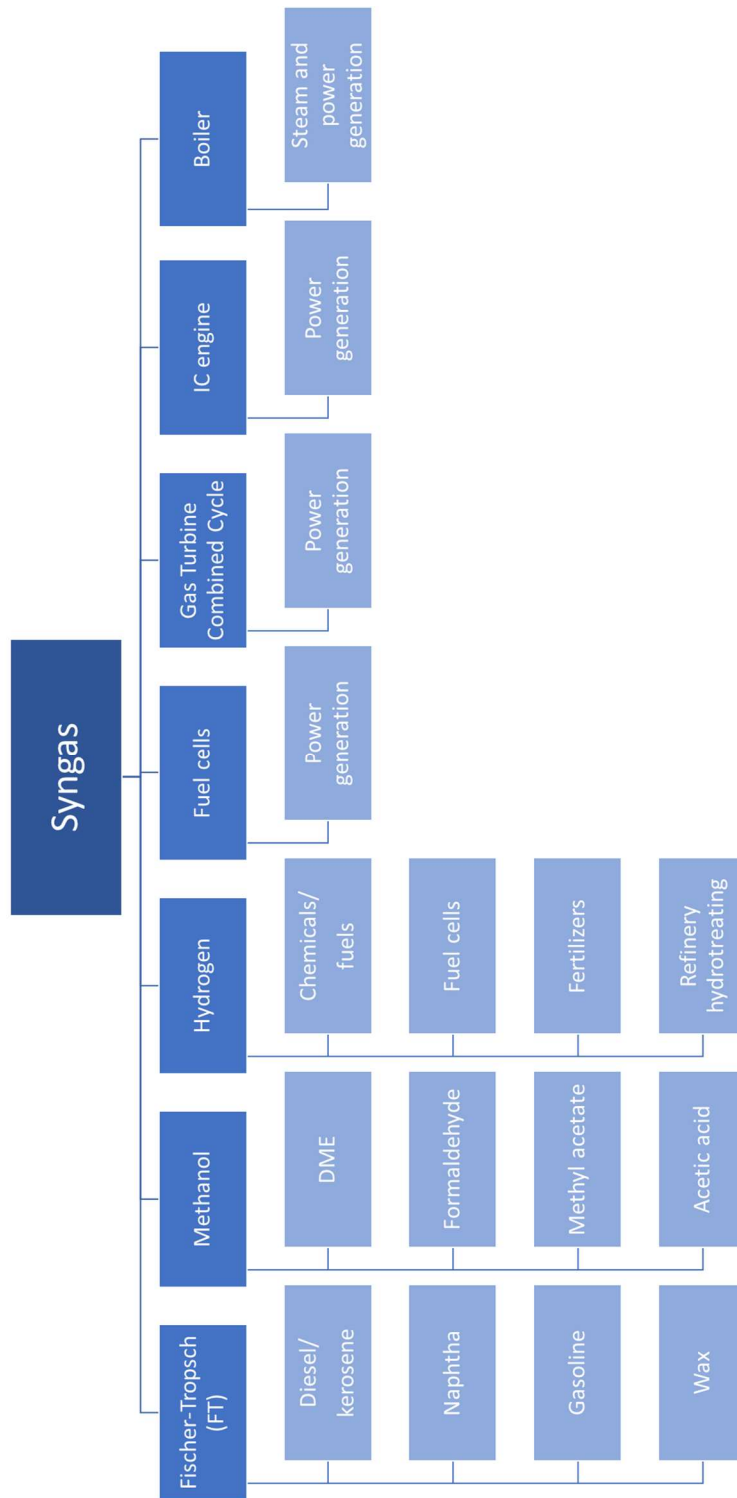


Figure 7 Gasification output pathways [34]

3.8.1 Power Generation from Syngas:

In order to recover the thermal energy of the system as heat, steam, or electricity, the syngas is directly burned. This heat can be utilized for heating or cooling applications during the gasification process. The steam produced as a result of recovered heat is used for the production of electricity. The facilities that usually support this idea of electricity generation are known as “combined heat and power plants (CHP)” [21].

3.8.2 Fuel Generation:

The resulting syngas can be utilized in Fischer-Tropsch synthesis along with other chemical and refining processes to create liquid fuel products. It is possible to refine this liquid fuel further to create commercial fuels like kerosene, diesel, and crude oil [21], [36].

3.8.3 Chemical Generation:

A variety of significant compounds, including hydrogen and methanol, can be produced from syngas gas. A number of significant industrial compounds, including ethanol, acetic acid, formaldehyde, and methyl acetate, are produced using methanol as a precursor [21].

3.9 Saudi Arabia: Towards a Cleaner Energy

Saudi Arabia, a fossil fuel-rich nation, uses fossil fuels to produce the heat and power needed to meet national needs. The sector's greenhouse gas emissions are increased due to the widespread usage of fossil fuels for power generation. According to the statistics, power and heat sector in KSA is the most polluting sector with over 270 million ton of CO_{2e} emissions in 2020 as shown in Figure 8. Due to its high GHG emissions and dominance in the energy sector, power generation process is selected in this study to be fueled using the waste plastic feed to reduce the environmental emission of the sector. This can be achieved

by converting the waste plastic into syngas which is further used as a fuel to generate power.

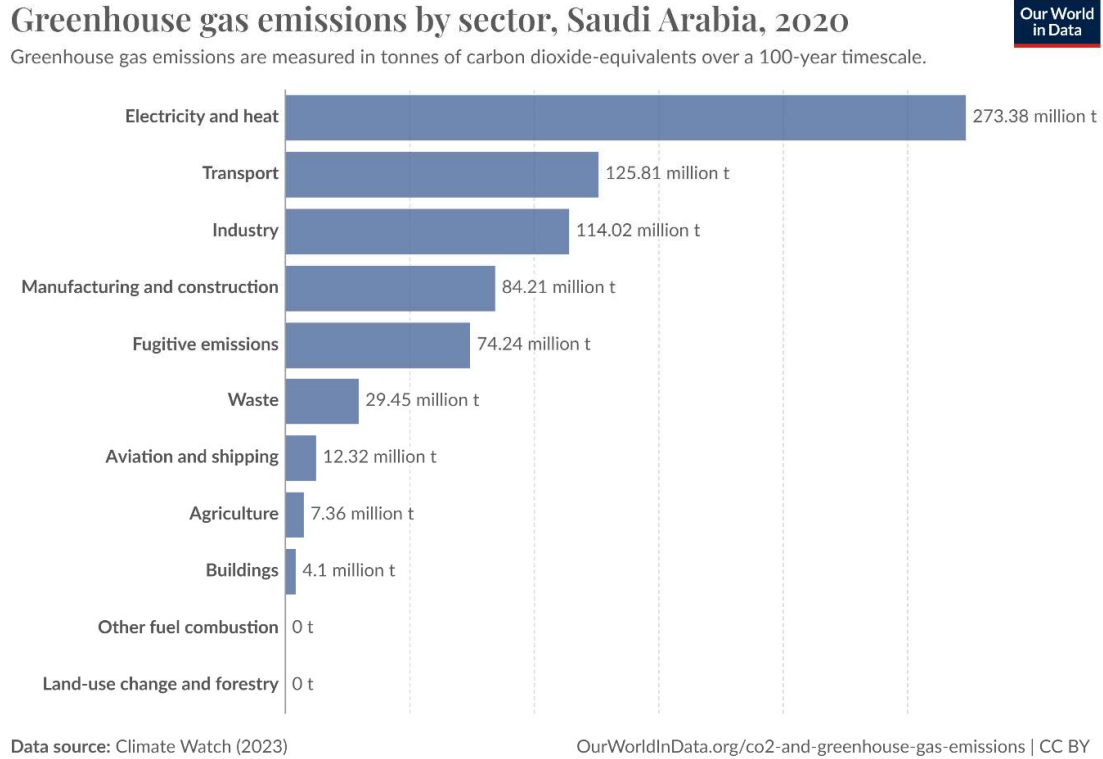


Figure 8 Greenhouse gas emissions by sector for Saudia Arabia in 2020 [37]

3.10 Power Generation from Syngas

The power generation process comprises of several stages including:

3.10.1 Oxygen Production

In industry, oxygen is frequently employed for a variety of purposes. There are numerous established and cutting-edge methods that can produce oxygen, which is used primarily as an oxidizing agent. The oxygen generation systems that are most widely used include the vacuum swing adsorption (VPSA), pressure swing adsorption technology (PSA), cryogenic air separation unit (ASU), and vacuum pressure swing adsorption (VPSA). Also,

there has been a lot of recent research on the chemical looping combustion (CLC) technique's ability to produce oxygen [6].

A. Cryogenic Air Separation Technique:

It's a mature method for producing high quality nitrogen and oxygen gasses. Since Carl Von Linde created this procedure in 1885, not many changes have been made to the system. The primary concept behind cryogenic technology, which relies on the distillation principle, is the liquefaction of oxygen gas. It is based on the idea that the elements in air can be divided according to their respective boiling points because they all have different melting points. The idea behind cryogenic separation is that by lowering the temperature, oxygen and nitrogen might be separated based on their respective boiling points. The system's oxygen is further cooled down using the nitrogen that was initially chilled and separated. It is estimated that 58.4 kWh/ton of energy requirement is needed to extract the oxygen from the air, according to thermodynamic research conducted by ASU. In order to make this method technologically and economically viable, it is best to lower the energy needed for the oxygen separation [6].

B. Pressure Swing Adsorption:

Based on operating pressure, the PSA is divided into PSA, VSA, and VPSA. The kind of PSA technology being used affects the amount of adsorbent as well. The order in which the adsorbent is used is as follows:

$$PSA < VPSA < VSA \quad (3.1)$$

The volume of adsorbent that is inserted into these systems determines the size of the PSA equipment as well. For operation, PSA systems require compressed air. It is necessary to

boost pressure in order to carry out the adsorption process, whilst ambient pressure is excellent for facilitating the desorption process. The gas that needs to be separated by PSA typically has the following elements: water, nitrogen, oxygen, and hydrogen. In PSA systems, zeolites are utilized as adsorbent materials. At higher working pressures, zeolites have a tendency to distort by forming a dipole. The dipole tends to gather nitrogen, but because of the difference in densities, it also lets oxygen pass. The PSA is pressurized to 1.5 atm operating pressure to improve oxygen purity, and after some separation, the zeolite surface is loaded with nitrogen. Depressurization of PSA causes the zeolite to regenerate, returning its polarity to its original state. The use of PSA yields oxygen purity levels of around 95% [6].

C. Vacuum Swing Adsorption:

Adsorption is also utilized in vacuum swing adsorption to separate oxygen from air. In this instance, the operating pressure fluctuates and a specific kind of zeolite is utilized to carry out selective nitrogen adsorption on the surface, allowing oxygen to pass. VSAs have a small design and require very little training to use. Because the working pressure of VSA is nearer to 1 atm than that of PSA, the turbo blower is utilized in place of the compressor to pressurize the equipment, which results in a decrease in power usage. To lower the gas pressure and replenish the adsorbent surface, a vacuum pump is employed [6].

D. Vacuum Pressure Swing Adsorption:

Prior to being fed into the adsorber vessel in VPSA, the air is compressed and filtered with the help of an air filter. After the zeolite is loaded with nitrogen, it is used as an adsorbent, and the air stream is then sent to the subsequent adsorber. The first adsorber is once again prepared for use in this manner. A vacuum pump is utilized to speed up the desorption

process and regenerate the zeolite surface for the following cycle once the VPSA has been depressurized in order to get rid of the nitrogen from the surface of the material. After being compressed by a blower, the regenerated adsorber is prepared to start producing oxygen once more. When the air stream is supplied to the second adsorber, an oxygen flow buffer is employed to prevent variations. In this system, the oxygen pressure is inadequate, hence a compressor is needed to raise the pressure. Compared to PSA and VSA, this approach generates cheaper oxygen without sacrificing oxygen purity.

3.10.2 Integrating Chemical Looping Combustion into Combined Cycle Power Generation

A combined-cycle power plant (CCPP) employs a gas turbine and a steam turbine to generate energy and electricity. The energy generated when the fuel, in this case syngas, burns is what drives the gas turbine's shaft to produce electricity. The combustion process generates steam, which operates the steam turbine in addition to electricity.

This system makes use of or integrates the Brayton and Rankine cycles to maximize electricity output and minimize heat losses, which increases energy recovery and performance [29]. The "heat recovery steam generation section (HRSG)" is made up of a steam turbine and a gas turbine that are connected to a heat recovery system. Gas turbine and syngas turbine designs are obviously different from one another since a natural gas turbine requires more low-calorific gas in order to operate. The CCPP is regarded as the segment of the integrated gasification combined cycle (IGCC) with higher operating costs. By improving the turbine design, it may be possible to boost the process efficiency. Compressor in a combined cycle unit are associated with the highest operating cost, according to a study done for the techno-economic assessment of the process [6].

The CCPP with chemically looping combustion (CLC) technology may be very successful in reaching the objectives of increased efficiency and carbon dioxide elimination. In the CLC process, an intermediate agent, usually a metal oxide, is present while the fuel is burning. Two fluidized bed reactors cycle this oxidized fuel and intermediate in a reduction reactor, which reduces the metal oxide through an endothermic process. After that, the fuel undergoes an exothermic reaction in the oxidation reactor [30]. In figure 9 below, the chemical looping combustion principle is depicted;

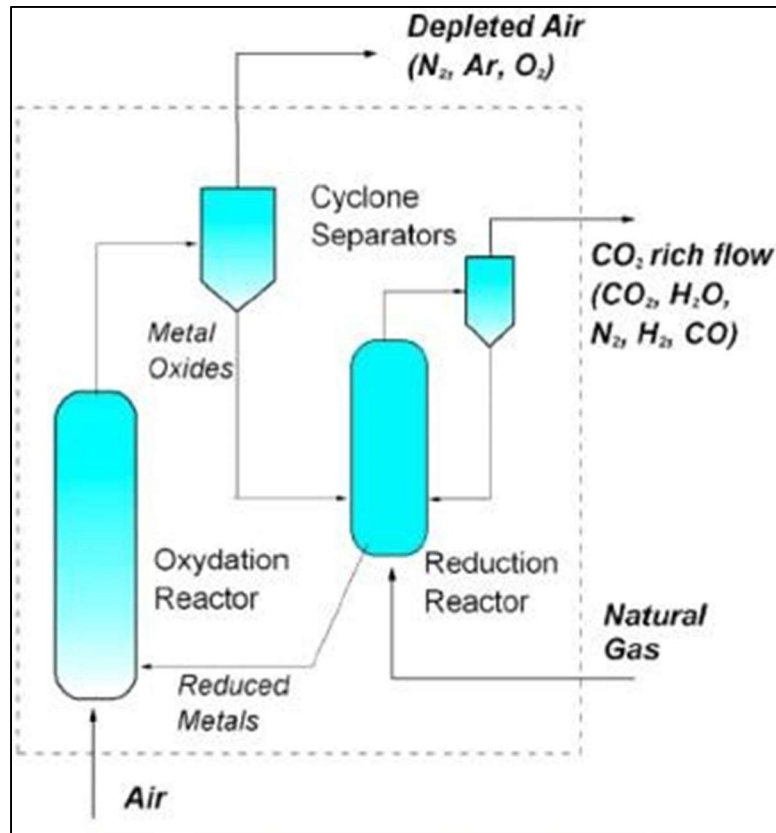


Figure 9 Chemical looping combustion concept [6]

The feedstock is fed into the reactor during the CLC process, and oxygen is added to it to reduce the metal oxide. This oxygen then undergoes oxidation, resulting in the formation of CO₂. Then, with the use of a cyclone, the oxygen-rich metal oxide and leaving products

are separated. Metal oxide is removed and then utilized again. The system has the advantage of having byproducts in flow that don't contain diluents over the typical combustion process.

The oxidation and reduction reactions are pressure-dependent due to the chemical looping mechanism associated with the oxygen carrier. The compressed CO₂ and H₂O's thermal energy is transformed into work by the secondary generator of the CCPP. The syngas was produced in a reduction reactor, and in this case, heating this stream at a higher temperature source—an oxidation reactor—gets the job done most efficiently. While gasifying natural gas, a variety of metal oxides have already been utilized as catalysts; however, when producing syngas, which includes CO₂ and H₂O, these resultant gases are oxidized. *Fe₂O₃/FeO* are the catalysts that will oxidize the synthesis gas in this situation the best. Here is a list of the reaction that was performed [13];



The potential for combined CCPP-CLC plants to produce energy more efficiently and with no emissions is very great. This is because contaminated air is used as the working medium to perform thermodynamic cycles, and it is then separated from the combustion products. Given that the effectiveness of the CCPP depends on it, the heat of the depleted air is significant. The two options that are most likely to occur are fired systems, in which polluted air is added and fed to the gas turbines at a temperature of 1000–1200°C, and unfired systems, in which CH₄ is sent directly to the CLC system. The practicality of these

two choices were studied, and it was shown that unfired systems with a maximum operating temperature of 850–1050°C may have an efficiency of 43–48%. The greatest efficiency that can be achieved in fired systems with a maximum temperature of 850–1200°C is 52%. In both the burned and unfired scenarios, there are no net CO₂ emissions. In comparison to unfired systems, fired systems have higher capital and running costs, but they also save more money on fuel. In the graphic below [13], the CCPP-CLC system diagram is displayed;

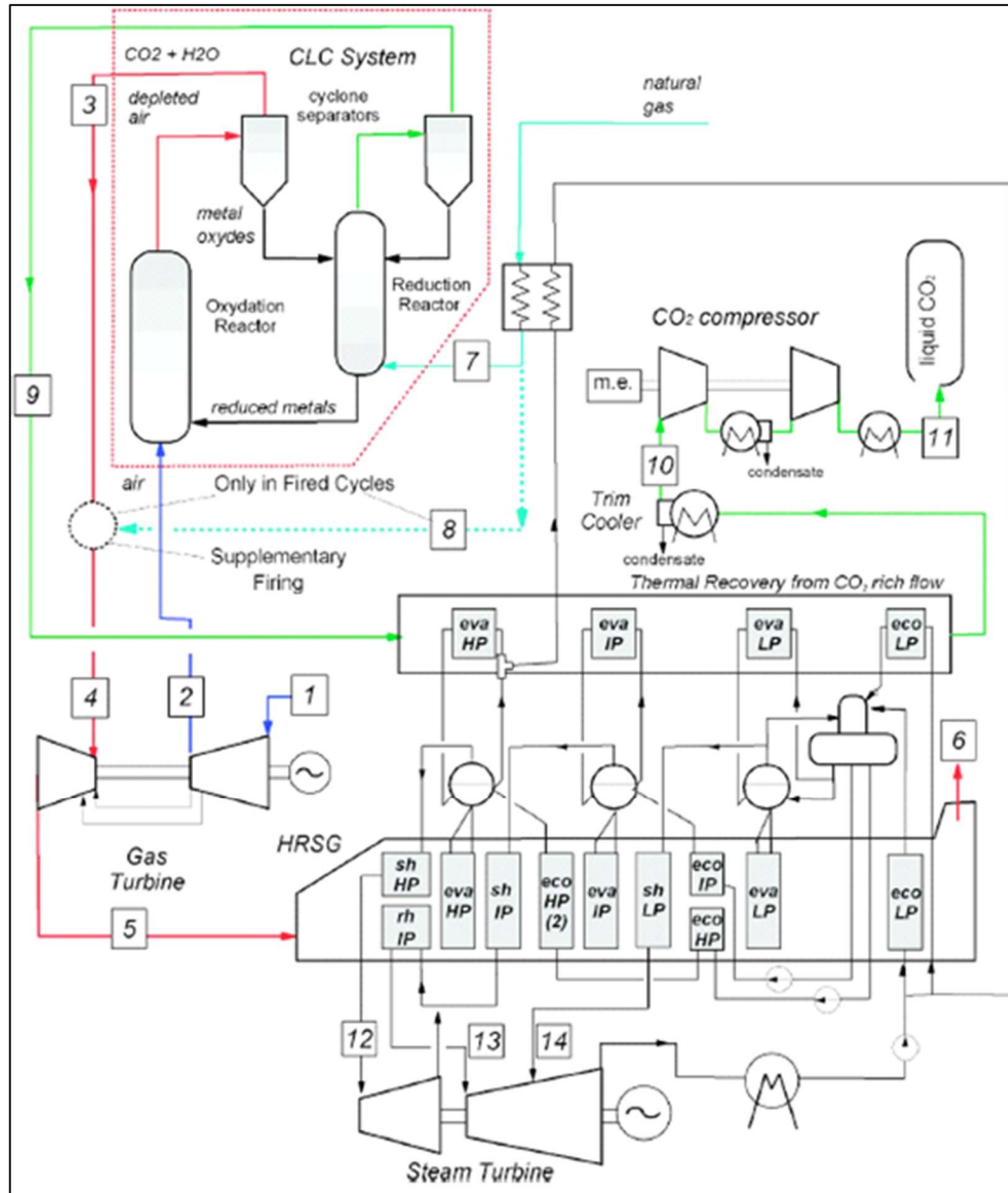


Figure 10 Combined cycle integration with CLC system [13]

Unfired steam cycles have a cheaper starting cost and are frequently employed when a single shaft system is installed as a unit. Fired systems can be utilized with any type of gasifier, but their running costs are higher than those of unfired systems, leading to higher efficiency.

There are specific consequences for the types of interactions between the CLC system and the power cycle when it is combined with CCPP. The gas turbine combustor, which compresses air at a greater pressure using a gas turbine compressor, is replaced by this integrated system with the CLC system. Direct discharge of compressed air into the oxidizing reactor, where an exothermic process is occurring, and discharge of vitiated air into the turbine expander. The method is feasible because the reducing unit in this system recovers energy from the compressed CO₂. Reducing the irreversibility of heat transmission and boosting waste heat recovery and energy production inside the system are imperative. The enhanced CO₂ is heated by three cooperating evaporators. Every one of these evaporators is connected to an LP economizer and HRSG drums. With the use of a knockout drum, the temperature of the CO₂ enhanced flow is lowered to room temperature in order to eliminate the extra water. In this manner, the CO₂ purity is increased to 94–96 vol% before being dried (the moisture content is removed) and then compressed to an increased pressure (85 bar) using an intercooled compressor for long-term storage. Aftercooler removes the compression heat before the intercooled compressor compresses the chilled air. The dryer is then equipped to use this compressed air. Syngas is delivered directly to the reduction reactor in an unfired system, and vitiated warm air is fed back to the turbine. The combustion chamber, which is positioned between the gas turbine expander and the CLC in a heated system, is where the syngas enters. Three evaporation pressure vessels are used in the steam cycles, which could be complicated and are particularly well suited for industrial high-performance combination cycles. When the efficiency, pressure-drops, thermal, and temperature losses are determined to be in

accordance with the cutting-edge technology accessible on a commercial scale, the parameters that affect CCPP power generation are estimated.

The relevant factors are [13];

1. Turbine inlet temperature (TIT).
2. Gas turbine pressure ratio.
3. Pressure of CO₂ enriched flow.
4. Steam cycles evaporation pressures.

TIT is crucial in assessing how effective the CCPP process is. The TIT and CCPP efficiency are directly related, meaning that if the TIT is higher, the CCPP efficiency will likewise be higher.

3.10.3 Carbon Capture Utilization & Storage

The most widely used technologies for producing electricity at the time were integrated gasification and combined cycle (IGCC), natural gas combined cycle (NGCC), and ultra-supercritical pulverized coal (USPC) power plants. This syngas is produced by each of these processes and is primarily composed of H₂ and CO₂ [31]. Yet, the system's total effectiveness of power stations is significantly decreased as a result of the integration of CCS technology. For example, the efficiency of NGCC, USPC, and IGCC power plants decreases by 7.2%, 11.5%, and 9.1%, respectively, when CCS technology is added [13]. The performance of power plants that are already in operation that have CCS added to them is affected, and the cost of electricity goes up as a result [13]. With the addition of CCS technology, USPC, IGCC, and NGCC each represent a cost increase of power of 38%,

44%, and 79%, respectively [31]. Notwithstanding the best operating performance offered by NGCC power plants with CCS, their widespread deployment is constrained by the rising of COE and the unpredictability of natural gas prices. Unlike the widespread utilization of coal-fired power facilities. Compared to other coal-based energy facilities, the IGCC has been demonstrated to have better control over greenhouse gas emissions and less of an adverse environmental impact, making it one of the most promising options. It offers higher overall process efficiency while also reducing COE emissions when compared to normal USPC power stations which are outfitted with CCS technology. Due to this, research has been conducted for many years to improve the efficiency of IGCC operations in the interest of rendering them more trustworthy and environmentally benign [38].

There are several strategies to reduce greenhouse gas emissions, including as improving power plant efficiency, decreasing the electricity demand, or integrating sources of clean energy that are suitable with CCS systems. The carbon capture and storage (CCS) system suite has been highlighted as a potential resource for reducing environmental carbon emissions. On the other hand, the implementation of CCS entails energy penalties for the generation of electricity in addition to an increase in financing and implementation costs [33].

The carbon capture and storage (CCS) method removes carbon dioxide from fixed-point sources, transports it to a storage region, and stores it there for an extended period of time. There are numerous approaches of CO₂ capture, depending on the kind of solvents and procedure employed. Due to the chemical makeup of solvents, chemical solvent absorption activities can be carried out at standard atmospheric pressures, reducing the dependence of solution loading and circulation on the partial pressure of acidic gases. Because of this, the

vast majority of pulverized coal (PC) power plants make every effort to remove CO₂ using chemical solvents. On the other extreme, Henry's law governs how physical solvents work. Greater pressures are used during the adsorption step for these solvents, and flash drums with multiple stages are used for the desorption process. The main energy consumers in these cases are auxiliary power consumption units, oxygen production units (OPUs), oxidant compression units, air/ CO₂ compression units for controlling combustor temperature, CO₂ compression units for compressing CO₂ to transit pressure 81 bar, and CO₂ compression systems [39].

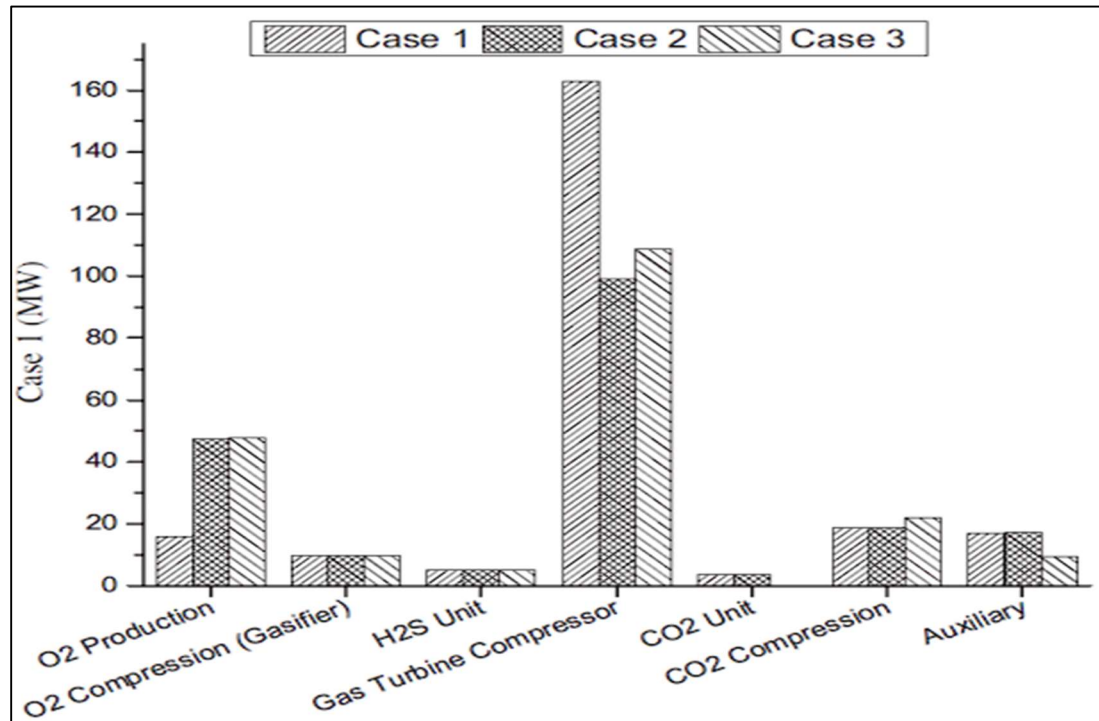


Figure 11 Power consumption breakdown in an IGCC power plant [32]

The main energy consumption categories for each of the three cases are shown in Figure 11. According to Ahmed et al.'s (2016) findings, their cases 2 and 3 have a lower compression energy requirement than the basic case design since they are built on the ideas

of an oxy-fuel ignition process [32]. As a result, less energy is needed overall and the reactor temperature can be controlled by CO₂ that has been recovered.

If SMR is to become a significant low-carbon supply of hydrogen, the capture and storage of carbon are a must. It is estimated that between 71% and 92% of the carbon dioxide can be absorbed during steam methane reforming. The technique has to capture more CO₂ if it is to be employed in the long run, though [6].

The biggest barrier to CCS in the immediate to medium term is cost, but there are other issues with the technology as well. One of the most recent developments for turning methane into low-carbon hydrogen through steam reformation and carbon dioxide absorption is sorption-enhanced SMR. The system has two main advantages. Firstly, the created H₂ can be 98% pure with only a small amount (ppm level) of CO and CO₂, which eliminates the need for filtration. However, the reaction is pushed to completion and produces noticeably greater yields of hydrogen when the generated CO₂ is continuously removed from the system by a solid absorbent [40].

Techniques for Carbon Capture and Storage

The terms "oxyfuel," "pre-combustion," and "post-combustion" refer to the three modes of carbon capture and storage (CCS) that are discussed as follows [13]:

A. Oxyfuel Carbon Capture

Burning coal, oils, or shale gas causes the generation of waste gas, which contains up to 15% of the carbon dioxide that was originally present [13]. Furthermore, removing CO₂ from the gas mixture is difficult and requires a lot of energy. In contrast, burning fuel with only pure oxygen as opposed to air is referred to as "oxyfuel" burning. The advantage of

this is that almost all of the water and carbon dioxide generated are vapors rather than solid or liquid forms. This made it possible to condense the vapor, which then allowed carbon dioxide to be transferred via pipes to undergo further processing [13].

Ordinary air only holds 21% oxygen, however oxygen-rich air is comprised of 95% oxygen, which is one benefit of using oxyfuel. As a result, the amount of volume used in the oxyfuel combustion process will be significantly reduced, which will have an important impact on the equipment's size. Moreover, during this process, NO_x production will be completely halted. Also, because the process would produce less nitrogen dioxide, there will be a larger concentration of carbon dioxide, which will simplify the separation procedure.

Using oxy-fuel combustion has the drawback that the machinery can have high capital costs. This is due to the necessity of liquefying oxygen and dissolving air into its constituent parts. Both of these procedures call for specialized equipment. As a result, up to 15% of the plant's overall power capacity will be consumed by this method. Also, as the purity of the oxygen in the environment increases, so does the degree of flammability. There are inherent dangers associated with the storage of oxygen because pure oxygen is infamous for its high level of reactivity [41].

B. Pre-Combustion Carbon Capture:

This technique extracts the CO_2 from the fuel prior to the combustion event. The interaction between both the fuel and water vapor or oxygen gas occurs during this phase. As a product of this reaction, a synthetic gas composed of CO and H_2 will be created. The syngas is subsequently combined with steam, leading the CO to be transformed into CO_2 . Whereas redirecting the hydrogen flow through a generator would allow it to be used to produce

power. Conversely, the pre-combustion method's drawback is that it can't be applied to older power plants that generate electricity by burning coal or other fossil fuels. This was the outcome of the technique. It may be used in a particular class of plants called "natural gas plants" to create synthetic gas, which is also hydrogen and carbon dioxide-containing. At these facilities, methane and steam are combined to create synthetic gas. Pre-combustion is not economically superior to post-combustion; hence this claim cannot be stated at this time because it has not been demonstrated [39].

C. Post Combustion Carbon Capture:

In this case, combustion will be followed by the separation phase. A simpler explanation would be that in order to extract CO₂, the flue gas is treated in an absorber column that is heavily supplied with the amine-based solvent. The CO₂ that had been contained in the flue gas is released when superheated steam heated to 120°C is employed to reheat the boiler of the absorption column, after which the boiler may be transported for storage. Many techniques can be used to capture carbon after burning, a few of which are illustrated below:

i. Sorption to Solids

Pressure swing adsorption started to be used in industrial plants in the 1970s. This process is used to extract carbon dioxide when producing hydrogen on a large scale [42]. Flue gas is introduced into a filled bed reactor during the adsorption to solids phase. The adsorbents in this procedure are the reactor's packing material. Two examples of the several adsorbents that can be utilized to selectively extract CO₂ from the flue gas stream are zeolites and metal-organic frameworks (MOF). The carbon dioxide is subsequently eliminated from the fine materials utilizing adsorption techniques like pressure swing or temperature swing.

However, a large number of reactors are needed to finish this process, which could result in higher initial investment. The duty needs, however, was found to be from 1.7 to 2.8 GJ per metric ton of carbon dioxide that was collected in a test on a smaller-scale design [43].

ii. Separation Through Membranes

Alternative applications using membrane technology for carbon capture using membrane are also common [29][36]. The separation of hydrogen is seen to be a potential application for palladium or palladium alloy membranes. The first study on hydrogen adsorption and dispersion across Pd membranes was published by Graham (1866), although membrane reactors (MRs) were not well known in science until the 1950s. Palladium-based membranes can work in highly hot settings, in contrast to pressure swing adsorption techniques. Moreover, they can keep the hydrogen at high pressures, which lowers the operational costs related to compression [44].

As a separating agent, it is feasible to make a screen that is sensitive to carbon dioxide. In this step, the exhaust gas passes through a thin layer of film that divides the material into different sizes based on molecules. The usage of membranes has a lot of advantages, such as versatility and a diminished impact on the environment. For processes that make use of high-pressure flue gas, this strategy is appropriate. In addition, compared to all other post-combustion processes, which may need as much as 9 GJ/ton of CO₂ under the same process conditions, the energy requirement is the minimum, ranging from 0.5 to 6 GJ/ton of CO₂. Yet, additional costs will be paid when installing compressors which are necessary in such relatively low pressure processes. The primary obstacles impeding the widespread use of this discovery are the membranes' low productivity and their high acid gas permeability [43].

iii. Cryogenic Capture

The distillation method, in which substances are sorted based on variations in their boiling points, and the procedure of cryogenic capture are relatively comparable. This method enables the recovery of CO₂ at the highest purity level possible at 99.99%, in contrast to all other post-combustion processes, which span from as little as 60% to a maximum of 95% in purity. However, if air is utilized instead of a combustion reaction with pure oxygen, the process might not be economically advantageous for the remediation of dilute CO₂ systems. Also, the substantial duty needed for cooling has an effect on the facility's overall running costs. At the beginning of the process, water must also be eliminated from the exhaust gases to avoid the formation of ice, which could choke the pipes [43].

iv. Amine-Based Solvents

One of the most widely used conventional methods for separating hydrogen from streams is chemical absorption, also recognized as the abolishment of carbon dioxide (CO₂) through the application of amine solvents. Large-scale processes are used to absorb carbon dioxide onto a range of amine-based solvents [29]. In a tall, tray-packed column known as an "absorber," the flue gas generated as the result of the burning process is processed with an amine-based solvent. The CO₂ is subsequently collected into the solvent as a result. The amount of CO₂ that is absorbed by the solvent can be increased by employing a counter-current flow mode. The solvent is then recovered by feeding the water-solvent mixture back to absorption after a stripper column separates CO₂ from it. However, maintaining the same solvent to CO₂ ratio throughout the process is vital to keep it operating at its peak efficiency. As a result, a make-up stream of the solvent is necessary [41]. This method has

the ability to capture one metric ton of carbon dioxide annually at the Boundary Dam power facility in Canada [41]. In figure 12 below, the CO₂ removal from MEA is depicted;

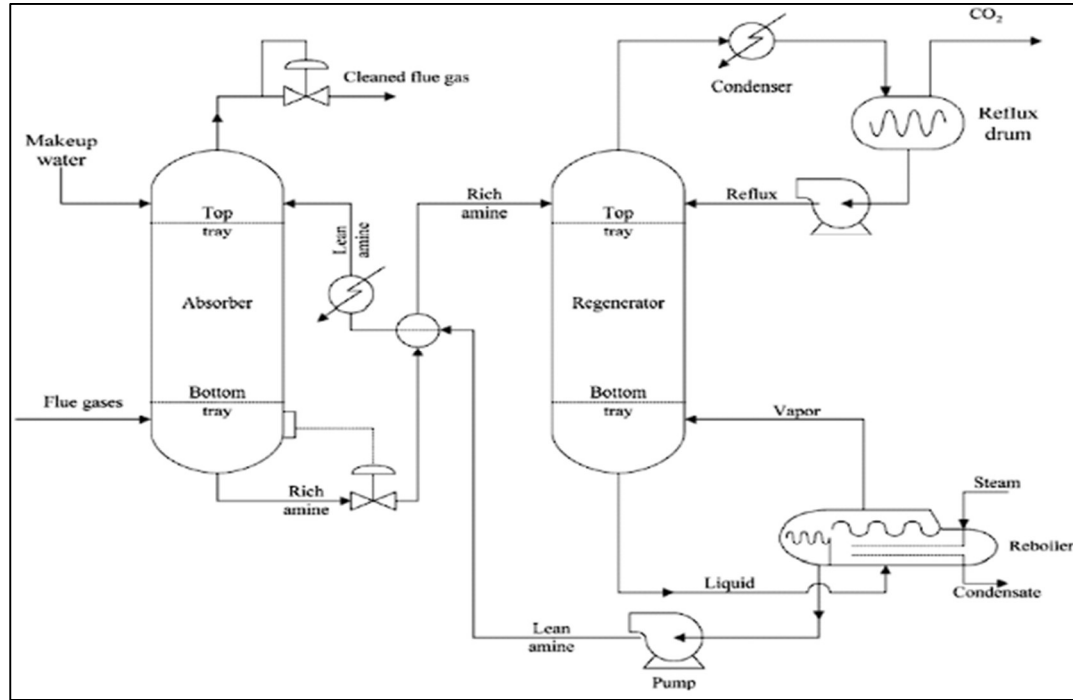


Figure 12 PFD of basic stripping process/chemical adsorption for amine-based CO₂ capture [13]

The two main downsides of the solvent absorption method are the reclamation of the solvent and the cost of maintaining the operation. It's important to understand that employing chemical solvents to remove CO₂ does not produce hydrogen with a high grade level. One technique for removing target gas species from gas mixtures at pressure and temperatures close to ambient is pressure swing adsorption [28].

It is better to run the absorption column at a lower temperature because CO₂ uptake is an exothermic process. The exhaust gases stream must therefore be chilled to between 10-15°C before entering the absorber column, so a cooler must be installed [13]. It has been demonstrated that this temperature range offers the best balance between optimizing savings and reducing makeup load. The stripper column's ideal temperature was found to

be 15°C higher than that of the absorption column. A low-duty blower is essential for assisting the flue gas stream to deal with pressure dips even though the system as a whole may run at atmospheric pressure [13].

The energy needed for absorption operations nowadays ranges from 2.3 to 9.2 GJ, which is significantly less than the 4.1 GJ per metric ton of absorbed CO₂ required by the prior CO₂ capture systems. Excluding the costs of the amine-based solvents, some laboratory tests have demonstrated that the power needed to process a metric ton of CO₂ can be decreased to 1.1 GJ. Industry uses a variety of amine-derived solvents, such as methyl diethanolamine (MDEA), diethanolamine (DEA), amino methyl propanolamine (AMP), and monoethanolamine (MEA). It is crucial to remember that amine-based solvents are corrosive materials, especially when carbon dioxide and oxygen are present. Three actions can be taken to address this problem. The system firstly by including corrosion inhibitors. Secondly, by using materials like stainless steel that are resistant to corrosion. Thirdly, adding corrosion-resistant materials is still necessary here even after lowering the solvent content to 40 wt% [13] [52]. Nonetheless, this retrofit strategy excels because it requires essentially no modifications to the plant's existing process machinery. When compared to alternatives, this method has also been found to offer significant cost savings.

Selected Carbon Capturing Process

Both pre and post combustion processes involve solvent-based or membrane carbon capture which is not being successful so far except in exploration industry [45]. Whereas the use of oxyfuel combustion eliminates the need of the carbon separation. This means that the only process required to isolate water from CO₂ will be a straightforward knockout

drum procedure. Therefore, the study employs the two oxyfuel combustion routes involving the use of ASU and CLC.

Although ASU is regarded as the energy intensive process but as it replaces the energy penalty due to the elimination of solvent-based carbon capture process it can become viable option providing high purity oxygen for the combustion. Also, ASU can be easily integrated into the existing power generation plant thus providing possible solution in terms of practicality [46]. On the other hand, CLC is a low-cost alternative for the oxyfuel combustion but can involve difficulties due to solid handling [46], [47].

CHAPTER 4

OBJECTIVES

1. This study's main goal is to create a novel and feasible model for producing electricity from waste plastic. The majority of the research being done now only looks at syngas production; however, further processing of the syngas is needed to generate electricity. When employing a combination of gas and steam turbines to produce energy and electricity, a combined-cycle power plant (CCPP) is necessary. First of all, four different cases were developed which comprise of (1) gasification+SMR with ASU, (2) gasification+SMR with CLC (3) gasification+POX with CLC, and (4) gasification+POX+DMR with CLC. Aspen Plus V12 was used to do thorough process simulations in order to determine the material and energy balance for these operations.

2. Based on these balances, equipment sizing is performed which provides the necessary information for the costing of plant. Once the required information is gathered, a technical analysis for sensitivity analysis, energy and efficiency calculations will be done. For that, initially, key technical parameters like syngas production, HCR, heating value (HHV), efficiency etc. are estimated. Based on these parameters, a comprehensive comparison among the developed cases is performed. The performed sensitivity analysis for key process parameters will then be useful to evaluate their impact on important output parameters like syngas quality and power generation.

3. In contrast, a cash flow model is developed for economic analysis in order to produce the project's cash flows and ascertain the levelized cost of electricity (LCOE), a crucial economic parameter. For that, installed equipment costs and manufacturing costs for the developed cases are calculated through equipment sizing and material and energy

balance, respectively. Using these parameters, CAPEX and OPEX are calculated by employing factor-based approach. Subsequently, cashflows are generated to estimate various economic parameters to evaluate the economic performance of the processes.

CHAPTER 5

METHODOLOGY

5.1 Process Modeling and Simulation

Figure 13 shows the methodological framework of the research adopted in this study. The property package utilized for the simulations, known as the RKS-BM property approach, uses the Redlich-Kwong-Soave (RKS) cubic equation of state with Boston-Mathias (BM) alpha function to derive thermodynamic properties [48]. In all temperature and pressure ranges, the property approach can be applied to mixes that are nonpolar or mildly polar. It is advised for use in petrochemical, refinery, and gas processing applications. The RKS-BM property approach is consistent throughout the critical region.

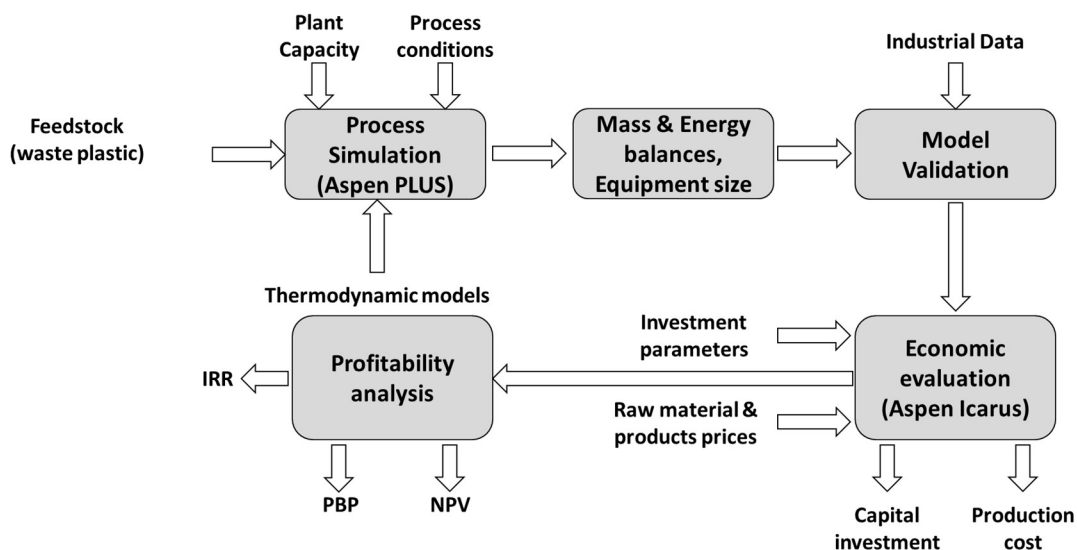


Figure 13 Schematic overview of the research methodology.

5.1.1 Design Information

The operating parameters for various equipment are sourced from the published literature, with slight variation to customize the given processes, as mentioned in Table 10.

Additionally, the Aspen plus models used to represent the actual process are mentioned. Gasifier is modelled with the combination of RYield and RGibbs reactors. Pyrolysis and gasification are the two phases that make up gasification, and they are carried out independently in RYield and RGibbs, respectively. In contrast, the solid waste plastic component in RYield is broken down into the appropriate elements by breaking molecular chains. In accordance with the most thorough and accurate assessment of the waste plastic, it generates the elements (H₂, carbon, N₂, S, etc.) and other components (ash). After reaching phase and chemical equilibrium at 1300°C and 25 bar, the products are then let to react with the steam in RGibbs to produce syngas. On the other hand, for all the reforming stages, RGibbs is used with the respective operating temperature and pressure sourced from the literature.

In terms of the power generation portion of the process, a combined gas and steam turbine is used in the CCPP. The study's gas turbine, referred to as "Auxiliary gas turbine unit," is connected to an HRSG in order to recover heat energy from GT's flue gases (FG) and convert it into medium-pressure steam. On the other hand, as the steam turbine under consideration is a condensing turbine, a condenser is needed downstream of the turbine. The turbine's output steam is supplied to the condenser, which runs at 0.1 bar of vacuum pressure. Due to the operation at the lower pressure and temperature, the condensing turbine is supposed to extract more enthalpy from the condensing steam thus producing more power.

Table 10 Key equipment design information

Equipment	Description	Aspen Model	Reference
Feed Crusher	Temp = 25 °C	Gyratory Crusher	[8]

	Press = 1 atm Size reduction = 200 microns		
Gasifier	Temp = 1300°C Press = 25 bar Steam: feed (plastic) = 1.25	RYield + RGibbs	[49]
HDS	Temp = 400 °C Press = 1 bar H2 feed = 1 kg/hr	RStoic	[62]
SMR	Temp = 900°C Press = 3 bar Steam: feed (CH ₄) = 1.83	RGibbs	[50]
POX	Temp = 700°C Press = 25 bar	RGibbs	[51]
DMR	Temp = 800°C Press = 25 bar CO ₂ recycle = 17mol%	RGibbs	[11]
CLC	Temp = 1400°C Press = 1 bar Oxygen carrier = Iron oxide Excess oxygen = 5mol%	RStoic	[46]
GT + HRSG	FG outlet temp = 90°C FG outlet press = 1 bar Steam temp = 300°C Steam press = 30 bar	RStoic + Turbine	[52]

	Polytropic efficiency = 0.9 Type = Auxiliary power unit		
Steam turbine	Condensate Temp = 49°C Outlet Press = 0.1 bar Polytropic efficiency = 0.9 Type = condensing turbine	Turbine	[52]

5.2 Development of Case Studies

Four distinct case studies were designed in order to thoroughly examine the production of power from waste plastic by means of different oxygen delivery methods. Brief description and process flow diagrams of each case are detailed below:

5.2.1 Case 1 (Base Case): SMR Coupled With ASU

The process for converting waste plastics and natural gas into energy and high-purity carbon dioxide consists of multiple integrated unit operations designed for efficient feedstock preparation, gasification, syngas production, and energy generation.

Feed Preparation and Size Reduction: The process begins with waste plastic feedstock, polyethylene (PE) and polypropylene (PP), entering a feed preparation unit. Here, the feed undergoes dehumidification and shredding to reduce moisture content and prepare the material for further processing. A crusher further reduces the size of the shredded plastic to particles with a maximum size of 200 microns, facilitating uniform heating and efficient gasification in subsequent stages.

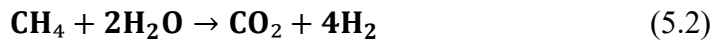
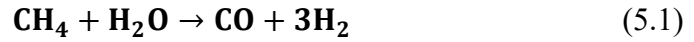
Steam Gasification Unit: The prepared plastic feed is fed into an entrained flow gasifier where it undergoes steam gasification at a temperature of 1300°C and a pressure of 25 bar

[49]. This process converts solid plastic into gaseous components, primarily producing syngas (a mixture of H_2 , CO , and CO_2) along with a small amount of solid residue. The gasification reaction utilizes a steam-to-plastic ratio of 1.25:1 [50], which aids in breaking down the polymer chains and converting them into useful gaseous products.

Hydrodesulphurization of Natural Gas Feed: The CH_4 feedstock for the process is typically sourced from natural gas, biogas, or synthetic methane production, and its composition can vary significantly based on the source. Natural gas, which is the most common source, predominantly contains methane (70-95%) along with other hydrocarbons, sulfur compounds, CO_2 , and water [62]. For the purpose of this study, it is assumed that the CH_4 feed contains 5% sulfur compounds, such as H_2S , which are common impurities in methane sources.

The presence of sulfur compounds in the methane feed poses significant challenges, as these impurities can poison catalysts in downstream processes, such as reforming. Therefore, a desulfurization unit is essential to remove sulfur compounds and reduce their concentration to acceptable levels, typically below 1 ppm, to ensure catalyst protection and process efficiency [62]. In this study, a desulfurization unit, such as an H_2S scrubber or hydrodesulfurization (HDS) unit, is included in the process flow diagram. Before entering the steam methane reformer (SMR), the natural gas feed is treated to remove impurities such as Sulphur containing compounds. This is done using a hydrotreating unit, where the feed is mixed with hydrogen and passed over a catalyst (typically nickel-molybdenum or cobalt-molybdenum) at elevated temperatures and pressures [62]. The process converts sulfur compounds into hydrogen sulfide, which is subsequently removed, ensuring a clean feed for the reforming process.

Steam Methane Reforming (SMR): The treated natural gas is preheated using heat recovered from the high-temperature outlet stream (1300°C) from the gasification unit. The SMR process operates at a temperature of 300°C and a pressure of 3 bar. In the reactor, methane reacts with steam in the following reactions:



The SMR process produces syngas consisting primarily of hydrogen (H₂), carbon monoxide (CO), and carbon dioxide (CO₂). The steam-to-methane ratio is maintained at 1.83:1 for consistent syngas production [50]. The outlet stream of SMR is compressed to 25 bar for subsequent power generation in combined cycle where gas turbine is used.

Air Separation Unit (ASU): The air separation unit (ASU) uses cryogenic distillation to separate oxygen from air at temperatures as low as -196°C. The produced pure oxygen is supplied to the combined cycle unit, minimizing the need for air and eliminating the energy penalty associated with post-combustion carbon capture [53], [54].

Combined Cycle Unit: The combined cycle unit utilizes the produced syngas as fuel and combusts it using pure oxygen to drive a gas turbine (GT). This combustion step generates substantial thermal energy, which is converted into electrical energy. To maximize energy recovery, the hot flue gases exiting the GT are directed through a heat recovery steam generator (HRSG), which produces superheated steam at 30 bar and 300°C [52]. The superheated steam generated by the HRSG powers a steam turbine, producing additional electricity. This integration of gas and steam cycles enhances the overall energy efficiency of the process.

CO₂ Separation and Purification: The process byproduct, carbon dioxide, is cooled and separated using a knock-out drum. The CO₂ is then compressed, dried, and purified to 99.5% purity, making it suitable for sale or other applications.

In cases 1-4, the CO₂ purification process was set to operate at 25 bar, a pressure carefully selected to balance CO₂ purity and recovery. This pressure is a result of optimizing the trade-off between these two key performance metrics: increasing the pressure tends to improve CO₂ purity but decreases recovery, while lowering the pressure enhances recovery at the expense of purity. Since the objective of the process is to achieve a high-purity CO₂ stream while maintaining reasonable recovery levels, 25 bar represents a compromise that ensures both targets are met efficiently.

At pressures higher than 25 bar, the system experiences an increase in the separation efficiency of CO₂ from water, resulting in higher purity. However, this comes with a trade-off: CO₂ recovery diminishes due to the higher energy requirements and altered phase behavior during separation. Conversely, reducing the pressure below 25 bar increases CO₂ recovery but compromises the purity of the CO₂ stream, as water separation becomes less effective. Such a drop in purity could affect downstream applications of the CO₂ or reduce its market value. The choice of 25 bar was guided by process simulations and engineering analysis, which indicated that this pressure achieves a practical balance between purity and recovery for the specific operating conditions of Case 1-4.

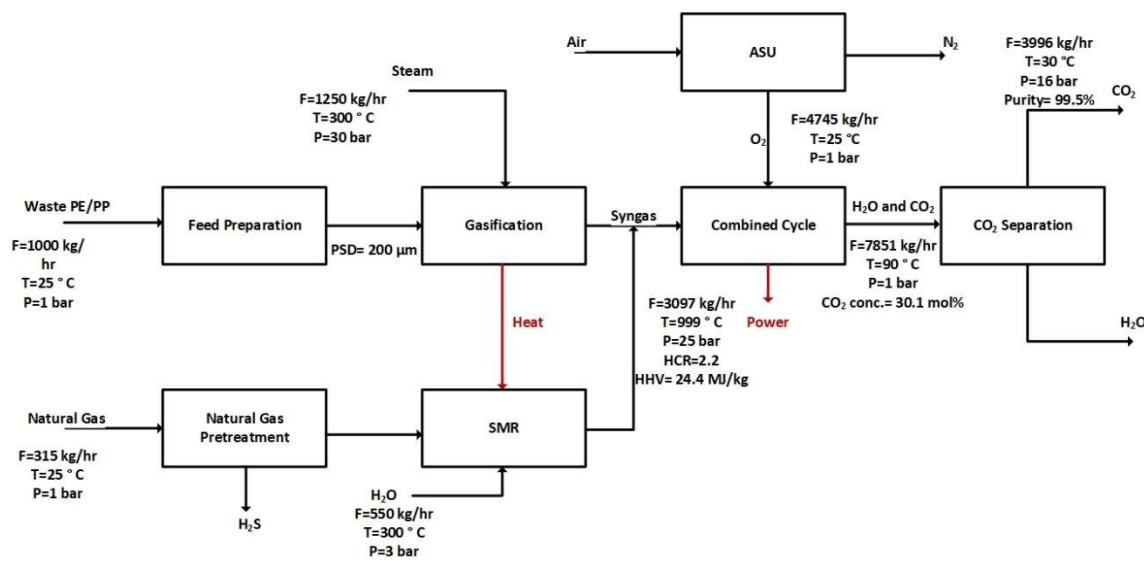


Figure 14 Block flow diagram of case 1 for power generation from waste plastic utilizing steam methane reforming (SMR) and air separation unit (ASU)

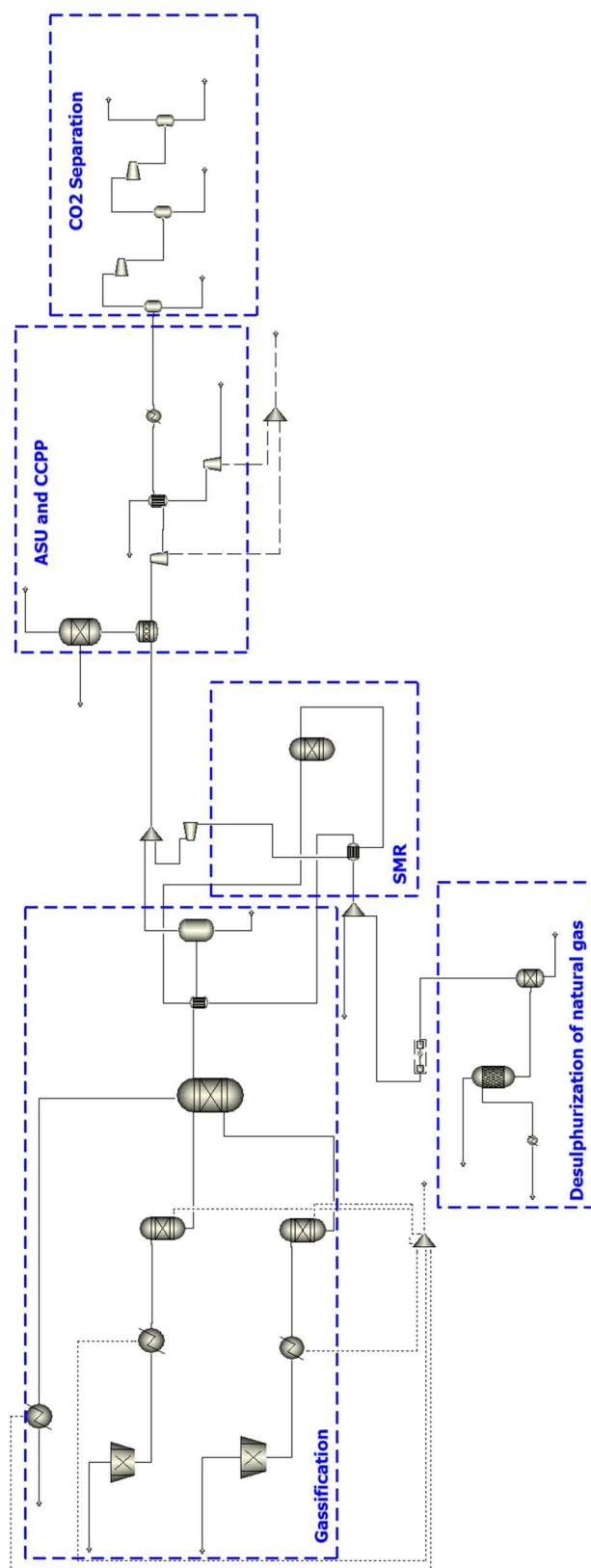


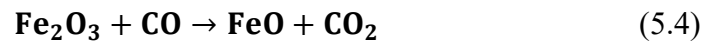
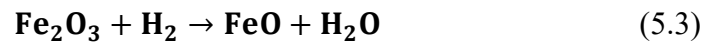
Figure 15 Aspen plus simulation flowsheet for case 1

5.2.2 Case 2: SMR Coupled With CLC

Case 2 presents an alternative configuration for the process, differing from case 1 by using chemical looping combustion (CLC) as the source of oxygen for the gas turbine (GT) combustion instead of an air separation unit (ASU). The remaining process steps remain unchanged, maintaining consistency in feed preparation, gasification, and syngas processing.

The process begins with the prepared plastic feedstock, which is charged into the steam gasification equipment. Operating at a temperature of 1300°C and a pressure of 25 bar, the gasifier converts the solid plastic into gaseous components (syngas) and a small amount of solid residue [49]. To enhance energy efficiency, the feed for the steam methane reformer (SMR) is preheated using the hot outlet stream (1300°C) from the gasifier. The SMR is carried out at 3 bar pressure and the outlet of SMR is then compressed to 25 bar pressure for power generation in combined cycle.

Syngas Combustion Using Chemical Looping Combustion (CLC): Once syngas is produced, it is combusted in a combined cycle unit, utilizing oxygen provided by the CLC system. The CLC approach uses iron oxide (Fe_2O_3) as a source of oxygen. During combustion, Fe_2O_3 reacts with the fuel (H_2 and CO in the syngas), resulting in the following reduction-oxidation reactions [46], [55]:



This process occurs in two fluidized bed reactors where the metal oxide acts as an intermediate oxygen carrier, alternating between reduced and oxidized states. The oxygen-rich metal oxide is separated from the resultant CO₂ using a cyclone separator and then recycled for further use in the CLC cycle. Unlike conventional combustion, this approach produces CO₂ that is undiluted with nitrogen, simplifying the separation and purification of the CO₂ [16], [17].

The use of CLC as an oxygen source eliminates the need for a separate ASU, thereby reducing operational complexity and associated energy costs. CLC relies on the natural reactivity of metal oxides, Fe₂O₃, to provide oxygen for combustion without direct contact with air, making the process inherently simpler and more environmentally friendly [18]. It also enables operation with contaminants in the feed, as it simplifies CO₂ separation and purification without the need for extensive upstream or downstream removal processes [19].

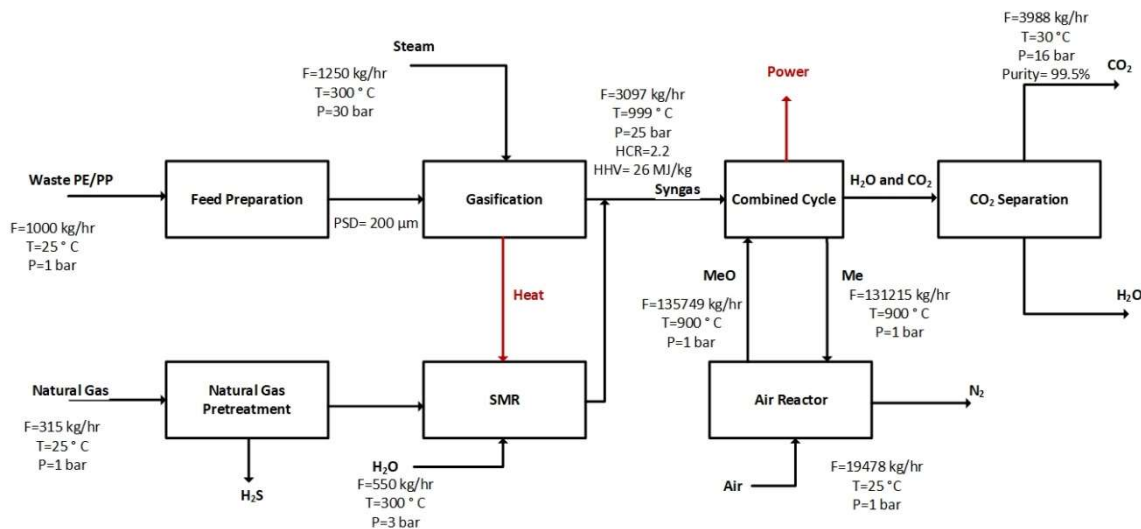


Figure 16 Block flow diagram of case 2 for power generation from waste plastic utilizing steam methane reforming (SMR) and chemical looping (CLC)

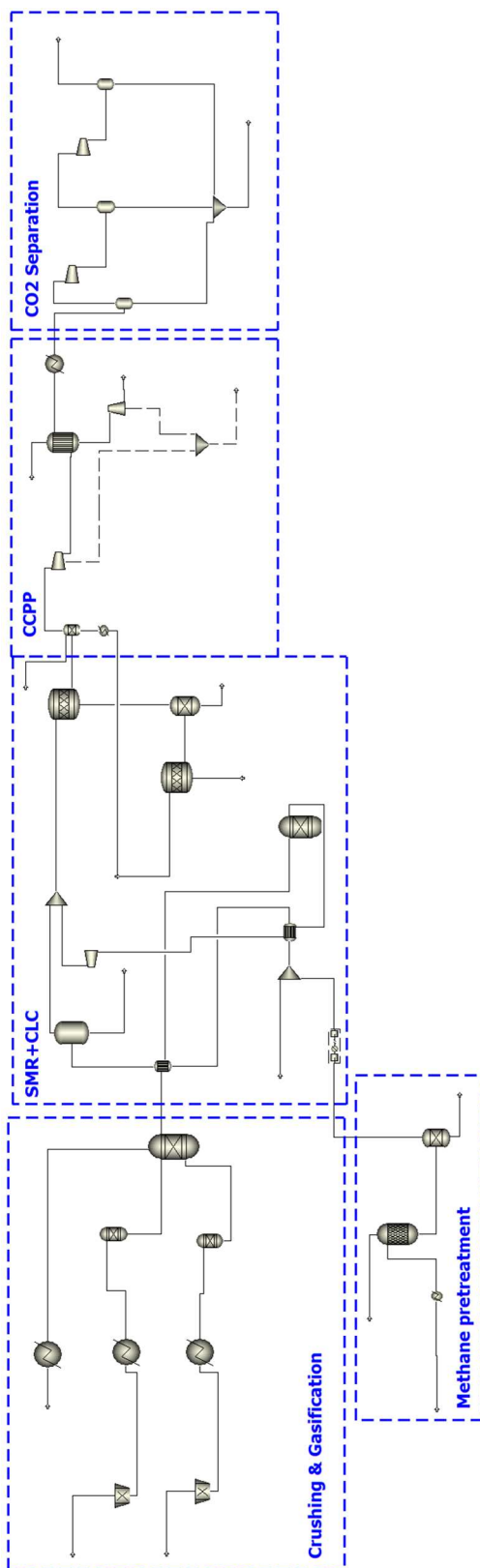


Figure 17 Aspen plus simulation flowsheet for case 2

5.2.3 Case 3: POX Coupled With CLC

Case 3 introduces a variation in the process by replacing the steam methane reformer (SMR) used in cases 1 and 2 with a partial oxidation (POX) reaction unit. This change eliminates the need for the heat stream from the gasifier, as the POX process is inherently exothermic. The reaction proceeds as follows:



In this process, methane reacts with a controlled amount of oxygen to produce syngas (a mixture of carbon monoxide (CO) and hydrogen (H₂) through partial oxidation. This exothermic reaction releases heat, which contributes to the thermal energy balance of the process and reduces external heating demands.

Adoption of Chemical Looping Reforming (CLR) System: In addition to using POX, this case incorporates a chemical looping reforming (CLR) system. The air separation unit (ASU) used in previous cases is replaced with a reaction chamber that supplies the necessary oxygen for both the POX process and the combustion of syngas. The CLR system is based on the principles of chemical looping combustion (CLC), utilizing metal oxides (Fe₂O₃) as an oxygen carrier. The key reactions involved are:

1. Reduction Reaction:



In this step, the metal oxide (Fe₂O₃) reacts with methane, producing reduced metal (Fe), syngas (CO and H₂), and water (H₂O).

2. Oxidation Reaction:



The reduced metal (Fe) is then re-oxidized by reacting with oxygen, regenerating the Fe_2O_3 for continuous operation of the system.

Key Advantages and Process Implications

The use of CLR in conjunction with the POX process offers significant advantages:

- **Efficient Oxygen Supply:** The CLR system provides a continuous and efficient supply of oxygen for the POX reaction and the combustion of syngas, without the need for an ASU.
- **Thermal Efficiency:** The exothermic nature of the POX reaction reduces the need for external heating, simplifying the energy integration within the process.
- **Sustainability and Environmental Benefits:** By incorporating a chemical looping system, the process minimizes energy consumption and allows for efficient CO_2 separation and capture, contributing to an overall more sustainable operation.

This configuration highlights the innovative integration of partial oxidation and chemical looping reforming to enhance process efficiency, reduce energy demands, and optimize oxygen utilization. The changes made in case 3 provide a distinct approach compared to the previous configurations using SMR and CLC.

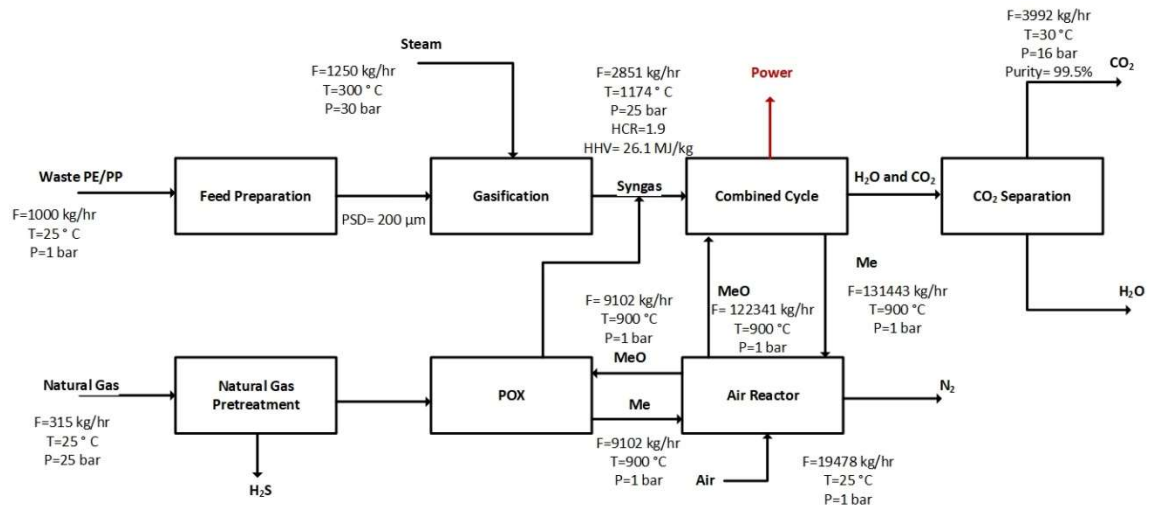


Figure 18 Block flow diagram of case 3 for power generation from waste plastic utilizing Partial oxidation (POX), and Chemical looping combustion (CLC)

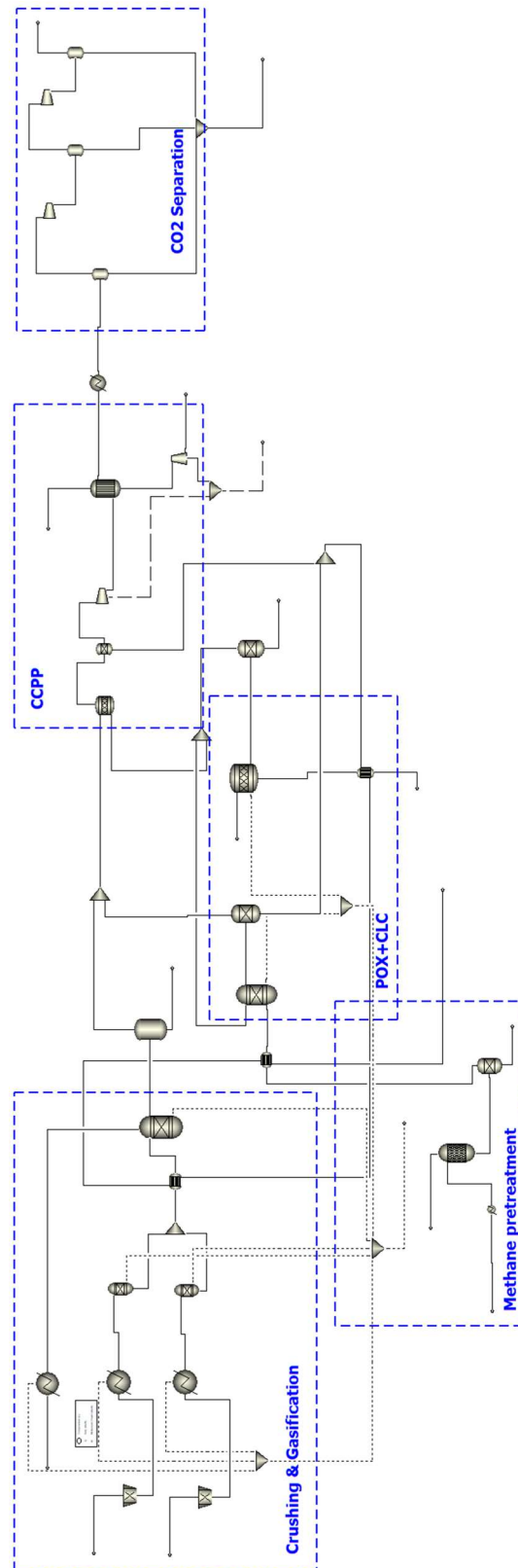


Figure 19 Aspen plus simulation flowsheet for case 3

5.2.4 Case 4: DMR And POX Coupled With CLC

The process flow diagram in case 4 illustrates an integrated system designed to convert waste polyethylene/polypropylene (PE/PP) and natural gas into syngas, power, and high-purity CO₂. This system uniquely combines dry methane reforming (DMR) and partial oxidation (POX) to optimize syngas production while enhancing process efficiency and sustainability. The integration of these two processes, DMR and POX, offers several advantages.

DMR is an endothermic process that uses CO₂ as a reactant to convert methane into syngas (a mixture of CO and H₂), effectively utilizing captured CO₂ and contributing to carbon recycling. However, the endothermic nature of DMR requires a significant energy input, which can be challenging to sustain efficiently on its own. On the other hand, POX is an exothermic process that partially oxidizes methane to produce syngas, releasing heat that can offset the energy requirements of DMR. By combining these two processes, the system achieves a thermal balance, reducing external energy demands and improving overall process efficiency.

In this system, oxygen for the POX process is supplied by the chemical looping combustion (CLC) unit. CLC technology enhances process sustainability by using solid oxygen carriers to deliver oxygen for combustion, eliminating the need for direct air separation and reducing energy consumption. This also prevents nitrogen from entering the process, ensuring the production of high-purity syngas.

For the DMR process, CO₂ is provided through the recycling of CO₂ captured from the system. This closed-loop approach not only supplies the necessary reactant for the

reforming reaction but also enhances carbon utilization, aligning with sustainability goals by reducing net CO₂ emissions. The synergy between POX and DMR, supported by the CLC unit and CO₂ recycling, creates a highly efficient and environmentally friendly process for converting waste plastics and natural gas into valuable products.

Dry Methane Reforming (DMR) and Partial Oxidation (POX): The combined DMR and POX unit operates at 850°C and 4 bar. In this unit, methane (CH₄) reacts with recycled CO₂ in the DMR reaction:



The CO₂/CH₄ molar ratio of 1.05 ensures an optimal balance of reactants to maximize syngas production while reducing the reliance on methane. By recycling 17% of the captured CO₂, the process utilizes CO₂ as a carbon source, lowering the consumption of fossil fuels and reducing overall CO₂ emissions. The choice to recycle 17% is based on achieving a sustainable and efficient carbon balance within the process, providing just enough CO₂ to maintain the reforming reaction without overwhelming the system or requiring excessive CO₂ capture downstream.

The POX process provides additional heat to drive the endothermic DMR reaction, balancing the thermal requirements and enhancing syngas yield. The reaction with oxygen is facilitated by a chemical looping reforming (CLR) system that supplies oxygen in the form of metal oxides, thereby eliminating the need for an air separation unit (ASU).

Air Reactor (Chemical Looping Combustion - CLC): An Air Reactor operates based on chemical looping combustion (CLC) principles. Metal oxides (Fe₂O₃) cycle between

reduced and oxidized forms, supplying oxygen for the POX reaction and syngas combustion. In this reactor:

- Reduction Reaction: Fe_2O_3 reacts with methane to produce reduced Fe, CO, and H_2 .
- Oxidation Reaction: Reduced Fe is re-oxidized with air, regenerating Fe_2O_3 and releasing nitrogen (N_2) as a byproduct.

This oxygen-carrying loop supports a continuous and efficient oxygen supply to the POX and combustion processes without the need for conventional oxygen separation.

CO₂ Capture and Recycling: The combined cycle unit generates flue gases with a CO₂ concentration of 38.4 mol%. These gases undergo CO₂ capture, where the CO₂ is separated, compressed, and purified to a 99.5% purity. Out of the captured CO₂, 17% is recycled back to the DMR process, while the remaining CO₂ is either stored or sold. Recycling 17% of CO₂ balances the carbon input, reduces reliance on fossil methane, and promotes a circular use of CO₂ within the system.

Rationale for 17% CO₂ Recycling and CO₂ to CH₄ ratio: Recycling 17% of the captured CO₂ ensures a stable and sustainable process. This percentage provides sufficient CO₂ to sustain the DMR reaction without overloading the reforming unit with excessive CO₂, which could negatively impact syngas composition and downstream operations. Additionally, this level of recycling supports carbon utilization while minimizing external CO₂ emissions, aligning with environmental goals and enhancing process efficiency.

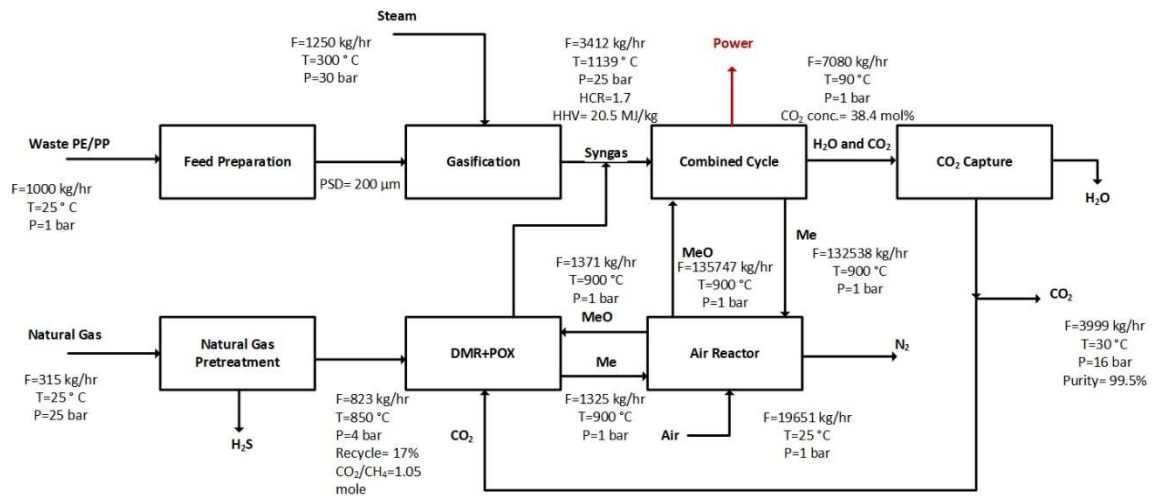


Figure 20 Block flow diagram of case 4 for power generation from waste plastic utilizing Partial oxidation (POX), Dry methane reforming (DMR) and Chemical looping combustion (CLC)

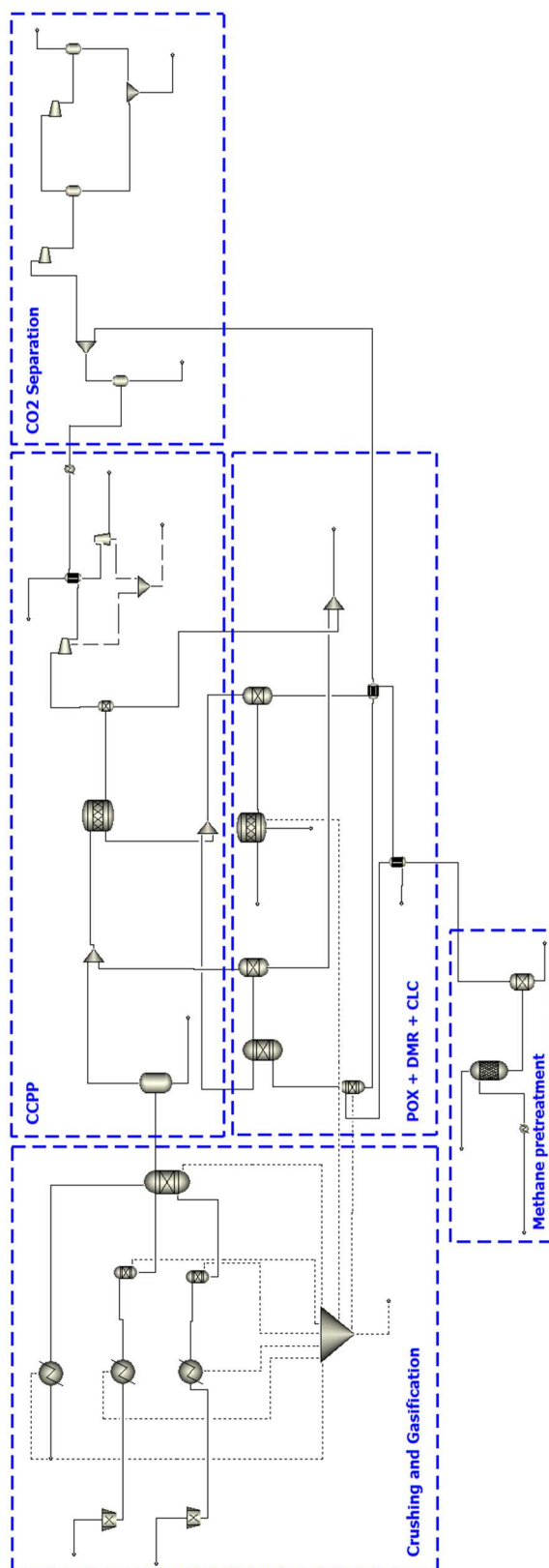


Figure 21 Aspen plus simulation flowsheet for case 4

5.3 Process Simulations

Aspen Plus V12 was used to simulate the proposed models in order to make them come to life. As the non-conventional components, PP and PE were introduced by specifying their ultimate and proximate analyses, which are shown in Table 11 [50]. Phase equilibrium computations and component property estimations are performed using the RKS-BM equation of state model. The gasification process that creates the basic components from the waste polymers is simulated using a combination of RYield and RGibbs. The SMR is modeled using RGibbs, which computes the products by minimizing the Gibbs free energy. HeatX, turbine, and Rstoic were used to provide a realistic depiction of the gas turbine HRSG.

The reactors used in this study, including the gasifier and reformers, were modeled using yield and Gibbs reactors to simulate the conversion processes. This modeling choice is driven by specific considerations regarding the nature of waste plastic gasification and the objectives of the study.

Rationale for Using Yield and Gibbs Reactors

The decision to utilize yield and Gibbs reactors stems from the lack of detailed reaction kinetics available for the gasification of waste plastics and reforming processes. Waste plastic feedstocks are inherently complex, consisting of a diverse range of polymers and additives. This complexity leads to numerous reaction pathways, intermediate species, and possible reactions during thermal conversion. Developing a comprehensive kinetic model for such a process is not only highly challenging but also nearly impossible due to:

- The variability in the composition and properties of waste plastics, which can vary significantly between different sources.
- The intricate nature of chemical reactions involved in the thermal decomposition and gasification of plastics.
- The limited availability of detailed kinetic data in existing literature for waste plastic gasification and reforming processes under a wide range of operating conditions.

As this study focus on the technological assessment and economic feasibility of different technological routes, the use of yield and Gibbs reactors provides an effective approach for estimating the potential performance and syngas composition under specified conditions. Unlike kinetic-based models, yield and Gibbs reactors rely on thermodynamic principles to determine the equilibrium state of the system. This approach offers several advantages:

- **Thermodynamic Consistency:** By assuming that the reactions proceed to equilibrium, these reactors provide an upper-bound estimate of achievable syngas yield and composition based on energy and mass balances.
- **Feasibility Assessment:** The use of equilibrium models allows for a robust evaluation of different process configurations without requiring extensive kinetic data. This is particularly valuable for feasibility studies where the focus is on comparing and assessing the potential of different technologies.
- **Simplification without Compromising Validity:** While equilibrium models may not capture detailed transient or kinetic behavior, they provide a reliable basis for

assessing overall system performance, making them suitable for high-level feasibility analyses.

Justification for the Modeling Approach

Given the complexity and lack of detailed kinetics for waste plastic gasification and reforming, using yield and Gibbs reactors is a practical and justified approach. It enables the evaluation of process feasibility, energy efficiency, and syngas production potential under different configurations without requiring extensive and often unavailable kinetic data. This modeling strategy aligns with the overall scope and objectives of the study, providing meaningful insights into the performance and viability of the technological cases analyzed.

Table 11: Ultimate and proximate analysis of Polyethylene and Polypropylene used in the study [50]

Components	Polyethylene	Polypropylene
Proximate analysis (wt%)		
Moisture	0.02	0
Ash	0.15	0.70
Volatile matter	99.83	99.30
Ultimate analysis (wt%)		
Carbon	85.81	86.23
Hydrogen	13.86	12.28
Nitrogen	0.12	0.62
Sulfur	0.06	0.17
Ash	0.15	0.70
LHV	38.04	44.70

5.4 Techno-Economic Analysis

A techno-economic analysis of the four examples was undertaken to assess the major performance criteria of the processes and their feasibility in the current market. The techno-economic analysis includes technical analysis like energy and efficiency calculations, and economic analysis like cashflow analysis to estimate levelized costs. Initially, key technical parameters like syngas production, HCR, heating value (LHV), efficiency etc. are estimated using material and energy balances. Sample calculations for all key parameters are provided for case 1. Efficiency of the process is calculated using the following formula:

$$\text{HCR} = \frac{\text{Moles of hydrogen in syngas}}{\text{Moles of Carbon monoxide in syngas}} \quad (5.9)$$

$$\text{HCR} = \frac{183.8}{84.8} = 2.2$$

$$\eta = \frac{\text{Electricity produced}}{\text{LHV}_{\text{PP}} + \text{LHV}_{\text{PE}} + \text{LHV}_{\text{NG}} + \text{Utilities consumed}} \times 100 \quad (5.10)$$

$$\eta = \frac{3.78}{5.9 + 4.7 + 4.2 + 7.9} \times 100$$

$$\eta = 16.6\%$$

Based on these parameters are comprehensive comparison among the developed cases is performed.

Conversely, a cash flow model is constructed for economic analysis in order to establish the levelized cost of electricity (LCOE), an important economic parameter, and to generate the project's cash flows. To do it, the process's CAPEX and OPEX were first determined. Table 11 provides a summary of the analysis's underlying assumptions:

Table 12 Assumed parameters for the economic analysis [58]

Assumed parameters	Value
Costing year	2023

Plant operational life	20 years
Construction period	2 years
Operating hours/year	8000
Depreciation of general plant	7 years
Discount rate	8% per year
Tax rate	15% per year

CAPEX includes the equipment installed costs, direct and indirect fixed costs. While the costs of specialized equipment, such as reactors, were scaled using prior research, the installed costs of common equipment were determined using Aspen Plus Cost Analyzer. Below is the used formula for determining the installed cost of specialized equipment [56], [57], [58], [59]:

$$\text{Installed Cost} = \left(\frac{\text{New Capacity}}{\text{Reference Capacity}} \right)^n \times \text{Reference installed Cost} \times \left(\frac{\text{CEPCI}_{2023}}{\text{CEPCI}_{\text{Ref}}} \right) \quad (5.11)$$

$$\text{Installed Cost (ASU)} = \left(\frac{114.1}{533.3} \right)^{0.7} \times 25.4 \times \left(\frac{803}{525.4} \right)$$

$$\text{Installed Cost (ASU)} = 4.21 \text{ Mil. \$}$$

for which the scaling factor is n. The overall capital investment is then determined by multiplying these installed costs by a number of variables, as shown in Table 13:

Table 13 Factors used for calculation of total capital investment. ISBL = Inside battery limit [58]

Economic Factors	Values
Total direct cost (TDC)	
Warehouse	4% of ISBL
Site development	9% of ISBL

Additional Piping	4.5% of ISBL
Total indirect cost (TIDC)	
Prorateable costs	10% of TDC
Field expenses	10% of TDC
Home office and construction	10% of TDC
Project contingency	10% of TDC
Other costs	10% of TDC
Total capital investment (TCI)	
Fixed capital investment (FCI)	TDC+TIDC
Working capital (WC)	5% of FCI
Land	6% (of installed costs)

The total installation cost (TIC) will be estimated as follows:

$$\mathbf{TIC} = \sum_e \mathbf{IC}_e. \quad (5.12)$$

The total direct installed cost (TDC) will be estimated as follows:

$$\mathbf{TDC} = \mathbf{TIC} + \mathbf{W} + \mathbf{SD} + \mathbf{AP}, \quad (5.13)$$

where, W is the warehouse cost taken as 4% of TIC, SD is the site development cost taken as 9% of TIC, and AP is the additional piping taken as 4.5% of TIC.

The total indirect installed cost (TIDC) will be estimated as follows:

$$\mathbf{TIDC} = \mathbf{TDC}(\mathbf{PC} + \mathbf{FE} + \mathbf{HOCF} + \mathbf{PC} + \mathbf{OC}), \quad (5.14)$$

where, PC are the proratable costs taken as 10% of TDC, FE are the field expenses taken as 10% of TDC, HOCF is the home office and construction fee taken as 20% of TDC, PC is the project contingency taken as 10% of TDC, and OC are the other costs including start-up, permits, etc. taken as 10% of TDC.

The fixed capital investment (FCI) will be calculated as follows:

$$\mathbf{FCI = TDC + TIDC.} \quad (5.15)$$

Eventually total capital investment (TCI) will be estimated as follows:

$$\mathbf{TCI = FCI + Land + Working\ capital,} \quad (5.16)$$

where land is land cost taken as 6% of TIC and WC is working capital taken as 5% of FCI.

The cost of manufacturing (COM) is sum of raw material costs (RMC), utility costs (UC), wastewater treatment cost (WWTC) and labor costs (LC), and defined as follows:

$$\mathbf{COM = RMC + UC + WWTC + LC} \quad (5.17)$$

The RMC and UC will be estimated based on material balance obtained from simulations, WWTC for per m³ is \$0.041 and LC is calculated as:

$$\mathbf{COL = f_1 + (f_2 \times SHE^2) + (f_3 \times LHE)^{0.5}} \quad (5.18)$$

where SHE, LHE, f1, f2, and f3 are solid handling equipment, liquid handling equipment, and f1–f3 are factors (6.29, 31.7, and 0.23) to estimate LC, respectively.

The net present value (NPV) and the rate of return on investment will be estimated as follows [58]:

$$\mathbf{NPV = \sum_{n=0}^{20} (Discounted\ cashflow)_n} \quad (5.19)$$

$$\mathbf{NPV = -22.68\ Mil.\ \$}$$

where n is the project life. In order to compare the suggested approach with the traditional power generation process, the LCOE is finally calculated. The following is the equation for LCOE:

$$\mathbf{LCOE = \frac{NPV\ of\ total\ cost\ over\ life\ time}{NPV\ of\ electrical\ energy\ produced\ over\ life\ time}} \quad (5.20)$$

$$LCOE = \frac{\sum_{n=0}^{20} \frac{\text{Annual cost}}{(1-r)}}{\sum_{n=0}^{20} \frac{\text{Electricity produced}}{(1-r)}} = 211.34 \text{ \$/MW}$$

where r is the discount rate.

The cost estimations used in this analysis were obtained using the Aspen Economic Analyzer, which provides preliminary capital and operational cost data based on standard design parameters and built-in equipment cost databases. While Aspen Economic Analyzer is a widely used tool for estimating costs in process engineering projects, it is important to acknowledge that its cost estimation accuracy can vary and is generally better suited for conceptual or comparative studies rather than for detailed budgeting or final investment decisions. The estimates provided are approximate and are influenced by factors such as regional cost indices, equipment scaling, and default cost models within the software.

In this context, the CAPEX values reported in this study are primarily intended for comparative purposes. The focus is on understanding the relative economic performance of different cases rather than obtaining absolute cost figures. However, the primary objective remains the comparison of relative costs across different process configurations, where the Aspen Economic Analyzer's estimates are sufficiently consistent and appropriate for drawing meaningful conclusions.

5.5 Impact of Key Process Parameters on Process Outputs

Subsequently, sensitivity analysis was performed for key process parameters to evaluate their impact on important output parameters like syngas quality, captured carbon and power generation. A suitable range of deviation in these parameters is selected to conduct the analysis forming comprehensive trend to provide process performance over a varying

parameter. These parameters help to narrow down the efforts in the identified hotspots of the process. Thus, further improving the process in terms of energy or material consumption, efficiency, economics and GHG emissions. Using sensitivity analysis, the processes can be optimized. Specifically, two parameters are selected for the sensitivity analysis including methane input and gasification temperature.

5.6 Pinch Analysis

In relation to energy integration, pinch analysis offers a wonderful way to lower a process's energy requirements through process-to-process heat transfer, significantly lowering utility costs. The same technique is used to design a more economically feasible chemical production processes and construct a heat exchanger network diagram. The study aims to perform a pinch analysis of the proposed processes, estimate the minimum heating and cooling requirements, and then design a heat exchanger network.

There are two methods to perform the pinch analysis, i.e., (1) graphical and (2) problem table method. But to carry out the analysis manually can be complex and time-consuming job. Therefore, professional software are available which can perform comprehensive analysis with some basic input and software knowledge. A description of the process is provided in this section. The data required for the analysis were (1) hot and cold streams in the process, (2) their inlet and outlet temperatures, and (3) their mass heat capacities which are taken from the process simulation performed.

The steps for performing pinch analysis in aspen energy analyzer are as follows. First, the minimum temperature difference between hot and cold streams is specified, i.e., $\Delta T_{min} = 10^{\circ}\text{C}$. Then, the inlet and outlet temperature of all the hot and cold stream of the process

are inserted. After that, the specific heat capacity of each stream to calculate the heat duty for each stream are entered. Subsequently, the required and suitable utilities to satisfy the heating and cooling requirements are specified. Now, the software will be able to perform the calculations and the minimum heating utility will be shown as the non-overlap on the right, and the minimum cooling utility is the non-overlap on the left.

5.6.1 Heat Exchanger Network (HEN)

A comprehensive approach for the HEN is presented in Figure 22. Same rules for HEN are described as follows. First, the external heating and cooling should only be used above and below the pinch, respectively. For the below pinch logical diagram, the hot stream's C_p should be higher than the cold stream's, and vice versa. If not, the cold stream has to be divided so that they are below the pinch and above the pinch, satisfying the criterion.

For below the pinch, there should be more hot streams than cold streams, and vice versa. If not, divide the hot streams so that they are below the pinch and vice versa, satisfying the requirement.

The minimum temperature differential between the hot and cold streams for each match should be equivalent to $\Delta T_{min} = 10\text{ }^{\circ}\text{C}$.

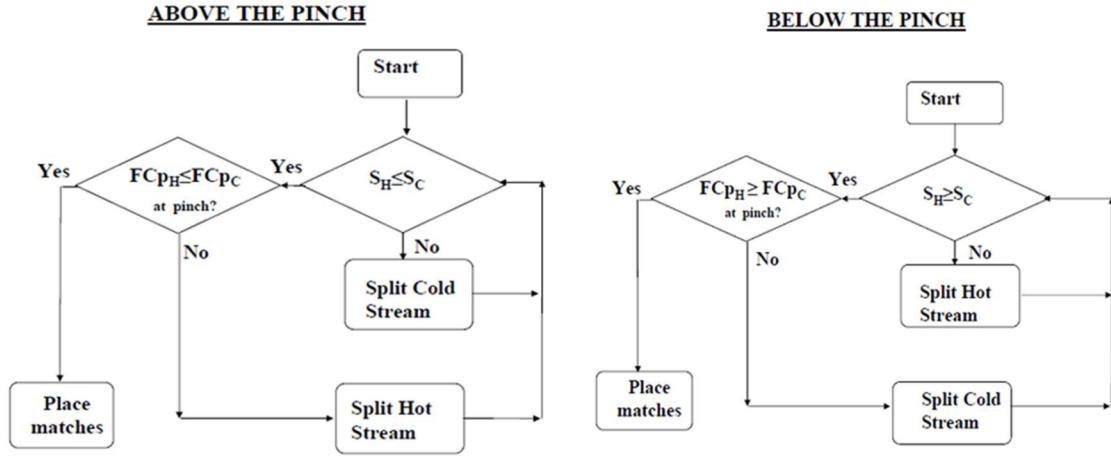


Figure 22 Design approach for heat exchanger network [60]

5.6.2 Heat Transfer Area

For each heat exchanger marked in HEN, calculate the enthalpy, inlet, and outlet temperatures of hot and cold temperatures, and log mean temperature and area using the following formulas:

$$\Delta T_{lm} = \frac{(Th1 - Tc2) - (Th2 - Tc1)}{\ln \frac{(Th1 - Tc2)}{(Th2 - Tc1)}} \quad (5.21)$$

$$A = \frac{Q}{U * \Delta T_{lm}} \quad (5.22)$$

Add areas for all the heat exchangers to calculate the total area of heat transfer required.

CHAPTER 6

RESULTS AND DISCUSSION

6.1 Material and Energy Balances

The simulations conducted on Aspen Plus yielded a comprehensive material and energy balances for each case, as indicated in Table 14:

Table 14 Operating parameters and stream compositions at the exit of key unit operations

	Plastic Feed	Steam to Gasifier	Gasifier Outlet	Methane to Reformer	Reformer Outlet				Gas Turbine Outlet (Flue Gas)			
	Case 1-4	Case 1-4	Case 1-4	Case 1-4	Case 1	Case 2	Case 3	Case 4	Case 1	Case 2	Case 3	Case 4
T (°C)	25	300	1300	25	900	900	700	700	756	728	758	773
P (bar)	1	30	25	3	3	3	25	25	1	1	1	1
Mass Flow (kg/hr)	1000	1250	2250	300	850	850	604	1830	7633	7633	7080	8610
Mole (%)												
PE	50	-	-	-	-	-	-	-	-	-	-	-
PP	50	-	-	-	-	-	-	-	-	-	-	-
H ₂	-	-	63.3	-	67.2	67.2	66.5	38.6	-	-	-	-
CO	-	-	34.1	-	19.1	19.1	33.2	50.2	-	-	-	-
CO ₂	-	-	0.11	-	2.4	2.4	-	-	29.4	29.4	34.5	41.6
H ₂ O	-	100	0.57	-	11.2	11.2	-	-	68.2	68.2	65.4	57.9
O ₂	-	-	-	-	-	-	0.26	11.1	2.3	2.3	0.06	0.3
CH ₄	-	-	1.84	100	-	-	-	-	-	-	-	-
	Steam Turbine Outlet (Steam)				CO ₂ Purification				Methane Feed to HDS			
	Case 1	Case 2	Case 3	Case 4	Case 1	Case 2	Case 3	Case 4	Case 1-4			
T (°C)	49	49	49	49	30	30	30	30	25			
P (bar)	0.1	0.1	0.1	0.1	30	30	30	30	1			
Mass Flow (kg/hr)	3070	2870	2870	3170	3996	3996	3992	3999	315			
Mole (%)												
PE	-	-	-	-	-	-	-	-	-			
PP	-	-	-	-	-	-	-	-	-			
H ₂	-	-	-	-	-	-	-	-	-			
S	-	-	-	-	-	-	-	-	5			
CO ₂	-	-	-	-	99.5	99.5	99.5	99.5	-			
H ₂ O	100	100	100	100	0.3	0.3	0.3	0.3	-			
O ₂	-	-	-	-	0.2	0.2	0.2	0.2	-			
CH ₄	-	-	-	-	-	-	-	-	95			

To create equal comparison grounds, the same amounts of waste plastic (1000 kg/hr) and natural gas (300 kg/hr) were fed to the gasifier and reformers, respectively. Additionally, the ratios of steam to plastic (1.25) and steam to methane (1.83) were maintained throughout all instances for both procedures. Constant syngas production from the gasifier

was 2,250 kg/hr with an HCR of 1.86. Whereas, the syngas production from the reformers varied due to the utilization of different reforming techniques. While cases 1 and 2 used the identical SMR units and produced syngas at a rate of 850 kg/hr with an HCR of 3.5, case 3 produced the least quantity of syngas because it used oxygen as an oxidizing agent during the POX process. The use of oxygen in POX requires less amount of oxidizing agent as compared to that of steam in SMR which resulted in less output. Lower requirement of oxidizing agent reduced the utilities consumption by 27.4% and 11.7% as compared to case 1 and case 2, respectively, but this saving came at the cost of lower HCR of 2.0. Finally, case 4, utilizing DMR integrated with POX, generated 1,830 kg/hr of syngas having least HCR of 0.8.

The decreased HCR of the gasifier outlet stream, when paired with the syngas generated by the gasifier, resulted in a reduction of the HCR for every scenario. Interestingly, case 3 having the lowest HCR was estimated to have the highest HHV of 26.1 MJ/kg. Upon investigation it was identified that the case 3 has the highest concentration of combustibles (99.2%) which increased the HHV among all cases.

For the detailed insight of the energy requirement of the process, all utility requirements of each process are described in Table 15. Overall, case 1 exhibits the largest utility requirement of 11.92MW with 60% contribution of heating utility. Whereas the case 2 and 4 has approximately the same utility requirement of 10.4MW followed by the case 3 having the least requirement of 8.93MW. Among all, case 2 has the highest heating requirement of 70% which can lead to the highest utility cost due to the costly nature of heating utility. But the exact evaluation is performed in the economic analysis.

Table 15 Detailed summary of utility requirements for the equipment used in each case without heat integration (all values are in MW)

	Case 1			Case 2			Case 3			Case 4		
	Cool	Heat	Power	Cool	Heat	Power	Cool	Heat	Power	Cool	Heat	Power
PP heater	-	0.09	-	-	0.09	-	-	0.09	-	-	0.09	-
PE heater	-	0.09	-	-	0.09	-	-	0.09	-	-	0.09	-
PP Crusher	-	-	0.01	-	-	0.01	-	-	0.01	-	-	0.01
PE Crusher	-	-	0.015	-	-	0.015	-	-	0.015	-	-	0.015
Gasifier Feed water heater	-	1.06	-	-	1.06	-	-	1.06	-	-	1.06	-
Pyrolysis + Gasification	-	4.74	-	-	4.74	-	-	4.74	-	-	4.74	-
Methane compressor	-	-	0.025	-	-	0.025	-	-	0.06	-	-	0.06
Syngas Compressor	-	-	0.3	-	-	0.3	-	-	-	-	-	-
POX	-	-	-	-	-	-	0.68	-	-	2.11	-	-
SMR feed heater	-	-	-	-	-	-	-	-	-	-	-	-
ASU	1.76	-	-	-	-	-	-	-	-	-	-	-
DMR	-	-	-	-	-	-	-	-	-	-	0.10	-
Air reactor	-	-	-	-	0.16	-	-	0.16	-	-	0.16	-
HDR pre-heater	-	0.09	-	-	0.09	-	-	0.09	-	-	0.09	-
SMR	-	1.16	-	-	1.16	-	-	-	-	-	-	-
GT + HRSG	-	-	-	-	-	-	-	-	-	-	-	-
Steam turbine	-	-	-	-	-	-	-	-	-	-	-	-
CO ₂ cooler	2.49	-	-	2.51	-	-	2.10	-	-	2.13	-	-
CO ₂ separator	-	-	-	-	-	-	-	-	-	-	-	-
CO ₂ Purification (Compression and dehydration)	0.26	-	0.22	0.26	-	0.27	0.32	-	0.27	0.26	-	0.23
Total	4.51	7.22	0.57	2.77	7.49	0.62	3.10	5.59	0.355	4.51	5.69	0.315

6.2 Model Validation

Model validation of the designed processes is essential to check the authenticity of the process which is performed by checking the used process parameters and resultant outlet parameters against the published literature. For that, such parameters are validated for all the main unit operations and processes employed in the study as shown in Table 16 for reforming. The gasification model developed shows good correspondence with the literature except being utilizing a slightly lower temperature of 1300 for the gasification.

Whereas among the reforming technologies, SMR had the best resemblance with the published literature but employed a high-pressure reforming rather than a lower pressure due to lack of data. Whereas for the POX, the operating conditions were in line with the published values but the syngas results varied slightly from our results. Lastly, the DMR is validated with literature data and found a good agreement with simulated data.

Table 16 Process model validation for gasification and reforming

		Gasification			SMR			
		Case 1-4	Ref [49]	%Diff	Case 1	Case 2	Ref [50]	%Diff
	T (°C)	1300	1500	-13.3	900	900	894	0.67
	P (bar)	25	25	0	3	3	3	0
Mole Composition (%)	H ₂	63.3	64.14	-1.3	67.2	67.2	68.3	-1.6
	CO	34.1	34.37	-0.78	19.1	19.1	20.6	-7.3
	CO ₂	0.11	0.02	450	2.4	2.4	2.0	250
	H ₂ O	0.57	0.13	338.4	11.2	11.2	8.9	25.8
	O ₂	0	0.00	0.00	0	0	0	0
	CH ₄	1.84	1.25	47.2	0	0	0.1	-100
		POX			DMR			
		Case 3	Ref [51]	%Diff	Case 4	Ref [11]	%Diff	
	T (°C)	700	700	0	800	1000	-20	
	P (bar)	25	25	0	25	30	-16.6	
Mole Composition (%)	H ₂	66.5	60.3	10.3	38.6	35.3	9.3	
	CO	33.2	29.7	11.8	50.2	55	-8.7	
	CO ₂	0	3.2	-100	0	0	0	
	H ₂ O	0	5.5	-100	0	0	0	
	O ₂	0.26	0	100	11.1	9	23.3	
	CH ₄	0	1.3	-100	0	0.2	-100	

6.3 Process Performance

Technical analysis and process simulation are used to determine key performance characteristics, as shown in Table 17. The syngas and its HCR were the same in cases 1 and 2, as they both used the same input flows and syngas generation processes. The difference was in the way the oxygen was burned; while 5 mol% more oxygen was used in both scenarios; the resulting flue gases were produced at different temperatures. When oxygen is used through ASU in case 1, an outlet stream with a temperature of 1500°C is produced; in case 2, the temperature is marginally lower at 1454°C. The presence of metal in the CLC, which absorbs some combustion heat during the process, is most likely the cause of this discrepancy.

Table 17 Summary of key process parameters obtained from process simulation and technical analysis (HI = heat integration)

Process parameters	Case 1	Case 2	Case 3	Case 4
Syngas production (kg/hr)	3,097	3,097	2,851	3,756
Syngas HCR	2.2	2.2	1.9	1.7
Syngas HHV (MJ/kg)	25.9	25.9	26.1	20.5
Combustion temperature (°C)	1500	1454	1487	1487
Minimum utility required (MW)	7.915	6.115	4.915	6.415
Power production (MW)	3.78	3.65	3.15	3.66
Power production (MW) after HI	3.78	4.46	3.96	4.22
Process Efficiency (%)	16.6	17.45	15.97	17.3
Process Efficiency (%) after HI	16.6	21.3	20.0	19.9
CO ₂ captured (kg/hr)	3,996	3,996	3,992	3,988

To achieve full combustion, the combined syngas was burned in the gas turbine with 5 mol% extra oxygen. Once again, the output of the gas turbine was not straight forward or depending upon any single factor. Case 1 generated the highest power of 3.78 MW whereas, the power output of case 2 and case 4 were almost 4.9% lower than earlier. To investigate the underlying reason, power output was plotted against the syngas flow and HCR as shown in Figure 23. The results showed that the power output didn't vary along a single factor, rather it was a result of combination of both the factors. Results showed that syngas flow and HCR both decreases for case 3, thus decreasing the overall power production. Conversely, the two parameters increase simultaneously for case 2 thus increasing the power output. Whereas for the case 4, although the flow rate of the syngas increases but it provides lower HCR. The combined effect of these two parameters keeps the power output in between the cases 1 and 3. However, due to the heat integration of high temperature nitrogen stream released from the CLC, provided opportunity to produce extra steam which resulted in higher power production thus increasing the overall process efficiency for case 2-4.

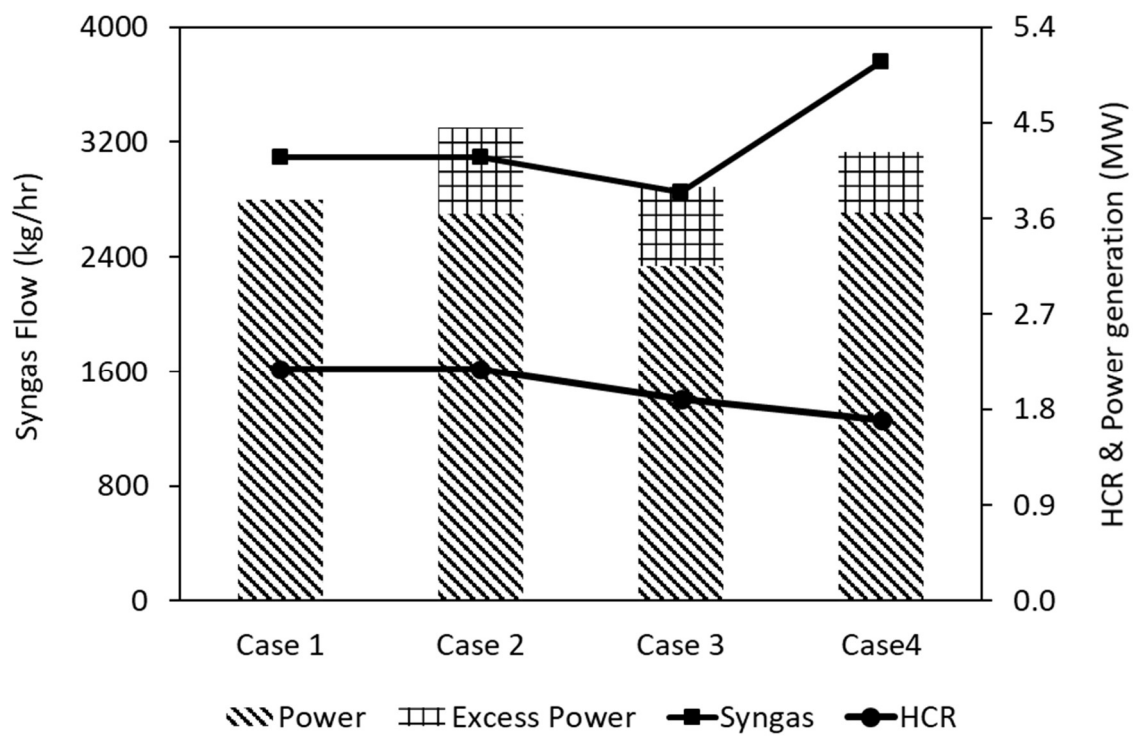


Figure 23 Co-relation of power generation with syngas and H₂/CO (HCR) for the developed cases

Furthermore, because the two processes' methods for reforming and providing oxygen differed, so did their utility requirements. In this case, the process's power production has a greater influence on process efficiency than utility requirements do. Figure 24 illustrates how the utility requirement for Case 3 is much lower than for Case 1, yet the process's efficiency still declines as a result of the process's lower power output. Although the power output decline is far smaller than the utility drop, the efficiency is still in line with the previous pattern, indicating that power output is what drives efficiency. On the other hand, case 1 had the greatest utility requirement among all at 7.9 MW due to utilizing ASU, which is an energy-intensive unit. However, cases 2-4, which made use of CLC, used 23.1% less power to function. Because of the interaction of these variables, case 1 produced the most electricity; nonetheless, cases 2 and 4 demonstrated efficiency increase of 4.2% and 5.4%, respectively. Whereas the efficiency of the case 3 was affected by the lower power production and syngas flow, least among all the cases and resulted in 3.6% lower efficiency as compared to that of case 1. Similar to earlier discussion, the excess steam and power generation as a results of heat integration further improved the process efficiency for case 2-4.

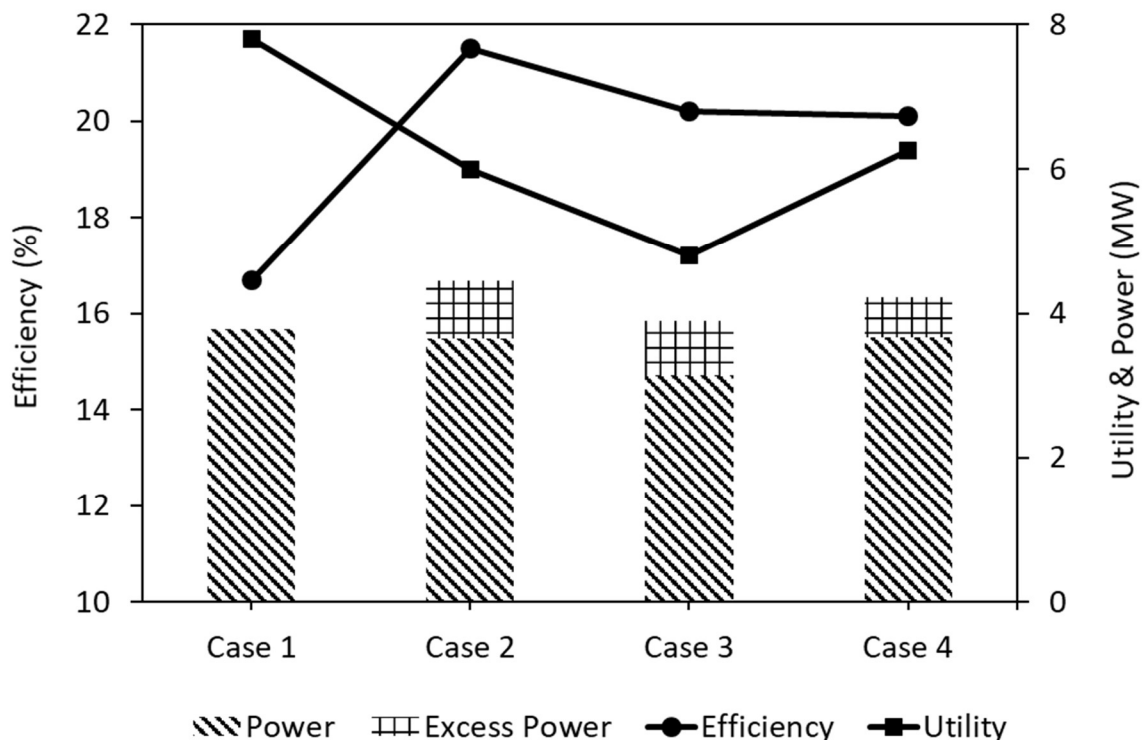


Figure 24 The relationship between process efficiency and power generation and utility requirements for the given cases

6.4 Impact of Key Process Parameters on Process Outputs

6.4.1 Impact of Methane Flow Rate on Power Generation and Captured/Uncaptured CO₂

First, a sensitivity analysis was done to evaluate the power output versus CH₄ input. For all cases, Figure 25 illustrates a linear increase in power generation as the CH₄ input in SMR increases. According to the linear trend, process economics and efficiency should not change as capacity increases. The steepest trend, however, is shown by case 2, which begins with a lower power generation output but eventually reaches case 1's capacity at a 500 kg/hr CH₄ intake. This suggests that if methane input increases, the efficiency gap between the scenarios would get wider. Whereas the slope of case 3 and 4 are similar, and least steep among the cases showing lower dependency of power output for the cases on methane input.

Furthermore, the impact of increasing CH₄ flow is also examined for the captured and uncaptured CO₂. All cases demonstrated same flow for the captured and uncaptured CO₂, therefore a single result is shown for all cases in Figure 26. The result shows linear growth in emissions with increasing input. Interestingly, again the two parameters tend to converge with the increasing CH₄, as in previous sensitivity. This shows that not only the individual flow of captured CO₂ increases with the increasing capacity but its relative contribution increases as well, implying that operating at a higher capacity favors the process in terms of environmental sustainability.

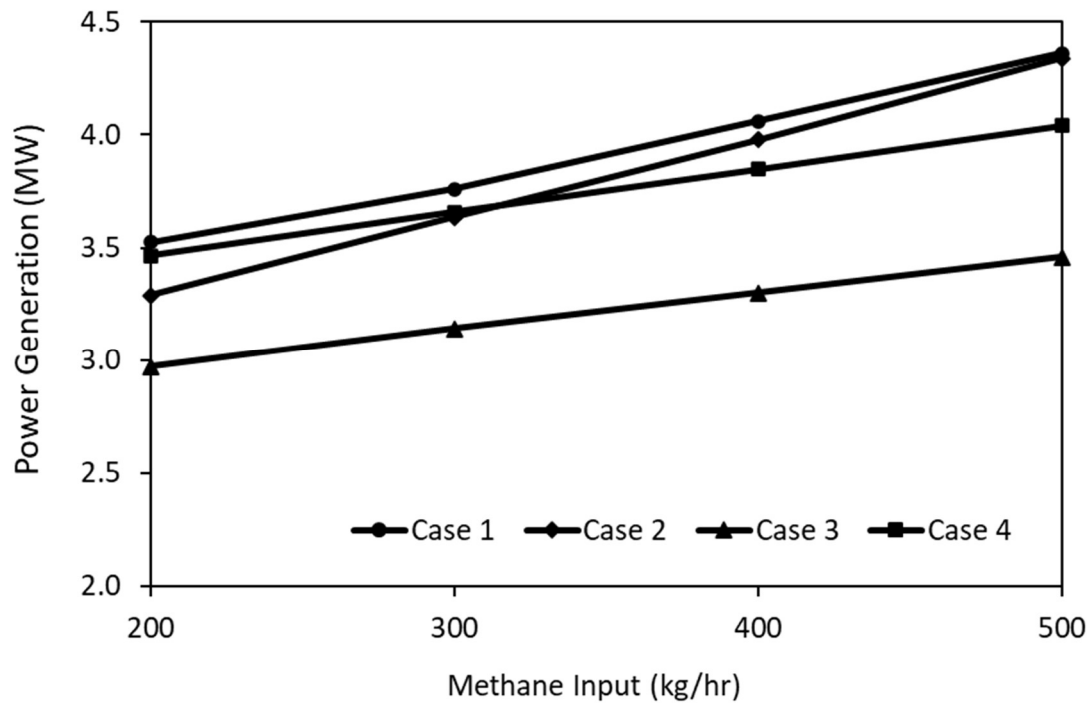


Figure 25 Impact of methane input on power generation

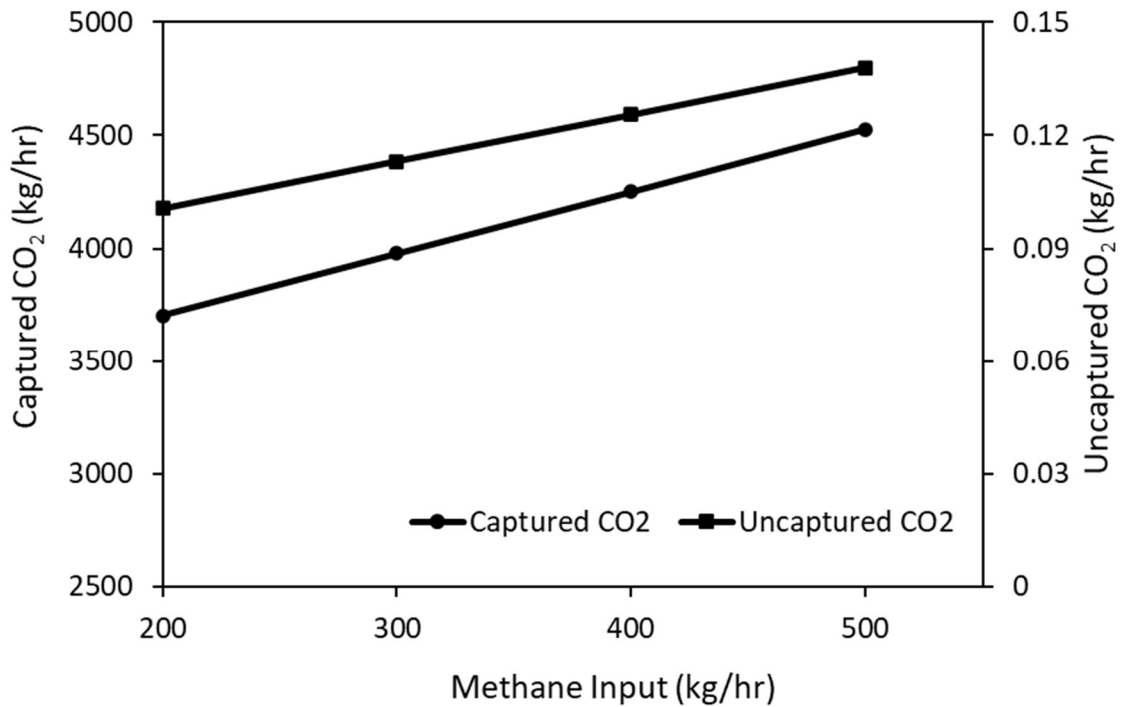


Figure 26 Impact of methane input on captured and uncaptured CO₂

6.4.2 Impact of Gasification Temperature on Power Generation and Syngas Composition

The effect on electricity generation was then noted in relation to the change in gasifier operating temperature. Figure 27 illustrates that the total power production remains constant for all cases when the temperature of the gasifier increases. This is explained by the fact that the total amount of combustible components remains constant even when the temperature rises and the CH₄ produced during gasification is transformed into H₂ and CO. As a result, the energy generated when these combustibles burn remains constant. However, it is crucial to note that the form of combustibles (either H₂ or CH₄) is not impacting the material and energy balance but retains significance in design phase when the pipes, equipment, burners etc. are specifically constructed for a specific chemical

component. As when a component is added for which the apparatus was not made, there will be risks and problems during operation.

The impact of gasifier temperature on the composition of syngas was then assessed. Figure 28 illustrates how the required components, H_2 and CO , in the syngas coming out of the gasifier increase exponentially until they reach about $1300^{\circ}C$, and vice versa for CH_4 and CO_2 . The trend stabilizes beyond $1300^{\circ}C$, highlighting the fact that raising the gasifier temperature further will not increase the desirable components. Once more, the findings were the same in every case, hence the graph only displays one outcome.

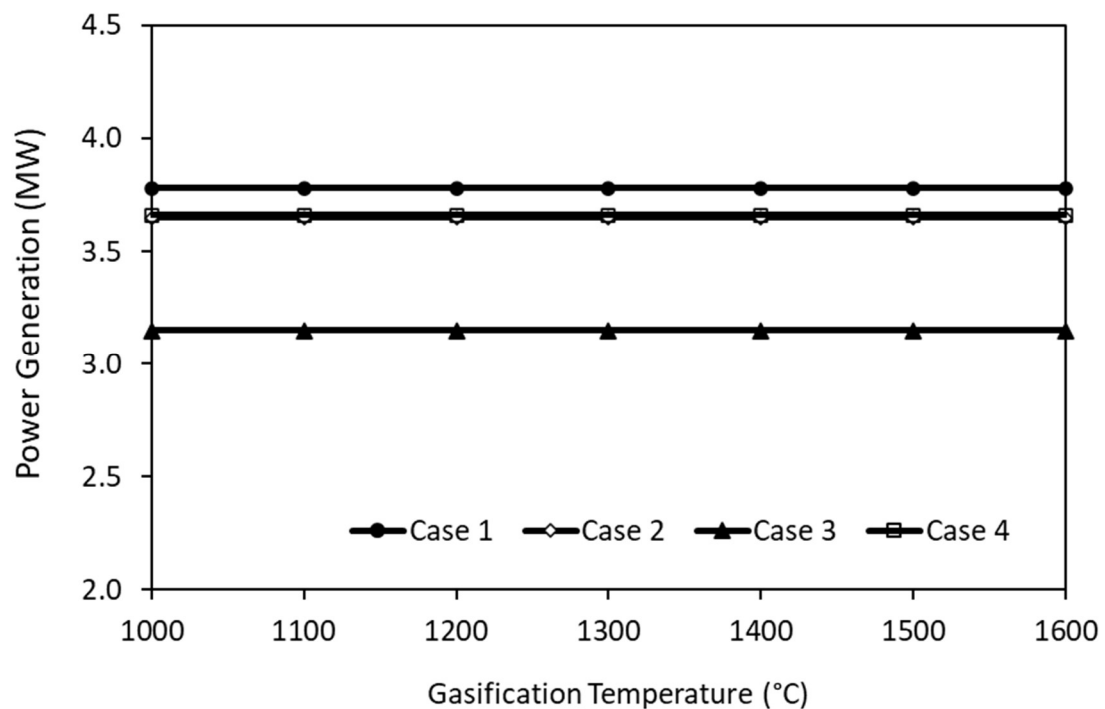


Figure 27 Impact of gasification temperature on power generation

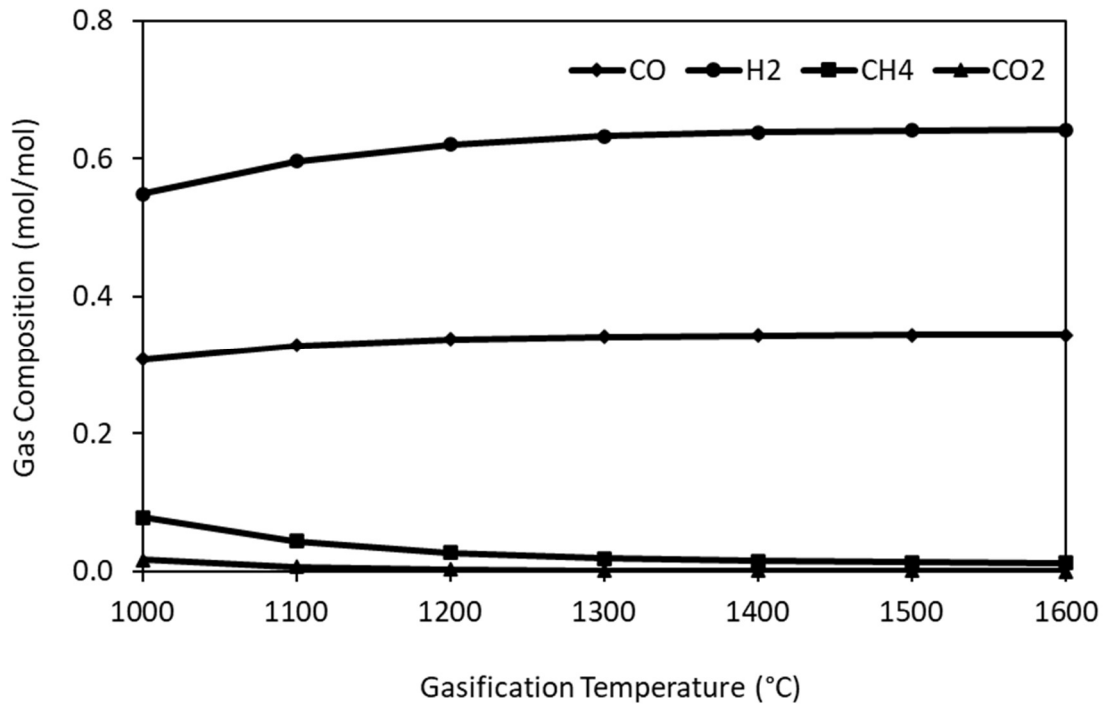


Figure 28 Impact of gasification temperature on syngas composition

6.5 Pinch Analysis Results

Pinch analysis provides a minimum heating and cooling utilities required against which the operating cost for the facility was calculated after heat integration. The total heat transfer area was also calculated based on which capital investment was calculated. Table 18 presents the results of the pinch analysis. Surprisingly only case 1 provided the pinch temperature whereas the other cases didn't. The reason that was found was that the pinch temperature is the point in heat exchanger network where the temperature difference is the smallest or where the two curves tend to converge. But in the given cases, the cold and hot utility curves did not tend to converge, due to which the pinch temperature was not estimated by the Aspen energy analyzer as shown in **Error! Reference source not found.-**

32.

6.5.1 Composite Curves

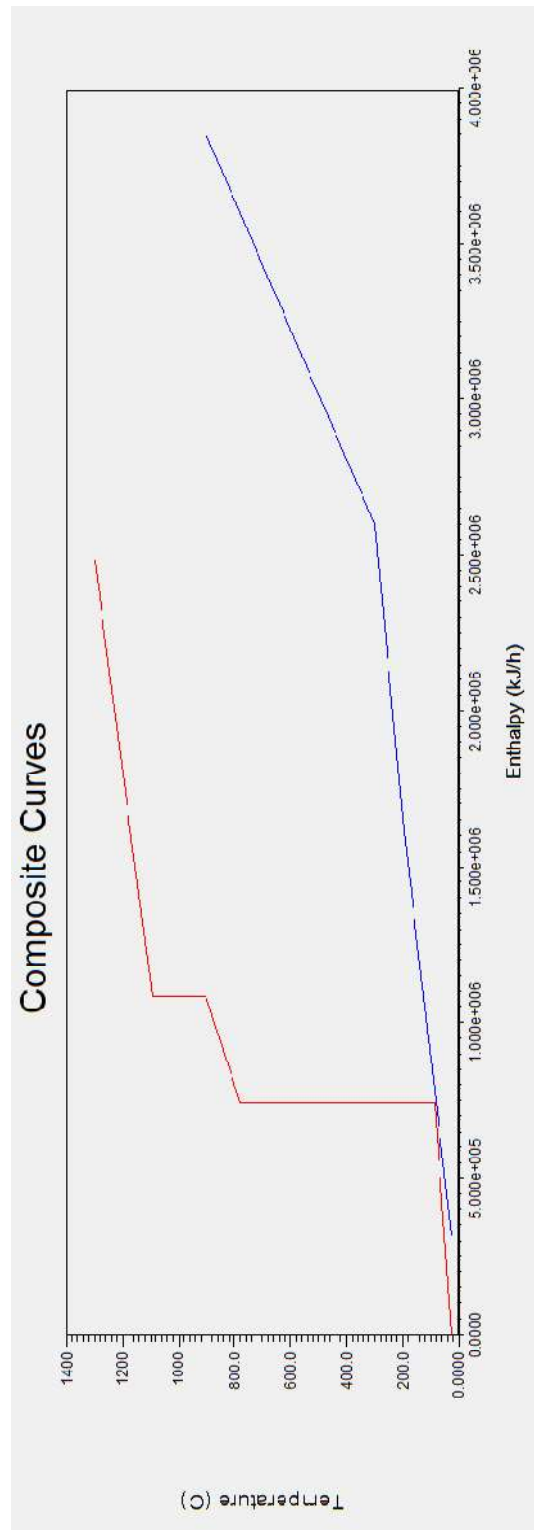


Figure 29 Figure 29 Composite curve for case 1 (base case)

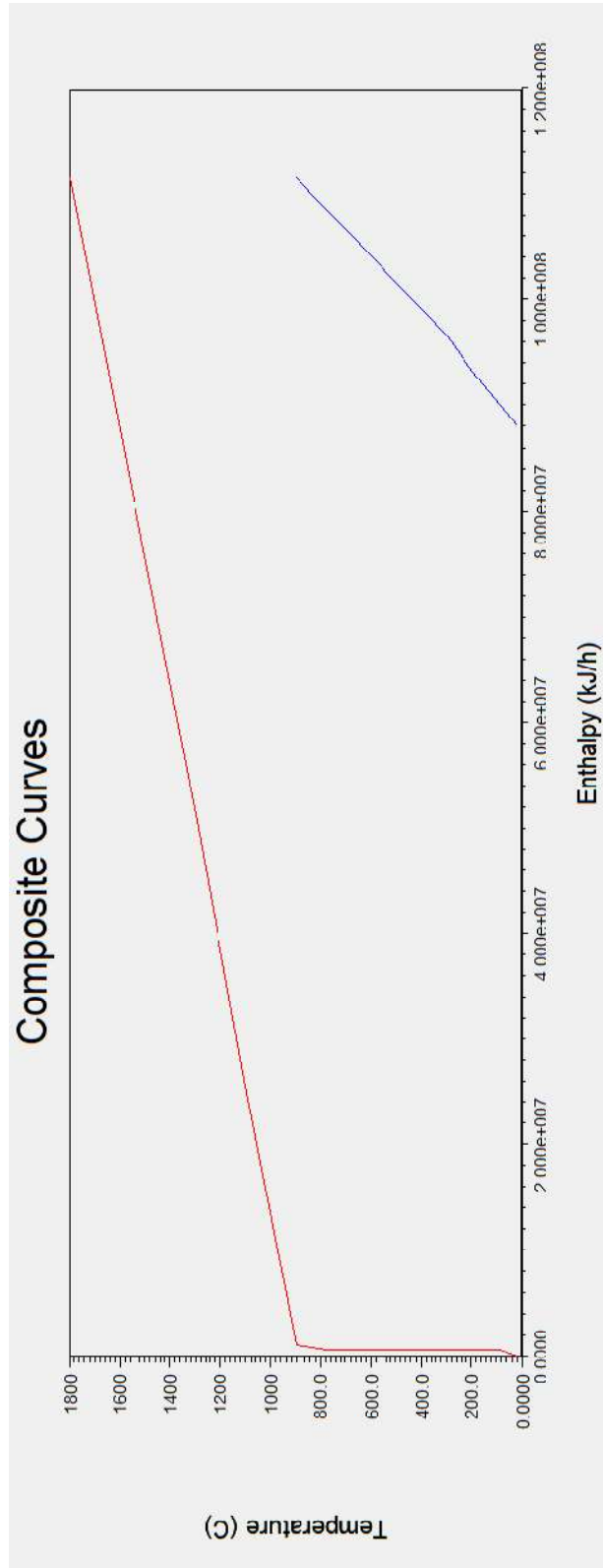


Figure 30 Composite curve for case 2

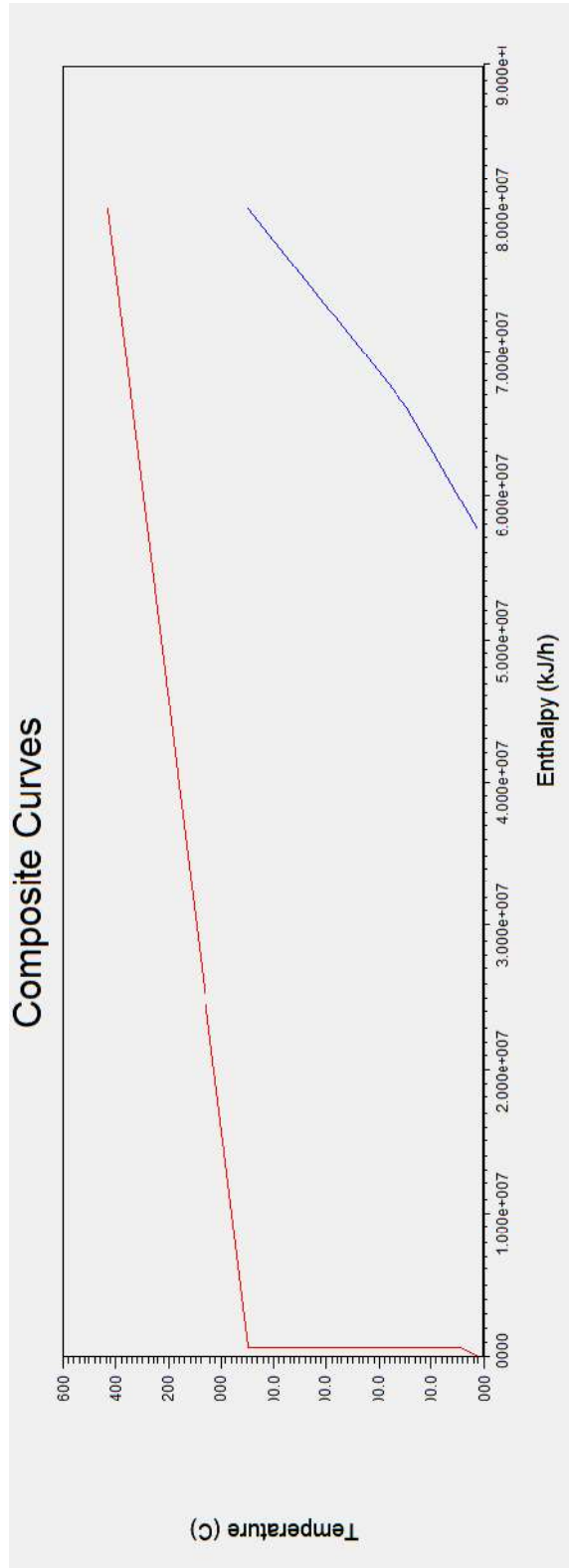


Figure 31 Composite curve for case 3

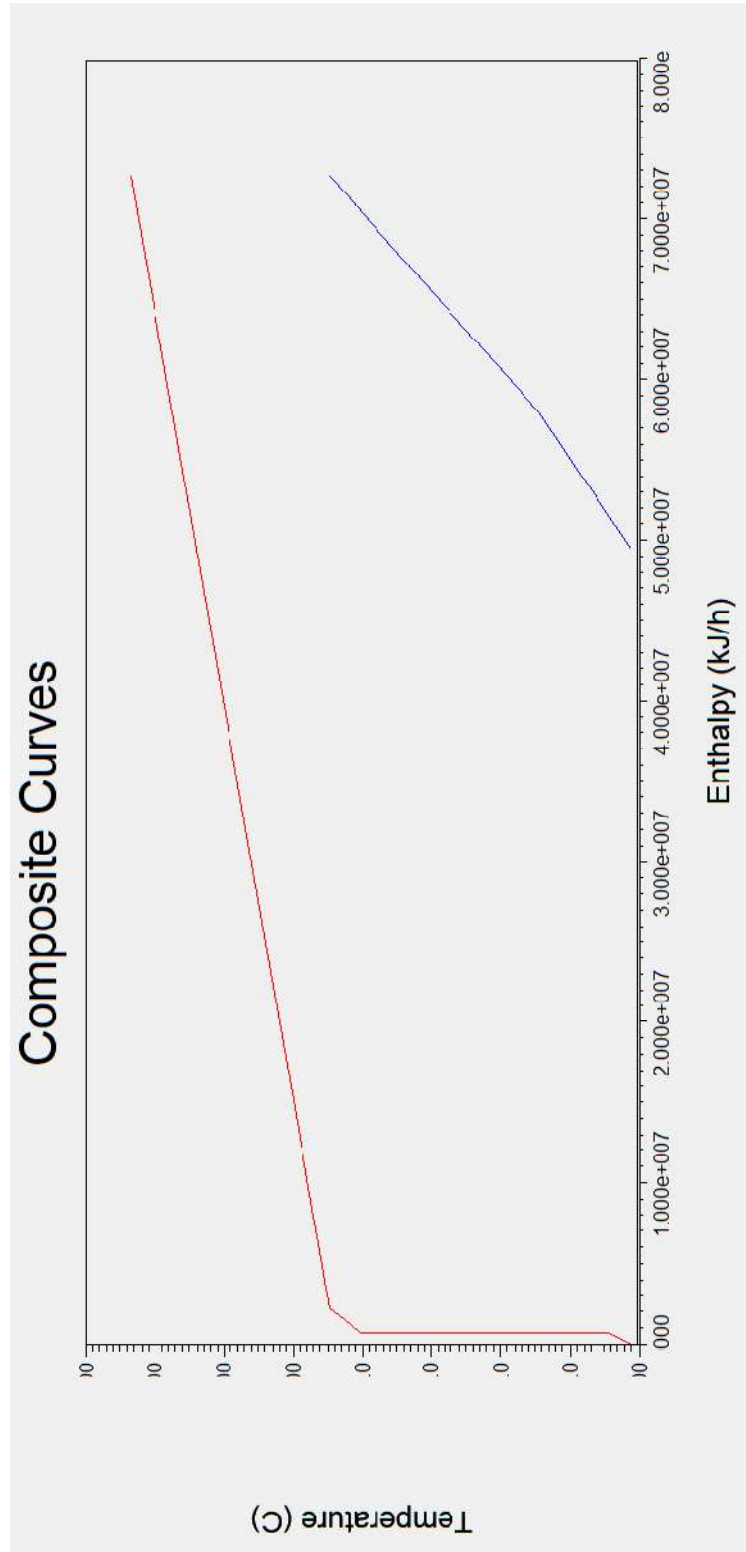


Figure 32 Composite curve for case 4

6.5.2 Heat Exchanger Network Results

Furthermore, cases 2-4 only possesses cold utility requirement to cool the hot metal to its inlet temperature of the air reactor. Such processes do not have the pinch point and regarded as the threshold problems. For such problems, pinch point can be introduced by adding multiple utilities such as HP and LP steam for heating and cooling water and steam generation for cooling purposes [61]. The highest cold utility requirement is for the case 2 resulting in operating expense of \$0.15 mil/yr. As suggested by the composite curves Figure 33-36, there is a considerable recoverable heat which can be recovered by adding a second cooling utility in the form of HP steam generation.

Figure 41-43, there is a considerable recoverable heat which can be recovered by adding a second cooling utility in the form of HP steam generation.

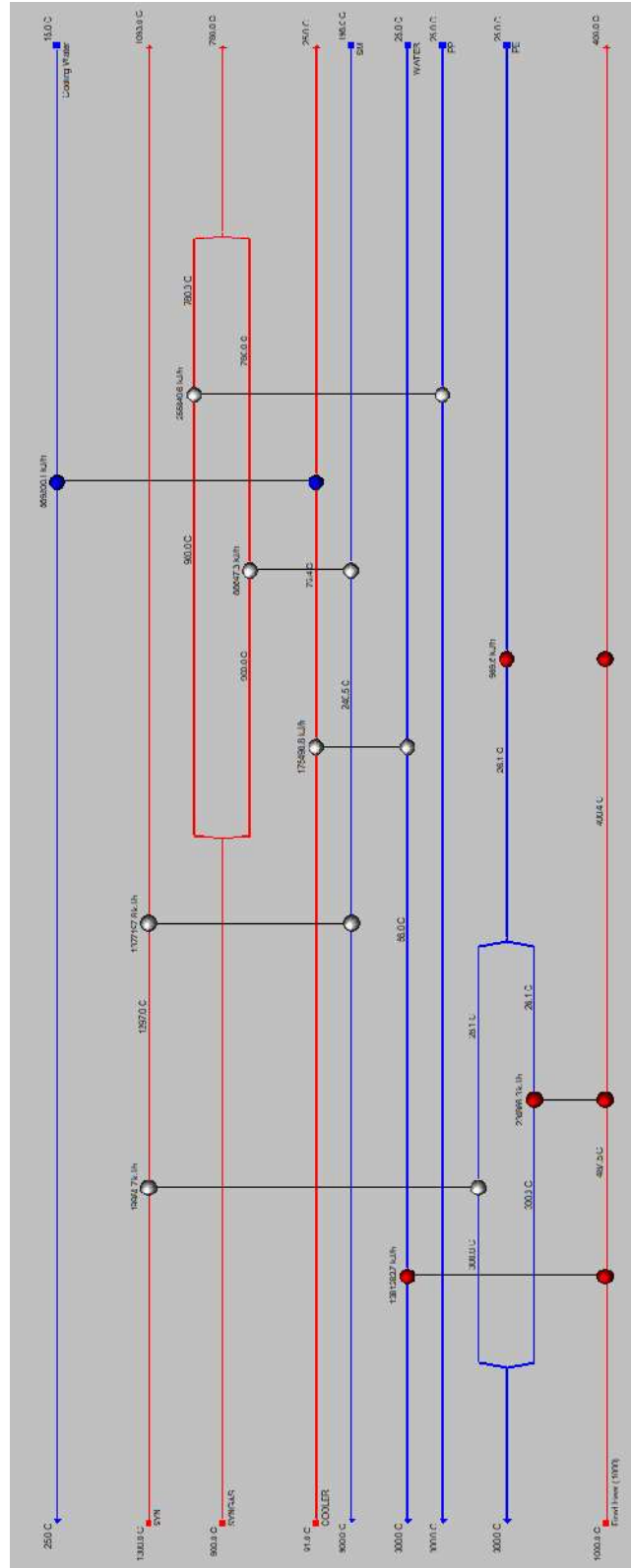


Figure 33 Heat exchanger network diagram for case 1 (base case)

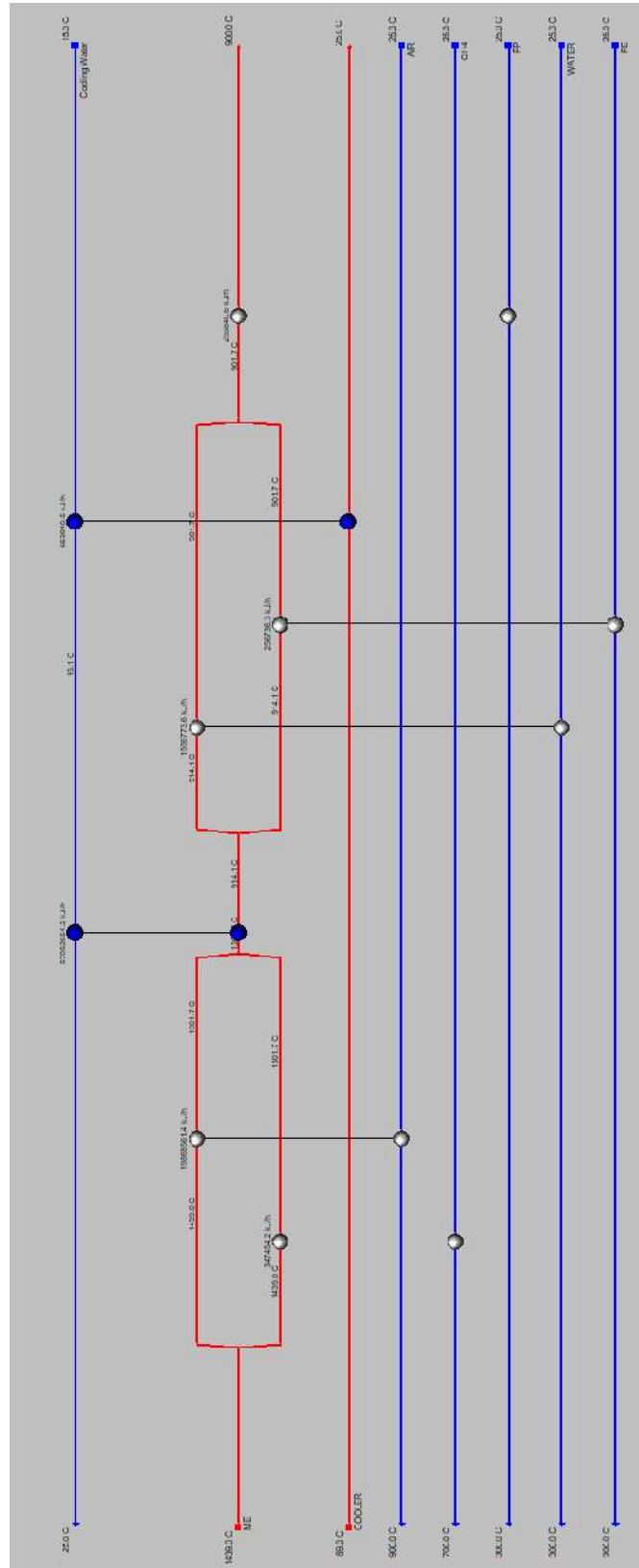


Figure 35 Heat exchanger network diagram for case 3

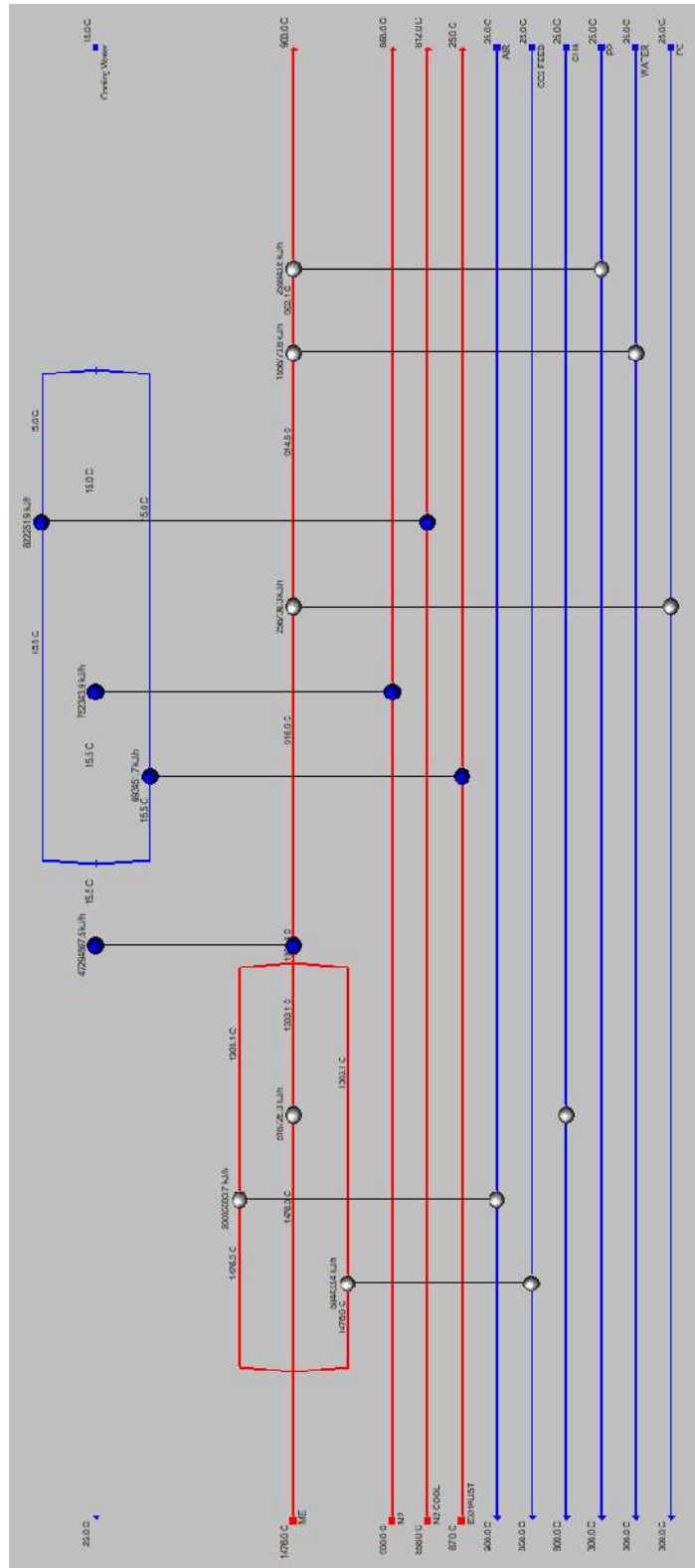


Figure 36 Heat exchanger network diagram for case 4

The pinch analysis results before and after heat integration are tabulated in Table 18.

Figures 37-39 illustrate the energy requirement, utility cost, and additional capital cost for four different cases before and after heat integration. Starting with energy requirements (Figure 37), case 1 shows a steady demand of 1694 MJ/hr both before and after heat integration, indicating no significant change in energy efficiency due to limited or no heat recovery potential. In contrast, case 2 exhibits a substantial reduction from 88110 MJ/hr before heat integration to 6188 MJ/hr after integration. This marked improvement highlights the effective use of heat integration to enhance energy efficiency. Case 3 sees a decrease from 57720 MJ/hr to 5725 MJ/hr, but the reduction is modest, suggesting less effective heat recovery compared to case 2. Case 4 shows a reduction from 49560 MJ/hr to 4449 MJ/hr, achieving better energy optimization than cases 1 and 3 but not as pronounced as case 2.

In terms of utility costs (Figure 38), Case 1 experiences a decrease from 0.087 Mill \$/yr to 0.054 Mill \$/yr after heat integration. Although this reduction is beneficial, it is relatively modest. Case 2, however, demonstrates the most significant utility cost savings, decreasing from 0.21 Mill \$/yr to 0.12 Mill \$/yr due to highly effective heat integration. Case 3 reduces its utility cost from 0.097 Mill \$/yr to 0.09 Mill \$/yr, reflecting a smaller but still positive impact. Case 4's utility cost drops from 0.083 Mill \$/yr to 0.06 Mill \$/yr, indicating a good level of savings.

The additional capital cost graph (Figure 39) shows that case 1 incurs a small increase from 0.18 Mill \$ to 0.22 Mill \$, indicating a relatively minor investment for heat integration. On the other hand, case 2 requires the highest additional capital cost, increasing from 0.17 Mill \$ to 0.34 Mill \$, suggesting a substantial investment for the level of energy and utility cost

improvements achieved. Case 3 shows a moderate increase in capital costs from 0.15 Mill \$ to 0.28 Mill \$, while case 4 sees a rise from 0.17 Mill \$ to 0.29 Mill \$, demonstrating a more balanced increase in costs compared to the benefits gained.

Overall, case 2 offers the most substantial improvements in energy efficiency and utility cost reductions but comes with the highest additional capital costs, making it highly effective if the initial investment is justified by long-term gains. Case 4 presents a better balance between energy savings, utility cost reduction, and moderate additional capital costs, making it potentially the most well-rounded option. Cases 1 and 3 show limited improvements in energy and cost savings, making them less favorable compared to the other cases. This comprehensive comparison indicates that while case 2 excels in maximizing energy efficiency, case 4 offers a more cost-effective solution with good performance benefits.

Case 3 generates 3,022 kg/h of HP steam and 0.75 MW of extra electricity (Figure 40). While slightly lower than case 2, the difference is marginal, and the associated capital cost increase is less than that of case 2. The moderate capital investment, coupled with effective energy recovery, makes case 3 an attractive option, as it strikes a balance between energy recovery and cost-efficiency.

In conclusion, case 3 appears to offer the best balance between energy recovery and economic considerations, with high HP steam and electricity generation capabilities combined with a more manageable capital investment compared to case 2. While case 2 has the highest energy recovery, the cost may not be justified unless maximum recovery is a priority. Case 4 remains a good compromise but falls short of the energy recovery

potential of case 3. Therefore, based on the provided data, case 3 seems the most beneficial in terms of balancing energy recovery, cost savings, and investment considerations.

Table 18 Key results of energy requirements before and after heat integration

Parameters	Case 1		Case 2		Case 3		Case 4	
	Before HI	After HI	Before HI	After HI	Before HI	After HI	Before HI	After HI
Energy requirement (MJ/h)	1694	1694	88,110	6188	57,720	5725	49,560	4449
Operating cost (Mil. \$/Y)	0.087	0.054	0.21	0.12	0.097	0.09	0.083	0.06
Capital investment (Mil. \$)	0.18	0.22	0.17	0.34	0.15	0.28	0.17	0.29
HP steam generation (kg/h)	-	2900	-	3,272	-	3,022	-	2,266
Extra electricity generation (MW)	-	0.72	-	0.81	-	0.75	-	0.56

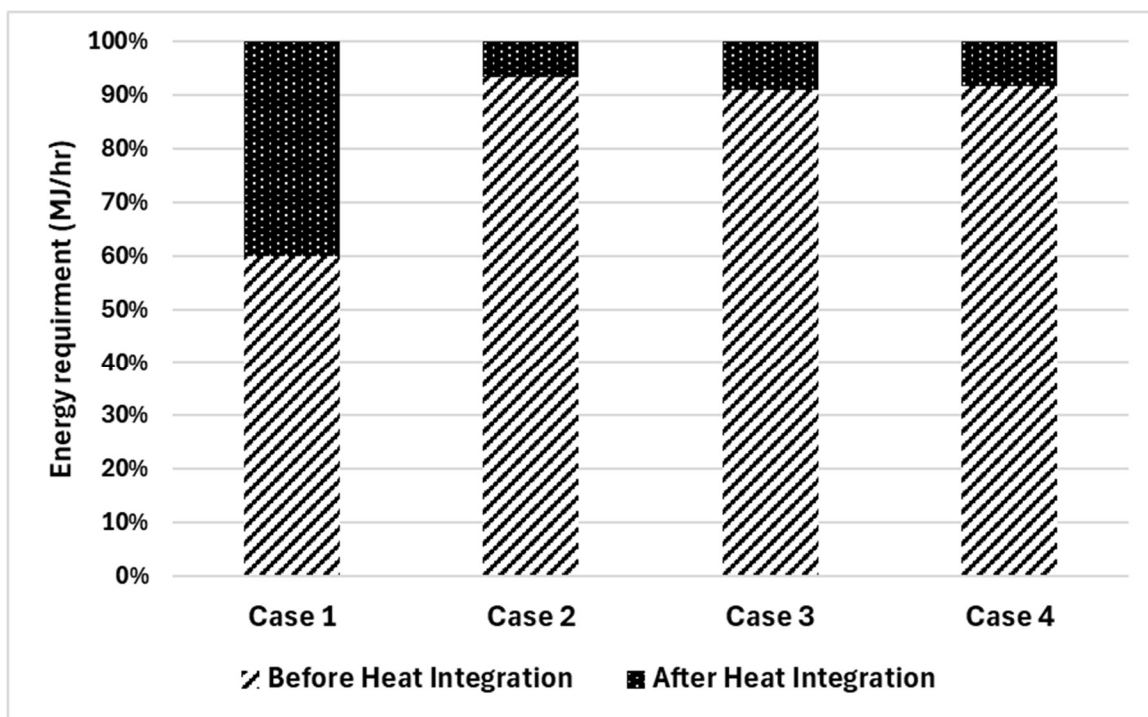


Figure 37 Comparison of energy requirements before and after heat integration.

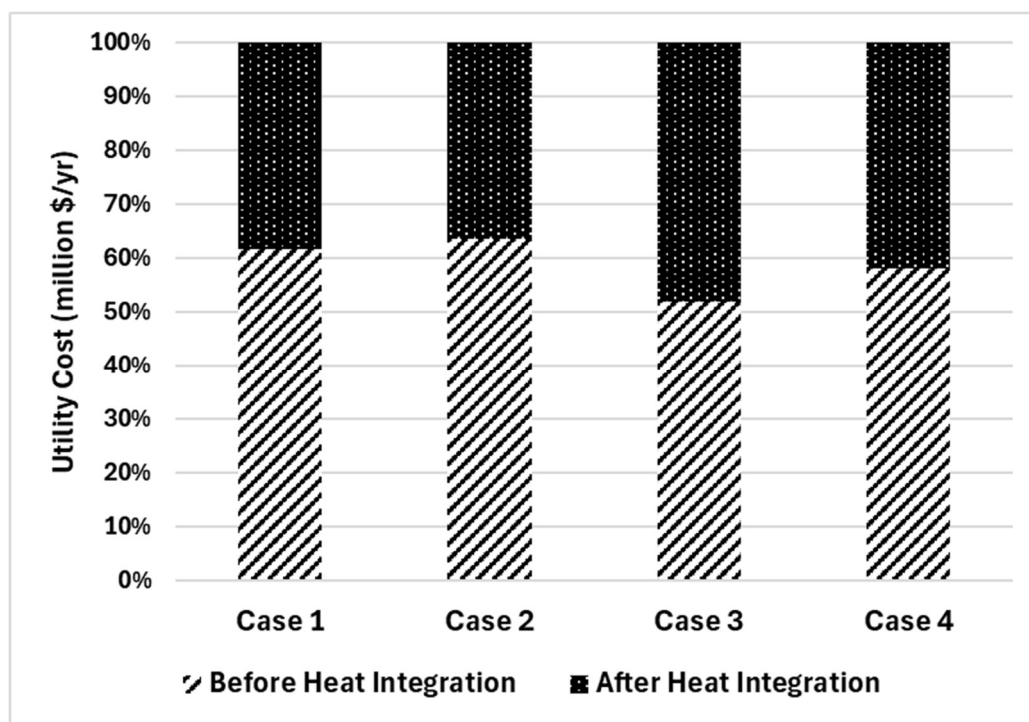


Figure 38 Comparison of utility cost before and after heat integration.

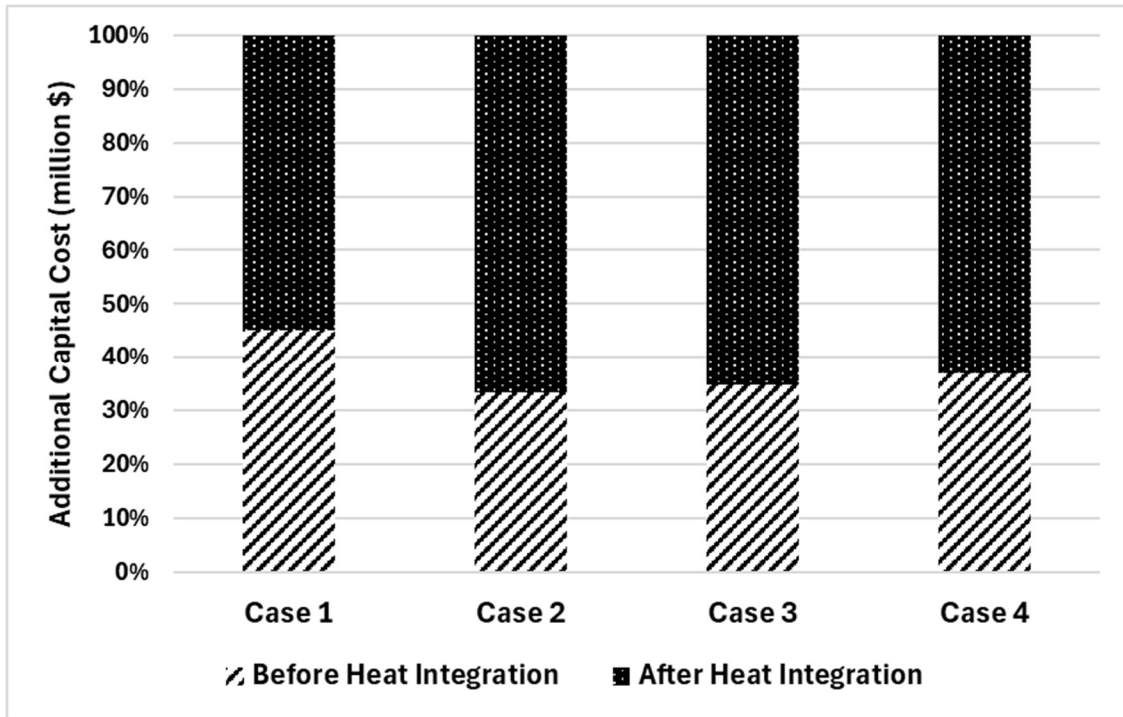


Figure 39 Comparison of additional capital cost requirements before and after heat integration.

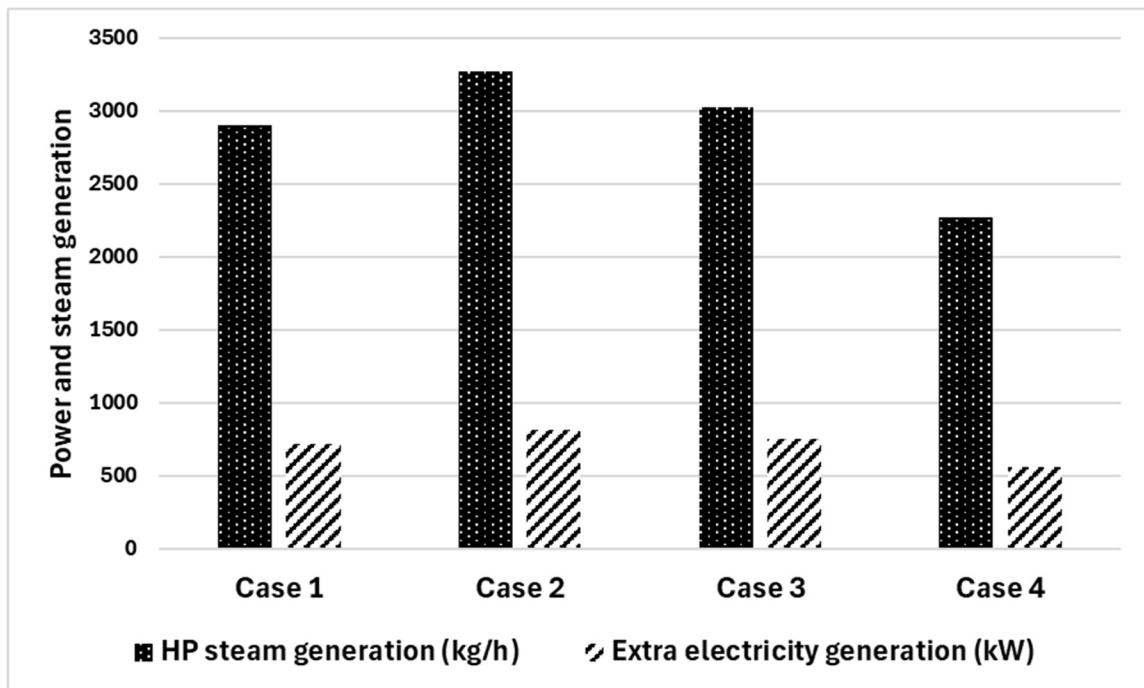


Figure 40 Steam and electricity generation after heat integration.

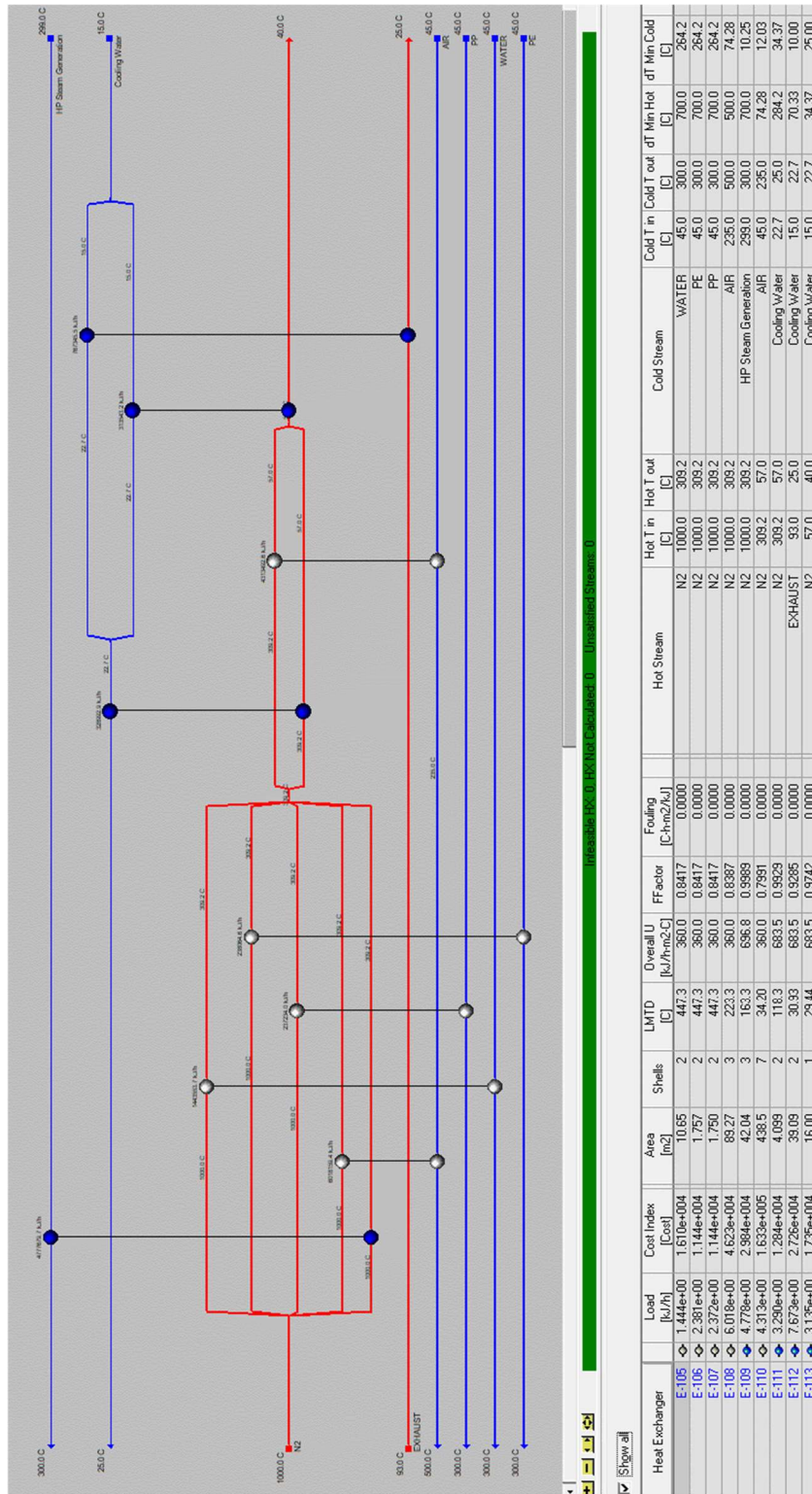


Figure 41 Improved heat exchanger network and system details for case 2

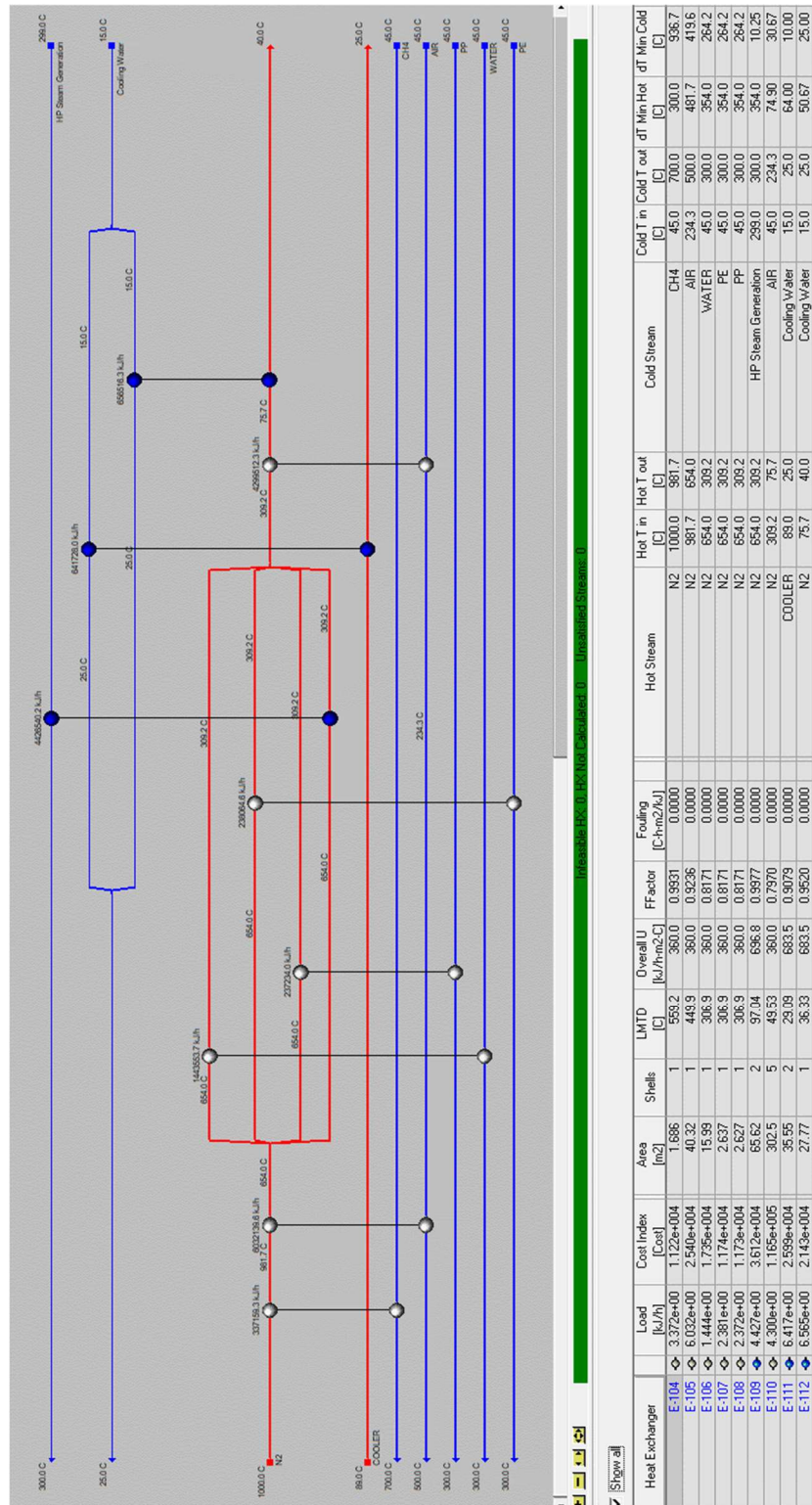


Figure 42 Improved heat exchanger network and system details for case 3

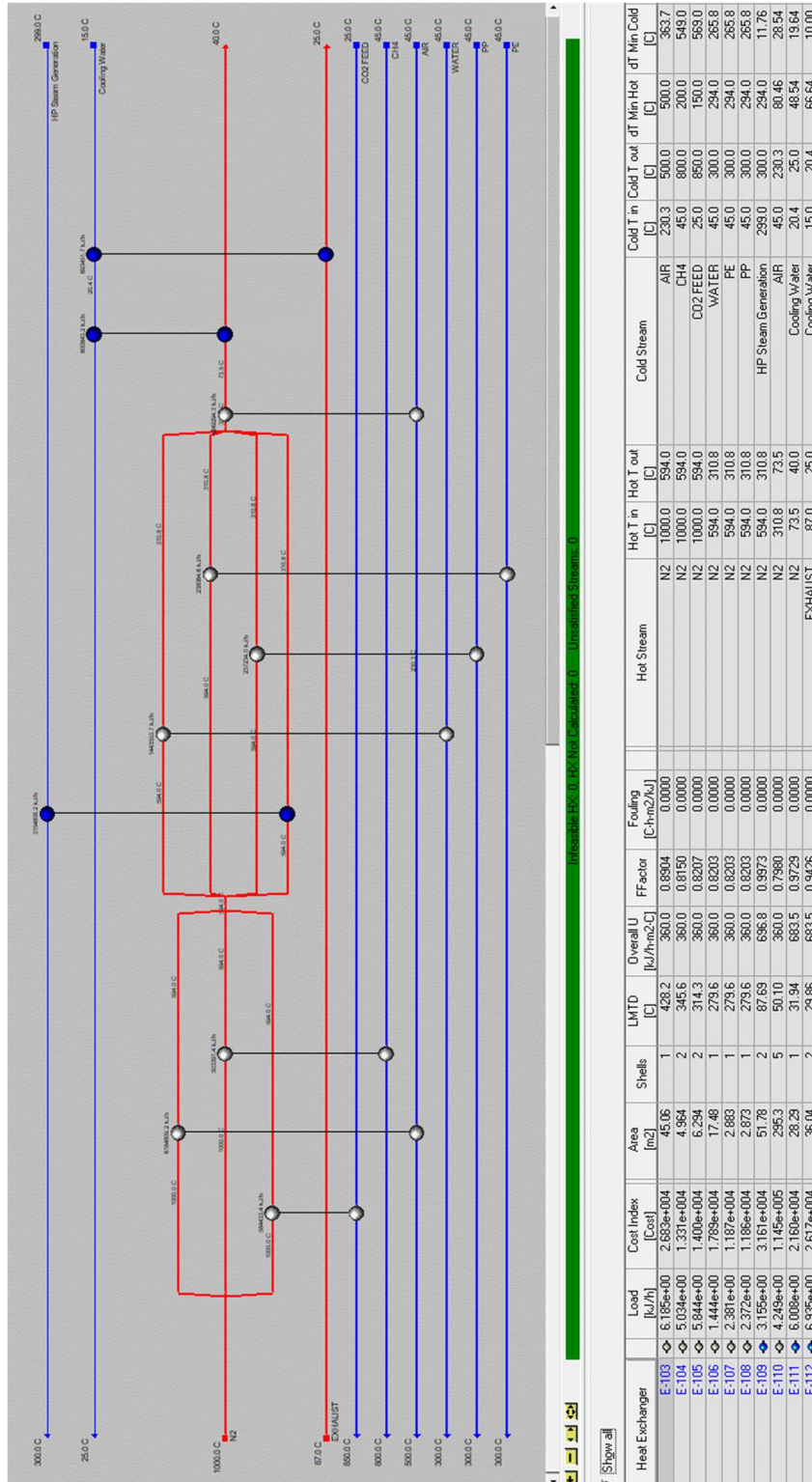


Figure 43 Improved heat exchanger network and system details for case 4

CHAPTER 7

ECONOMIC ANALYSIS

7.1 CAPEX and OPEX

The economic analysis of the four cases provides a detailed assessment of their feasibility and profitability. While none of the cases demonstrate full economic feasibility due to relatively high payback periods (PBP), which indicate slower recovery of initial investments, Case 3 emerges as the most economically viable option among the evaluated scenarios. However, it is crucial to emphasize that these processes are not solely intended for profitability. Their primary purpose lies in addressing critical environmental challenges, that is eliminating the waste plastic in eco-friendly sustainable practices. Consequently, the implementation of such initiatives often requires governmental support, policy-driven incentives, or external funding from charitable and environmental organizations to bridge the financial gap and ensure their long-term success.

To evaluate the power production processes' economic viability in the current market, an economic analysis of the cases under study was conducted. First, as shown in Table 19, CAPEX and OPEX were calculated for each example.

Starting with the capital expenditure (CAPEX), case 1 has the highest CAPEX at \$35.69M as shown in Figure 44. This is primarily due to the inclusion of an Air Separation Unit (ASU), which is a major cost driver in the process. The ASU is a complex system comprising several compressors, heat exchangers, and multiple distillation columns, which collectively cost around \$4.21M. This makes equipment and indirect costs the dominant components of case 1's CAPEX structure. Case 2 significantly reduces the CAPEX to

\$20.89M by replacing the ASU with Chemical Looping Combustion (CLC), which effectively lowers both equipment and indirect costs. Case 3 further lowers the CAPEX to \$19.96M by using Partial Oxidation (POX) instead of Steam Methane Reforming (SMR), reducing the overall complexity and associated costs. Case 4, which integrates POX with Dry Methane Reforming (DMR) and CO₂ recycling, has a slightly higher CAPEX of \$20.97M compared to Case 3 due to the additional process complexity. While Case 2 achieves the lowest CAPEX, it is not sufficient to make it the most viable when considering all factors. Case 3 offers a good balance between cost reductions and operational efficiency.

The operational expenditure (OPEX) analysis reveals notable differences among the cases as shown in Figure 45. Case 1 and Case 2 have higher OPEX values of \$1.68M and \$1.70M per year, respectively, driven by high utility costs due to energy-intensive ASU and CLC operations. On the other hand, cases 3 and 4 demonstrate lower OPEX at \$1.41M per year, attributed to efficiency improvements from POX and reduced utility demands enabled by CO₂ recycling. While case 2 offers the lowest CAPEX, its OPEX is higher than cases 3 and 4, which makes case 3 the more operationally efficient option when both capital and operational costs are considered.

Table 19 Summary of capital and operating expenditures

	Case 1	Case 2	Case 3	Case 4
CAPEX Calculation (M\$)				
Equipment, and installation cost	14.4	11.41	10.92	11.45
Direct cost	4.09	1.82	1.74	1.83
Indirect cost	13.74	6.12	5.83	6.14
Working and land cost	3.46	1.54	1.47	1.55

Total CAPEX	35.69	20.89	19.96	20.97
OPEX Calculation (M\$/year)				
Labor Cost	0.59	0.59	0.59	0.59
Raw Material Cost	0.01	0.01	0.01	0.01
Utility Cost	1.07	1.09	0.80	0.80
Total OPEX	1.68	1.70	1.41	1.41

Figures 44 and 45 also present a breakdown of the developed cases' CAPEX and OPEX. The components of CAPEX that contribute to the overall value are shown in Figure 44, with equipment installation expenses showing up as the largest contribution at 52.3% across all cases. On the other hand, as Figure 45 illustrates, using waste plastic as a raw material reduces the cost of raw materials when compared to other OPEX components. As a result, it establishes utilities prices as the primary economic driver of OPEX, with labor costs following closely behind.

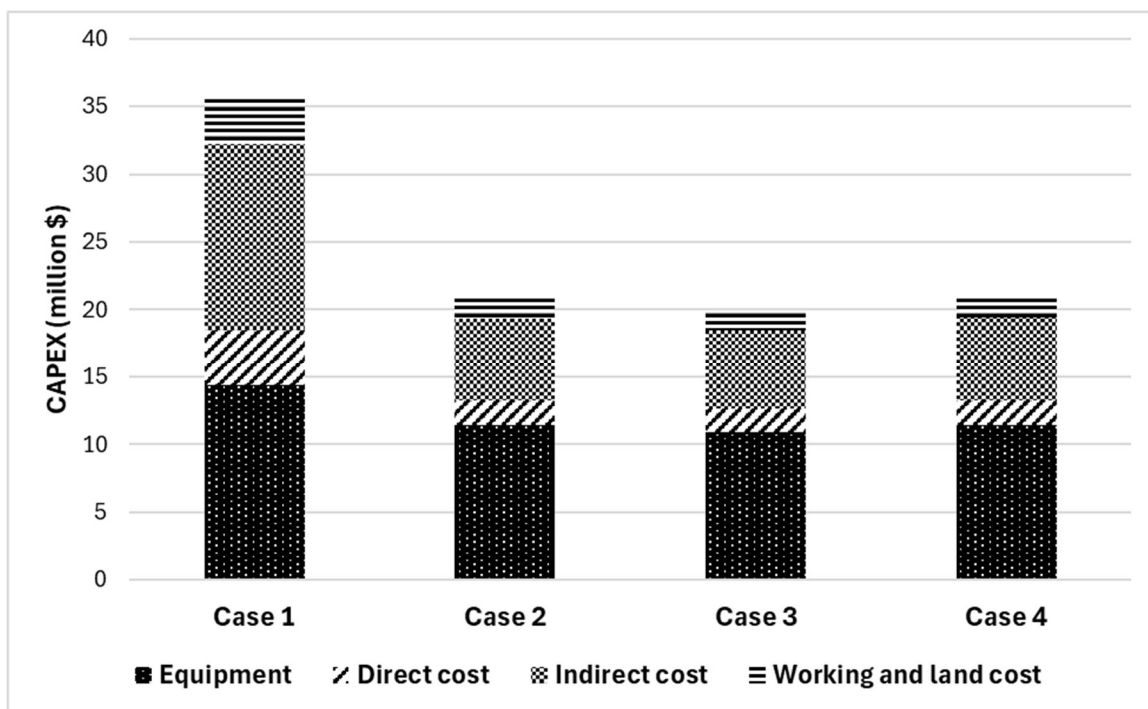


Figure 44 Breakdown of capital expenditures

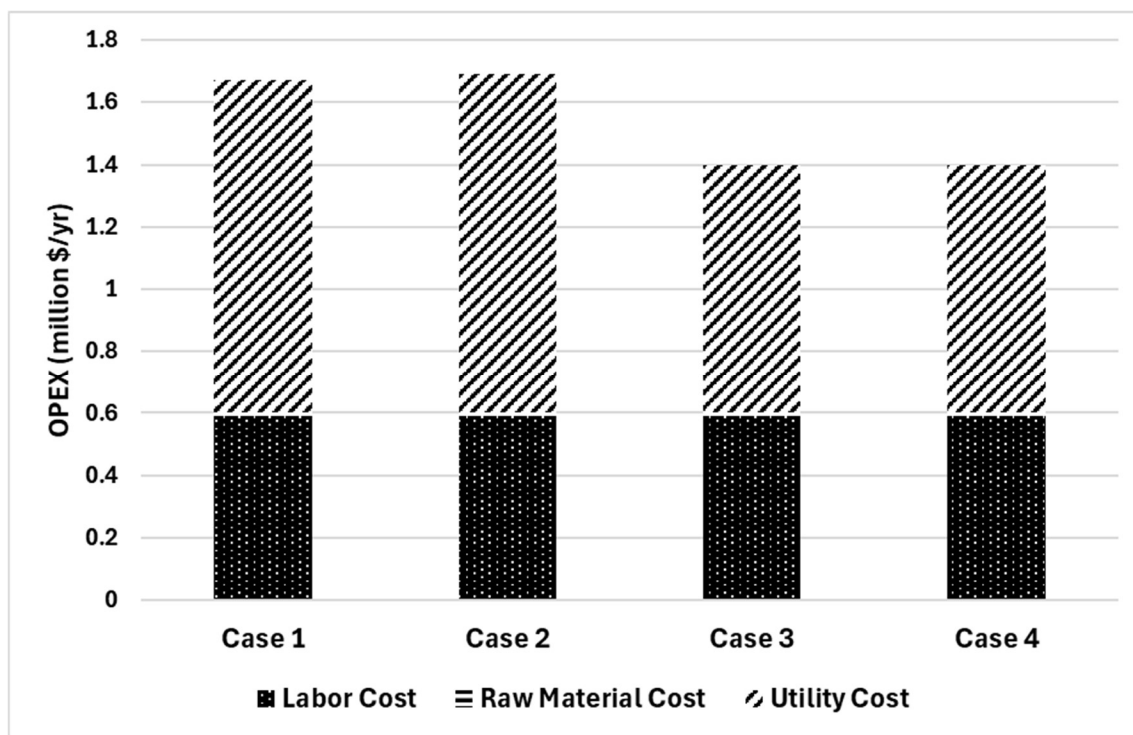


Figure 45 Breakdown of operating expenditures

For further insights, breakdown of equipment cost is presented in Table 20, showing the individual installed cost of all the equipment utilized in each process. The table shows that case 1 incurs the highest equipment cost due to the inclusion of ASU, leaving the other cases with almost the same equipment cost. Among other equipment, GT and HRSG together contribute the highest costs of approximately 44.6% in the total equipment cost followed by the by the CO₂ dehydration unit. The CO₂ dehydration unit consist of series of compressors and separators which compresses and separates the condensed water from the CO₂ stream.

The similarities in the CAPEX of cases 2–4 can be attributed to the integration of chemical looping combustion (CLC) in all three cases. The CLC technology inherently requires lower capital investment compared to the air separation unit (ASU) used in Case 1, as it eliminates the need for large-scale oxygen production infrastructure. Furthermore, the reforming methods in Cases 2–4 share similar equipment requirements and operating scales, leading to comparable capital expenditures across these cases.

Case 3 stands out as better primarily due to its lower CAPEX and LCOE. Additionally, the higher concentration of combustibles in the syngas generated in Case 3 results in the highest higher heating value (HHV), making it the most energy-efficient scenario among the cases.

For operating expenses (OPEX), the distinctions arise from the varying utility consumption and efficiency of the processes. Case 1 incurs higher OPEX due to the energy-intensive ASU operation, while Cases 2–4 benefit from the inherent efficiencies of CLC, which reduces utility costs. Case 3, in particular, achieves further OPEX reductions due to its

optimized reforming process and lower requirements for utilities, contributing to its overall better economic performance. These factors highlight the interplay of capital and operational efficiencies in determining the economic viability of the cases.

Table 20 Breakdown of the equipment installed cost

	Case 1	Case 2	Case 3	Case 4
Crushers	0.087	0.087	0.087	0.087
Pyrolysis + Gasification	0.18	0.18	0.18	0.18
Hydrodesulphurization	0.071	0.071	0.071	0.071
Methane compressor	1.34	1.34	1.34	1.34
POX	-	-	0.33	0.40
SMR feed heater	0.23	0.23	-	-
ASU	4.21	-	-	-
DMR	-	-	-	0.02
Air reactor	-	0.38	0.38	0.40
SMR	0.51	0.51	-	-
GT + HRSG	4.73	4.56	4.23	4.73
Steam turbine	0.17	0.17	0.17	0.17
CO ₂ separator (prior to purification)	0.10	0.12	0.12	0.12
CO ₂ Dehydration (KODs)	2.80	2.80	2.91	2.80
Additional CAPEX for integration	0.13	0.34	0.28	0.29
Total	14.4	11.41	10.92	11.45

Revenue Analysis and Profitability Indicators

Revenue analysis highlights (Table 21) that case 1 generates the highest power revenue of \$6.8M annually. However, as the power production efficiency decreases in cases 2, 3, and 4 due to process modifications, their power revenues fall within a range of \$4.58M to \$5.06M. The CO₂ revenue remains consistent across all cases, ranging from \$1.11M to \$1.18M annually. While case 1 generates the highest revenue, its high CAPEX and OPEX overshadow these gains, making it the least viable.

The profitability indicators show further distinctions between the cases (Figure 46). Case 1 has the longest payback period (PBP) of 9.76 years and the lowest net present value (NPV) of \$0.79 M, confirming its financial infeasibility. Case 2 achieves a shorter PBP of 8.79 years and an NPV of \$5.63M, making it seem promising, but its high OPEX impacts its overall viability. Case 3 achieves a PBP of 8.86 years and an NPV of \$5.10M, supported by an efficient OPEX, which establishes it as the most economically feasible option despite its slightly longer PBP compared to case 2. Case 4 has a comparable PBP of 9.16 years and an NPV of \$4.27M, but its increased complexity and slightly reduced profitability make it less attractive than Case 3.

Table 21 Financial feasibility summary of the four cases investigated.

Profitability Indicator	Case 1	Case 2	Case 3	Case 4
Power revenue (million \$/yr)	6.8	5.06	4.58	4.59
CO ₂ revenue (million \$/yr)	1.18	1.12	1.13	1.11
Total revenue (million \$/yr)	7.98	6.18	5.71	5.7
PBP (Year)	9.76	8.79	8.86	9.16
NPV (million \$)	0.79	5.63	5.10	4.27

ROROI (%)	7	12	12	11
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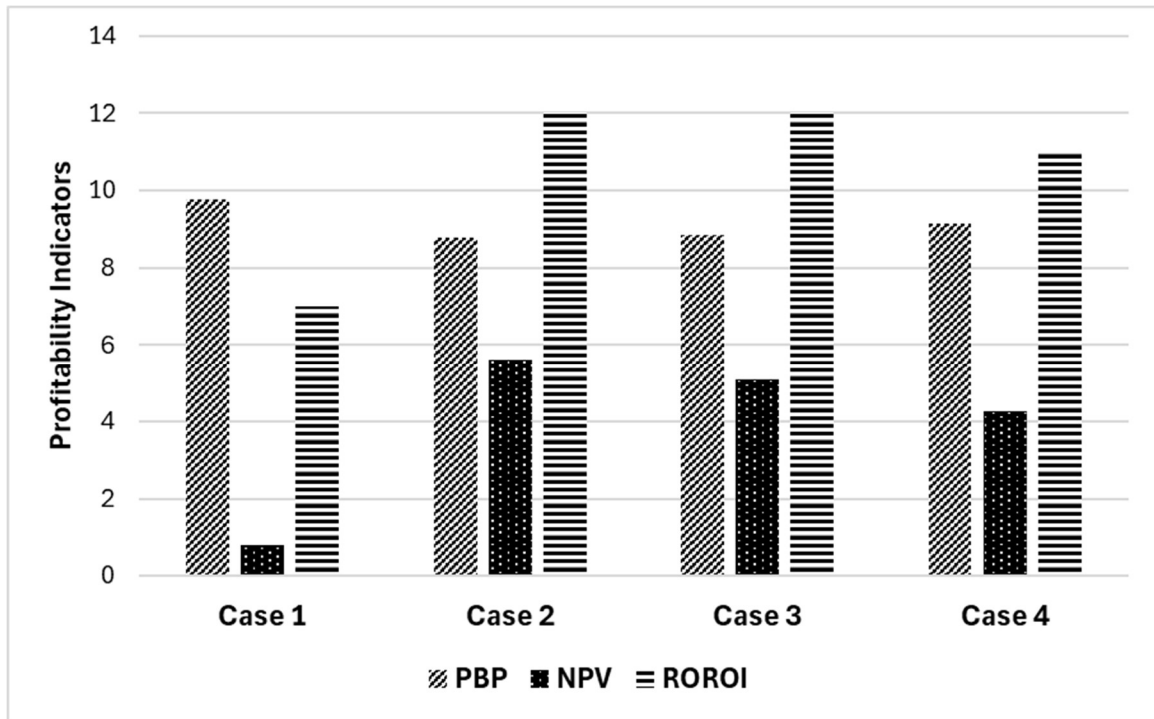


Figure 46 Summary of economic indicators.

7.2 Cashflow Analysis

A complete cashflow results of the study are presented in Figure 47. The cumulative discounted cash flow diagram provides valuable insights into the financial trajectories of the four cases over their respective project lifetimes. The chart illustrates how each case performs in terms of recovering the initial investment and generating net cash flow over a 22-year period.

Case 1 shows the steepest decline in cumulative discounted cash flow during the initial years, driven by its high CAPEX of \$35.69M, dominated by the cost of the Air Separation Unit (ASU) and associated equipment. The cash flow remains consistently lower than zero

throughout the project life, indicating that revenues from power and CO₂ production are insufficient to offset the high investment and operational costs. This results in a in a very low net present value (NPV) of \$0.79M, confirming that case 1 is not financially viable under the given assumptions.

Case 2, with a reduced CAPEX of \$20.89M and slightly higher OPEX by \$1.70M per year, shows a much more favorable trajectory. The initial investment is recovered relatively quickly, with cumulative discounted cash flow turning positive around the eighth year. By the end of the project life, case 2 achieves the highest NPV of \$5.63M, reflecting strong long-term profitability. However, the higher OPEX impacts its cash flow growth rate, slightly limiting its economic appeal despite its high NPV.

Case 3 demonstrates a similarly positive trajectory, achieving a breakeven point around 8.86 years. With a CAPEX of \$19.96M and lower OPEX by \$1.41M per year, the case shows steady and consistent cash flow growth after the initial investment is recovered. By the end of the project, it achieves an NPV of \$5.10M, only slightly lower than case 2. This strong balance of CAPEX, OPEX, and profitability makes Case 3 the most economically feasible option when compared to the other cases.

Case 4 starts with a slightly higher CAPEX of \$20.97M, which, combined with added process complexity, leads to a slower cash flow recovery. The breakeven point is reached slightly later than in case 3, but the cumulative discounted cash flow remains positive thereafter. By the end of the project, case 4 achieves an NPV of \$4.27M, reflecting its financial viability but showing reduced profitability compared to cases 2 and 3.

The cash flow diagram highlights that while none of the cases are entirely feasible due to their high payback periods, cases 2 and 3 show strong financial performance. Case 2 achieves the highest NPV, but its higher OPEX reduces its long-term cash flow growth rate. Case 3, on the other hand, offers a more balanced economic performance, with steady cash flow recovery and a competitive NPV. Case 4, although financially viable, is less attractive due to its slightly longer breakeven time and reduced profitability. Case 1 is clearly the least viable, with cash flows remaining negative during 20 years of the project life, confirming its financial infeasibility.

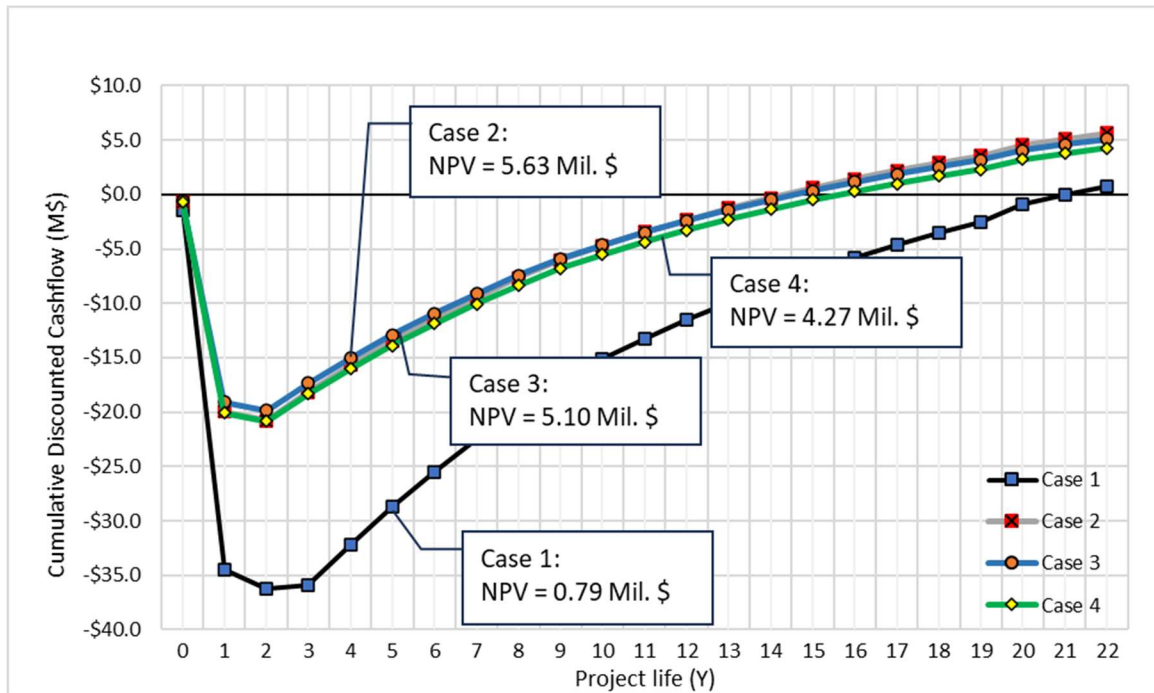


Figure 47 Cashflow analysis of all cases

7.3 LCOE

These economic analyses allowed for the assessment of levelized cost of electricity (LCOE) for each of the developed cases, providing a useful metric for direct comparison with the market prices for energy, as illustrated in Figure 48. The LCOE for scenario 1 is

substantially higher than the LCOE for the other examples due to a large variance in CAPEX, resulting in an absolute value of \$211.34/MW. Conversely, among rivals, case 3 (\$94.6/MW) turned out to be the most cost-effective.

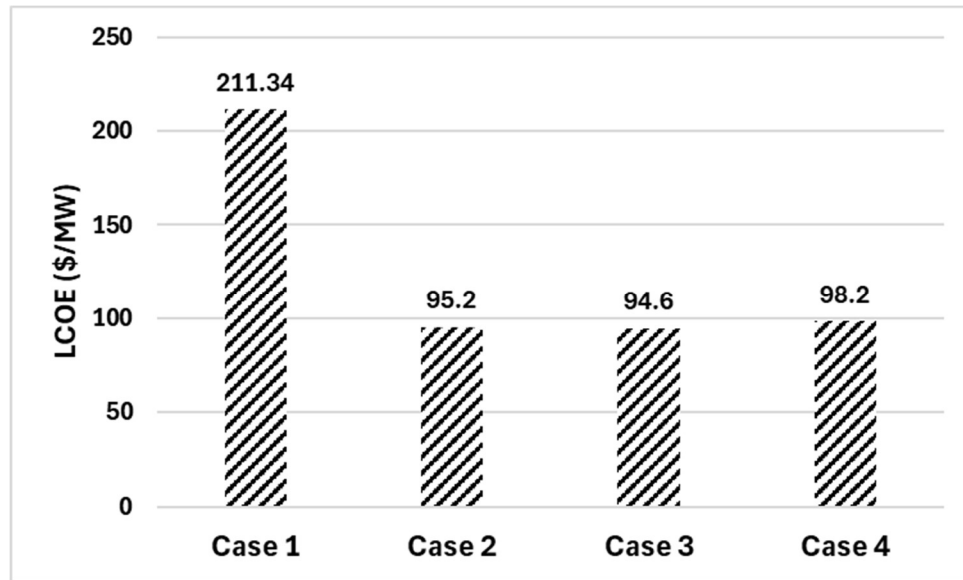


Figure 48 Levelized cost of electricity (LCOE) for the developed cases

CHAPTER 8

CONCLUSIONS AND RECOMMENDATIONS

This study provides a detailed and holistic evaluation of power generation from waste plastic, focusing on its potential to alleviate environmental burdens and promote sustainability through a circular economy approach. By utilizing waste plastic as fuel, the process not only addresses the challenge of plastic disposal but also reduces dependence on fossil fuels, leading to significant reductions in emissions associated with power generation. To achieve these dual benefits, four distinct scenarios were developed, employing various reforming and oxygen-supplying techniques. The scenarios included gasification with SMR and ASU, gasification with SMR and CLC, gasification with POX and CLC, and gasification with POX and DMR along with CLC. Each scenario was subjected to detailed process simulations to provide comprehensive material and energy balances, which served as the foundation for a thorough techno-economic analysis encompassing performance, sensitivity, and economic assessments.

The findings from the simulations demonstrated that the choice of reforming and oxygen-supplying techniques had a substantial impact on syngas production, energy efficiency, and utility consumption. Cases 1 and 2, employing SMR, produced identical syngas quantities of 850 kg/hr with a hydrogen-to-carbon ratio (HCR) of 3.5. However, the integration of CLC in Case 2 reduced utility consumption by 10.3% compared to Case 1. Case 3, which utilized POX with CLC, produced the least syngas at 700 kg/hr due to a lower requirement for the reforming agent but achieved the highest energy density with a higher heating value (HHV) of 26.1 MJ/kg. This high HHV was attributed to the highest concentration of combustible components (99.2%) among all scenarios. Meanwhile, Case 4, incorporating

DMR, achieved the highest syngas production at 1,830 kg/hr with the lowest HCR of 0.8, benefiting from the partial recycling of captured CO₂ in the reformer.

A comparison of the scenarios revealed that, although Cases 1 and 2 utilized the same input flows and syngas production processes, they exhibited differences in flue gas temperatures after combustion. This discrepancy was linked to the presence of metal in the CLC of Case 2, which absorbed some combustion heat. While Case 3 produced the least syngas due to a lower oxidizing agent requirement, its higher concentration of combustible components resulted in superior energy density. Case 4, on the other hand, maximized syngas production by leveraging partial CO₂ recycling alongside steam in the reformer.

The analysis of power output and efficiency revealed a complex interplay of factors. Power output was strongly influenced by syngas flow and HCR, whereas efficiency depended on power generation and utility requirements. Case 1 produced the highest electricity due to its favorable syngas flow and HCR, while Cases 4 and 2 showed efficiency improvements of 4.04% and 4.87%, respectively. In contrast, Case 3 exhibited 3.94% lower efficiency than Case 1 due to reduced power production and syngas flow. Nevertheless, the heat integration of high-temperature nitrogen streams released from the CLC in Cases 2–4 provided an opportunity to produce additional steam, enhancing power generation and increasing overall process efficiency.

Sensitivity analysis revealed that increasing methane input to the SMR resulted in a linear increase in power generation across all scenarios. This indicated that process efficiency and economics remained stable with increasing capacity. A similar linear trend was observed for captured and uncaptured CO₂ emissions, which also increased with higher

methane input. Notably, the relative contribution of captured CO₂ to total emissions grew with increasing capacity, suggesting that higher operational capacities improve the environmental sustainability of the process. Furthermore, variations in gasifier operating temperature demonstrated no significant impact on overall power output. However, the composition of syngas changed significantly with temperature, with the desired components, H₂ and CO, increasing exponentially up to 1300°C, while CH₄ and CO₂ decreased. Beyond 1300°C, the composition stabilized, indicating that further temperature increases would not yield additional benefits.

The economic analysis provided critical insights into the financial viability of the scenarios. A significant discrepancy was observed in the capital expenditures (CAPEX) among the cases, with Case 1 incurring the highest CAPEX due to the use of an ASU. In contrast, the construction expenses for CLC in other scenarios were comparatively lower. As a result, Case 1 exhibited the highest levelized cost of electricity (LCOE) at \$211.34/MW. On the other hand, Case 3 emerged as the most cost-effective solution, with the lowest LCOE of \$94.6/MW, due to its lower CAPEX and efficient resource utilization. Operating expenditures (OPEX) for Case 2 aligned closely with Case 1, while Case 4 had a similar OPEX to Case 3.

In conclusion, this study highlights the potential of power generation from waste plastic as a sustainable and economically viable approach to address both environmental and energy challenges. The scenarios developed and analyzed in this work exhibit diverse advantages, with some offering higher energy efficiency and others presenting superior economic attractiveness. Among the scenarios, Case 3 stands out as the most promising due to its high energy density, optimal utilization of resources, and lowest LCOE. Future research

should focus on optimizing the integration of circular economy principles, further minimizing environmental impacts, and advancing reforming technologies to enhance process efficiency and scalability.

In conclusion, it is unlikely that investors would take interest in and invest in such projects with the given economic parameters. The high capital expenditures (CAPEX) and operating expenses (OPEX) associated with certain cases, coupled with the levelized cost of electricity (LCOE) values that are not competitive compared to conventional or alternative energy sources, make these projects financially unattractive. Investors typically seek projects with clear profitability, competitive costs, and manageable risks, none of which are strongly demonstrated by these cases under the current market conditions. However, once government support and commitment are there, such project may receive traction from investors.

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