A comparative assessment of fast pyrolysis, hydrothermal liquefaction, and intermediate pyrolysis to produce value-added products from municipal solid waste

by

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Abstract

The increasing complexity in municipal solid waste (MSW) streams along with the growing waste-to-product market call for innovative and flexible technologies that can efficiently process MSW streams into value-added products. One of these value-added products is transportation fuels; producing them from MSW, can reduce greenhouse gas (GHG) emissions from both the transportation and waste sector. A new and emerging method that provides flexible, on-site generation of high-quality bio-oil and other value-added by-products from MSW is intermediate pyrolysis (IP). Other thermochemical conversion routes such as hydrothermal liquefaction (HTL) and fast pyrolysis (FP) are also adept at converting MSW into bio-crude or bio-oil. However, to reach compatible transportation grade fuels, it is necessary to upgrade the intermediate product (bio-crude or bio-oil) in all the processes; the extent of upgrading differs depending on the quality and quantity of the product. Unlike HTL and FP processes, the performance and economics of intermediate pyrolysis have not yet been investigated to identify areas in which investment and research can be focused on for the greatest impact. Thermo-catalytic reforming (TCR) is considered one of the intermediate pyrolysis pathways. In this study a comprehensive techno-economic assessment (TEA) of a decentralized IP plant of 500 kg h^{-1} (12 dry t d^{-1}) input capacity with MSW as the preferred feedstock was conducted. From the developed process model and economic evaluation, the plant's capital cost was estimated to be 3.9 million USD. The calculated IP bio-oil production cost was \$ 2.01 L^{-1}. It was also found that generating additional revenue by selling by-products (biochar and hydrogen) is an important means of reducing bio-oil production costs. Other financial benefits are in the form of gate fees and carbon credits from using MSW as a feedstock. The effects on bio-oil production cost of increasing the plant scale from 500 kg h^{-1} to 4000 kg h^{-1} were also investigated.

To understand better the limitations and opportunities of HTL and FP processes versus IP, a comparative techno-economic assessment of the three technologies was performed. The organic-dominant MSW stream was considered the primary feed for each process and transportation fuels (gasoline, jet fuel, and diesel) the final products. The effects of different production frameworks were also integrated into the study by evaluating these technologies in centralized and decentralized configurations. In a centralized system, MSW is transported to the facility where an intermediate is produced and upgraded (on-site upgrading), while in a decentralized system, an intermediate is produced from MSW at distributed regions and transported to an upgrading facility (off-site upgrading). In this study, four scenarios were developed to evaluate the production cost of gasoline, jet fuel, and diesel: 1) centralized HTL plant (C-HTL) at 2000 dry t d^{-1} with on-site upgrading, 2) centralized FP plant (C-FP) at 2000 dry t d^{-1} with on-site upgrading, 3) decentralized FP plant (D-FP) at 50 dry t d^{-1} with off-site upgrading; and 4) decentralized IP plant (D-IP) at 12 dry t d^{-1} with off-site upgrading. For decentralized scenarios, multiple plants were used to have an overall processing capacity of 2000 dry t d^{-1}. Jet fuel was considered to be the primary fuel for comparison and the production cost was calculated to be \$0.71 L^{-1}, \$0.80 L^{-1}, \$1.00 L^{-1}, and \$0.78 L^{-1} for C-HTL, C-FP, D-FP, and D-IP, respectively. Secondary products (gasoline and diesel) could be produced for \$0.96 L^{-1} - \$1.36 L^{-1}, and \$1.01 L^{-1} - \$1.43 L^{-1}, respectively, through the developed scenarios.

Preface

This thesis is an original work by Wasel Ur Rahman under the supervision of Professor Amit Kumar.

Chapter 2 of this thesis, titled "Techno-economic assessment of the use of thermo-catalytic reforming to produce high-value products from municipal solid waste in remote North American municipalities" by Wasel Ur Rahman, Madhumita Patel, Vinoj Kurian, and Amit Kumar will be submitted to a peer-reviewed journal.

Chapter 3 of this thesis, titled "A comparative assessment of fast pyrolysis, hydrothermal liquefaction, and intermediate pyrolysis of municipal solid waste for liquid transportation fuels production" by Wasel Ur Rahman, Madhumita Patel, Vinoj Kurian, and Amit Kumar will be submitted to a peer-reviewed journal.

I was responsible for the data collection, data analysis, model development, and manuscript composition. Madhumita Patel and Vinoj Kurian reviewed the developed models, assessed the results, and provided feedback on the research structure. They also corrected journal papers along with Astrid Blodgett, the editor of the research group. Professor Amit Kumar was the supervisory author and provided supervision on concept formulation, models and results validation, and manuscript edits.

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My M.Sc. journey has been an emotional experience, with the sudden demise of my father very early in my program. My mother has been my rock and if not for her psychological and emotional support, I doubt I would have come this far. My younger brother’s reassurance from time to time has been of immense support to me while I am living so far from home.

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For Abbu and Maa

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List of abbreviations

MSW	Municipal solid waste
HTL	Hydrothermal liquefaction
FP	Fast pyrolysis
TCR	Thermo-catalytic reforming
IP	Intermediate pyrolysis
AD	Anaerobic digestion
WTE	Waste-to-energy
WTP	Waste-to-product
CHP	Combined heat and power
GC-MS	Gas chromatography-mass spectrometry
TAN	Total acid number
SHC	Solid heat carrier
TEA	Techno-economic assessment
TPEC	Total project equipment cost
TIC	Total installed cost
IC	Indirect cost
TPI	Total project investment
C-HTL	Centralized hydrothermal liquefaction plant
C-FP	Centralized fast pyrolysis plant
D-FP	Decentralized fast pyrolysis plant
D-IP	Decentralized intermediate pyrolysis plant
HDO	Hydrodeoxygenation

Chapter 1 : Introduction

1.1 Background

Municipal solid waste (MSW) refers to recyclables and compostable materials, as well as garbage from homes, businesses, institutions, and construction and demolition sites [1]. MSW can be categorized into recyclables (paper, plastic, glass, metals, etc.), toxic substances (paints, pesticides, used batteries, medicines), compostable organic matter (fruit and vegetable peels, other food waste), and soiled waste (sanitary napkins, disposable syringes, household hygiene products, etc.) [2-4]. As global industrialization and population growth escalate at unprecedented rates, the effect of this growth is seen through the increase in the per capita generation of MSW. The world generates 2.01 billion tons of municipal solid waste annually from all sources (residential, commercial, and industrial), and this figure is expected to grow to 3.40 billion tons by 2050 [5].

The global impact of unrestrained waste generation and disposal is increasing. Based on the volume of waste produced and its management, it is estimated that 1.6 billion tonnes of carbon dioxide (CO_2) equivalent greenhouse gases were emitted in 2016, which represents 5 percent of global emissions [5]. In a business-as-usual case, waste-related GHG emissions are anticipated to reach 2.6 billion tonnes of CO_2 equivalent by 2050. Canada alone reported the disposal of 25 million tonnes of residential and non-residential waste into landfills in 2016 [6] or 17 megatonnes (Mt) of CO_2 equivalent. Moreover, emissions from landfills are the third largest contributor to Canada's overall methane (25 times more potent than CO_2) emissions, reported to be 17 Mt in 2017 [7].

Canada's geographic and economic diversity contributes to the variation in MSW composition as well as the per capita disposal. In addition, due to the complexity in the treatment and management of MSW, it is often difficult to establish a uniform strategy for waste minimization between local authorities and municipalities or between provincial and federal governments. MSW is regulated by provinces and territories and managed either by municipal authorities directly or contracted out to the waste management industry [8]. In jurisdictions that lack adequate resources, the federal government is responsible for regulating MSW-related activities. These activities, limited by technical (lack of technologies and infrastructures) and economic (lack of financial incentives) constraints, are not aligned with waste management hierarchy (i.e., reduce, reuse, recycle, recover and only then dispose of). This leads to local authorities being forced to dispose of MSW in landfills, some of which are close to exceeding designated capacities [9].

The Canadian province most challenged to reduce its waste disposal is Alberta. The province recorded the country's highest waste disposal per capita of 1034 kg in 2016 [6], 42% higher than the national average (729 kg). Efforts to reduce disposal rates have changed Alberta's waste management infrastructure over the years. For the collection phase, residents are now provided with well distributed curbside collection systems for garbage, recycling, and organics to initiate waste sorting from the source. Moreover, to recover energy and value-added products, the province promotes waste-to-energy (WTE) or waste-to-product (WTP) facilities in its energy strategy and has 420 registered small energy recovery facilities to treat organics (e.g., food, yard waste, manure) [8]. However, the registration of these facilities never expires, nor are their operations supervised or looked after by pertinent authorities; as a consequence, there is very limited data is available to evaluate how these facilities might improve landfill disposal rates [8]. To meet federal and provincial goals of reducing per capita disposal to 490 kg (a 30% reduction by 2030 from 2014), far more innovative and sustainable strategies are necessary.

These strategies could include shifting the perspective of MSW from merely a waste destined for landfills to an urban resource. For MSW to be considered a resource, it must be converted from a heterogeneous stream into a usable homogeneous form with many applications. Conversion routes can be broadly classified as biological and thermal. Biological treatments such as composting and anaerobic digestion (AD) are established pathways to process organic fractions of MSW. However, they cannot process inorganics (i.e., plastics, glassware, textiles, etc.) and so cannot completely replace regional landfills. Thermal treatment such as incineration, gasification, and direct liquefaction can be used to treat uncharacterized waste streams and are pathways for waste-to-energy (WTE) and waste-to-product (WTP) applications [10-12]. Incineration is discouraged in many jurisdictions because it emits large amounts of pollutants (NO_x, SO_x, and dioxins) and fine particles [13-15]. These pollutants, along with CO_2, are a significant threat to the environment and human health. The gasification of MSW produces synthesis gas (syngas), which requires a considerable amount of cleaning and conditioning prior to use and contributes significantly to operating expenses [16, 17]. Moreover, internally fired systems are not efficient in small-scale gasification systems and often not economically feasible, due to poor economies of scale [16, 17].

Direct liquefaction of feed is often favored over gasification due to the simplicity and high liquid yields it provides. Two techniques that can directly liquefy biomass are hydrothermal liquefaction (HTL) and pyrolysis. In general, HTL reactions occur at elevated temperatures of 250˚C to 380˚C and pressure between 5 and 30 MPa [18-20]. At these conditions, water exhibits enhanced solvent characteristics that facilitate the formation of bio-crude, gas, and water with dissolved organics [19]. Pyrolysis is the thermal degradation of organic substances in the complete absence of oxygen and at heating temperatures between 400 and 700˚C [21]. Pyrolysis techniques can be divided into fast, intermediate, and slow depending on feedstock heating rate and residence time in the reactor.

The operating parameters (heating rate and residence time) can be varied to optimize the yield of products, which mainly consist of pyrolysis gas, solid char, and an organic phase (bio-oil). Fast and intermediate pyrolysis techniques have received more interest than slow pyrolysis because they can produce a liquid-dominant product stream. Fast pyrolysis (FP) is carried out at high heating rates of 1000˚C/s with a residence time of 1-2 s at atmospheric pressure. Although the quantity of bio-oil produced from the FP process is higher than from other pyrolysis techniques, it is highly viscous, acidic, and contains many oxygenated compounds. These properties limit direct applications of the FP bio-oil, and the bio-oil requires further upgrading to transform it into a usable product.

Intermediate pyrolysis is performed at moderate temperatures (400-450˚C) and heating rates (100-500˚C/min), with a longer feedstock residence time in the reactor than conventional fast pyrolysis. A recently developed approach to convert biomass and waste streams through a combination of intermediate pyrolysis and reforming is thermo-catalytic reforming (TCR®) [22, 23]. This emerging thermochemical process can produce bio-oil with more enhanced physical and chemical properties. A wide variety of feedstocks including MSW has been tested under intermediate pyrolysis conditions, yet the produced bio-oil does not have the same low levels of moisture, oxygen, and aromatics present in TCR-derived bio-oils [24-27]. The TCR bio-oil is of higher quality because of the additional reforming stage, which promotes thermal and catalytic cracking of the initial pyrolysis vapors. Here, long-chained aromatics are broken down into lighter fractions because of secondary cracking reactions, allowing the TCR process to generate products with superior properties [28].

The direct liquefaction techniques (HTL, FP, and IP) described above have all been tested at lab scales using a wide variety of feedstocks including MSW [20, 23, 29-33]. The primary products from these techniques (either bio-crude or bio-oil) have limited applications on their own; however, they could be upgraded to useful products such as transportation fuels, given adequate research and funds. For their development from lab or pilot scale into commercial ventures, research and incentives are needed to resolve bottlenecks that limit the replication of lab or pilot scale results in large-scale units. Without scaling up production capacities, the technical and economic barriers will remain unexplored, whereas to secure investments these technologies must exhibit robustness and reliability at commercial scale. This has led to a catch-22 situation: investors are reluctant to fund large-capacity constructions due to the uncertainty in their techno-economic performance, and the lack of real production costs from established plants is a hurdle to gain additional knowledge. For decision-makers and policymakers including local government and authorities, a solution to this dilemma is needed to develop long-term waste reduction frameworks.

Techno-economic assessment (TEA), a method to assess an emerging technology in terms of its technical and economic performance, can be a possible solution. TEA uses process models and economic analysis to evaluate one or more specific process pathways. It allows us to quickly and inexpensively calculate missing information such as the capital investment required or the anticipated financial return upon that investment. While uncertainties with TEA results could be a high cause of the necessary simplification of complex processes, TEA facilitates comparisons at the early stages of commercialization under current and future market scenarios.

This study follows a TEA-based approach to identify the technical and economic benefits of converting MSW into urban resources or value-added products using the direct liquefaction techniques described above. Final products can be in the form of energy (heat or electricity) or material (fuels, diluents, mixed alcohols, etc.). In this research, the focus has been on the production of transportation fuels (blends of gasoline, jet fuel, and diesel). Other researchers have evaluated the performance of HTL and FP-based routes in the production of value-added products including transportation fuels [34-38]. A review and comments on those studies are presented in Table 1.1. Further details are included in subsequent chapters. TCR, however, is a recent technology whose techno-economic performance has not yet been evaluated. For a comprehensive comparison among the techniques, a TEA is conducted first on TCR technology and is presented in chapter 2.

Table 1.1: Literature review on TEA studies for HTL, FP, and IP plants

Plant capacity	Configuration	Process	Feedstock	Final product	Comment/References
2000 dry t d^{-1}	Centralized	FP	Corn stover	Diesel and naphtha	Wright et al. investigated two scenarios, hydrogen purchase and hydrogen production, and concluded the former to be the cheaper route to produce diesel and naphtha [39]. This work was expanded by Brown et al. [34], who integrated combined heat and power (CHP) units to combust char and non-condensable gases. The estimated total project investment (TPI) was \$79 million higher and the minimum fuel selling price (MFSP) (\$0.75 L^{-1}) was 22% higher than reported by Wright et al.
2000 dry t d^{-1}	Centralized	FP	Aspen hardwood	Renewable diesel and gasoline	The production costs of renewable diesel and gasoline were estimated through a TEA. The upgrading stages include two-stage hydrotreating followed by hydrocracking. Only the hydrogen purchase scenario is incorporated, with production costs estimated to be \$1.13 L^{-1} and \$1.08 L^{-1} for diesel and gasoline, respectively [40].
72 dry t d^{-1}	Centralized	FP	Pinewood	Gasoline and diesel	Two-stage hydrotreating with hydrocracking is used to produce gasoline and diesel, requiring capital investments of 23.7 M \$ and 9.5 M \$ annual operating costs. Bio-oil upgrading costs make up about 61% of the TPI, bio-oil production 39%.
2000 dry t d^{-1}	Centralized	HTL	Woody biomass	Diesel range fuels	This study assessed two scenarios, the state of technology (SOT) and the goal case. Two-stage hydrotreating and single-stage hydrotreating followed by

					hydrocracking were assumed for the SOT and the goal case, respectively. The MFSPs were estimated to be \$1.27 L^{-1} and \$0.72 L^{-1} [37].
2000 dry t d^{-1}	Centralized	HTL	Defatted microalgae	Transportation fuels	A TPI of 504 M \$ and an annual operating cost of 158 M \$ were estimated for co-existing hydrogen generation and an upgrading plant. Sensitivity analysis indicated that the primary bio-crude yield from HTL and the feedstock cost were the most influential parameters on the MFSP [38].
10 and 50 dry t d^{-1}	Decentralized	FP	Conifer tree	Bio-oil	Labor and capital equipment costs were found to have the most influence on bio-oil production cost. Selling biochar as a by-product could improve the overall economic performance [41].
50 dry t d^{-1}	Decentralized	FP	Forest residue	Bio-oil/Torrefied wood	Forest residues were converted to bio-oil or torrefied wood for ease of transportation to a biofuel facility from a distributed facility. This study found that relocating mobile facilities can lower delivery costs by reducing the transport distance of raw biomass [42].
100 dry t d^{-1}	Decentralized	FP	Pine wood chips	Bio-oil	The production cost of bio-oil was estimated to be \$1.12 L^{-1} with capital investments estimated to be 7.06 M \$ USD and operating expenses 3.86 M \$ y^{-1}. Feedstock cost, feedstock quality, and pretreatment of biomass (ash content) drastically influence bio-oil cost [43].
2000 dry t d^{-1}	Decentralized	FP	Canadian hardwood	Bio-oil	Multiple pyrolysis units of 10-100 dry t d^{-1} were used to reach the processing capacity. Bio-oil production costs of \$0.363 L^{-1} and \$0.423 L^{-1} were calculated

for two allocation scenarios (radial and truncated). Labor costs increased by 40-75% when plant capacities decrease from 100 to 10 dry t d^{-1} [44].

HTL and FP-based plants from the studies presented in Table 1.1 were analyzed based on a "cost-generating feedstock," i.e., corn stover, defatted algae, pinewood, aspen hardwood, etc. Feedstock costs from these plants range from \$48-87 t^{-1}, and these values significantly influence production cost [34, 37, 38, 40]. If feedstock costs can be shifted to a revenue source (by using MSW, for instance), production costs can be reduced. Other TEA studies that have explored the use of a revenue-generating feedstock focused on producing a different product (i.e., diluents, mixed alcohols, cellulosic ethanol, or energy through electricity), which limits the ability for a direct comparison.

Production costs also vary depending on the overall production configuration. Several studies have found that production cost is affected by the process (centralized or decentralized) [45-48]. Centralized or large-scale facilities are generally taken as fixed processing plants where a bulk amount of feedstock is transferred to be processed to an intermediate (bio-oil or bio-crude) and upgraded to the desired product. The conversion and upgrading facilities co-exist. This can offer significant savings through economies of scale, the multiplicity of operating units, and consistency in production. However, these facilities require large capital investments, are inflexible, and face difficulties such as feedstock unavailability and high feedstock transportation costs [45, 48]. Decentralized and distributed facilities, on the other hand, are of a smaller scale, mobile, and can use local waste streams [44]. Because of their relatively smaller size, these plants can be located at feedstock sites, requiring minimum costs in feedstock transportation and handling. That said, these facilities are often restricted by processing capacities and lack the benefits of economies of scale [42-44]. Upgrading operations do not co-exist with decentralized plants and the intermediate product needs to be shipped to a centralized upgrading plant.

In studies on the production of transportation fuels via direct liquefaction techniques, the scope is limited to centralized facilities, as seen in the review presented in Table 1.1. Other TEA studies have looked at the economics of deploying mobile pyrolysis units, but the studies' boundaries are limited to the production of primary bio-oil, not transportation fuels [41, 42, 44, 49]. Investigating both configurations – centralized and decentralized – to produce transportation fuels can help identify significant cost contributors and cost reducers. For instance, in a centralized configuration, because operations are immobile, the feedstock must be hauled over large distances to the facility, while in a decentralized scenario, plants can be located at feedstock sites and this transportation is minimized. Nevertheless, producing the intermediate product (bio-oil or bio-crude) in small distributed facilities would require the product to be transported to an upgrading facility. This could lead to significant differences in transportation costs due to distinct densities of feedstock and bio-oil, as shown in some studies [45, 50, 51].

Therefore, this research sets out to determine which alternative technology and configuration to minimize landfilling and reduce GHG emissions by converting MSW into transportation fuels is the most cost-competitive. Here, transportation fuels consist of blends of gasoline, jet fuel, and diesel. Besides investigating the technical aspect of each conversion process, the study also develops frameworks for production configurations and analyzes their impact on overall economics. In the process, this work also conducts the first TEA of TCR technology.

1.2 Research objectives

The contribution of this research is in the use of MSW to produce renewable transportation fuels. The overall aim of this research is to study the production technology, scale, and economics to identify the best strategies to establish feasible MSW conversion routes. To reach the aim, the specific objectives are to:

- Develop process models and process equipment scale factors to estimate capital costs of hydrothermal liquefaction (HTL), fast pyrolysis (FP), and intermediate pyrolysis (IP) based process plants.
- Conduct a TEA of TCR technology to better understand its limitations and opportunities for commercial use.
- Develop supply chain networks for the centralized and decentralized production of transportation fuels using HTL, FP, and IP techniques and MSW as a feedstock.
- Calculate production costs of transportation fuels (gasoline, jet fuel, and diesel) by converting MSW in commercial HTL, FP, and IP-based plants.
- Conduct sensitivity and uncertainty analyses to identify the key parameters that affect the economics of each process and assess the volatility of transportation fuel production.

1.3 Scope and limitations

1. The study used MSW composition data for Edmonton as a reference for Alberta. However, with a change in geographical location, the composition is likely to change. Therefore, to better estimate the local composition of MSW, data should be collected from multiple sources and authorities presiding at the location of interest.
2. The process models were developed under the assumption that the yield and quality of the products obtained at the lab scale could be replicated at commercial units. The developed models were validated by comparing the results from the models to the published experimental results from lab-scale set-ups.
3. In this study, the emphasis was on production costs and yield of transportation fuels from the processes analyzed. The quality index of the intermediates (bio-oil and bio-crude) was

considered based on the primary conversion techniques; however, final gasoline, jet fuel, and diesel qualities were considered to be similar.

1.4 Organization of the thesis

This thesis consists of four chapters along with a table of contents, list of tables, list of figures, and references. The thesis is presented in a paper-based format. Chapters 2 and 3 are expected to be published in peer-reviewed journals. Hence, the reader should expect some repetition of concepts and information.

Chapter 1, the current chapter, describes the background, objectives, scope, and limitations of the study, as well as the organization of the thesis.

Chapter 2 presents the developed process model and results of the economic evaluation of a decentralized intermediate pyrolysis plant. The effect of using biochar as a soil amendment on overall bio-oil production cost was also investigated.

Chapter 3 is a comparative techno-economic assessment of hydrothermal liquefaction (HTL), fast pyrolysis (FP), and intermediate pyrolysis (IP) plants' conversion of MSW to transportation fuels. The quality index of the primary bio-crude or bio-oil was considered and the extent of upgrading to reach transportation fuel specifications was compared. The chapter also describes the various influencing parameters and how plant configurations affect the overall production costs of fuels.

Chapter 4 is the conclusion and includes recommendations for future work.

Chapter 2 : Techno-economic assessment of the use of thermo-catalytic reforming to produce high-value products from municipal solid waste in remote North American municipalities

2.1 Introduction

Rapid urbanization, population growth, and changes in daily consumption habits have all contributed to the generation and diversity of municipal solid waste (MSW) across the world. Increases in the volume and non-uniformity of MSW have made its management a specialized and intricate activity. Hence, national, local, and regional authorities are challenged to develop frameworks and regulations for innovative waste management approaches. Authorities are encouraged to follow approaches aligned with the waste management hierarchy, i.e., reduce, reuse, recycle, recover, and only then dispose. However, due to evolving lifestyles, differing MSW characteristics, and lack of sustainable MSW conversion technologies, disposal to landfills remains the dominant option in many parts of the world [52, 53].

Canada, one of the North American country, alone reported the disposal of around 25 million tonnes of residential and non-residential waste into landfills in 2016 [6]. Due to the country's geographic and economic diversity, it is difficult to develop a uniform approach to recover energy or value-added products from MSW. Hence most of the municipalities follow the status quo of disposing their waste in assigned transfer stations or directly into community landfills, which are close to exceeding disposal capacities [8, 9]. A feasible solution could be to treat MSW locally and in small capacities through decentralized or distributed waste conversion facilities. The benefits of this approach include (a) reduced disposal and transportation costs associated with landfilling; (b) low investment requirements compared to large-scale waste conversion facilities; (c) making MSW an urban resource for social and economic benefits; (d) less operational pollution than large-scale conversion facilities [47]. However, before municipalities convert existing practices into decentralized ones, conversion routes, based on social, economic, and environmental impacts, need to be evaluated.

Waste conversion routes can be broadly divided into biological and thermal treatment. Biological treatments such as composting and anaerobic digestion (AD) are established pathways to process organic fractions of MSW. However, these methods cannot process inorganics such as plastics, glassware, textiles, etc., and so cannot completely replace regional landfills [4, 11, 54]. Thermal treatment such as incineration, gasification, and pyrolysis can be used to treat uncharacterized waste streams and are pathways for waste-to-energy (WTE) or waste-to-product (WTP) applications [10-12, 55]. Incineration is discouraged in many jurisdictions because it emits a high amount of pollutants (NO_x, SO_x, and dioxins) and particles [13-15]. These pollutants, along with CO_2, are a significant concern to the environment and human health. On the other hand, the current technologies for gasification of MSW produce synthesis gas (syngas), which requires a considerable amount of cleaning and conditioning prior to use and contributes significantly to operating expenses [16, 17]. Moreover, current gasification systems are not efficient in small-scale and often not economically feasible, given poor economies of scale [16, 17].

Pyrolysis, unlike gasification, yields several products, mainly pyrolysis gas, solid char, and an organic phase (bio-oil) when feedstock is heated in absence of air. Pyrolysis techniques can be divided into fast, intermediate, and slow depending on feedstock heating rate and residence time in the reactor. Moreover, pyrolysis-based plants are scalable, which increases their potential in small to moderate-sized municipalities [41-43, 56]. Fast and intermediate pyrolysis techniques have received more interest than slow pyrolysis because they can produce a liquid-dominant product stream. Although fast pyrolysis yields more organic bio-oil, this bio-oil is highly oxygenated, corrosive, and viscous, preventing direct applications as fuels into diesel or dual fuel engines [57, 58]. That said, bio-oil derived through intermediate pyrolysis has showcased improved and more nearby physical and chemical properties to diesel and biodiesel [24, 25, 59]. Intermediate pyrolysis is performed at moderate temperatures (400-450˚C) and heating rates (100-500˚C/min), with a longer feedstock residence time in the reactor than conventional fast pyrolysis [60, 61].

A recently developed approach to convert biomass and waste streams through a combination of intermediate pyrolysis and reforming is the thermo-catalytic reforming (TCR®) process [22, 23]. This emerging thermochemical conversion route can produce bio-oil with high carbon content, low water content, low oxygen content, and a high heating value [23, 62, 63]. A wide variety of feedstocks including MSW has been tested under intermediate pyrolysis conditions, yet the produced bio-oil does not have the same low levels of moisture, oxygen, and aromatics present in TCR-derived bio-oils [24-27]. Qualities such as high stability, low acidity, and higher carbon content of TCR bio-oil is because of the additional reforming stage, which promotes thermal and catalytic cracking of the initial pyrolysis vapors. Here, long-chained aromatics are broken down into lighter fractions because of secondary cracking reactions, allowing the TCR process to generate products with superior properties. This has been demonstrated in several-lab scale tests using various types of feedstock as described below.

Several lignocellulosic biomass and waste streams have been trialed to investigate their potential for TCR. These include woody biomass and waste streams such as oak tree chips, vine shoots, digestate, MSW, sewage sludge, co-form® rejects, paper sludge, etc. [22, 23, 29, 30, 64-67]. Conti et al. [65] tested woodchips, digestate, sewage sludge, and paper sludge in a laboratory TCR setup of 2 kg h^{-1} throughput. The generated bio-oil showed a higher heating value (HHV) of 35-38 MJ

kg^{-1}, irrespective of the feedstock, and the total acid number – a measure of acidity of the oil – was from 1.8-5.4 mg KOH g^{-1}. The ability of the process to convert a high-plastic waste stream was demonstrated by Ouadi et al. [67]. Co-form reject pellets, composed of 30% plastic polypropylene and 70% wood pulp fiber, were successfully processed in a lab-scale TCR 2 kg h^{-1} unit. The most abundant product was synthesis gas (71 wt.%); 12 wt.% in bio-oil was generated. The HHV of the bio-oil (39.36 MJ kg^{-1}) is comparable with biodiesel, and the bio-oil can be completely blended with diesel without phase separation. Ouadi et al. [30] obtained a similar highly stable and non-corrosive bio-oil in a lab-scale TCR test of uncharacterized MSW pellets. The authors reported the bio-oil to be completely miscible with diesel as well when they blended in 50:50 vol% ratios. The low oxygen content (6.7 wt.%) and total acid number (TAN) of 2.9 mg KOH g^{-1} allow the bio-oil to be stored for long periods in carbon-steel vessels instead of expensive stainless-steel storage vessels required for fast pyrolysis bio-oil. In addition, the process generated useful by-products such as hydrogen-rich (36 vol %) synthesis gas and biochar with good combustible properties.

Other waste-derived feedstocks such as digestate, pulper rejects, and sewage sludge have produced similarly enhanced quality products [22, 23, 68, 69]. Bio-oil yields of 4-10 wt.%, depending on the feedstock, were found; however, the deficiency of oxygenated and acidic components was evident. Compared to intermediate pyrolysis-produced bio-oil, the oxygen content in TCR-derived bio-oil and the corrosivity, which is measured by TAN, decreased by 67% and 92-95%, respectively [26, 27, 30, 33]. Therefore, despite the relatively low liquid yield compared to other pyrolysis-based conversion routes, TCR can be a promising conversion route if the quality of the generated products (bio-oil, biochar, and synthesis gas) is considered when the technology is evaluated.

The product quality index has been discussed in the literature, and several direct applications of the products, without the need for post-processing, are mentioned. TCR bio-oil can be directly blended with diesel or biodiesel for dual-fuel engines or directly used in boilers, combustors, and stationary CHP engines [30, 63, 67]. One of the by-products, biochar, can be used as soil amendment, filtration material (activated carbon), in the metallurgical industry, or combusted for energy production [63, 66]. Hydrogen from the other by-product, synthesis gas, can be separated to be used as a green fuel or for hydro-processing the primary TCR bio-oil. The ability of the TCR technology to process waste streams as well as generate multiple directly usable products makes it a promising candidate for decentralization in small municipalities.

Despite these advantages, TCR's technical and economic performance is yet to be demonstrated on a commercial scale. A detailed techno-economic assessment (TEA) of the technology in a decentralized application may lead to a solution. To the best of the authors' knowledge, there have been no TEAs done of a decentralized TCR plant, and no capital cost estimates have yet been reported for a North American jurisdiction. Earlier TEA studies focused on decentralized pyrolysis-based plants cannot be used here as those TEAs are based on solitary fast pyrolysis or intermediate pyrolysis technique [43, 44, 70-72]. Given the lack of TEAs on decentralized TCR, techno-economic models were developed to conduct an economic evaluation of a TCR plant with a capacity of 500 kg h^{-1} and located in a municipality in Alberta, Canada. The overall aim of this study is to determine the economic competitiveness of the TCR process through current and future market scenarios. The specific objectives are to:

- Develop a process model and determine the capital costs of a TCR plant with a capacity of 500 kg h^{-1}.
- Develop scale factors of the process equipment at various scales of a TCR plant.

- Estimate the production cost (\$ L^{-1}) of TCR-derived bio-oil and investigate the effects on bio-oil production cost when biochar is sold as a soil amendment.
- Investigate the effect of increasing plant capacity on TCR bio-oil production costs.
- Conduct sensitivity and uncertainty analysis to identify the key parameters that affect the economics of the process.
- Conduct a case study to estimate the feasibility for Western Canada.

2.2 Decentralized thermo-catalytic reforming (TCR) facility

For this study, a decentralized facility was defined as a convenient, easy to assemble and use, and portable plant [42, 44, 49, 73]. The terms decentralized and distributed facilities are used interchangeably in this work. It is assumed that a decentralized facility is mobile and can be located in rural and remote locations, requiring minimal regional infrastructure to implement. However, the facility will likely face capacity limits depending on the technology used. Patel et al. [44] reported mobile pyrolysis unit capacity limits in decentralized applications of between 10 and 100 dry t d^{-1}. The capacity of a TCR unit of 500 kg h^{-1} (12 dry t d^{-1}), hereafter referred to as TCR-500, is at the lower end of that range and is hence assumed to be suited for a distributed application.

A TCR-500 plant based on its ability to process disposed MSW for a municipality of 5000 residents is considered the base case and has been evaluated. The rationale for the municipality size and amount of MSW available is given in Table A-1 and Figure A-1 of the appendix A. The facility is assumed to replace an existing local landfill. Road networks and infrastructure required to transport MSW from households to transfer stations and from thereon to the TCR-500 facility are already established, and so this analysis does not include transportation costs. As-received MSW is first sorted from recyclables, electronics, and household hazards at transfer stations to meet the composition required to produce pellets. MSW can only be fed to the TCR unit in the pellet form.

The MSW undergoes additional manual sorting in the plant to remove any bulky objects and is inspected for any remaining electronics or hazardous material before being passed into the grinder. The overall schematic flow diagram is given in Figure 2.1.

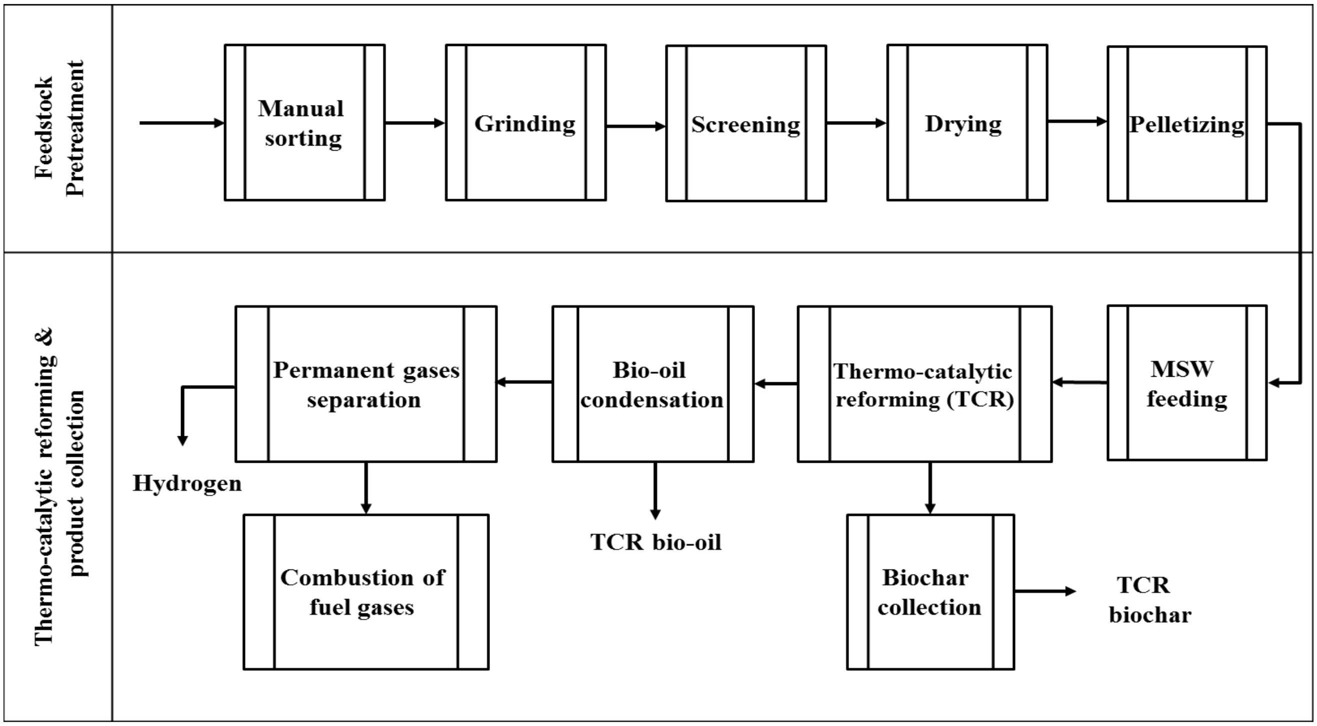


Figure 2.1: Schematic diagram of a TCR-500 plant

2.3 Method

2.3.1 Feedstock

Municipal solid waste is used as the feedstock in the TCR plant. As described above, the received MSW stream is assumed to be free of recyclables and have an initial moisture content of 25% [74]. The as-received MSW feed undergoes several pretreatment stages (manual sorting, grinding, screening, drying, and pelletizing) to make it compatible for reactor feeding. The proximate and ultimate analysis values used to define the feedstock were adopted from Edmonton waste management facility and are presented in Table 2.1.

Table 2.1 Proximate and ultimate analysis of municipal solid waste samples [74]

Parameter	Unit	Value
Moisture	wt.%	24.8
Proximate analysis		
Ash	wt.%	16.5
Volatile	wt.%	70.1
Fixed carbon	wt.%	13.4
Ultimate analysis		
Nitrogen	wt.%	1.1
Carbon	wt.%	45.8
Hydrogen	wt.%	7.5
Sulfur	wt.%	0.4
Oxygen	wt.%	28.7

2.3.2 Process modeling

A schematic of the process model developed to evaluate the thermo-catalytic reforming of MSW pellets is shown in Figure 2.2. The model boundaries are from the reception of the sorted MSW to the production of primary TCR products, i.e., bio-oil, biochar, and hydrogen. MSW and biochar are defined as non-conventional solids, and a particle size of 8 × 35 mm is defined by Ouadi et al. [30]. The process model has three major sections: 1) feedstock pre-treatment, 2) thermo-catalytic reforming, and 3) product separation and collection.

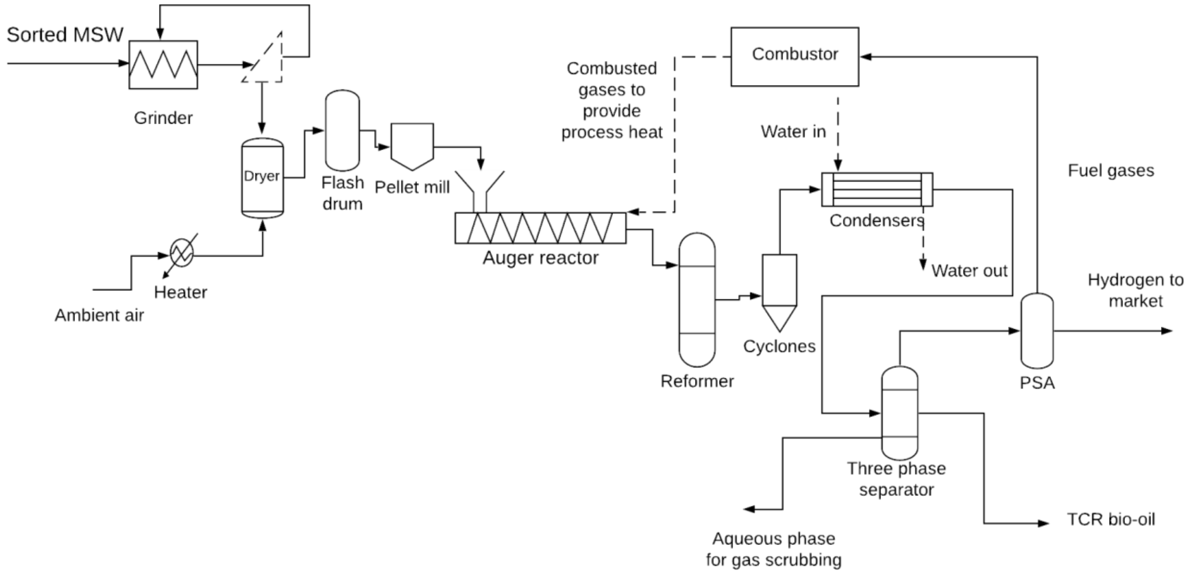


Figure 2.2: Process model flowsheet of TCR-500 plant

2.3.2.1 Feedstock pre-treatment

Incoming MSW is assumed to be sorted from recyclables and household hazardous waste at transfer stations. The feed stream undergoes a secondary manual sorting at the plant site to remove large bulky objects such as glass bottles and concrete blocks. The stream is then passed through a gyratory crusher (grinder) with a mechanical efficiency of 95% to grind materials to 20 mm – 25 mm particle sizes. A circular vibratory separator is used to screen oversized materials and recycle them to the grinder. The outlet from the grinder is connected to a counter rotary drum dryer for thermal treatment. The feedstock is dried at 270°C using air to achieve a final moisture content (MC) of 10% by wt. from the as-received state (25% MC). A flash separator is used to isolate steam from dried MSW. The dried MSW stream is pelletized in dimensions of 8 mm × 35 mm prior to being fed to the reactor using a pellet mill.

2.3.2.2 Thermo-catalytic reforming

Thermo-catalytic reforming (TCR) is the integration of two core processes; intermediate pyrolysis and post-reforming. Intermediate pyrolysis takes place in an auger reactor operated between 400˚C and 450˚C in inert conditions and atmospheric pressure [22, 23, 29]. An internal screw mechanism allows the user to adjust the pellet feed rate and residence time in the reactor and to avoid blockages. The MSW pellets are conveyed along the reactor length by means of internal auger screws. At these high pyrolysis temperatures and a moderate residence time of 15-30 minutes thermally degrade the pellets. Several decomposition and chain reactions occur simultaneously to break down cellulose, hemicellulose, and lignin constituents [32, 33, 75]. This is followed by the initiation of secondary cracking reactions that occur because of the longer residence time of the pellets inside the reactor. The resultant pyrolysis vapors and biochar are transferred to the reforming unit. The auger reactor is modeled using mass and compositional yields from the gas chromatography-mass spectrometry (GC-MS) results presented in Yang et al. [27]. Minor adjustments are made to establish mass balance closure.

The reforming unit is a cylindrical vessel operated at elevated temperatures of 600°C – 700°C and at atmospheric pressure. The unit allows pyrolysis vapors and biochar to stay in contact for long periods and promotes thermo-catalytic activity. Secondary cracking is maximized in the reforming unit due to presence of char [28]. During the TCR operation, biochar accumulates at the bottom of the vessel to form a packed biochar bed. As the pyrolysis vapor passes into the biochar bed, it is exposed to further thermal degradation and catalytic cracking through the high temperatures and the presence of char. Long-chain hydrocarbons are broken down into lighter fractions of C_6-C_9 by secondary cracking; therefore, the organic fraction (bio-oil) does not have any waxy material (usually formed by aromatic and long-chained compounds) [32, 33]. Moreover, reforming reactions, which include steam methane reforming and water-gas shift reactions, favor high synthesis gas yields. The yield and composition of products are adopted from GC-MS results presented in an earlier study [30] to model the product distribution from the reformer block. While the GC-MS results represent many compounds, the relevant compounds considered for simulation purposes are shown in Table 2.2. Minor adjustments to the yield of components were made to establish mass balance closure.

Table 2.2: Product composition and yield from the thermo-catalytic reforming of MSW pellets [30]

Compound	Mass fraction (wt.%)
Bio-oil composition	
Ethylbenzene	0.24%
Xylene	0.16%
Styrene	0.97%
1,3,5,7-cyclooctatetraene	0.26%
Benzene	0.36%
Phenol	0.54%
1h-indene,-3-methyl-	0.08%
1-phenylnaphthalene	0.15%
Cis-1-propenylbenzene	0.17%
1-propynylbenzene	0.10%
Naphthalene	1.56%
Diphenyl	0.12%
Fluorene	0.19%
Phenanthrene	0.36%
Pyrene	0.31%
Methyl-stearate	0.20%
Acenaphthalene	0.15%
Indane	0.08%
Permanent gases composition	
Hydrogen	2.04%
Carbon-dioxide	26.13%
Carbon-monoxide	9.50%
Methane	6.33%
Water	19.00%
Biochar	31.00%

2.3.2.3 Product separation and collection

2.3.2.3.1 Biochar

The hot reformed outlet stream is immediately cleaned of any dust and particles with a series of cyclone separators at a solid separation efficiency of 95% [76, 77]. Micro-sized dust or particles can pollute the bio-oil and affect downstream operations. The biochar is continuously extracted from the bottom of the reforming unit using a screw conveyor system. The separated biochar is collected and allowed to cool to prevent unanticipated combustion of the hot char. Several studies have described the material use of char as a soil amendment, filtration material, and feed additive and estimated a selling price based on its use [66, 78-80]. For the current study, the biochar collected is considered a revenue-generating by-product; therefore, no further processing is modeled.

2.3.2.3.2 Bio-oil

Clean reformed vapors, free of char and dust, are quenched immediately to separate permanent gases and condensables. The quenching system, equipped with cooling water as the medium, is a shell and tube heat exchanger train that collectively reduces the vapor temperature from 700˚C to a minimum of 30˚C in three stages. The temperature of the successive stages can be changed to separate the heavy and light fractions of the produced bio-oil. The final stage operates at the minimum temperature to phase out any remaining condensables from the permanent gases. Fractions of oil are collected from all stages and vary in density and property depending on the compounds present. All fractions of bio-oil are mixed to maximize overall yield. The deficiency of oxygenated and acidic compounds in the bio-oil makes it suitable for long-term storage without any phase separation or aging effects [23, 27, 30]. The enhanced properties indicate that TCR bio-oil qualifies for a wide variety of applications, including blending with biodiesel for CHP applications, blending with diesel for diesel engine fuels, or deoxygenating the bio-oil to transportation grade fuel [22, 23, 30, 63, 71, 72]. In the current study, the TCR bio-oil is considered to be used as a drop-in fuel in diesel-engine community vehicles such as farming trucks or dual-fuel engine applications.

2.3.2.3.3 Hydrogen

The composition of the permanent gas stream is highly dependent on post-reformer operating temperatures. At higher temperatures, the water-gas shift reaction, which converts water and CO to form H_2, is favored. The steam methane reforming reaction is also favored at high temperatures due to its endothermic nature. In this reaction, methane is broken down into CO and H_2 in the presence of steam at high temperatures. Both of these reactions contribute to the overall hydrogen yield in the permanent gases. Hydrogen from biomass or other renewable sources is of interest because it can be used as a clean fuel and in refineries for crude oil upgrading [81-88]. A pressure swing adsorber (PSA) with 80% recovery efficiency is used in downstream operations to adsorb the generated hydrogen [37, 89]. The remaining flue gas is recycled to provide heat to the overall process.

2.3.3 Techno-economic analysis

The economic model was developed using spreadsheet-based calculations. The key assumptions taken for the model are presented in Table 2.3.

Table 2.3: Key assumptions for techno-economic model development

Parameter	Value	Comments/Remarks
Plant lifetime (years)	20	Assumed

Plant operation (hours)	8000	[90]
Base year	2018	
Currency	USD	
Plant construction factor		Taken relative to factors used for large-scale plants [39, 91, 92]
Year -2	40%	
Year -1	60%	
Plant startup factor		Taken relative to factors used for large-scale plants [39, 91, 92]
Year 0	0.85	
Year 1	0.90	
Year 2	0.95	
Internal rate of return (%)	10	Assumed
Maintenance cost ($)	2.5 % of TPI	[71, 72]
Operating charges ($)	25% of operating labor cost	Values are obtained from multiples sources [40, 90, 93]
Plant overhead	2% of TPI	[71, 72, 91]
Subtotal operating cost, SOC ($)	Sum of all operating costs including raw material and utility cost	
General and Administrative (G&A) cost ($)	5% SOC	[94, 95]

A techno-economic model was developed. A discounted cash flow analysis (DCFA) was conducted to calculate the breakeven price of the TCR bio-oil. The breakeven price can be defined as the amount of money for which a product or service must be sold to cover the costs of manufacturing or providing it. Revenue sources to the plant include the sale of products (biochar and hydrogen), gate fees, and carbon credit received for the use of MSW as the feedstock. It is estimated that using one ton of MSW is equivalent to abating two tons of CO_2 from the atmosphere [96]. The plant's expenses include total capital costs, operating and labor costs, maintenance costs, utilities, plant overhead, and general and administrative (G&A) costs.

2.3.3.1 Capital cost

The total capital cost was calculated for the TCR plant with the method developed by Peters and Timmerhaus [97]. In this approach, the baseline equipment cost for each unit operation is first calculated to determine the total project equipment cost (TPEC). Investment factors are then used to integrate installation costs, indirect costs, contingency, and location factors, etc., to determine the total project investment (TPI). The investment factors used are provided in Table 2.4. Baseline equipment costs are estimated by using process model-based cost estimates for conventional process equipment (i.e., shell and tube heat exchangers, three-phase separator, PSA, etc.) and derived from the literature using the cost-capacity equation (1) for new equipment (i.e., the intermediate pyrolysis reactor and reformer). A mass and energy network balance is first established for the developed process model. Unit blocks are then mapped and sized to commercial process equipment based on process model results, design, and operating parameters to estimate the baseline equipment cost. For new equipment, baseline costs are estimated by the cost capacity method shown by Equation 1:

$$C = C_o \left(\frac{S}{S_o}\right)^f \qquad (1)$$

where C, the estimated cost, is determined by multiplying C_o, the reference cost, with the ratio of sizes (S/S_o) raised to the power of f, an empirical scaling factor. For each piece of equipment, the scale factor is derived from either published sources or vendor quotes. The equation is then used to scale the costs to the desired capacity used in the plant. The scale factors used for individual pieces of equipment can be found in Table A-2 in the appendix. Installation factors for instrumentation, piping, electrical, buildings, and service facilities are incorporated with the TPEC to estimate the total installed cost (TIC) of the plant. Other indirect costs, such as engineering and supervision, construction expenses, and legal expenses are integrated as a function of the TPEC as well.

Table 2.4 Factors for estimating plant capital cost [97]

Parameter	**Value**
Total purchase equipment cost (TPEC)	100% TPEC
Total installed cost (TIC)	207% TPEC
Indirect cost (IC)	30% TPEC
Total direct and indirect cost (TDIC)	TIC + IC
Contingency	20% of TDIC
Fixed capital investment	TDIC + contingency
Location factor (LF)	10% FCI

Parameter	**Value**
Total capital cost	FCI + LF

2.3.3.2 Operating and maintenance cost

Among the major contributors to the operating and maintenance costs are labor and utility. The labor cost was calculated based on the total yearly operational hours and the number of shifts. It was assumed that each shift has two plant operators and used an hourly Canadian wage of $32 h^{-1} ($ 24 USD h^{-1}) to calculate annual operating labor costs [98]. The maintenance cost for the plant was assumed to be 2.5 % of the capital cost given its low capacity compared to other conventional biomass plants [71, 72, 91]. The plant operating charges were taken as 25% of the operating labor cost, whereas the yearly plant overhead was projected to be 2% of the total capital cost.

Utility costs include electricity and water consumption in the TCR plant. The electricity use in the plant is for office space, lighting, and electrical appliances. Unlike in a lab-scale TCR setup, the TCR-500's reactor and reformer are equipped with channels that are thermally heated, which significantly reduce the plant's overall electricity consumption. The high fuel gas fraction is combusted and recycled to provide the required heat to the reactor and reformer during continuous plant operation. The equilibrium stage is reached through an initial ramp-up by combusting an alternative fuel such as propane [44, 49]. The electricity consumption is taken to be 28 kWh t^{-1} of wet treated MSW [72]. This assumption is used due to the close similarity in arrangements between intermediate pyrolysis and TCR-based plants. Cooling water demand is estimated as a function of the feedstock processed, as found in economic assessments of commercial pyrolysis plants [71, 72, 91]. In this work, the cooling water consumption is assumed to be 20 m^3 t^{-1} of dry MSW and a commercial rate of \$0.95 m^{-3} is used [99].

2.3.3.3 Feedstock cost

Using MSW as a feedstock qualifies the TCR plant to receive gate fees. A gate fee is a charge levied on the waste coming into a waste conversion or waste management facility [8, 100]. An average gate fee of \$80 t^{-1} of MSW was assumed based on the disposal fees of Edmonton's Waste Management Centre [101]. An operating charge of \$20 t^{-1} was used to account for the sorting of recyclables and manual removal of bulky objects at the plant site. An overall revenue stream of \$60 t^{-1} was taken. The gate fee is likely to vary depending on the location of the plant and the effect of this variation was investigated in the sensitivity analysis.

2.3.4 Product revenue

Products from the thermo-catalytic reforming (TCR) process are of a higher quality than those from other thermochemical routes and so have several direct applications and can generate revenue. Using data from the literature, each product stream was allocated a selling price in current and future market scenarios. The TCR bio-oil production cost is the variable to be calculated each time the model is run and therefore no fixed price is assigned.

Using biochar to generate energy through combustion is widely reported in the literature [39, 102-104]. However, many researchers are interested in the material use of biochar [66, 78-80]. Applications of biochar as a soil amendment, filtration material, feed additive, and raw material for fertilizer production can lead to more revenue than combusting it for energy production. Moreover, although the standards for land applications are stringent, biochar generated through TCR is suitable because of its high carbon content and low concentrations of heavy metal and organic pollutants [66, 105]. Despite the advantages of the direct application of biochar, standard prices are not yet set and thus vary significantly according to their end use and the location of application. The price of biochar used as filtration material to absorb fine bio-solids is reported to be from (CAD) \$1500 to \$2000 t^{-1} in 2019 [78, 106]. The price of biochar, as a soil amendment (the application assumed in this study), is reported to be between (USD) \$80 t^{-1} and \$500 t^{-1} for the year 2015 [78-80]. The discrepancy and uncertainty in Canada's current biochar market do not allow a fixed price to be set on biochar. Therefore, a base biochar price of \$400 t^{-1} was assumed to integrate both the quality of TCR-generated biochar and its material use. However, to incorporate the likelihood of a developing and mature biochar market where prices might fluctuate, values from \$100 t^{-1} to \$700 t^{-1} were investigated.

The hydrogen recovered from produced synthesis gas can be used as fuel for hydrogen fuel cells or used to further upgrade the TCR bio-oil through hydro-processing. The extracted hydrogen can be introduced in the biomass-derived hydrogen (bio-hydrogen) market, as it is derived from a renewable source. However, for it to compete in other bio-hydrogen routes, the TCR-hydrogen should be sold at prices comparable to the production prices of bio-hydrogen from alternative routes. Studies on the production of hydrogen from biomass or bio-oil report production costs of \$1- \$5 kg^{-1} of H_2 [84, 107-111]. For the current study, we assumed a hydrogen selling price of \$5 kg^{-1}.

2.3.5 Sensitivity and uncertainty analysis

A global sensitivity analysis was performed using the Regression, Uncertainty, and Sensitivity Tool (RUST) to determine the influence of parameters on the production costs of the TCR bio-oil. Using the same tool, a Monte Carlo simulation was performed using 10,000 iterations to measure the uncertainty in the production cost.

2.4 Results and discussion

2.4.1 Process model results for TCR-500

The technical process model was integrated with the spreadsheet-based calculations to complete the flowsheet presented in Figure 2.2. The model for the TCR-500 was developed and validated by linking each block operation and establishing mass and energy balance between them. Simulation runs were conducted and the results were compared and validated with Ouadi et al.'s experimental results [30] and are given in Table 2.5. The pyrolysis unit and the reformer were identified as the main energy consumption units due to their high temperature maintenance and continuous operation. Fuel gases stripped of hydrogen are combusted and recycled to provide adequate heat to maintain operating temperatures.

Table 2.5: Experimental and simulation results

Product yield	Mass distribution ($kg\ h^{-1}$)	Experimental results [30]	Process simulation results
Biomass	499.6		
Organic (Bio-oil)	30.71	6%	5.9%
Char	154.87	31 %	30.1%
Permanent gases*	209.53	44%	43.6%
Hydrogen	10.17		
Aqueous phase	94.29	19%	16.8%

* excluding hydrogen

The overall composition of the pyrolysis vapors indicates heterocyclic, benzene-based compounds, and long-chain aliphatic compounds. It is understood that, due to the prolonged solid and vapor residence time in the reactor and the reformer, secondary reactions such as cracking and reforming are favored, and these result in the formation of more stable chemicals in TCR bio-oil compared to other pyrolysis oils [27, 30]. At higher reforming temperatures, the liquid yield was observed to decrease, and the formation of synthesis gas is highly favored. This is due to the increased catalytic activity of char and the existence of favorable conditions for steam methane reforming and water-gas shift reactions [28]. Moreover, the compositional analysis of the char-reformed pyrolysis oil shows a great number of light hydrocarbons in C_6-C_9 and fewer components with a carbon number greater than 20.

A three-phase separation unit is used to isolate the TCR products into streams of permanent gases, bio-oil, and an organic-deficient aqueous phase by keeping temperatures at a minimum. The bio-oil in this case is collected as the top layer of the two immiscible liquid fractions, that is, the bio-oil and aqueous phase; however, this may change depending on the feedstock used and the density of the produced oil. The permanent gas stream is passed into the PSA to separate the pure hydrogen. Concentrated sulfuric acid can be used as a dehydrating agent for any moisture present in the flue gas; however, no cleaning agent was assumed to be required in the PSA for this low-capacity unit.

2.4.2 Economic analysis

Important outcomes of the economic analysis of a TCR-500 plant are given in Table 2.6. The total capital investment required was calculated to be 3.9 M \$ (USD) and the operational charges were 1.7 M \$ (USD) y^{-1}. A capital cost breakdown is given in Figure 2.3. As small-scale pyrolysis plants do not benefit from economies of scale, the capital investment is affected mostly by high purchase equipment costs and installation. This observation was found in other cost assessment studies of small-scale pyrolysis systems [44, 71, 72]

Table 2.6: Summary of costs for TCR-500 as developed through the techno-economic model

Parameters	Value
Total purchase equipment cost	\$1,245,899
Total capital cost	\$3,897,670
Total operating cost (y^{-1})	\$1,733,478
Operating labor cost (y^{-1})	\$950,567
Maintenance cost (y^{-1})	\$77,953
TCR bio-oil production ($L\ y^{-1}$)	394,322
Production cost (\$ L^{-1})	\$2.01

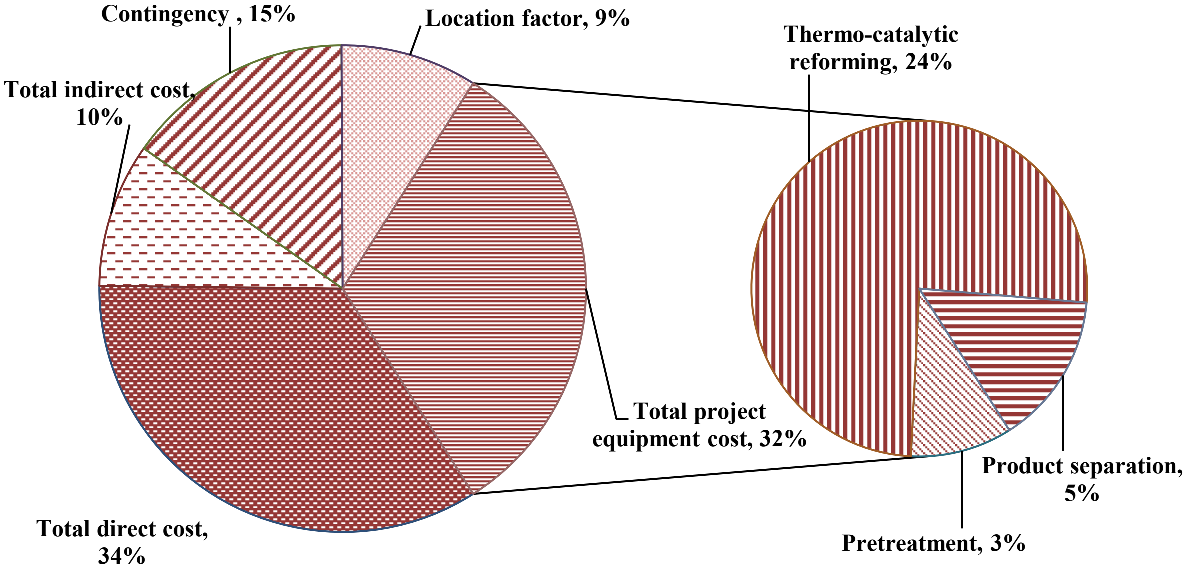


Figure 2.3: Capital cost breakdown of TCR-500 plant

The most significant contributor to the total equipment cost is the auger pyrolysis reactor. Commercial auger reactors are still under development at the time of this study, and improvements in heat transfer mechanisms were identified as critical pathways of transitioning to larger scale reactors [71, 112, 113]. Studies to overcome the heat transfer barriers are being conducted and some solutions have been proposed. These include using internal solid heat carriers (SHC) such as steel balls, using biochar along with the feed, and integrating a secondary internal auger to carry SHC [112, 113]. These solutions are expected to be developed and applied in commercial auger reactors in the near future, and manufacturing costs are expected to fall considerably.

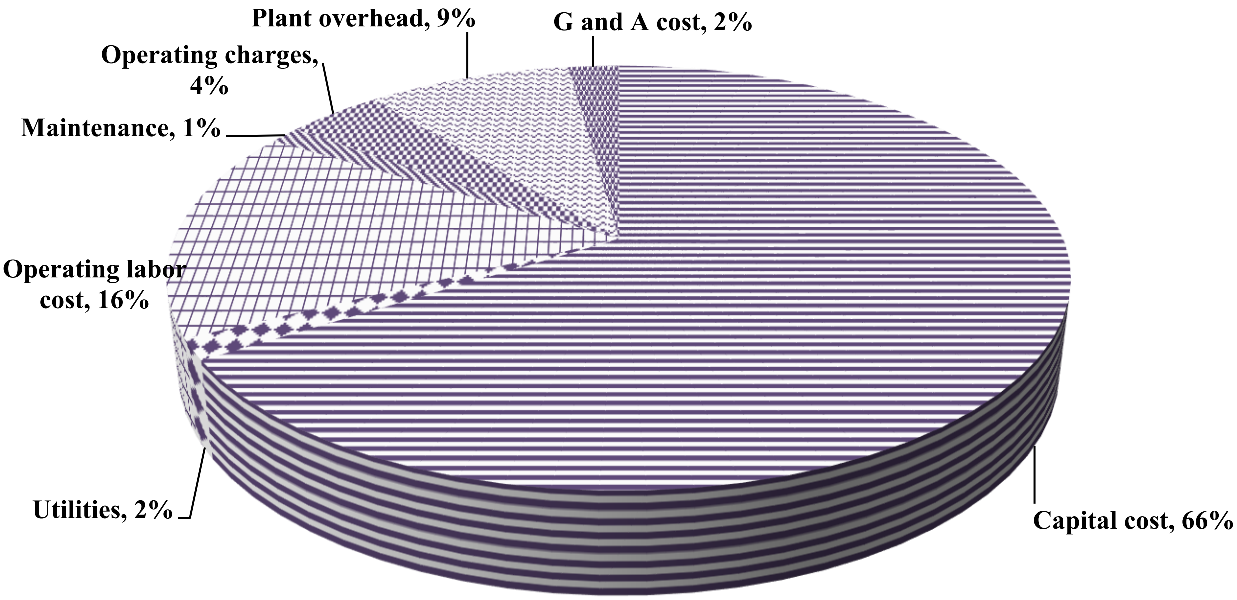


Figure 2.4: Production cost distribution for TCR-500 Plant

For the base case TCR-500 plant, the bio-oil production cost was calculated to be \$2.01 L^{-1}. The overall breakdown of various cost contributors to the production cost of bio-oil (see Figure 2.4) suggests that TCR plants can benefit from reducing capital costs. A reduction in manufacturing and fabrication costs of new equipment used in this process can drive this positive change, as discussed above. Other cost reduction opportunities can be found in operating expenses, such as labor costs. Although wage rates differ from one location to the next, the contribution of labor cost is likely to remain significant at smaller scales. Other studies of small-scale pyrolysis systems report similar (significant) contributions of labor costs to overall operating costs [44, 71]. Advancements in automated systems and auxiliary equipment, which may require less manual supervision, can lead to savings in small-capacity decentralized plants. Using MSW or other waste streams as feedstock ensures considerable operational savings, as feedstock-related expenses are limited to pretreatment processes. However, for a biomass-based feedstock such as wood chips, operating costs for TCR plants are higher due to associated raw materials costs. The utility cost share of overall production is considerably low with respect to the use of flue gases to provide process heat, which reduces electricity or natural gas use and improves the overall energy efficiency of the process.

2.4.3 Sensitivity analysis

2.4.3.1 Influence of plant capacity

The benefits of economies of scale are that they can reduce average production costs when large equipment is used. However, large-scale facilities compromise mobility, simplicity, and decentralized applications. Pyrolysis-based plants are considered to be mobile at capacities of 10 to 100 dry t day^{-1} [41, 43, 44, 49, 73]. Although TCR plants have similar arrangements, when considering scale up an account must be taken of the upper limit of reactor capacity of a single unit. Auger reactors, the type of reactor used in TCR, are promising for their low or no requirements of carrier gas and low energy [112]. However, studies have raised concerns over the technical feasibility of scaling them up. The desired heat transfer rate is difficult to achieve in larger reactors as the ratio between reactor surface area and reactor volume continues to decrease [71]. Dimensional analysis also concluded that there is no straightforward scaling up strategy [114]. 500 kg h^{-1}, therefore has been assumed as the upper limit of a single TCR reactor. For TCR plants exceeding this capacity, multiple units are used.

The effect of increasing plant capacity was investigated from 500 kg h^{-1}(12 dry t d^{-1}) to 4000 kg h^{-1} (100 dry t day^{-1}), by multiplying the number of TCR units. The results are shown in Figure 2.5.

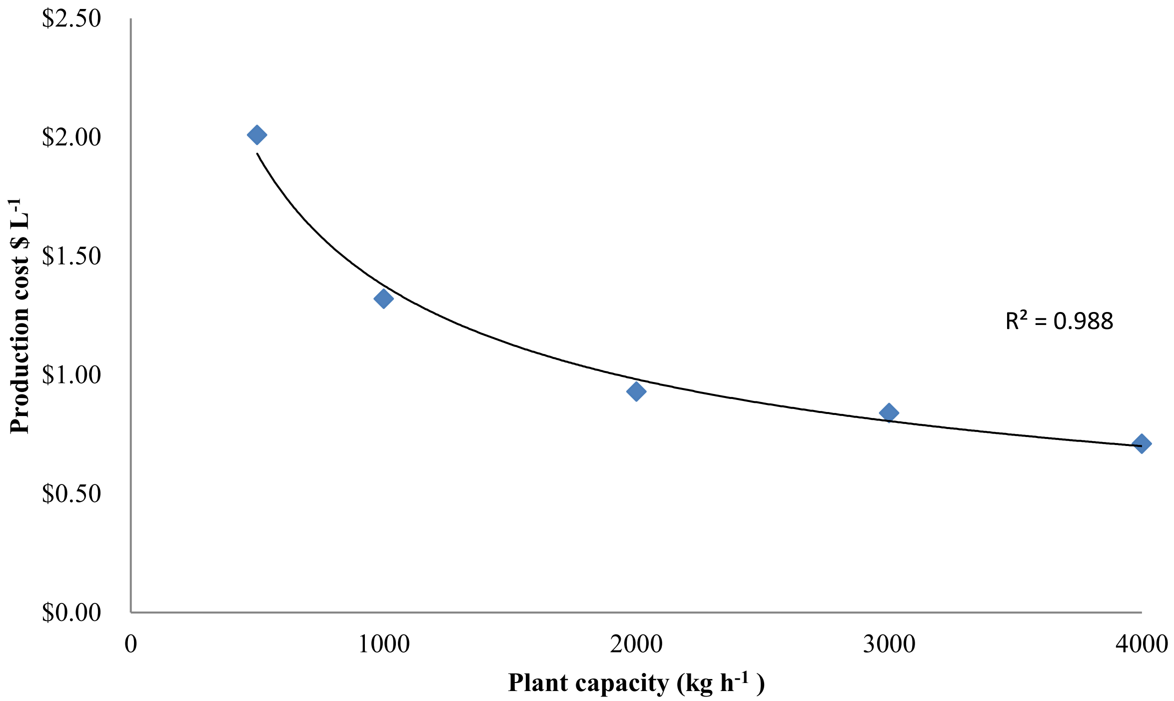


Figure 2.5: Influence of plant capacity on the production cost of bio-oil

The decrease in the production cost of TCR bio-oil was found to be strongly correlated with increased processing capacity. Production costs were found to decrease by 65% when overall production capacity was scaled up by combining multiple TCR units to process feed at 4000 kg h^{-1}. The reductions in production costs, however, were due to the nonlinear relationship between labor requirements and plant scale. Although multiple reactors within the same plant may share some common auxiliary equipment, the capital cost reduction was mild due to the absence of economies of scale. Higher revenue from increasing product sales and gate fees due to the processing of more MSW further contributed to the reduction. Labor requirements per unit of product can be reduced at larger scales as automated equipment is integrated into the plant. However, any further increase in plant capacity, although it might reduce production costs, will likely limit its decentralization capability. As it does not align with the scope of the study, plant capacities greater than 4000 kg h^{-1} were not looked at.

2.4.3.2 Exploration of key parameters

To better understand the effect on production cost from each parameter, a sensitivity analysis was performed using the Regression, Uncertainty, and Sensitivity tool (RUST) [115]. The results are shown in Figure 2.6. A basic one-at-a-time (OAT) sensitivity analysis evaluates the effect at the maximum and minimum values of each parameter, while the remaining is kept at their base values. As a result, it fails to capture any interaction between the input parameters. To overcome this limitation RUST integrates the Morris method to account for the primary and interaction effects between the parameters. As shown in the Morris plot (Figure 2.6), the horizontal axis representing the Morris average (μ) is the average change in the output as the input varies from is minimum to a maximum value. A higher μ represents the parameter being sensitive to the output, in this case, the production cost of bio-oil. The vertical axis represents the Morris standard deviation (σ) of output when input varies from its minimum to maximum values. If an input has an SD of zero, then a change in its value will have the same response across the entire parameter space Sensitivity increases as you move to the upper right corner of the plot.

The accuracy of an estimate depends on the availability of critical information such as the amount of design detail, accuracy of cost data, firm quotes for equipment, etc. Preliminary estimates or feasibility studies such as produced in this work are typically accurate to ±30% [90]. Other TEA studies have shown, that variation of parameters in the range of ±30% has been sufficient to determine key cost-driving parameters [39, 92, 116]. Parameters except the biochar price were varied by ±30%. Biochar price was varied by ±75% due to the large price uncertainty in the current North American market.

Based on the screening of the input parameters, bio-oil yield, gate fees, labor costs, biochar, and hydrogen prices were found to be most interactive with the production cost. Improvements and optimization in process design, operating conditions, and equipment efficiency are expected as the technology matures which could boost the bio-oil yield. However, attention should be given to not compromise bio-oil quality over quantity, as it would limit the firsthand applications. Other ways to increase bio-oil yield include using refined feedstock with low ash content; however, this may incur additional processing costs.

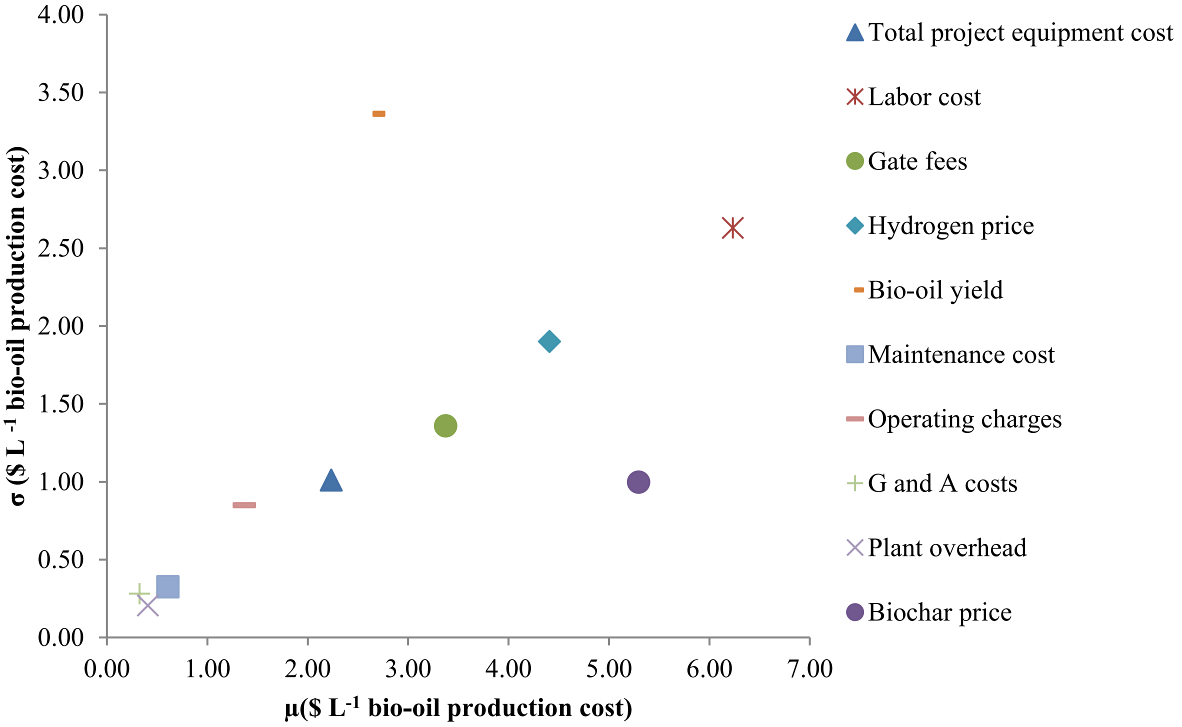


Figure 2.6: Sensitivity analysis of the key parameters in the production of TCR-500 bio-oil

The gate fees received by a TCR plant vary across locations. For remote municipalities, gate fees are likely to be higher than in cities due to the limited infrastructure and accessibility of waste disposal.

Labor cost was found be to the most sensitive operating expense in TCR bio-oil production in Western Canada. It was assumed that a minimum of 2 operators for a small-scale plant. As the number of labor requirements does not follow a linear correlation with the overall plant scale, the effect of plant operator wages and salaries is likely to be less sensitive with increased base plant capacity. Hence the contribution of labor costs to the overall production cost of bio-oil is likely to decrease.

2.4.4 Uncertainty analysis

The uncertainty analysis was performed for the TCR-500 base case by running a Monte Carlo simulation using the RUST tool. In a Monte Carlo simulation, random values are chosen, in this case from all the parameters that impact the production cost of bio-oil. Parameters were not screened out from the previously done sensitivity analysis as all inputs showed interactions with the output, i.e., production cost of the bio-oil. Cost parameters considered include capital cost, operating labor cost, maintenance cost, utilities, operating charges, general and administrative cost, and plant overhead. The limits of uncertainty for each parameter are taken the same as in sensitivity analysis.

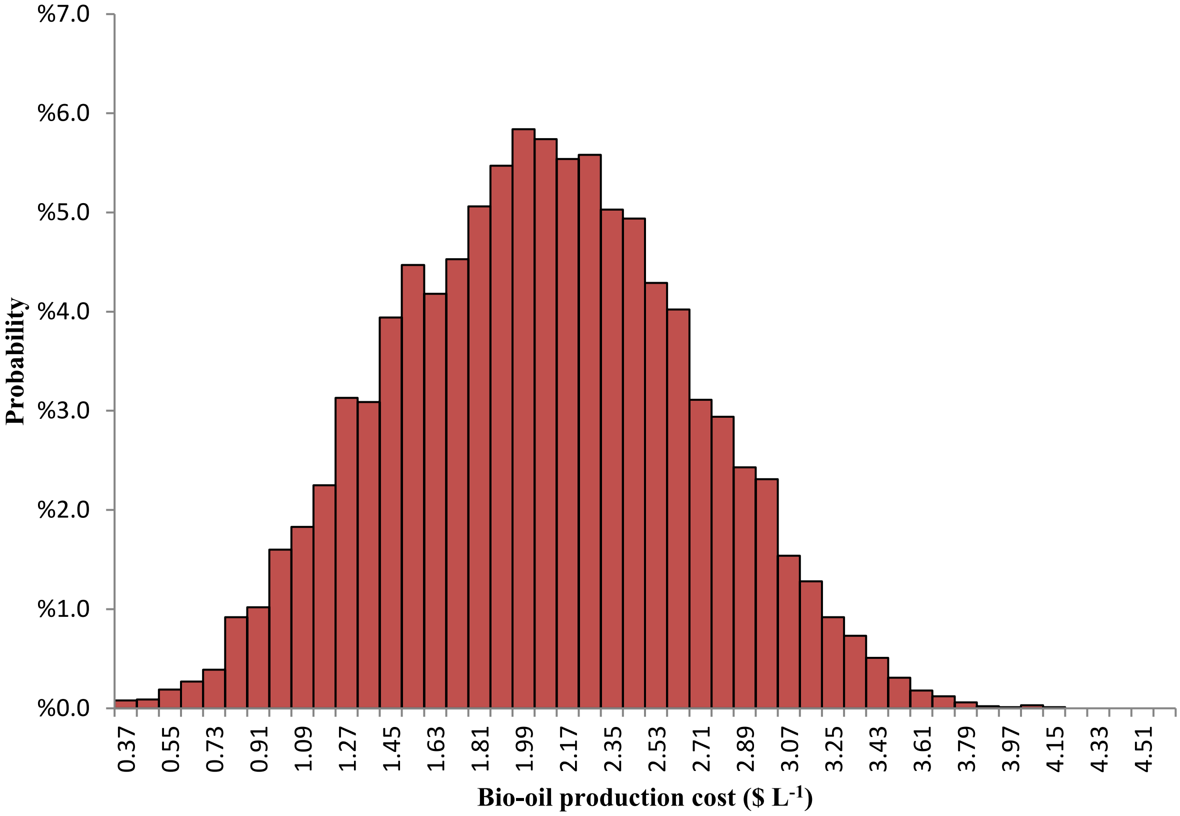


Figure 2.7: Uncertainty analysis of TCR bio-oil production cost

The results of the Monte Carlo simulation are presented in the form of a histogram in Figure 2.7. The production costs of bio-oil in the base case scenario were calculated to be \$2.01±0.79 L^{-1}.

2.5 Comparison of production cost with other studies

This techno-economic study is the first of its kind conducted on the concept of thermo-catalytic reforming. TEA studies conducted on intermediate pyrolysis systems so far are centralized and integrated with combined heat and power (CHP) to combust the primary bio-oil, therefore, bio-oil production costs are not disclosed [71, 72]. However, production costs in decentralized pyrolysis plants, predominantly using fast pyrolysis, have been reported in several studies [41, 43, 44, 49, 73]. Badger et.al [43] reported bio-oil production costs of (USD) \$0.30 L^{-1} (adjusted for inflation to the year 2018) for a 100 dry t d^{-1} mobile fast pyrolysis plant processing pine wood chips. The total pyrolysis oil yield considered was 60% of the feed. The production cost was drastically influenced by bio-oil yield, plant capacity, and feedstock cost as suggested by the authors. Another study that investigated the economic feasibility of decentralized pyrolysis units in New Mexico concluded that a base plant scale of 50 dry t d^{-1} performed significantly better than 10 dry t d^{-1}. The production costs were estimated to be (USD) \$0.26 L^{-1} for 50 dry t d^{-1} and (USD) 0.81 L^{-1} for 10 dry t d^{-1} plant scales, respectively. Palma et al [73] analyzed the economic feasibility of a mobile pyrolysis system (40 dry t d^{-1} capacity) in 15 scenarios which included the plant being stationary or frequently being relocated throughout the year. The projected production costs of bio-oil were reported to be in the range of (USD) \$1.05-1.23 L^{-1}. Bio-oil production costs were reported to range from (USD) \$1.48 – 1.49 L^{-1} for 10 dry t d^{-1} mobile pyrolysis units when relocated once a year [44]. The bio-oil yield obtained was taken to be 50%. The authors concluded labor costs to be the key cost driver since a threshold number of operators is required irrespective of the plant size. The influence of labor costs decreased from 76% to 46% when base plant scales were increased to 100 dry t d^{-1}.

Based on the current scenario of bio-oil production costs from mobile pyrolysis units, our estimate of the bio-oil cost lies at the higher end for a base TCR plant of 12 dry t d^{-1} (500 kg h^{-1}). The major difference-makers in the production cost are identified to be primary bio-oil yield and absence of economies of scale. In comparison to the mean bio-oil yield of 50-60% produced through fast pyrolysis units, the base case study is performed for a conservative bio-oil yield of 6%. Moreover, when considering scale-up, attention was given to the technical feasibility and heat transfer limits of large-scale commercial auger reactors which were found to be limiting factors to gain benefits of economies of scale. Nevertheless, production costs reduced when multiple TCR units were combined to increase the processing capacity. The integration of multiple reactors might limit its mobility and add complexity in operation; that is a dilemma that requires further research to be resolved. Other factors that should be taken into account before financing TCR plants are the market maturity for its by-products. For this study, a biochar selling price of \$400 t^{-1} is taken which might be optimistic for the location of interest. Nevertheless, it is imperative to be able to generate revenue from by-products for TCR plants to be competitive. Although projected bio-oil production costs are high in the Western Canada context for IP-plants, opportunities for viable use may lie if the quality index of the bio-oil is considered. Its lower oxygen content, lower acidity, and improved thermal stability compared to fast pyrolysis or hydrothermal liquefaction derived bio-crude make it a promising candidate to be upgraded into higher-grade fuels with minimum hydroprocessing stages [62]. Even though large-scale TCR plants might yet be distant and require further research and development, the stability of TCR bio-oil allows it to be transported from distributed facilities to an upgrading plant under the current state of technology. Moreover, there are prospects of using the H_2 –rich synthesis gas (by-product in this study) for upgrading operations which can lead to lowering operational costs significantly.

2.6 Conclusion

A process model was developed to conduct a detailed techno-economic assessment of an emerging technology based on intermediate pyrolysis called thermo- catalytic reforming (TCR). The base case plant capacity was set to 500 kg h^{-1} and evaluated in a scenario where the plant replaces an existing landfill site at a local municipality. MSW was used as the preferred feedstock to produce a directly applicable bio-oil. In this study, high-value applications of the by-products, biochar, and hydrogen were assumed. Scale factors were developed for new equipment used in the process, and the capital investment required for a commercial application was estimated. The potential benefits of a distributed facility include the mobility of the facility, cost savings in feedstock transportation, opportunity for social and economic benefits from local urban resources, and increased self-sufficiency. A small-scale TCR plant faces challenges with respect to economies of scale, as production costs are high. For the base case, the cost of bio-oil is \$2.01 L^{-1} from a plant with a capacity of 500 kg h^{-1}. Although post-processing and storage costs are low, product yields need to be increased and manufacturing costs decreased. Production costs are likely to decrease if plant capacity increases through economies of scale. However, an increase in plant capacity would likely affect other benefits such as mobility, ease of assembling, and implementation in remote regions.

Following a sensitivity analysis, it was found that improving bio-oil yield and pricing biochar effectively influenced production cost the most. High value applications of biochar, i.e., as soil amendment, activated carbon filters, or raw material for fertilizer production, may lead to the highest revenue from the by-product stream. Monte Carlo simulations showed production cost to change by \$±0.79 L^{-1} with respect to the base value calculated. A techno-economic assessment was conducted by evaluating the quality index of the TCR products. The results provide insights on pathways and key factors that require further attention to develop the technology competitively.

Chapter 3 : A comparative assessment of fast pyrolysis, hydrothermal liquefaction, and intermediate pyrolysis of municipal solid waste for liquid transportation fuels production

3.1 Introduction

The need for alternative, affordable, and renewable sources to produce transportation fuels (i.e., gasoline, diesel, and jet fuel) has received increased attention and awareness in recent years. With advancements in conversion processes and technologies, complex feedstocks such as municipal solid waste (MSW) are becoming attractive options to contribute to the cause. Over the last decade, several countries have explored better ways to use waste streams by recovering energy or producing value-added products from them [117-119]. This has helped to shift the global perspective on MSW from "just a waste" to an urban resource. In many jurisdictions around the world, MSW is landfilled and there is a cost associated with its landfilling. The relative abundance and low or negative costs associated with using MSW make it a promising feedstock to produce transportation-grade fuels. However, to obtain the desired product, the heterogeneous MSW stream needs to be converted into a usable homogenous form.

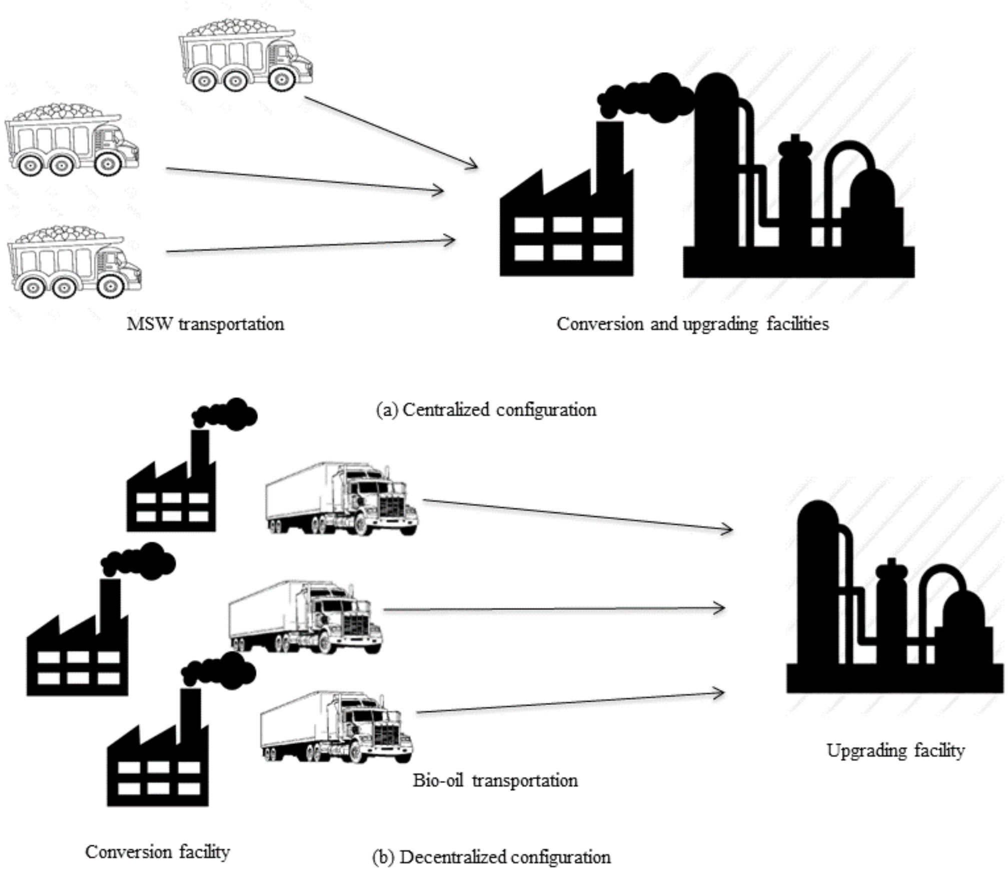
Municipal solid waste (MSW) refers to recyclables and compostable materials, as well as garbage from homes, businesses, institutions, and construction and demolition sites [1]. Routes to process MSW can be broadly divided into thermochemical and biochemical. Thermochemical treatment is preferred over biochemical routes such as landfilling and anaerobic digestion due to its fast processing rates, low operating costs, and the ability to process unsorted MSW streams [120-122]. Thermochemical conversion processes follow two basic approaches gasification and direct liquefaction. Incineration can also be used to thermally treat MSW streams but is discouraged in many jurisdictions because it emits large amounts of pollutants (NO_x, SO_x, and dioxins) and particles [10, 118, 123]. Direct liquefaction is often favored over gasification because of its simplicity and high liquid yields [124-126]. Moreover, if the desired goal is to produce transportation fuels such as diesel, kerosene, or gasoline, it is advantageous to select a processing route that directly yields a liquid-dominant product stream. Hydrothermal liquefaction (HTL), fast pyrolysis (FP), and intermediate pyrolysis (IP) are three direct liquefaction techniques analyzed in this study to process the MSW received from municipality households. The 'black bags' received from households are sorted out of recyclables, household hygienes, and hazardous wastes to form an organic dominant final stream to be compatible with the techniques. The primary product from all the processes is a highly viscous and thermally instable liquid fraction, which is often referred to as bio-crude or bio-oil [20, 127]. Bio-crude is generated through pressurized solvent liquefaction of biomass, and frequently contains 5 to 15% oxygen, which is more than petroleum but less than bio-oil [128]. It is the intermediate produced when the primary conversion process is HTL and although viscous, is thermally more stable over time than pyrolysis bio-oils. Bio-oil or pyrolysis oil is the result of rapid heating and quenching of biomass in the absence of air to form a liquid. The physical and chemical properties of bio-oil differ upon the pyrolysis technique used. Fast pyrolysis oil compounds contain up to 40% oxygen by weight and acidic and reactive compounds [77, 129]. Intermediate pyrolysis is performed at slower heating rates compared to FP and can produce more thermally stable bio-oil with less oxygen and acidic content [24, 26, 60]. Nevertheless, to be converted to transportation fuels, both bio-crude and bio-oil require extensive chemical modifications to increase volatility, reduce viscosity, and gain higher thermal stability. These modifications include removing heterogeneous atoms such as oxygen and reducing molecular weight, and the extent of the transformation required depends on the quality of the liquid originally produced [122, 130]. Therefore, although all techniques may produce liquid products, the route to upgrade bio-crude or bio-oil into transportation fuels varies [37, 40].

In addition to the limitations on the direct use of bio-oils and bio-crudes, the commercial feasibility of HTL, FP, and IP plants is affected by the overall system configuration. Whether a conversion plant is located as a centralized or decentralized facility affects the overall economic performance, as several studies have shown [45-48]. Centralized or large-scale facilities are generally assumed to be fixed processing plants where the bulk of the feedstock is transferred, processed to an intermediate (bio-oil or bio-crude), and upgraded to the desired product. The conversion facility and upgrading facility co-exist. They can offer significant savings through economies of scale, multiplicity of operating units, and consistency in production. However, they require large capital investments and have high feedstock transportation costs [45, 48]. Decentralized and distributed facilities, on the other hand, are of smaller scale, mobile, and can use local waste streams [44]. Due to their relatively smaller size, these plants can be implemented at feedstock sites, requiring low to minimum costs in feedstock transportation and handling. However, these facilities are often restricted by processing capacities and do not benefit from economies of scale [44]. Upgrading operations do not co-exist with decentralized plants and the intermediate product needs to be shipped to a centralized upgrading plant. The benefits and drawbacks present in both configurations, along with the lack of information on the most economic direct liquefaction process for MSW, make it difficult to determine which configuration is most suitable. Large capital investments are withheld due to concerns over the market competitiveness of these processes for transportation fuel production. Investors are reluctant to fund large capacity projects due to uncertainty in the production cost, yet without real production costs from established plants we cannot gain additional knowledge. This has led to slow development [131].

Techno-economic assessment (TEA) is a means of assessing an emerging technology in terms of its technical and economic performance and can be a possible solution to this dilemma. TEA uses process models and economic analysis to evaluate the competitiveness and sustainability of technologies and production pathways under current and future market scenarios. Several TEAs have been conducted to evaluate the economic performance of centralized and decentralized facilities producing transportation fuels via a primary bio-crude (HTL-based) or bio-oil (FP-based or IP-based) [34, 37-39]. These studies have investigated scenarios of centralized bio-crude or bio-oil production plants with co-existing upgrading operations. Though some FP-based studies have evaluated decentralized systems, they estimated the production costs of bio-oil only [41-44]. For IP-based plants, the economic performances have been measured through estimating the levelized cost of electricity in centralized plants by combusting all the products including bio-oil in combined heat and power units [71, 72]. Although production costs of bio-oil from IP has been estimated by the same author in previous work, transportation fuel costs are not yet reported. A list of the studies that use one of the direct liquefaction techniques as the primary feedstock conversion process is presented in Table 1.1. Publications on the production of road or air transportation fuels via other thermochemical or biochemical techniques can be found elsewhere [132-136]. All monetary values are converted to 2018 US dollar values.

Following a technical and economic review of the HTL, FP, and IP plant TEAs presented in Table 1.1, few knowledge gaps were identified: first, the TEA studies conducted on HTL and FP-based plants used a cost-generating feedstock, i.e., corn stover, pinewood, aspen hardwood, etc., and these plants incurred feedstock costs in the range of \$48-87 t^{-1}, which greatly influenced overall production cost [34, 37, 38, 40]. There are opportunities to reduce the overall production costs of the desired product if the feedstock costs incurred can be shifted to a revenue source using a revenue-generating feedstock. Second, when evaluating transportation fuel production pathways, the plant configurations considered has been limited to centralized facilities. Decentralized facilities have been assessed until the production of bio-oil, and not upgraded products. Third, there is a scarcity of this kind of analysis, where technologies are compared on the same baseline (i.e., same feedstock to the same final products). Fourth, there is very limited literature on the utilization of MSW using these technologies for the production of liquid transportation fuels. Fifth, the technologies considered here are still in their early stages of development and hence there is a need to develop bottom-up models based on fundamental science for assessment of liquid fuel production. There is a limited focus from earlier research on this aspect and this paper aims at addressing these gaps.

MSW is considered a revenue-generating feedstock since its use allows the processing plant to receive gate fees. Gate fees are the payments taken by waste conversion facilities for waste received [137, 138]. Several TEA studies have used MSW as a feedstock to explore the benefits of receiving gate fees; however, since the final products in these earlier studies [71, 72, 76, 139] are cellulosic ethanol, mixed alcohols, and electricity, a direct comparison with transportation fuels production cannot be made. The difference in plant configurations could affect the overall production cost in a number of ways. For instance, in a centralized configuration, given the immobility in operations, feedstock needs to be hauled long distances to the facility, while in a decentralized scenario, plants can be located at feedstock sites, and thus this transportation is reduced. Nevertheless, production in small distributed facilities would require that the intermediate product (bio-oil or bio-crude) be transported to an upgrading facility (see Figure 3.1). This could lead to significant differences in capital and operating costs, as shown in the earlier studies [45, 50, 51]. Another parameter that could affect operating costs is labor costs. Because the relationship between labor requirements and production rate is not always linear, scaling up a plant does not always involve hiring more operators. Large-scale plants generally provide a higher scope for automation, and maintenance requirements are low, as auxiliaries such as pumps, motors, etc., can be associated with multiple units. Identifying these key differences due to changes in system configurations while considering the same baseline for all processes can help policymakers and authorities to select the best suited method for transportation fuel production in different environments and locations.



(a) Centralized configuration

(b) Decentralized configuration

Figure 3.1: Overview of (a) centralized and (b) decentralized configurations

Despite the ability of HTL, FP, and IP to process MSW, concerns remain over their technical and economic performance in different configurations. A detailed comparative TEA could provide insight. To the best of the authors' knowledge, neither there has been a study comparing the cost of producing transportation fuels from HTL, FP, and IP with a revenue-generating feedstock such as MSW, nor has the production cost of transportation fuels through upgrading the bio-oil produced through distributed facilities been investigated. Although decentralized FP plants have been evaluated by some researchers [41, 43, 44, 49], the scope of these studies was limited to the calculation of the production cost of bio-oil. The overall objective of this research, therefore, is to conduct a comparative TEA of HTL, FP, and IP plants that process MSW into transportation-grade fuels, that is, gasoline, jet fuel, and diesel. The effect of the system configuration in the production cost of these fuels was evaluated in four scenarios: 1) a centralized 2000 dry t d $^{-1}$ HTL plant (C-HTL) with on-site upgrading, 2) a centralized 2000 dry t d^{-1} FP plant (C-FP) with on-site upgrading, 3) a decentralized 50 dry t d^{-1} FP (D-FP) with off-site upgrading, and 4) a decentralized 12 dry t d^{-1} IP (D-IP) with off-site upgrading. Multiple decentralized plants are used to meet an overall processing capacity of 2000 dry t d^{-1}. Hereafter the scenarios are referred to by their acronyms and are described in the next sections. The final products from all the pathways are gasoline, jet fuel, and diesel. This study has the following specific objectives.

- The development of supply chain networks for centralized and decentralized production of transportation fuels using MSW as a feedstock.
- The development of process models for techno-economic assessment of each scenario (C-HTL, C-FP, D-FP, and D-IP processing 2000 dry t d^{-1} of MSW).
- The estimation and comparison of transportation fuel production costs from all centralized and decentralized scenarios.

- The investigation of the key parameters that affect overall production cost and the assessment of the volatility of gasoline, jet fuel, and diesel cost in each scenario.

3.2 Materials and method

3.2.1 Materials

3.2.1.1 MSW properties

The physical and chemical properties of raw MSW streams differ significantly depending on the source from which they are collected. For efficient conversion, a feed with uniform composition is desirable. The primary MSW stream is sorted from recyclables, hazardous material, and non-combustibles such as hygiene products to form an organic-dominant MSW stream prior to thermochemical conversion. In centralized production scenarios, the removal of recyclables and the non-organic fraction is completed at the transfer stations, from which the sorted MSW stream is transported to the centralized facility. However, in decentralized production scenarios, the use of transfer stations is not considered as plant capacities are low. The as-received MSW stream is classified to meet the process requirements at the plant site. In every scenario, the sorted MSW stream is assumed to have moisture content (MC) of 50%. Any further treatment required for a specific conversion technique such as grinding, drying, or pelletizing is performed at the plant site. The properties of MSW are presented in Table 3.1.

Table 3.1: Properties of sorted MSW [74]

Parameter	Unit	Value
Moisture	wt.%	50.0
Proximate analysis		
Ash	wt.%	16.5
Volatiles	wt.%	68.5
Fixed carbon	wt.%	15.0
Ultimate analysis		
Nitrogen	wt.%	1.1

Parameter	Unit	Value
Carbon	wt.%	45.8
Hydrogen	wt.%	7.5
Sulfur	wt.%	0.4
Oxygen	wt.%	28.7

3.2.2 Method

The following section is divided into two main parts, one describing the system configurations of the centralized and decentralized production networks and the other, the techno-economic assessment. The two centralized configurations evaluated are C-HTL and C-FP, both with on-site upgrading. The two decentralized configurations are D-FP and D-IP, both with off-site upgrading.

3.2.2.1 Process configurations/scenarios

3.2.2.1.1 Scenario 1: C-HTL with on-site upgrading

A C-HTL plant with co-existing upgrading operations is considered in this scenario, shown in Figure 3.2. In recent years, HTL plants have received attention for their ability to produce moderate yields of bio-crude from very moist feedstocks [18, 19, 89]. The techno-economic assessments of large-scale HTL plants have been conducted at capacities of 2000 dry t d^{-1} (2205 short tons) [37, 38, 89, 140]. This is a baseline capacity widely used in other assessments of biorefinery concepts as well [34, 40, 131]. Since the technology is still under development, these assessments are carried out under assumptions such as reactor scalability, optimized operating conditions, and 'nth-plant' designs to incorporate future improvements and matured processes. The overall processing capacity of the plant is assumed to be 2000 dry t d^{-1} for ease in comparison with earlier studies. HTL reactions are performed at temperatures from 250-380˚C and elevated pressures between 5 and 30 MPa. The residence time of feed in the reactor is from 5 to 60 minutes, during which the feed is converted to bio-crude in the presence of water at supercritical conditions [19, 38, 141]. At these operating conditions, the ionization of water is promoted while its polarity decreases, favoring dehydration, deoxygenation, and decarboxylation reactions. The bio-crude obtained from HTL has been observed to have a lower oxygen content (10-20 wt. %) and a higher heating value (35 MJ kg^{-1}) than FP bio-oil [19, 20, 141].

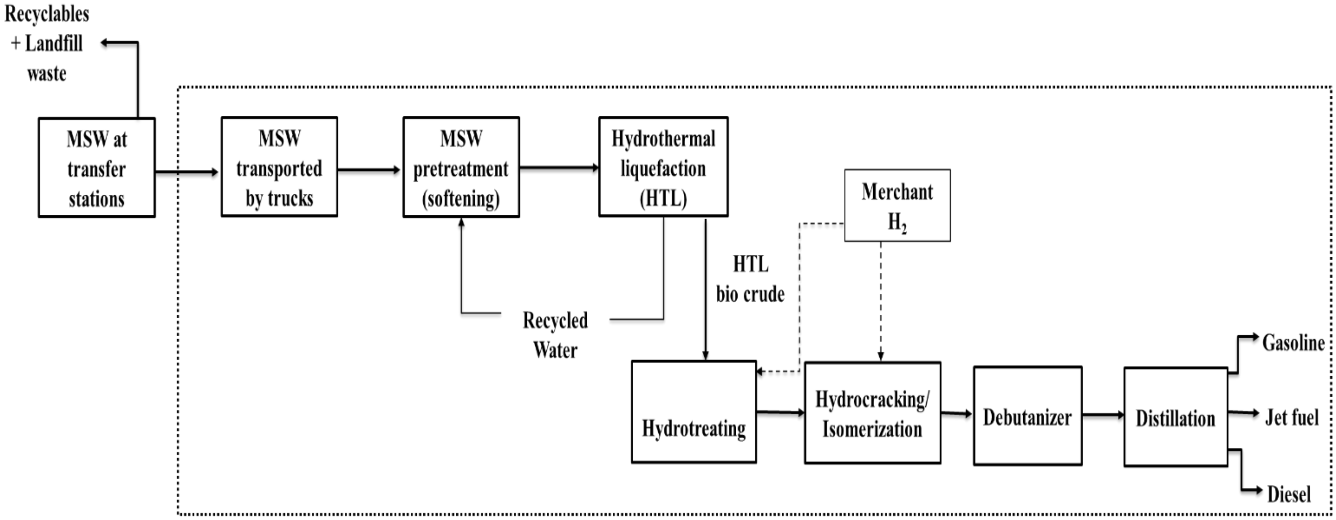


Figure 3.2: Overview of the C-HTL scenario

The pre-sorted MSW from several transfer stations is transported to the C-HTL plant site by truck. The organic-dominant and very moist MSW stream do not undergo any further pre-treatment at the plant site, and it is stored in tanks. The hot water from the HTL reactor is recycled to the storage tanks to soften the MSW feed and form a pumpable slurry. The slurry is then directly pumped in stages to the operating pressure of the HTL reactor. The HTL reactor is simulated as a plug flow type reactor due to its economic benefits over continuous stirred tank reactors (CSTRs) and is operated at a temperature and pressure of 355˚C and 20.3 MPa, respectively. The hot feed is converted to HTL bio-crude, an aqueous stream, and gas in the reactor. The filtered effluent is then cooled and depressurized to separate it into streams of liquids and non-condensable gases. The liquid stream consists of bio-crude, the prime product, and an aqueous phase carrying mainly organics. The yield of HTL bio-crude has been reported to be 35-40 wt. % for a variety of moist feedstocks including wet waste streams [20, 31, 141, 142]. The rest is made up of permanent gases, an organic-rich aqueous phase, and residual solids in the form of ash. The produced bio-crude is pumped directly to the upgrading operations.

HTL bio-crude upgrading has not yet been demonstrated commercially; however, it is assumed to be similar to FP upgrading, as reported in other studies [37, 38, 89, 140]. Two-stage hydroprocessing is employed to first deoxygenate the bio-crude and then further crack the long-chained hydrocarbons into desired fuel specifications. In both hydrotreating and hydrocracking operations, compressed merchant hydrogen is used. Other studies have reported similar arrangements for upgrading raw HTL bio-crude into transportation-grade fuels [37, 38]. The hydrotreater is operated at conditions of 150˚C and 115 atm pressure to deoxygenate the more stable bio-crude in the presence of H_2 and a catalyst through several hydrodeoxygenation (HDO) reactions [125, 129]. The amount of H_2 required on a per gram basis of dry bio-crude is not completely known. In this study, the hydrogen consumption is assumed to be similar to that required in the FP bio-oil hydrotreating stages. This is taken as a conservative estimate as the presence of heterogeneous atoms like oxygen and nitrogen is low in HTL bio-crude; hence, the hydrogen required is likely to be low. In the hydrocracker, the deoxygenated bio-crude is cracked into lower molecular weight fragments to maximize C_9-C_{15} hydrocarbons and meet the desired jet fuel range. Alongside the cracking, several isomerization reactions simultaneously rearrange straight-chained alkanes into branched structures. The hydrocracking stage is operated at atm pressure of 170 and a temperature of 430˚C. The upgraded bio-crude is pumped into a debutanizer to remove off-gases and light components before being distilled into gasoline, jet fuel, and diesel ranges based on component boiling points [37, 38, 89].

3.2.2.1.2 Scenario 2: C-FP with on-site upgrading

The C-FP plant receives pre-sorted MSW in a similar arrangement to the C-HTL discussed above (Figure 3.3). Unlike in HTL, the MSW in an FP plant undergoes several pretreatment stages before it is fed into the reactor. The processing capacity of the C-FP plant is taken as 2000 dry t d^{-1}. The initial moisture content of the received MSW is too high to achieve a thermally efficient pyrolysis process. A counter rotary drum dryer is used to reduce the moisture content of the feedstock to 5-7% to minimize energy demand in the pyrolyzer. The dried feed is then passed into a gyratory grinder to reduce the particle size to less than 2 mm [40]. The reduction in size promotes interactions between the sand in the pyrolysis reactor and the feedstock, thereby optimizing bio-oil yield. The grinder is modeled with a mechanical efficiency of 95% and some oversized materials remain at its outlet [40]. A vibratory screen is used to separate the oversized particles, which are recycled to the grinder to gain uniformity in size throughout the feed stream and minimize any feed loss.

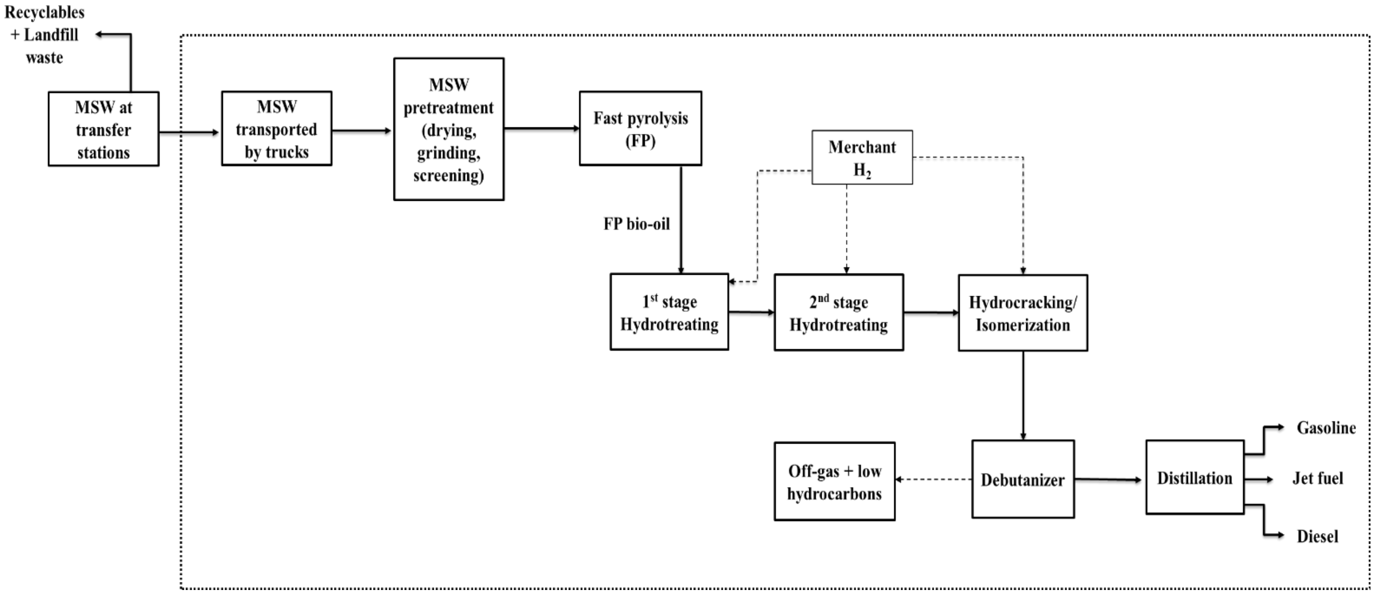


Figure 3.3: Overview of the C-FP scenario

Pyrolysis, unlike HTL, is performed at atmospheric pressure and high temperatures (400-600˚C) in the absence of oxygen. High heating rates (500-1000˚C/s) actively break down the cellulose, hemicellulose, and lignin constituents in the MSW feed at a very low residence time (1-2 s) [32, 33, 75]. Nitrogen is used to create an inert atmosphere and minimize any unwanted oxidizing of the feedstock. In the C-FP plant, a fluidized bed reactor, which uses sand as the fluidizing medium, is operated at 500˚C and in atmospheric pressure to disintegrate MSW under high heating rates and low residence time. A series of decomposition and chain reactions occurring simultaneously converts the MSW into pyrolysis vapors and char. The resultant pyrolysis vapor is freed from any micro-sized sand and char particles by being passed through a series of cyclone separators. This reduces the risk of bio-oil contamination and prevents dust interference in downstream operations. The clean pyrolysis vapor is passed to the quenching column, where it undergoes three stages of condensation. FP bio-oil is unstable and tends to deteriorate over time due to polymerization and condensation [143, 144]. To avoid detrimental effects on the generated oil, the bio-oil is immediately sent to hydro-processing units, where it undergoes dual-stage hydrotreating followed by hydrocracking to treat the highly corrosive and oxygenated raw FP bio-oil, as shown in Figure 3.3. Other studies investigating pathways to convert raw FP bio-oil to transportation fuels have suggested the use of similar hydroprocessing units [34, 40, 77]. First, mild hydrotreating, at 240˚C and 150 atm pressure in the presence of a sulfided nickel-molybdenum catalyst, is performed to remove acidic compounds and stabilize the oil for further upgrading. Further removal of heteroatoms, mainly oxygen, takes place in the second hydrotreater. It is operated at 300˚C and 170 atm pressure in the presence of the same catalyst. High pressure hydrogen is passed into the reactor to initiate several hydrodeoxygenation (HDO) reactions. The hydrocracking unit transforms the hydrocarbon composition into lower weight compounds, some of which are simultaneously isomerized into the range present in transportation fuels. The unit operates at 430°C and 170 atm pressure [40]. Merchant hydrogen is compressed in two stage-compressors to meet the respective pressure requirements in hydrotreater and hydrocracking units. The extensive hydroprocessing stages are followed by a debutanizer and a fractionation column to first remove off-gases and low hydrocarbon molecules before separating gasoline, jet fuel, and diesel with respect to boiling points.

3.2.2.1.3 Scenario 3: D-FP with off-site upgrading

Compared to centralized configurations, decentralized plants are modeled to have lower processing capacities and be distributed in small-sized municipalities. Here, the as-received MSW is sorted at the plant site, and transfer stations are not included in the network. Small pyrolysis plants for similar decentralized applications have been evaluated at scales ranging from 10 dry t d^{-1} to 100 dry t d^{-1} [41, 42, 44, 49]. For scenario 3, the processing capacity for a single D-FP plant is considered to be 50 dry t d^{-1}. Therefore, to reach an overall goal of processing 2000 dry t d^{-1} of feedstock, the plants are distributed in 40 locations.

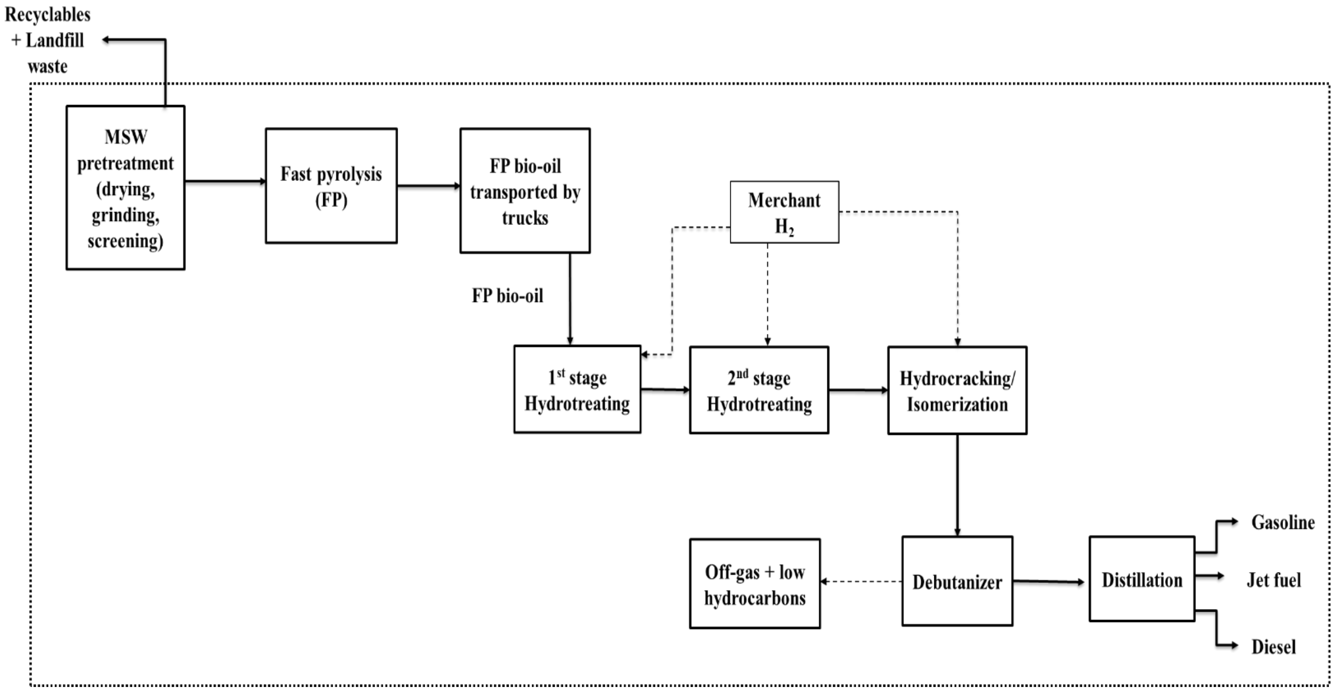


Figure 3.4: Overview of the D-FP scenario

The sorted MSW feed requires similar pretreatment stages (drying, grinding, and screening) as in the centralized facility, but on the plant site and at a smaller scale, before being fed for pyrolysis as shown in Figure 3.4. For D-FP plants, the pyrolysis reactor is assumed for simplicity and flexibility to be an auger reactor. Auger reactors are horizontal screw-fed reactors and have been demonstrated to fast pyrolyze a variety of feedstocks, including MSW [145-148]. The use of augers reduces the need for carrier gas and requires less energy in operation. A feeding screw is placed before the auger reactor to minimize disruption such as clogging and backflow of the ground MSW stream. Initially, the auger reactor is heated to FP temperatures by combusting propane until an equilibrium stage is reached, after which non-condensable gases (NCGs) generated in the D-FP plant are combusted to provide the process heat. As MSW is decomposed to produce pyrolysis vapors and biochar, cyclone separators are used to clean the pyrolysis vapors from any residual dust or char. Biochar is extracted using a secondary screw connected at the end of the auger reactor. The bio-oil is separated from permanent gases using a series of shell and tube condensers [71, 149].

The upgrading plants for decentralized configurations are located off-site at a central location to which the bio-oil is transported from distributed plants. Given the low volumes of bio-oil from a single plant, trucks were considered over pipelines as the most economical mode of transportation [51]. The acidic nature of FP bio-oil makes it corrosive for conventional truck tank material (i.e., carbon steel). The trucks used for transporting the bio-oil to the upgrading facility are considered to have stainless steel tanks to avoid any corrosion or leakage to the tanks [51, 70]. The bio-oil from D-FP plants are received at the upgrading facility site and stored in large stainless-steel storage tanks prior to being pumped into hydroprocessing units. The hydroprocessing stages are assumed to be similar to those of C-FPs (i.e., two-stage hydrotreating followed by hydrocracking). The final products are separated through the debutanizer and the distillation column.

3.2.2.1.4 Scenario 4: D-IP with off-site upgrading

For the production of transportation fuels using decentralized intermediate pyrolysis (IP) plants, a similar configuration to that of scenario 3 is used and is shown in Figure 3.5. The as-received MSW is sorted from the recyclables, household hygiene, and hazardous materials at the plant sites, and hence no feedstock transportation costs are associated in this scenario. Thermo-catalytic reforming (TCR) which is one of the technologies based on the principle of IP is an emerging conversion process that integrates intermediate pyrolysis and reforming. Further details on TCR can found in earlier studies [22, 30, 64]. Intermediate pyrolysis is another technique of pyrolysis performed at moderate heating rates (100-200˚C min^{-1}) with a feed residence time of 5-10 minutes [60, 61]. The different operating conditions compared to other pyrolysis techniques produce a higher quality bio-oil at the cost of overall bio-oil yield. In TCR, the intermediate pyrolysis reactor is combined to a reforming unit to enhance secondary decomposition of the initial pyrolysis vapors formed. This results in further breakdown of long-chain hydrocarbons into lighter fractions of C_7-C_{15} by secondary cracking reactions. TCR-derived bio-oil is reported to have very low oxygen content as well as low viscosity and corrosivity from a wide variety of feedstocks [22, 23, 29, 30, 64-67, 69]. However, commercial TCR plants are still under development. Currently, a TCR unit can only process 12 dry t d^{-1} of feed [150, 151]. For this study, it was assumed that four TCR units are combined and are referred to as a D-IP plant. A single decentralized D-IP plant can process 48 dry t d^{-1} of feed. For ease of calculation, it was assumed that 40 D-IP plants were distributed to meet an overall processing capacity of 2000 dry t d^{-1}.

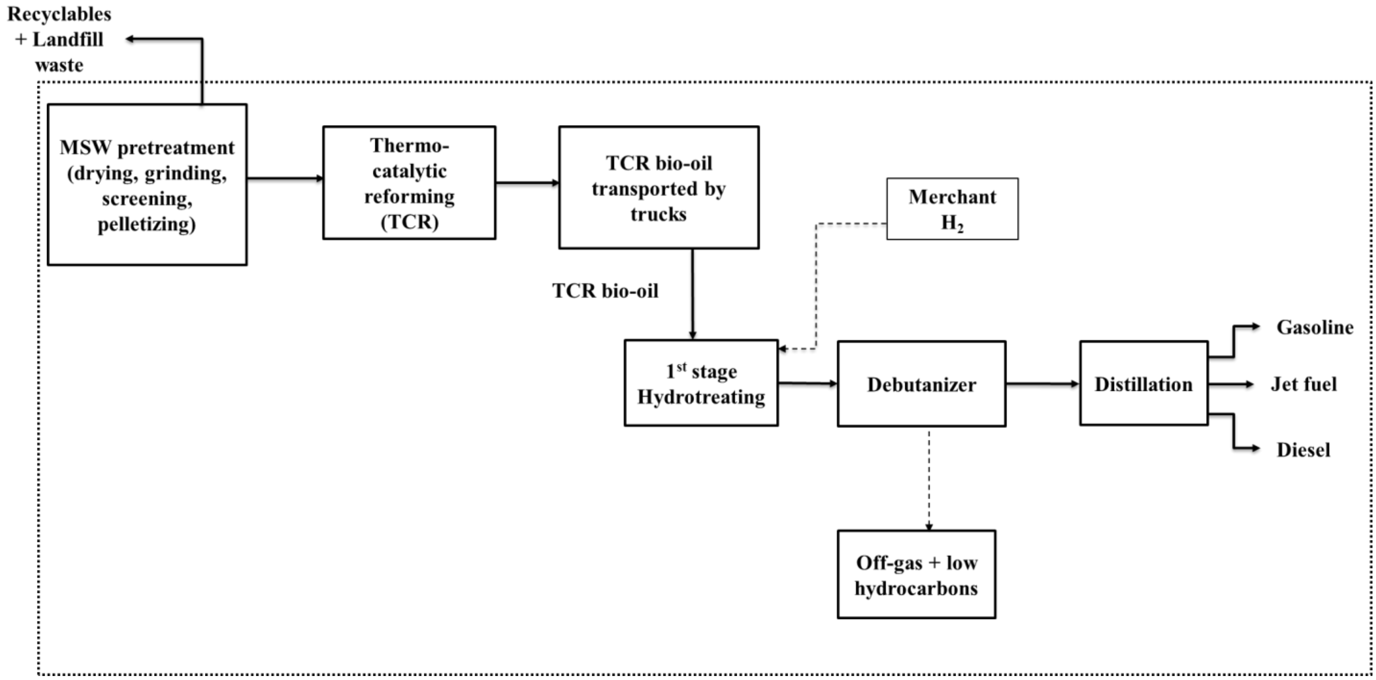


Figure 3.5: Overview of the D-IP scenario

In the D-IP plant, the local MSW, after being categorized, undergoes several pretreatment stages to reduce it to the desired composition. First, a counter rotary drum dryer is used to reduce the moisture content of the incoming feed by 40%. The stream is then passed through a gyratory crusher (grinder) to reduce the particle size to 20-25 mm; the particles are then passed through a vibratory screen to gain a uniform size distribution. An additional treatment performed in TCR is pelletization. The pelletized feed is used to avoid distortion due to its longer exposure time to high temperatures in the reactor, compared to the conventional pyrolysis process. A pellet mill is integrated into the pretreatment stage to form MSW pellets that are then screw-conveyed into the reactor.

Intermediate pyrolysis takes place in the auger reactor, which is operated at 400-450˚C [60, 61]. Here the feed pellets undergo thermal degradation and secondary cracking in moderate residence time. Like D-FP plants, a flex-fuel such as propane is used for the initial ramp-up to reach pyrolysis temperatures, and no carrier gas is used in the reactor. As pyrolysis vapors and biochar forms, all products are immediately passed to the vertical reforming vessel maintained at higher temperatures (600-700˚C). The biochar forms a packed bed at the bottom of the vessel and promotes thermo-catalytic activity on the incoming pyrolysis vapors. The long contact times between the pyrolysis vapors and biochar maximizes secondary cracking [28]. Moreover, several reforming reactions, such as steam methane reforming and water gas shift reactions, are favored, which yields an H_2-rich synthesis gas. In downstream operations, dust and residual particles are removed through a series of cyclones, and clean vapors are passed through a train of condensers to separate bio-oil from non-condensable gases (NCGs). The NCGs are passed through a pressure swing adsorber (PSA) with a recovery factor of 80% to separate the hydrogen. The remaining gases are combusted to provide process heat. Biochar, on the other hand, is simultaneously extracted from the bottom of the reforming unit using screw conveyers and cooled to avoid undesired combustion.

The stable and noncorrosive bio-oil is transported to an upgrading facility (located off-site) from each D-IP plant via truck. The truck tank's material is conventional carbon steel. At the upgrading facility sites, storage tanks are used to collect and store bio-oil prior to upgrading. Hydrodeoxygenation (HDO) of TCR bio-oil has been investigated through lab-scale experiments [62]. Compared to bio-crude or bio-oil derived from HTL and FP processes, the oxygen content and presence of long-chain hydrocarbons are low in TCR bio-oil. This is due to the thermal and catalytic cracking the bio-oil undergoes at the initial reforming stage at elevated temperature and in the presence of biochar. The catalytic activity of biochar has been shown to limit the maximum carbon number of components in the range of C_6-C_{20} [22, 28, 30, 65]. In this study, a single-stage hydrotreater is considered to deoxygenate the bio-oil. The hydrogen consumption per unit mass of bio-oil treated is assumed to be equal to that reported by Neumann et.al [62]. No additional hydrocracking stage is integrated as the bio-oil is already cracked into the desired component range. The resulting off-gas and low weight compounds are separated using a debutanizer prior to distillation.

3.2.2.2 Techno-economic Assessment

The techno-economic assessment for all scenarios was conducted through validated process models that were integrated into detailed spreadsheet-based calculations. The economic performance of each pathway was determined by calculating the minimum fuel selling price (MFSP) in \$ L^{-1}. The MFSP was determined by estimating the total project investment (TPI) in each scenario. The TPI is calculated by accounting for the total equipment costs, installation costs, indirect costs, contingency, and location factor. To estimate total revenue generation, selling prices are assigned to the products based on a comprehensive market analysis. All plants were assumed to use "nth" plant design; therefore, special financial needs, technical barriers, and long start-up times were not considered. The essential parameters and assumptions made in the techno-economic assessment are presented in Table 3.2. A discounted cash flow analysis (DCFA) was performed in which the TPI was offset by the revenue streams (the product and by-product sales and the gate fees) to arrive at an MFSP. In a multiple product plant scenario, as in this case, the method to calculate the MFSP can be broadly classified into two general ways: (i) the price of the prime product is varied while the others are fixed at their market value and (ii) the price of every product is varied simultaneously to optimize the yield and price of each product, until a break-even scenario is reached. For this study, the first approach was modified to include the market shares of gasoline, jet fuel, and diesel in the Canadian fuel market. The average market values of transportation fuels in Canada over the last decade were used to develop the ratios between diesel, jet fuel, and gasoline. Calculating the production price of the prime product (jet fuel in this case) thus resulted in production prices for gasoline and diesel based on the market ratios assigned. It should be noted that from here onwards, primary fuel refers to jet fuel and secondary fuel to gasoline and diesel.

Table 3.2: Inputs and assumptions

Parameter	**Value**	
	Decentralized Plants	Centralized Plants
Plant lifetime (years)	20	20
Plant operation (hours)	8000	8000
IRR	10%	10%
Base year	2018	2018
Currency	USD	USD

Escalation rate of different components		
General and administrative	2%	3.5%
Project capital	2%	5%
Products	2%	5%
Feedstock cost	2%	3.5%
Operating and maintenance labor	2%	3%
Utilities	2%	3%
Plant startup factor		
Year 0	0.7	
Year 1	0.8	
Year 2	0.85	
Spread of construction cost		
Year -3	0%	20%
Year -2	40%	35%
Year -1	60%	45%

3.2.2.2.1 Raw materials cost

The raw materials cost consists of expenses incurred for all the input materials required for plant operation. These include feedstock costs, merchant hydrogen costs, sand costs, catalyst costs, etc. The cumulative raw materials costs differ between the scenarios as the input materials for each technology are different. The materials considered in the techno-economic assessment for each pathway with their unit prices are given in Table 3.2.

3.2.2.2.2 Feedstock cost

One of the benefits of using MSW as the primary feedstock is the receipt of gate fees. The gate fee is the charge levied on the waste received at a waste processing facility [8, 100]. In the techno-economic assessment, a gate fee of \$80 t^{-1} of dry MSW was considered; this is the rate used in Western Canada [101]. However, since the MSW is at a pre-sorting stage at the transfer stations (for centralized scenarios) and at the plant gate (for decentralized scenarios), an additional cost of \$30 t^{-1} (dry basis) was assumed for the sorting and discarding the unusable portion of MSW. This results in an overall feedstock cost of negative \$50 t^{-1} of MSW in the techno-economic model. Gate fees and sorting costs vary by region and regulation, and the effect of this is investigated in the sensitivity analysis.

3.2.2.2.3 Catalyst and hydrogen cost

The catalysts used by researchers to hydroprocess primary bio-crude or bio-oil can be broadly classified into noble metal catalysts and sulfided transition metal catalysts. While noble metal catalysts offer a better degree of hydrodeoxygenation (HDO), these can be easily poisoned by small amounts of sulfur and are expensive to use. [152]. The activity on a per gram basis catalyst for transition metal catalysts is lower than for noble metals by a factor of two [153]. However, transition metal catalysts are inexpensive and can offer economic benefits when used at large-scales. For this study, sulfided NiMo supported on alumina, a standard catalyst for hydroprocessing bio-oil is taken for upgrading the bio-crude and bio-oil. The catalyst cost was taken from an earlier study [37] and the catalyst lifetime was assumed to be one year. For upgrading operations, merchant hydrogen was used. The delivered cost of hydrogen was taken from two studies [154, 155].

Table 3.3: Raw materials and process utility rates

Material	Unit price	Comments/Sources
Feedstock ($ dry t^{-1}, including sorting cost)	50	[101]
Merchant hydrogen ($ kg^{1})	1.03	Hydrogen produced through SMR is used [154, 155]
Nitrogen ($ kg^{1})	0.63	[40]
Electricity ($ kWh^{1})	0.07	Commercial electricity rates taken for the plant [99]
Wastewater treatment ($ m^{-3})	0.72	Wastewater disposal between 10,000m^3 – 100,000m^3 considered [99]
Solid disposal ($ t^{-1})	42	[37]
Sand ($ t^{-1})	8.27	Cost of sand is only taken for C-FP scenario for the fluidized bed reactor [40]
Catalyst ($ kg^{1})	42.50	[37]
Cooling water ($ m^{-3})	0.095	[99]
Bio-oil transportation cost ($ m^{-3})	15.44	Fixed cost and variable cost is derived from the work of Pootkhanam et al. [51]. An arbitrary distance of 200 km is taken.

Material	**Unit price**	**Comments/Sources**
Propane ($ L^{-1})	0.465	[44]

3.2.2.2.4 Total project investment (TPI)

The total project investment is a function of the baseline equipment costs in the method developed by Peters and Timmerhaus and followed in this study [97]. For each production route, a process model cost estimation approach is used to determine the costs of the conventional process equipment involved. Unit blocks in the process models are mapped to process equipment such as reactors, grinders, dryers, cyclones, etc., based on design specifications and operating parameters. For more complex components involved in some of the processes, such as the auger reactor in the TCR process and the hydrotreater and hydrocracking units, scale factors specific to the equipment were derived from several studies [38, 71, 72]. With the scale factors, we scaled components to estimate their costs in the capacities used in the plants. The total project equipment cost (TPEC) is based on the direct and indirect cost factors developed by Peters and Timmerhaus; these factors are shown in Table A-3 in the appendix. These factors account for instrumentation and controls, piping, electrical connections, service buildings, yard improvements, and land development required. For C-HTL and C-FP, both the conversion and upgrading equipment costs are multiplied with factors assigned for centralized plants. For the D-FP and D-IP scenarios, factors for decentralized plants are assigned to the equipment costs for conversion plants, and upgrading facilities are taken as centralized plants.

3.2.2.2.5 Operating and maintenance cost

The operation and maintenance requirements of the process plants differ depending on their scale and the type of thermochemical process followed. The operating costs include utility costs, labor and maintenance costs, operating charges, general and administrative (G&A) costs, and plant overhead. For process utilities such as electricity and cooling water, Western Canadian rates were used and are provided in Table 3.3. Downstream costs such as wastewater treatment and solid disposal rates are presented in Table 3.3 as well. The associated labor cost was calculated based on the total yearly operational hours and the number of shifts. Each shift was assigned plant operators and supervisors with respect to plant capacity, and hourly Western Canadian wage rates for plant operators and supervisors were considered [98]. For each decentralized plant, 2 plant operators, 1 supervisor, and 1 administrative staff were considered for each shift. The plant maintenance costs are taken as a percentage of the TPI, 3% for large-scale commercial plants and 2% for small-scale distributed plants. Operating charges and plant overhead were taken as 25% of the operating labor cost and 50% of the total operating and maintenance cost, respectively, for plants of all capacities.

3.2.2.2.6 Transportation costs

In both centralized and decentralized configurations, the preferred mode of transportation for both MSW and bio-oil was by truck. The overall transportation cost is made up of two components: the fixed cost (FC) and the variable cost (VC). The fixed cost consists of truck loading cost, which depends on the setting-up time, loading time, and unloading time of the truck. It is expressed as \$ t^{-1} cost for MSW and \$ m^{-3} cost for bio-oil transportation. The variable cost is a function of the distance the truck needs to travel and is expressed as \$ t^{-1} km^{-1} and \$ m^{-3} km^{-1} for MSW and bio-oil, respectively. The overall cost for transportation was calculated by Equation 1:

$$Transportation\ cost = Fixed\ cost + Distance \times Variable\ cost \qquad (1)$$

For the transportation of MSW from transfer stations to the facility, trucks with a maximum payload of 17 tonnes were used. This is based on the maximum weight allowed on municipal roads in Western Canada [156]. At transfer stations, these trucks were filled using waste handling equipment such as hydraulic excavators [157]. At the facility site, the trucks dump the feed without requiring any handling equipment. The loading time was calculated based on the cycle time (travel time + total turn and dump time + load time) of the equipment and fill factor of the bucket (the percentage of an available volume in bucket or bowl that is used). Unloading time was assumed to be one-third of the loading time. The overall feedstock transportation cost was calculated to be \$56.4 t^{-1} based on Western Canadian truck rental rates and assuming an arbitrary distance of 200 km. For bio-oil transportation, costs were derived from Pootakham et al. [51]. Trucks of similar capacity (30 m^3) were assumed in this study, and costs of transportation were inflated to 2018 US dollars.

3.2.2.2.7 Product revenue

The sales of produced fuels along with other by-products were taken as revenue sources. The HTL process predominantly generates bio-crude. The yield of other minor streams such as residual solids and gases was not significant enough to be considered as a by-product and so was not included as a revenue stream for the plant. FP and IP processes, on the other hand, produce biochar and synthesis gas as by-products that can contribute to income. The most common application of biochar suggested is its combustion to produce energy [39, 102, 103]. However, several applications of biochar such as soil amendments, filtration material as activated carbon, and raw materials for fertilizer plants have also been reported in the literature [66, 78, 104-106, 112]. The material uses of biochar instead of combusting it for energy could generate higher revenue. The price of biochar used as filtration material to absorb fine bio-solids is reported to be from \$1500 to \$2000 t^{-1} [78, 106]. The price of biochar, as a soil amendment (the application assumed in this study), is reported to be between \$80 t^{-1} and \$500 t^{-1} [78-80]. Currently, the biochar market is developing in Western Canada and around the world; therefore, a price of \$100 t^{-1} is assigned as the selling price for the biochar from FP and IP plants.

An additional by-product recovered from the IP process is hydrogen. The extracted hydrogen can be used as a fuel for hydrogen fuel cells or in upgrading operations. In the D-IP scenario, the hydrogen is produced at the local plant site and upgrading operations are completed off-site. Rather than transporting the hydrogen for upgrading operations, it is introduced in the biomass-derived hydrogen (bio-hydrogen) market, as it is derived from a renewable source. However, for it to compete with other bio-hydrogen routes, the IP-hydrogen should be sold at prices comparable to those of bio-hydrogen from alternative routes. Studies on the production of hydrogen from biomass or bio-oil report production costs of \$1-\$5 kg^{-1} of H_2 [84, 107-111]. For the current study, we assumed a hydrogen selling price of \$5 kg ^{1}.

3.3 Results and discussion

3.3.1 Cost analysis

The summary of key cost estimates for the evaluated configurations is presented in Table 3.4. The difference in contributions from each cost component is described in separate sections.

Table 3.4 Summary of key estimates

Parameters	C-HTL	C-FP	D-FP	D-IP
Total project equipment cost (M \$)	\$114.1	\$122.1	\$157.6	\$165.5
Total direct costs (M \$)	\$334.4	\$357.9	\$378.4	\$356.3

Annual operating cost (M $)	$140.4	$183.5	$244.4	$128.1
Total project investment (M $)	$494.7	$529.4	$540.4	$504.1
Primary intermediate yield* (wt. %)	40.20%	43.50%	43.50%	15.00%
Jet fuel yield (ML y^{-1})	121.2	127.2	127.2	52.8
Diesel yield (ML y^{-1})	74.1	77.6	77.6	32.2
Gasoline yield (ML y^{-1})	50.6	53.6	53.6	22.3

*yields obtained after primary conversion of feed

3.3.1.1 Total project equipment cost

First, the total project equipment cost (TPEC) was estimated for each configuration and is shown in Table 3.4 at a plant capacity of 2000 dry t d^{-1}. The contributions of equipment costs to the primary bio-crude/bio-oil and for upgrading operations differ because of the type of thermochemical technique. This is highlighted in Figure 3.6. The C-HTL configuration was found to require the least overall TPEC among all the pathways. The equipment cost (73.6 M $) to produce the primary bio-crude is higher than the cost to upgrade it (40.6 M $). The reason for this is the use of the high pressure in the HTL reactor, which contributed to 55% to the TPEC for bio-crude production.

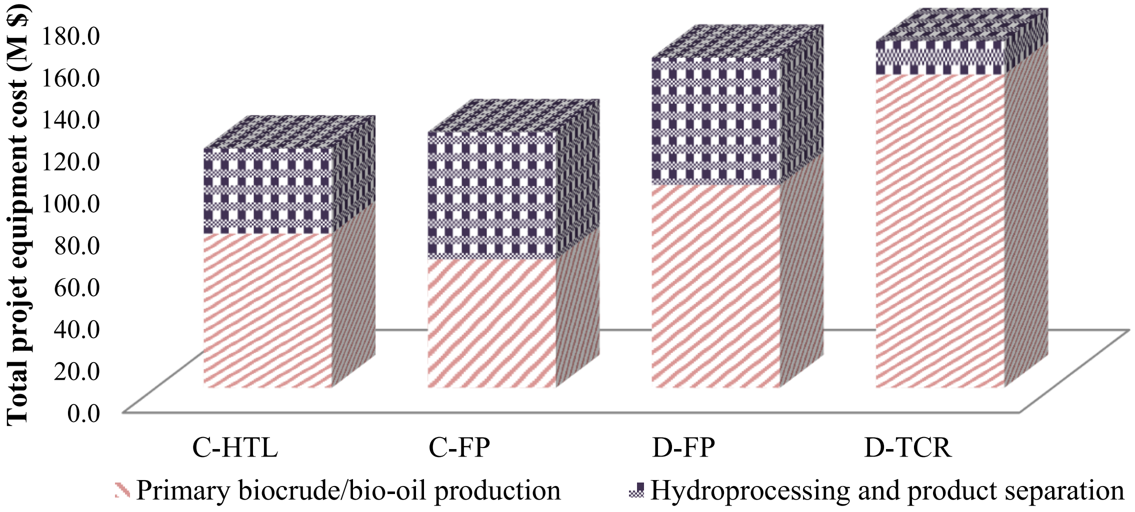


Figure 3.6: Total project equipment cost breakdown

In the C-FP configuration, economies of scale significantly reduce the equipment costs of primary bio-oil production. Using a fluidized bed reactor and operating pyrolysis at atmospheric pressure are the key difference makers. The reactor cost was 16.5% of the equipment costs calculated for producing the bio-oil, which were calculated to be 61.3 M $. This means the primary bio-oil could be produced with a 15.3% lower investment in equipment than in the C-HTL scenario. However, an additional 20.2 M $ more than the HTL-based route would be required to upgrade the bio-oil. The increment is primarily due to the additional hydrotreating stage along with supplementary equipment (compressors, valves, etc.) additions considered for upgrading the low-quality FP bio-oil.

While economies of scale lower equipment costs in the C-FP configuration, its effect is minimal in the D-FP scenario. In small-scale FP plants, auger reactors were considered for pyrolysis instead of fluidized bed reactors. Auger reactors are still under development at the time of this study and face limitations in heat transfer mechanisms toward their scale up. [112, 113]. Cost estimates for large-scale auger reactors in other studies have also shown minimal effects from economies of scale [71, 72]. The need for multiple small-scale plants increased the equipment costs in the D-FP configuration. Equipment costs to produce the primary bio-oil were found to be 28.8% higher than for C-FP in this configuration. The upgrading operations for both C-FP and D-FP produced bio-oil were assumed to be similar and hence the upgrading costs were comparable.

Auger reactors were used in the D-IP configuration as well. Although the plants were designed (48 dry t d^{-1}) for capacities similar to those of a D-FP plant (50 dry t d^{-1}), the system configurations within the plants were different. In D-IP, multiple auger reactors were combined to meet the overall processing capacity (48 dry t d^{-1}), and hence the effect of economies of scale was even lower than for D-FP. As a result, D-IP bio-oil production costs were the highest. However, the configuration benefitted from the lowest upgrading equipment costs. The equipment cost for upgrading the D-IP bio-oil was found to be 9.6% of the overall TPEC.

3.3.1.2 Production cost of transportation fuels

The production costs of diesel, jet fuel, and gasoline from all the configurations are presented in Figure 3.7. The figure also shows the contribution of each cost and revenue parameter to the jet fuel production cost. Here jet fuel is taken as the reference since it is the primary fuel in this work. The influence of each cost parameter was calculated based on the annual cost and total output of jet fuel from each configuration. Since the total jet fuel yields are different and depend on the intermediate yield of each process, as shown in Table 3.4, their contributions in \$ L^{-1} production cost varies as well.

3.3.1.2.1 Total project investment (TPI)

The TPI was highest for D-FP (540.4 M $) and lowest for C-HTL (494.4 M $). The contributions of the TPI to the jet fuel cost were \$0.17 L^{-1}, \$0.18 L^{-1}, \$0.18 L^{-1}, and \$0.40 L^{-1} for C-HTL, C-FP, D-FP, and D-IP respectively. Although the capital cost estimated for D-IP was lower than for the C-FP and D-FP configurations, the effect on production cost was higher because the yield of in D-IP was significantly lower than in the other processes.

3.3.1.2.2 Operating and maintenance cost

This major difference-maker in operating expenses was labor cost. The operating labor costs were found to be significantly lower in centralized approaches than in decentralized configurations. Because the labor requirement does not follow a linear correlation with the overall plant scale, fewer operators are required in large-scale plants. In decentralized configurations, even though a few personnel (4) are employed at a single plant, the total number is greater because multiple plants are used. This effect was also seen in other studies that evaluated decentralized scenarios [43, 44, 49].

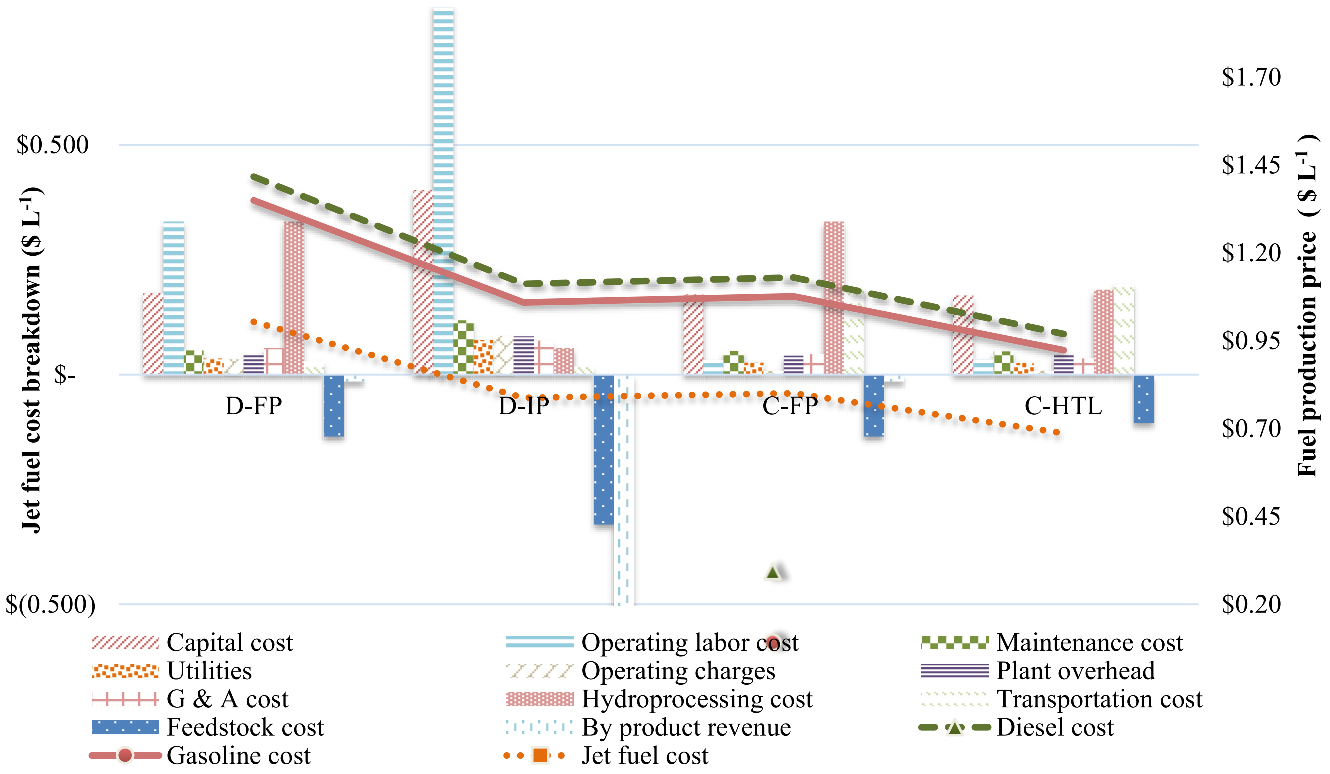


Figure 3.7: Breakdown of transportation fuels production cost

With an equal number of employees in D-FP and D-IP, the contribution of labor costs was \$0.33 L^{-1} and \$0.81 L^{-1}, respectively. Maintenance costs were slightly higher in D-FP and D-IP scenarios as the number of plant sites and equipment were greater. Maintenance requirements increased production costs by \$0.05 L^{-1} and \$0.12 L^{-1} for D-FP and D-IP, respectively.

3.3.1.2.3 Hydroprocessing cost

The hydroprocessing cost is influenced by the difference in the amount of catalyst and hydrogen consumption required for upgrading the intermediate in each process. The consumption differs as the quality aspects such as oxygen content, total acid number, and the thermal stability of the intermediates varies upon the primary conversion method. These \$ L^{-1} costs exclude the equipment cost of the hydroprocessing units as it is already integrated into the TPI breakdown. The C-FP and

D-FP configurations had similar hydroprocessing costs of \$0.33 L^{-1}. Based on the assumption that the quality and quantity of bio-oil from both configurations were similar, annual hydrogen and catalyst consumption were also assumed to be similar, and the highest among all the pathways. If bio-oil transportation loss from decentralized facilities and quality difference from using different reactors (the fluidized bed reactor for C-FP and the auger reactor for D-FP) are integrated, it may lead to a change in this cost. For C-HTL, the hydroprocessing cost was calculated to be \$0.19 L^{-1}. The subsequent upgrading stages of HTL bio-crude are in development, and few studies have been conducted to estimate hydrogen and catalyst consumption [125]. As the primary bio-crude is more deoxygenated and stable than FP bio-oil, the requirement of hydrogen and catalyst deactivation is likely to be lower, thus reducing the hydroprocessing cost. The hydroprocessing cost for D-IP was the lowest (at \$0.06 L^{-1}). The cost was low because of the reduced hydrogen consumption and catalyst. Besides the superior quality of IP bio-oil, for which the lowest amount of oxygen was removed, costs were further reduced due to the low volume of bio-oil that needed to be treated in this configuration.

3.3.1.2.4 Revenue

The revenue streams that offset the production expenses are gate fees and sales of by-products. All configurations qualified to receive gate fees for each t^{-1} of dry MSW feed. However, the impact of gate fees on the overall production cost was found to be highest in D-IP (\$0.33 L^{-1}). C-FP and D-FP shared a similar cost distribution (\$0.14 L^{-1}) because of the higher quantity of final products. The impact of fees was lowest in C-HTL production. First, it is lower than in D-IP because the C-HTL jet fuel yield is higher and therefore the cost is distributed over a larger product volume. Second, the revenue share of gate fees in C-HTL is low, because, given its ability to process moist feedstocks, the dry fraction of MSW it processes is lower than in the other processes.

Another source of revenue the plants could use to offset their production costs is by-product sales. The effect of these sales was significant in the D-IP configuration, largely because it included two by-products, biochar, and bio-hydrogen. For C-FP and D-FP, only biochar was considered. For C-HTL, the low yield of residual solids and ash content in it limits further application and is disposed of; therefore, no C-HTL by-product revenue stream was included in this analysis. The overall reduction in jet fuel cost for D-IP from the combined sales of by-products was \$0.69 L^{-1}. The major contributor to this cut was the sale of bio-hydrogen, which lowered the production cost by \$0.51 L^{-1}. This further emphasizes the need for a market for hydrogen and the ability to sell it. Although an equal selling price of biochar was assigned, the C-FP and D-FP scenarios generated less revenue because less biochar was produced in the FP process.

3.3.2 Sensitivity and uncertainty analysis

3.3.2.1 Exploration of key parameters

A global sensitivity analysis was performed using the Regression, Uncertainty, and Sensitivity Tool (RUST) for each configuration to explore the key parameters influencing overall fuel production cost [115]. A basic one at a time (OAT) sensitivity analysis examines the maximum and minimum values of a parameter while other parameters are kept at their base values. As a result, it fails to capture interactions between the input parameters. This limitation is overcome by RUST, as it integrates the Morris method to account for the interaction and interdependency between parameters. As shown in Figure 3.8, the results are represented in Morris plots, where the horizontal axis is the Morris average (μ) and the vertical axis is the Morris standard deviation (σ) of the output (jet fuel production cost) when input varies from its minimum and maximum values. Larger μ reflects the model to be more sensitive to the input, whereas, a large σ concludes the input to have non-linear and large interactions with other parameters. Sensitivity increases as we move to the upper right corner of the plot.

3.3.2.1.1 Centralized configurations (C-HTL and C-FP)

Centralized configurations were found to be highly sensitive to hydroprocessing costs and transportation costs. The hydroprocessing costs reflect the consumption of resources such as hydrogen and catalyst which were higher in C-HTL and C-FP plants due to extensive upgrading operations. Advancements in the quality of primary bio-crude or bio-oil from FP and HTL processes are understood to be the key pathway to significantly reduce the sensitivity of post processing in overall production cost. The transportation of MSW in centralized configurations contributed largely due to the difference in density between biomass and bio-oil. The effect can be minimized by assessing an optimal location for the centralized plant with respect to its feedstock sources which can reduce transportation distances. Total equipment cost closely followed as one of the most sensitive parameters in both C-HTL and C-FP. Labor cost and feedstock cost (recipient of gate fees) were not found to be influencing the jet fuel production cost significantly. In this analysis, an increase in feedstock cost implies a decrease in gate fees received by a plant. The increase (and corresponding decrease) could be a result of an increase in sorting costs or a change in location or regulations. Given that a cost-generating feedstock such as wood for FP or algae for HTL is used, the production cost is likely to increase and become more sensitive. However, using a more suitable feedstock for the process can increase the final product yield by increasing the yield of the intermediates (bio-oil or bio-crude)

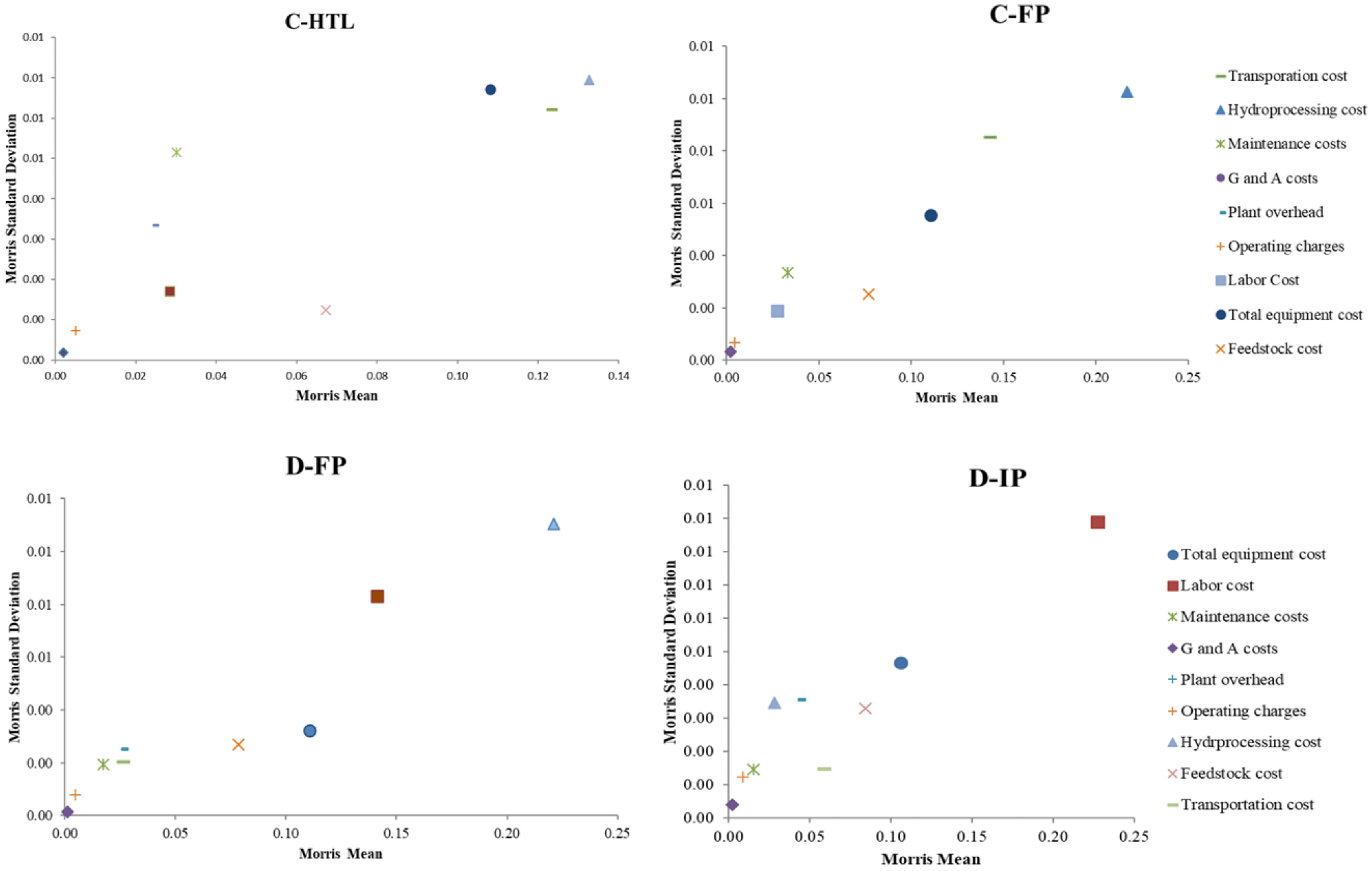


Figure 3.8: Sensitivity analysis using RUST

3.3.2.1.2 Decentralized configurations (D-FP and D-IP)

The production cost in the two decentralized configurations was dictated by operation labor cost. However, since the upgrading stages in D-FP were taken to be similar to C-FP, hydroprocessing costs remain as the most sensitive parameter in this configuration. High operating labor costs were incurred in these scenarios due to the minimum number of operators that are required irrespective of the plant size. Labor costs can be reduced by scaling up the base plant size or introducing automation. Nevertheless, the simplicity and mobility of a decentralized system must be retained when a scale-up is undertaken unless the product yield can be improved. Improving the yield requires that commercial auger reactors and process conditions be optimized. Attention must be given in order not to compromise the quality of the bio-oil in D-IP when these improvements are made as it would lead to an increase in hydroprocessing cost, which is one of the least sensitive parameters at the current scenario. The total project equipment cost was found to be the second most sensitive parameter in the D-IP model and is largely due to the number of auger reactors used. Being at an early stage of commercialization at the time of this study, auger reactors are yet to show strong benefits of economies of scale. Limited by the maximum scale of a single reactor, the D-IP scenario uses multiple reactors to meet the base plant capacity and although multiple reactors may share common auxiliary equipment, the savings are mild.

3.3.2.2 Uncertainty analysis

An uncertainty analysis was performed to examine the interdependency between sensitive parameters, as well as account for uncertainties from assumptions, price inputs, and process model parameters considered. A Monte Carlo simulation was performed using Regression, Uncertainty, and Sensitivity Tool (RUST) [115] – total equipment cost, operating labor cost, transportation cost, were varied by ±30% for 10,000 iterations. Hydroprocessing costs for all scenarios were varied by

±50% as it is yet to be demonstrated in commercial scales, therefore is understood to incorporate higher uncertainties. Preliminary estimates such as in TEA studies are considered to be accurate to ±30%, and similar ranges are reported in them [89, 92, 116]. The uncertainty analysis results are presented in the form of histograms in Figure 3.9. The associated uncertainties in the production cost of jet fuel in all scenarios were calculated to be \$0.71±0.06 L^{-1}, \$0.80±0.09 L^{-1}, \$1.01±0.08 L^{-1}, and \$0.78±0.07 L^{-1} for C-HTL, C-FP, D-FP, and D-IP, respectively. All scenarios evaluated in this study resulted in jet fuel prices higher than the current and 10-year average prices of jet fuel in Canada. In the last decade, the average market value of commercial jet fuel has been close to \$0.58 ± 0.18 L^{-1} with minimum and maximum price reached to be \$0.24 L^{-1} and \$0.86 L^{-1}, respectively [158, 159]. C-HTL, C-FP, and D-IP scenarios were able to reach production prices comparable to the market value; however, with the inclusion of profit margin the selling prices are likely to be higher.

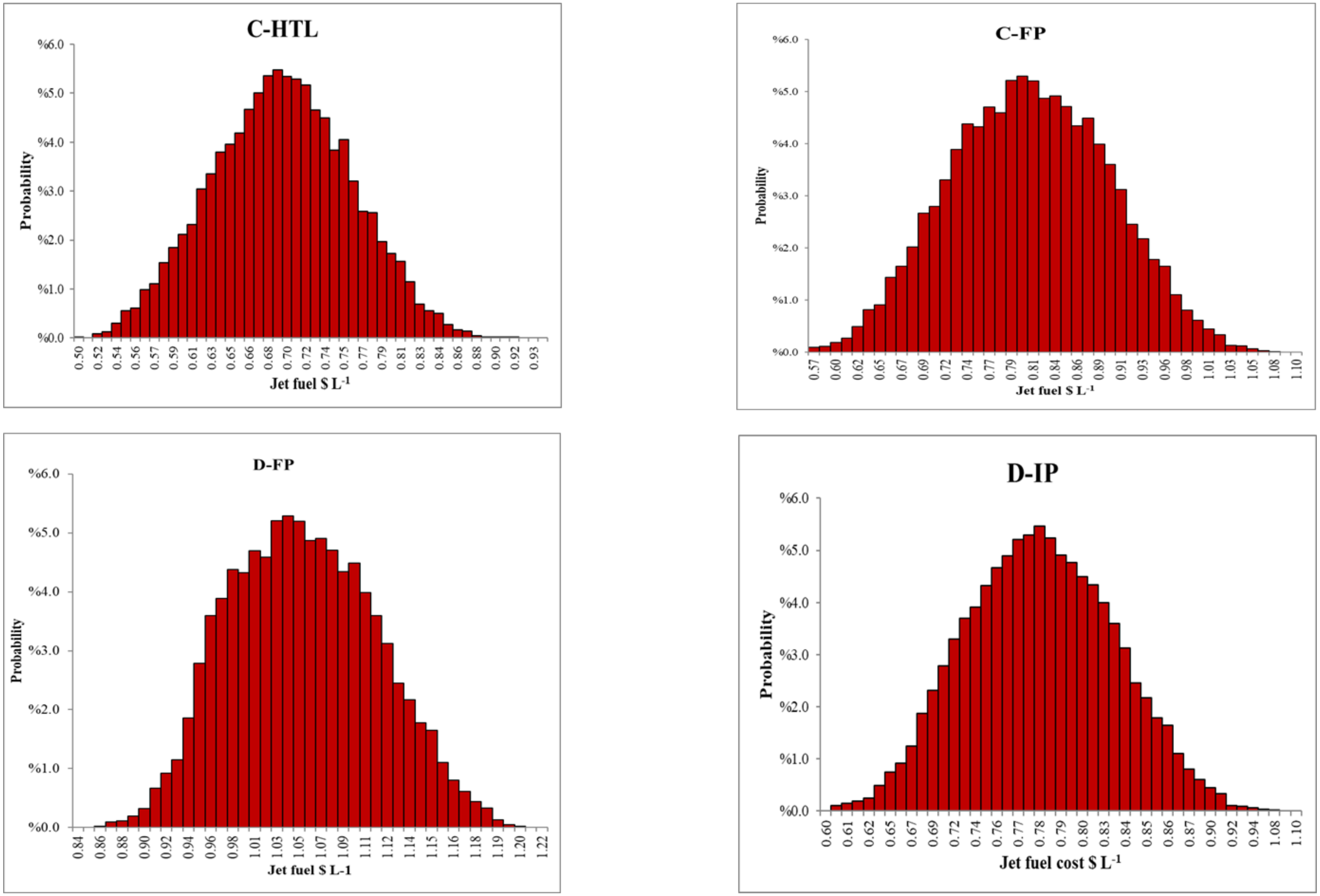


Figure 3.9: Histogram plots based on uncertainty analysis using RUST

3.4 Performance results and comparison with previous studies

This comparative techno-economic study was carried out for a proposed location in Western Canada. The focus of this research was to explore the most cost-efficient thermochemical technique and plant configuration to convert MSW into transportation fuels, with jet fuel as the primary product. Other studies that have focused on producing transportation fuels using HTL or FP processes have mostly reported the prices of gasoline and diesel but not jet fuel. Besides, the scope of these studies has been limited to centralized production routes where feedstocks that are known to be more compatible with these techniques (such as algae or woody biomass) are used. Zhu et al. [37] estimated the production cost of diesel range fuels to be \$ 1.27 L^{-1} and \$0.72 L^{-1} in the state of technology (SOT) case and goal case, respectively. The feedstock for the centralized plant was considered to be woody biomass. The reduction in production cost in goal case was a result of assumptions such as technology maturity, less organic loss to the aqueous phase, and advancements in hydroprocessing of bio-crude. A centralized HTL plant of 2000 dry t d^{-1} processing defatted microalgae to produce gasoline and diesel range fuels was evaluated by Ou et al. [38]. The diesel and gasoline production costs were reported to be in the range of \$0.52 L^{-1} - \$1.16 L^{-1}, with a 50% probability of gasoline prices being below \$0.79 L^{-1}. The upgrading stages consisted of two-stage hydrotreating, where hydrogen was provided from the hydrogen plant installed on-site. In comparison, the C-HTL plant in this study was able to produce diesel and gasoline at prices of \$0.96 L^{-1} and \$0.92 L^{-1}, respectively. The slight incremented prices are due to the added hydrocracking stage to intensify the cracking of primary bio-crude and boost jet fuel yield.

Several studies have evaluated centralized fast pyrolysis plants to produce transportation fuels [34, 39, 40]. The feedstock considered in these studies has been in the form of corn stover, aspen hardwood, pinewood, etc. Brown et al. reported the minimum fuel selling price of gasoline and diesel to be \$0.75 L^{-1} for a 2000 dry t d^{-1} fast pyrolysis plant using corn stover as the feedstock [34]. The production costs were found to be highly sensitive to the bio-oil yield, and varied between \$ 0.96 L^{-1} – \$ 0. 67 L^{-1}, given the yield was between 49 wt. % - 70 wt. %. Diesel and gasoline production prices were also reported in a similar range in the work by Patel et al. [40], where aspen hardwood was chosen as the feedstock. Renewable diesel and gasoline could be produced at \$1.13L^{-1} and \$1.08L^{-1}, respectively. The author concluded the production cost to be most sensitive to bio-oil yield and internal rate of return. The results from the C-FP configuration in this study show similar gasoline and diesel prices; \$1.14L^{-1} and \$1.07L^{-1}, respectively. The slight increment in production costs was due to the lower bio-oil yield (43.5%) and two-stage hydrotreating considered in this analysis, compared to 63% yield and single-stage hydrotreating reported in Brown et al [34]. Decentralized production routes have been mostly investigated using the FP technique as the technology is scalable [41, 42, 44, 49]. The emergence of IP has led to an alternative to FP, and its economic competitiveness has been explored in a previous study by the same authors [150]. However, to the best of our knowledge, these studies have been limited to calculate the production of bio-oil excluding the upgrading operations and therefore, a direct comparison between the prices of transportation fuels in this route cannot be made at this time.

3.5 Conclusion

A comparative analysis of the production of transportation fuels from organic dominant portion of the MSW using hydrothermal liquefaction, fast pyrolysis, and intermediate pyrolysis based pathways was conducted. The production routes were designed to incorporate the production of the intermediate (bio-crude or bio-oil) in a centralized or a decentralized configuration through four scenarios. Each scenario was designed to process 2000 dry t d^{-1} of sorted and organic-dominant MSW, and the production costs of diesel, jet fuel, and gasoline were calculated. Jet fuel was taken to the prime product and its production price was used to compare the scenarios. The upgrading costs of transforming a primary bio-crude or a bio-oil differed by technology because of the difference in the quality and quantity of the intermediates. The upgrading costs for a liter of jet fuel were the lowest (\$0.06 L^{-1}) when the intermediate was produced through intermediate pyrolysis and highest in the fast pyrolysis-based scenario (\$0.33 L^{-1}). If additional revenue can be generated by selling by-products from the process, production costs can be reduced considerably. However, it is necessary to have mature markets for the by-products. In conclusion, a centralized hydrothermal liquefaction plant was found to be the most cost-competitive route to produce fuels from MSW; however, it might not be preferred if large volumes of MSW are not available at the desired location. The jet fuel production costs from decentralized intermediate pyrolysis plants were similar to other configurations such as centralized fast pyrolysis plants because mature biochar and bio-hydrogen markets were assumed in this study. The outcomes of this study can be used to encourage local governments to strategize waste management approaches with respect to the technology that best suits the need. The placement of decentralized units can increase local employment, transform MSW into an urban resource, and further increase the self-sufficiency of localities.

Chapter 4 : Conclusions and recommendations for future work

4.1 Conclusions

This study was conducted to evaluate the techno-economic competitiveness of intermediate pyrolysis technology in Western Canada in the decentralized treatment of municipal solid waste (MSW). This study found that the commercial use of thermo-catalytic reforming (TCR) technology, an intermediate pyrolysis technology, relies on whether there are high-value applications for the by-products. A detailed process model was developed for an intermediate pyrolysis plant with 12 dry t d^{-1} processing capacity and assessed for a Western Canada municipality with a population of 5000. Using mass and energy balance calculations, the process model was validated by comparing the results to published lab-scale experimental results. Along with process model-based cost estimates, scale factors were developed from the literature for the new auger reactor used in the intermediate pyrolysis process. The capital cost for a decentralized intermediate pyrolysis plant processing MSW was calculated to be 3.9 M \$. Assuming there is a market for biochar and bio-hydrogen by-products, the production cost of bio-oil from intermediate pyrolysis was calculated to be \$2.01 L^{-1}.

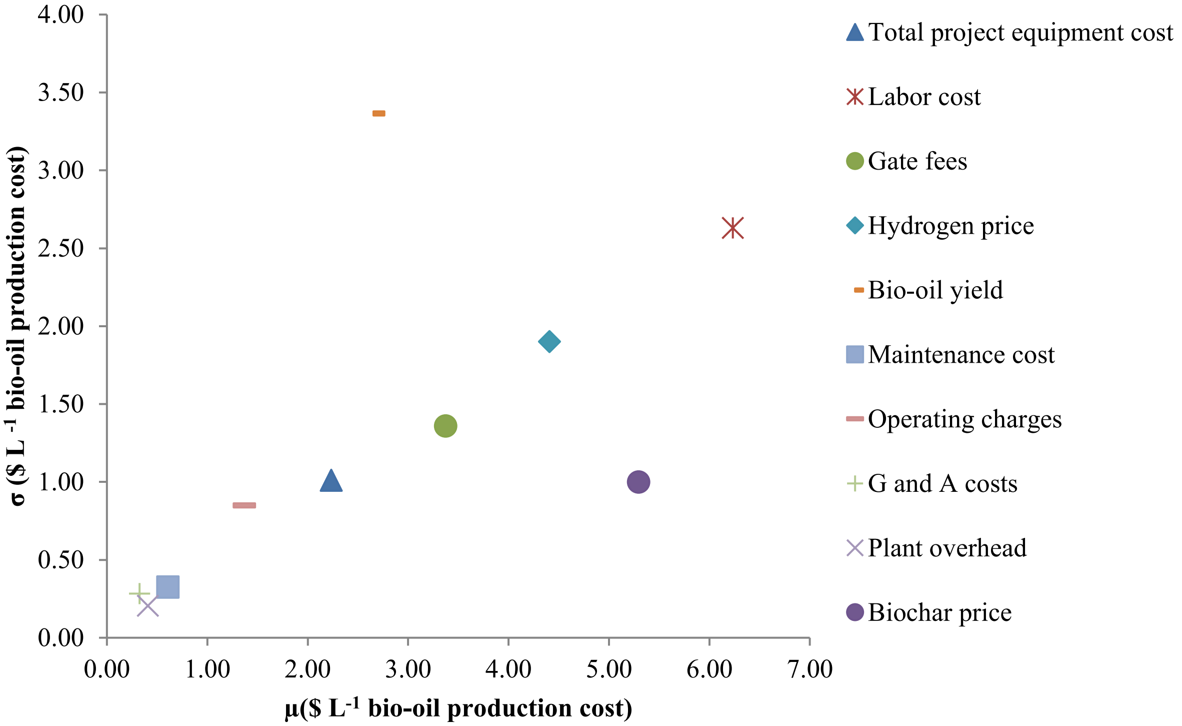


Figure 4.1: Sensitivity analysis to identify key parameters affecting the production cost of bio-oil

Through the sensitivity analysis, the bio-oil yield, biochar selling price, and increase in plant production capacities were identified as key parameters that could reduce the production cost, as shown in Figure 4.1. Product yields could be increased by optimizing process conditions and reactor developments. However, the quality of products should not be compromised in the process as it would limit their first-hand applications.

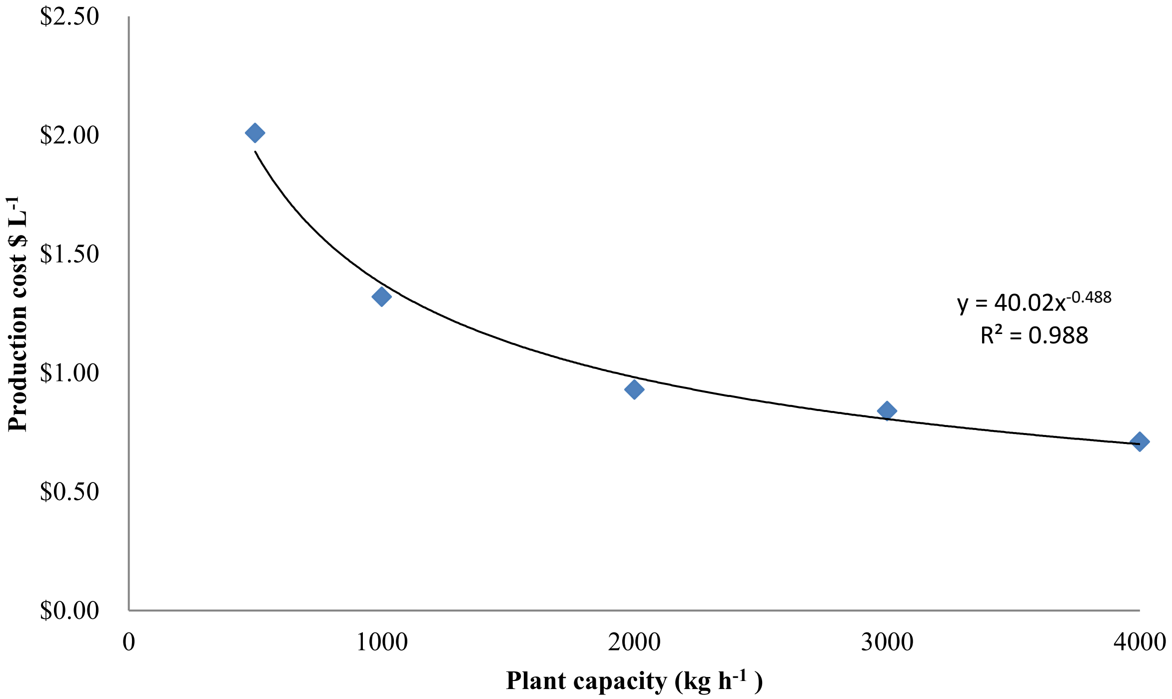


Figure 4.2: Influence of plant capacity on the production cost of bio-oil

Increasing plant capacities could significantly reduce production costs through economies of scale (shown in Figure 4.2). Production costs were found to decrease by 65% when overall capacity was scaled up by combining multiple TCR units to process feed at 4000 kg h^{-1}. However, increasing plant scales could require improvements in heat demand and heat transfer mechanisms in large-scale auger reactors (the main reactor used in intermediate pyrolysis). Using multiple auger reactors at their current capacity could boost production capacity and financial savings could be modest. On the other hand, adding multiple units could limit mobility and reduce the simplicity of the current approach, compromising the scope to implement it in a decentralized manner.

After conducting an independent analysis of the process, a comparative TEA of intermediate pyrolysis (IP) and the other direct liquefaction techniques; hydrothermal liquefaction (HTL) and fast pyrolysis (FP) was performed to understand the relative competitiveness of these techniques in converting MSW into a usable product. Every technology was evaluated at equal feed processing capacities of 2000 dry t d^{-1} with organic-dominant MSW feed and transportation fuels as the end product. The assessment further integrated frameworks of centralized and decentralized production routes. HTL, FP, and IP were evaluated based on four configurations, centralized HTL plant (C-HTL), centralized FP plant (C-FP), decentralized FP plant (D-FP), and decentralized IP plant (D-IP). The primary bio-crude or bio-oil, depending on the conversion technique used, was then upgraded into transportation fuels using hydroprocessing operations. Centralized facilities were modeled as if they co-existed with upgrading operations; these required the transportation of a large volume of MSW from various sources to the plant site. Decentralized plants were assumed to be located at the source of the MSW, usually small municipalities, and the primary bio-oil transported to upgrading facilities for hydroprocessing. The extent of hydroprocessing in each process differed depending on the quality of the primary bio-crude or bio-oil produced. Therefore, the number of hydroprocessing stages and the amount of resources such as catalyst and hydrogen differed among the configurations and affected fuel production cost. The final fuel composition from all configurations consisted predominantly of jet fuel as the primary fuel, with gasoline and diesel blends as secondary products.

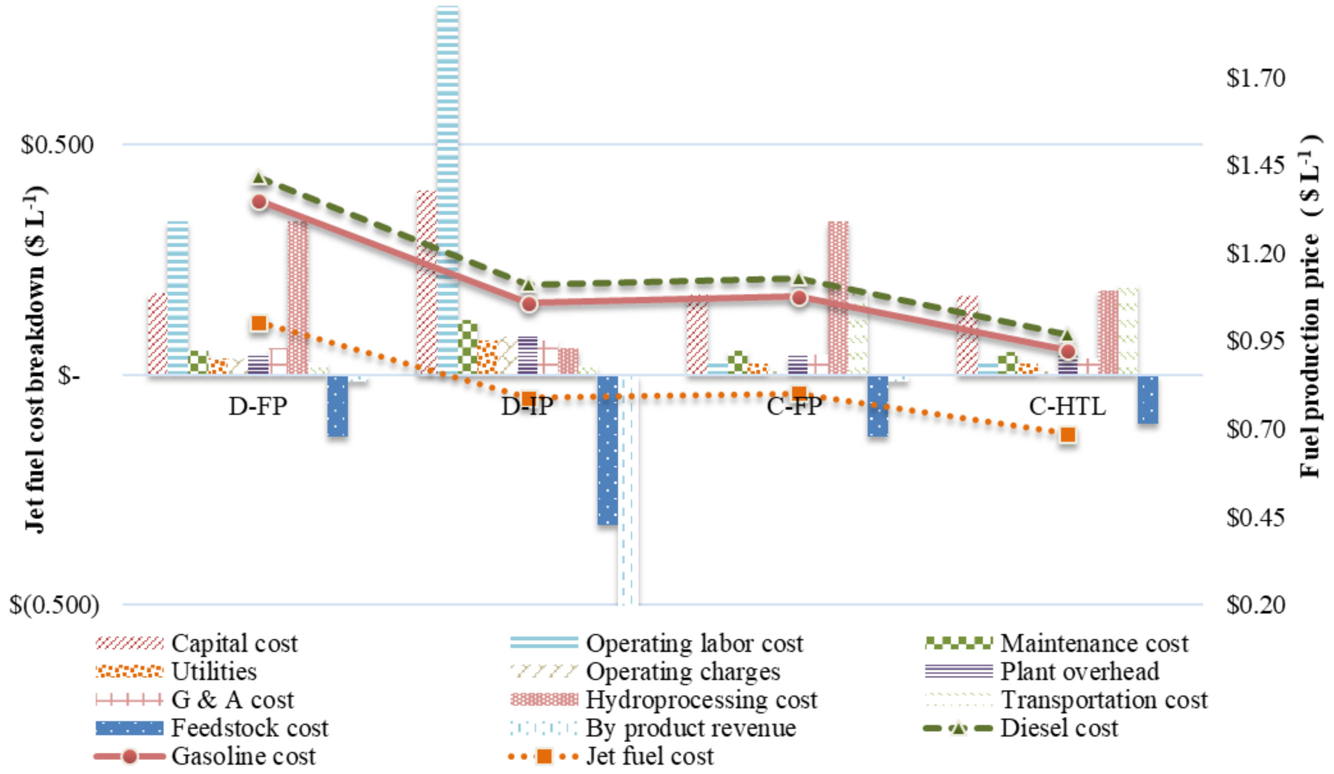


Figure 4.3:Breakdown of jet fuel production cost

The production cost of jet fuel was calculated to be \$0.71 L^{-1}, \$0.80 L^{-1}, \$ 0.78 L^{-1}, and \$1.00 L^{-1} from the C-HTL, C-FP, D-IP, and D-FP scenarios, respectively as shown in Figure 4.3. The capital investments estimated were comparatively higher for the decentralized routes (540.4 M \$ USD for D-FP) than the centralized routes (529.4 M \$ USD for C-FP) as multiple small-scale plants were required in the former to reach an overall capacity. However, contributions of cost and revenue sources to the overall jet fuel production cost varied significantly between the methods. For instance, D-IP plants could retain additional revenue through the sale of biochar and bio-hydrogen, whereas, C-FP and D-FP plants could only sell biochar. Hence, even though both D-FP and D-IP plants had significant operating labor costs per liter (\$0.33 L^{-1} and \$0.81 L^{-1}, respectively), the D-IP plants could offset a greater portion of it through its by-products. Higher hydroprocessing costs per liter of jet fuel were associated with hydrothermal liquefaction and fast pyrolysis based pathways (\$0.19 L^{-1} and \$0.33 L^{-1}, respectively) than in D-IP plants (\$0.06 L^{-1}). The parameters driving cost were identified through sensitivity analysis and differed by technology and configuration. For instance, production cost in the centralized routes was high because the cost of handling and transporting large volumes of MSW is high, whereas the contribution of transportation costs in decentralized routes was low because bio-oil can be shipped in smaller vehicles because of its higher density. Nevertheless, decentralized production routes were impacted by operating labor costs due to the high number of distributed plants. Through a comparative TEA, intermediate pyrolysis was found to be competitive in producing jet fuel among these alternative routes. Further development in large-scale auger reactors could lead to an increase in intermediate pyrolysis plant capacities and improve the liquid product yield at the same time. An increase in the liquid product can make intermediate pyrolysis an attractive option as it is likely to further reduce production costs.

The model developed in this work can be an important tool for local authorities, governments, and policymakers looking for alternative waste management systems. Whether one is planning a decentralized waste management system or a centralized route, one can use the results from this study as key pointers to understand the key challenges before large investments are made. With these results, research can be directed at more critical domains, which would result in the commercialization of these technologies.

4.2 Recommendations for future work

The following are recommendations for future research:

1. This study was conducted with MSW as a feedstock. Other widely available feedstocks such as woody biomass, agricultural residue, or digestates can be used to analyze the economic competitiveness. Although some might be cost-generating feedstocks, the product yield might be higher, which can lead to more opportunities to use intermediate pyrolysis.
2. This study evaluated intermediate pyrolysis as a stand-alone thermochemical technology. However, there may be further benefits of using intermediate pyrolysis as a complementary unit in a hybrid waste management system. For instance, in a fusion scenario in which an anaerobic digestion plant co-exists with an intermediate pyrolysis plant, technical and economic rewards can exist; these should be investigated.
3. Further development of the process model can help determine the optimum conditions for several feedstocks. Once optimized process conditions are found, the process can be modified to produce the best-valued product given the market conditions. In a mature biochar market, the operating conditions can be changed to promote biochar yield rather than bio-oil or vice-versa.
4. The intermediate pyrolysis process model and techno-economic model should be updated if provincial or waste management regulations change in the future.

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Appendix A

Table A-1: Alberta municipality population data

Acadia No. 34, M.D. Of	493
Athabasca County	7,869
Barrhead No. 11, County Of	6,288
Beaver County	5,905
Big Lakes County	4,103
Bighorn No. 8, M.D. Of	1,334
Birch Hills County 1	1,553
Bonnyville No. 87, M.D. Of*	11,661
Brazeau County	7,771
Camrose County	8,458
Cardston County	4,481
Clear Hills County	3,023
Clearwater County	11,947
Cypress County	7,662
Fairview No. 136, M.D. Of	1,604
Flagstaff County	3,738
Foothills No. 31, M.D. Of	22,766
Forty Mile No. 8, County Of	3,581
Grande Prairie No. 1, County Of	22,303
Greenview No. 16, M.D. Of	5,583
Kneehill County	5,001
Lac La Biche County**	9,531
Lac Ste. Anne County	10,899
Lacombe County	10,343
Lamont County	3,899
Leduc County	13,780
Lesser Slave River No. 124, M.D. Of	2,803
Lethbridge County	10,353
Minburn No. 27, County Of	3,188
Mountain View County	13,074
Newell, County Of	7,524
Northern Lights, County Of	3,656
Northern Sunrise County	1,891
Opportunity No. 17, M.D. Of	3,181
Paintearth No. 18, County Of	2,102
Parkland County	32,097
Peace No. 135, M.D. Of	1,747
Pincher Creek No. 9, M.D. Of	2,965
Ponoka County 9	9,806
Provost No. 52	2,205
Ranchland No. 66	92
Red Deer County 19	19,541

Rocky View County 39	39,407
Saddle Hills County 2	2,225
Smoky Lake County 2	2,461
Smoky River No. 130	2,023
Spirit River No. 133	700
St. Paul No. 19	6,468
Starland County 2	2,066
Stettler No. 6	5,526
Sturgeon County 20	20,495
Taber	7,173
Thorhild County 3	3,254
Two Hills No. 21	3,641
Vermilion River	8,267
Vulcan County 3	3,984
Wainwright No. 61	4,479
Warner No. 5	3,847
Westlock County 7	7,220
Wetaskiwin No. 10	11,181
Wheatland County 8	8,788
Willow Creek No. 26	5,179
Woodlands County 4	4,754
Yellowhead County 10	10,995

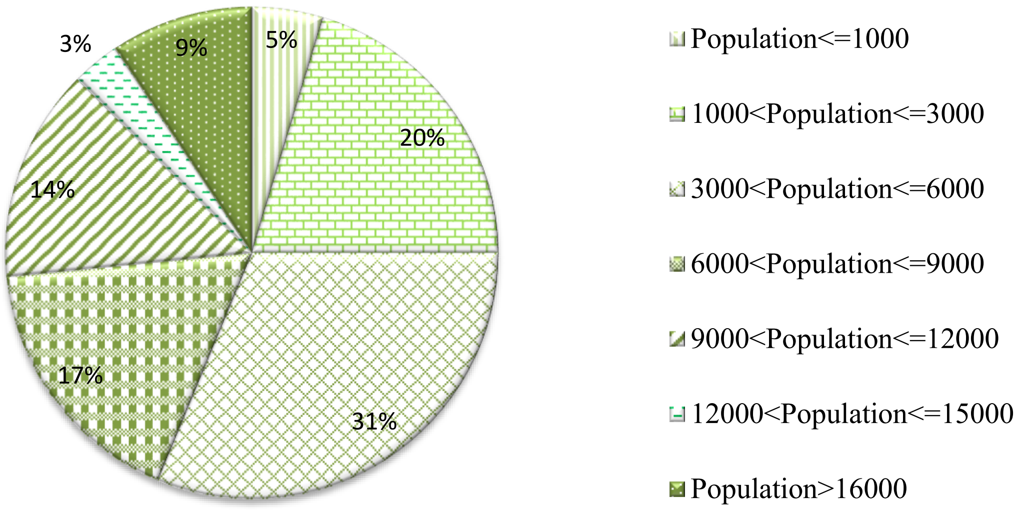

Figure A-1: Municipality size distribution for Alberta

Table A-2: Operating parameters and scale factors used for cost evaluation

Process	Block operation	Process equipment	Cost evaluation method	Parameters / Scale factor (f)
Grinding	Crusher	Gyratory crusher	Process model cost estimation	95% mechanical efficiency Grindability: 20-25 mm Flowrate: 0.5 TPH
Screening	Solid separator	Circular vibratory screen	Process model cost estimation	Screening efficiency: 90%
Drying	R- Stoic	Counter rotary indirect dryer	Process model cost estimation	Temperature: 270˚ C; Air as drying medium f=0.65
Pelletizing		Pellet mill	Scaling-exponent method	Solid capacity: 0.5 TPH f = 0.12 for solid capacity 0.35 - 45
Feeding		Screw conveyors	Scaling-exponent method	f = 0.46 for capacity (TPH) x length (m) = 5000
Intermediate pyrolysis	R- yield	Auger reactor	Scaling-exponent method	Temperature: 400-450˚C; Pressure: 1 atm, material stainless steel f = 0.76
Reforming	R-yield	Vertical process vessel	Scaling-exponent method	Temperature: 600-700˚C; Pressure: 1 atm f = 0.58, material stainless steel
Solid separation	Cyclone	Cyclone solid separator	Process model cost estimation	Gas flow rate: 814 cfm f = 0.35 for flow capacity of 9-1300 L/s
Three step condensation	Condensers	Shell and tube condensers	Process model cost estimation	Stage 1: 280-300˚C; Stage 2 stage: 90-110˚C; Stage 3: 30-40˚C
Hydrogen separation	Separator	Pressure swing adsorber	Process model cost estimation	Gas flow rate: 151.4 cfm Temperature: 303 K

Table A-3: Investment factors used for decentralized and centralized plants

Parameter	Decentralized plants	Centralized plants
Total purchase equipment cost (TPEC)	100% TPEC	100% TPEC
Total direct cost (TDC)	207% TPEC	293% TPEC
Indirect cost (IC)	30% TPEC	66% TPEC
Total direct and indirect cost (TDIC)	TDC + IC	
Contingency	15% of TDIC	
Fixed capital investment (FCI)	TDIC + contingency	
Location factor (LF)	5% FCI	
Total project investment (TPI)	FCI + LF	

Table A-4: Breakdown of total direct costs

Purchased equipment delivered (including fabricated equipment, process machinery, pumps, and compressors)	100%
Purchased equipment installation	25%
Instrumentation and controls (installed)	8%
Piping (installed)	10%
Electrical systems (installed)	10%
Buildings (including services)	10%
Service facilities and yard improvements	40%
Land	4%
Total Direct Cost (TDC)	207% of TPEC