FUEL FRACTIONS DERIVED FROM BIOCHAR THROUGH THE DIRECT COAL LIQUEFACTION PROCESS

Lappeenranta–Lahti University of Technology LUT

Master´s Thesis

Master's programme in Biorefineries

2025

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Examiners: Associate Professor (Tenure Track) Kristian Melin

Professor (tenured) Tuomas Koiranen

ABSTRACT

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Keywords: Sustainable liquid fuel, DCL process, biochar-based DCL process

This thesis evaluates the technical and economic feasibility of utilizing biochar as feedstock for direct coal liquefaction (DCL) process for sustainable transportation liquid fuel production. While DCL process has been utilized for conversion of fossil coal in the past it serves as interesting route to achieve sustainable liquid fuels if biomass-derived feedstock is used. The thesis combines literature review with technical and economic analyses to determine the biochar-based DCL process viability under current market conditions. A mass balance calculation and investment analysis for a 100.000 ton per year (biochar input) production facility reveal favorable economics. Sensitivity analyses demonstrate that the process maintains profitable when final product price is minimum 2,000 €/t. The thesis further evaluates how hydrogen economics and biochar markets could affect long-term viability. Alternative high-value end applications for liquid products are explored. Comparative analysis with Fischer-Tropsch technology indicates that biochar-based DCL process offers potential advantages and topics to further study. The thesis recommends to further study the process via pilot-scale validation. Key technical parameters identified for viability include achieving at least 33% conversion yield from biochar to liquid fuel as well as managing hydrogen consumption as one of the major variable cost element.

TIIVISTELMÄ

Lappeenrannan–Lahden teknillinen yliopisto LUT
LUTin insinööritieteiden tiedekunta
Kemiantekniikka

Harri Huttunen

BIOHIILESTÄ JOHDETUT POLTTOAINEJAKEET SUORAN HIILEN NESTEYTTÄMISPROSESSIN KAUTTA

Diplomityö

2025

69 sivua, 17 kuvaa, 19 taulukkoa and 1 liite
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Tuomas Koiranen

Avainsanat: Kestävä nestemäinen polttoaine, DCL-prosessi, biohiilipohjainen DCL-prosessi

Tässä diplomityössä arvioidaan biohiilen hyödyntämisen teknistä ja taloudellista toteutettavuutta suoran hiilen nesteytysprosessin (DCL) raaka-aineena kestävän liikenteen nestemäisten polttoaineiden tuotannossa. Vaikka DCL-prosessia on käytetty aiemmin fossiilisen hiilen muuntamiseen, se on mielenkiintoinen reitti kestävien nestemäisten polttoaineiden saavuttamiseksi, jos käytetään biomassaperäistä raaka-ainetta. Diplomityö yhdistää kirjallisuuskatsauksen teknisiin ja taloudellisiin analyyseihin biohiileen perustuvan DCL-prosessin kannattavuuden määrittämiseksi nykyisessä markkinatilanteessa. Massataselaskenta ja investointianalyysi 100 000 tonnin (biohiilen määrä raaka-aineena) tuotantolaitokselle osoittavat suotuisan talouden. Herkkyysanalyysit osoittavat, että prosessi säilyttää kannattavuuden, kun vetykustannukset, biohiilen kustannukset ja tuotesaanto ovat kohtuullisella alueella. Diplomityössä arvioidaan edelleen, miten vetytalous ja biohiilimarkkinat voivat vaikuttaa pitkän aikavälin elinkelpoisuuteen. Vaihtoehtoisia korkea-arvoisia loppusovelluksia nestemäisille tuotteille tutkitaan. Vertaileva analyysi Fischer-Tropsch-tekniikalla osoittaa, että biohiileen perustuva DCL-prosessi tarjoaa potentiaalisia etuja ja aiheita jatkotutkimukselle. Diplomityössä suositellaan prosessin jatkamista pilottimittakaavan validoinnin kautta. Kannattavuuden kannalta tärkeimmät tekniset parametrit ovat vähintään 33 % tuotesaanto biohiilestä nestemäiseksi polttoaineeksi sekä vedyn kulutuksen hallinta yhtenä tärkeimmistä muuttuvista kustannuksista.

SYMBOLS AND ABBREVIATIONS

Roman characters

p	pressure	bar, Pa
LHV	lower heating valye	MJ/kg
T	temperature	ºC, K
V	volume	m^3

Abbreviations

CAPEX	Capital expenditure
CTL	Coal-to-Liquids
DCL	Direct coal liquefaction
EDS	Exxon donor solvent
FCOP	Fixed cost of production
FT	Fischer-Tropsch
HTC	Hydrothermal carbonization
ICL	Indirect coal liquefaction
ISBL	Inside battery limits
LCA	Lice-cycle assessment
MP	Medium pressure
LP	Low pressure
OPEX	Operational expenditure
OSBL	Outside battery limits
RITSL	Reconfigured integrated two-stage liquefaction
RWGS	Reverse water-gas shift
SRTC	Solvent refined coal
VCOP	Variable cost of production
WGSR	Water-gas shift reaction

Table of contents

1 Introduction

The global transportation sector faces already existing challenge of reducing its environmental footprint while maintaining the operations that existing society demands. Direct coal liquefaction (DCL) is a process that was developed in the early 20th century and is an alternative solution to serve as an process to more sustainable transportation fuels with modern renewable feedstocks. This thesis investigates the technical and economical feasibility of replacing traditional fossil coal with biochar in DCL processes offering another route towards carbon-neutral liquid fuels without requiring complete change of existing fuel infrastructure.

Previous research has demonstrated promising results in the conversion of biomass-based biochars to liquid fuels. Studies have shown that biochar produced from agricultural rice straw can serve as a feedstock for liquefaction processes. Biobased feedstocks have low amount of sulfur making them attractive also from the total investment point-of-view. Sulfuric acid plant or flue gas cleaning for sulfur is not needed. Attractive path would be to reach direct route to marine fuel.

However, significant gaps exist in the current research situation. Most previous studies have focused on low-temperature liquefaction leaving the high-temperature studies unexplored. The high-temperature liquefaction process could be an attractive process to reach ultimately high quality liquid fuels. Moreover, while laboratory-scale studies have shown technical promise, full economic feasibility studies for industrial-scale implementation remain also unexplored. The critical questions of whether biochar-based DCL can be economically feasible with today´s market conditions and what scale would be required for commercial success have not been studied in detail.

This thesis aims to evaluate both the technical and economical feasibility of implementing a biochar-based DCL process in today´s environment with present costs. The research combines literature review with detailed mass balance calculations and investment analysis for a proposed 100,000-ton (biochar feedstock) production facility. The research questions highlight whether biochar has the appropriate technical properties to serve as DCL feedstock

and if such an process can achieve economical success at current market conditions. Limitations of this study include the correlation on literature values for technical and economical parameters and the uncertainties in scaling up laboratory processes or executed DCL plant values from the history. The thesis is structured in two main parts having a literature review first and followed by a commercial feasibility study including techno-economical analysis of the biochar-bsed DCL process.

2 Direct versus indirect liquefaction process

2.1 Direct liquefaction process

Coal direct liquefaction technology is a key method for efficient and clean use of coal resources (Song et al., 2025). Coal liquefaction is a process where coal is converted into a liquid fuel by using direct or indirect liquefaction process. Direct liquefaction process consists of pyrolysis, solvent extraction and catalytic liquefaction (Basha et al., 2016). Pyrolysis involves heating coal to approximately 400ºC to convert it into gases, liquids and char. During solvent extraction coal is mixed with a solvent that can donate hydrogen molecules at temperatures of approximately 500ºC and pressure approximately 340 bar. Target is to break down the coal into lower molecular weight products. Catalytic liquefaction therefore uses catalysts like metal sulfides. FeS and FeS_2 or acid catalysts like $FeCl_3$ and $ZnCl_2$ are typical catalysts to help to inject hydrogen into coal.

Direct liquefaction process produces liquids which still contains various contaminants. Contaminants like sulfur, nitrogen and possibly metals are required to remove in order to achieve clean fuels specification (IIT Roorkee, 2018).

Liquefaction process target to form smaller aromatic units that can exist as liquids. Aromatic polymer is first broken down into smaller coal macromolecules and then to separate aromatic units. Three level of fractions are formed as shown on Figure 1.

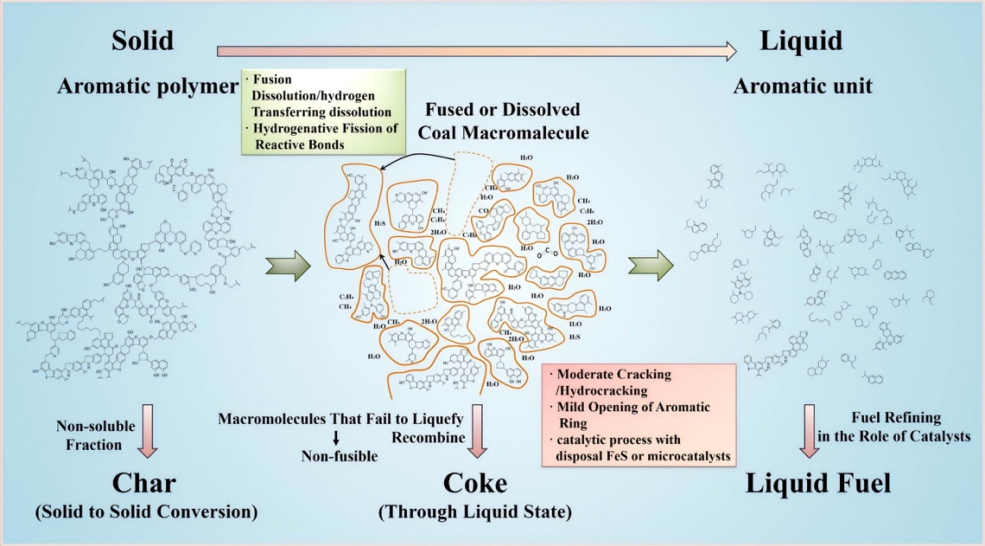


Figure 1. The chemical process of direct coal liquefaction (Song et al., 2025)

The coal molecules that resist liquefaction aggregate into larger structures or convert into gas molecules. Importance of sufficient hydrogen is high since hydrogen should stabilize the radicals and stop them to combine and form higher molecular weight semi-cokes or cokes.

The coal liquefaction process is good for the countries with lack of crude oil but with resources of coal (Song et al., 2025). Liquefaction process produces liquids by altering the coal structure chemically and increasing the content of hydrogen in relation to carbon. A significant portion of the process costs arises from the low hydrogen content in coal and demand for increased supply of the hydrogen during the process (Fuel and Energy Lecture 17 Liquefaction of coal ICL 2nd, 2021). Developing this technology and alternative energy sources are important steps to ensure both energy security and economic stability. Anyway, there are several issues to solve like the low conversion rate, limited oil yield, considerable coking, demanding reaction conditions and high energy demands. Research has been done for the issues, but still certain areas need more focus and understanding. These areas are understanding the pyrolysis of large molecules during the process, hydrogen activation, catalyst stability and performance, hydrogenation of solvents as well as interactions between free radicals. These parameters and variations within them make the simulation complex and time consuming.

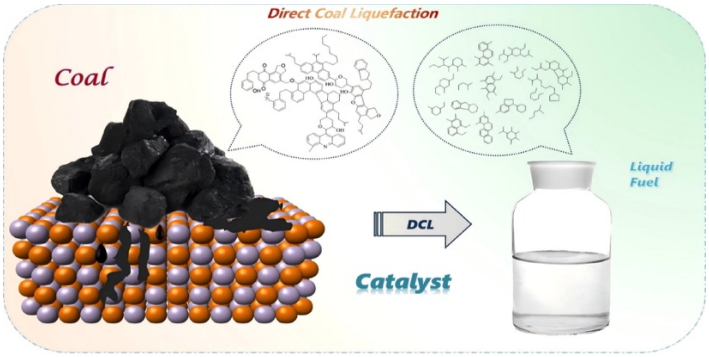


Figure 2. Direct coal liquefaction with catalyst (Song et al., 2025).

The use of coal can be mainly divided into two areas. First is direct combustion and second is conversion. Direct combustion is most often used in industrial boilers, home heating or in power generation. Conversions methods include coal gasification and other coal processing. Target of conversion is to produce cleaner gas or liquid fuels or different chemicals through the conversion process. Today direct coal liquefaction is mostly used for producing oil as an alternative product for petroleum. Alternative term for direct coal liquefaction (DCL) is CTL meaning Coal-to-Liquids.

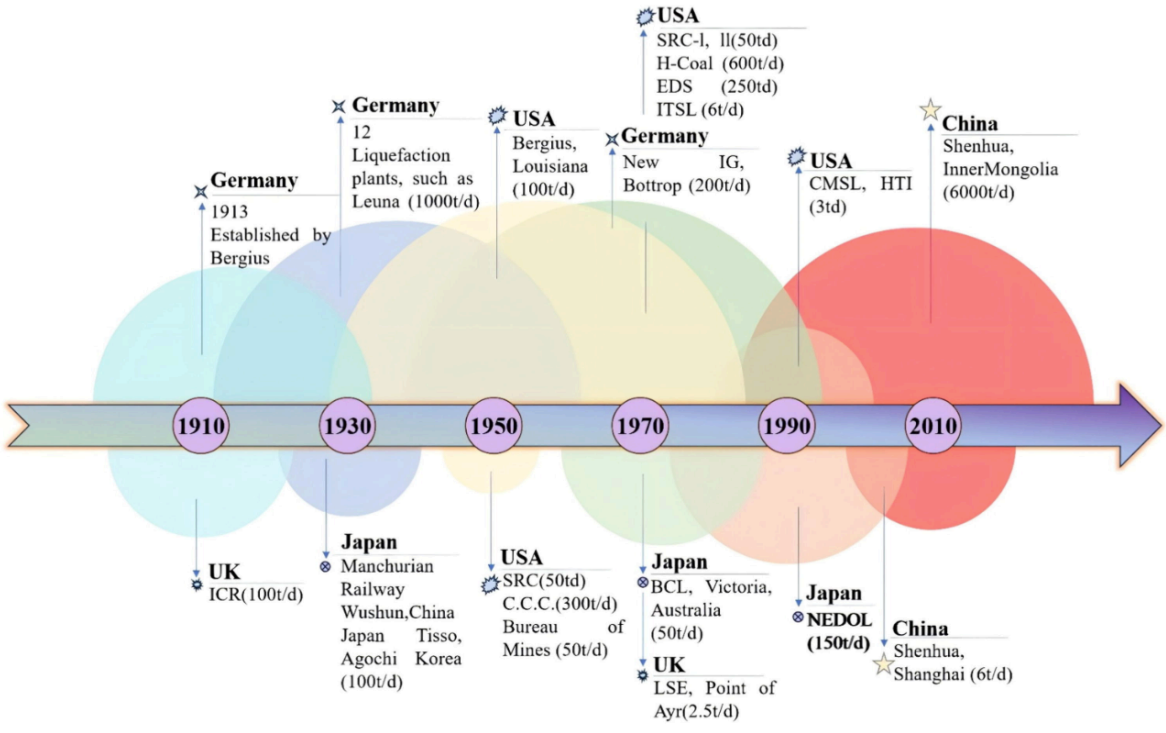


Figure 3. History milestones of DCL development (Song et al., 2025).

Direct coal liquefaction process was involved in Germany by Friedrich Bergius in 1913. Afterwards many plants have been started but also closed. Again between 1970s and 1990s more attention and focus was given to a process due to impact of oil crisis (Wang et al., 2024). Latest reference is Shenhua plant in China that operates using bituminous coal with process scale of 6000 t/d. Shenhua plant operates using synthetic Fe sulfide catalyst. This iron-based catalyst has been employed in the Shenhua plant since it has properties like high activity, low cost and minimal environmental pollution (Song et al., 2025).

2.2 Indirect liquefaction process

Indirect liquefaction processes (ICL) means reacting coal with steam and oxygen (Basha et al., 2016). The process requires gasifying the solid feedstocks into syngas first (National Energy Technology Laboratory, no date). In comparison to DCL process the ICL process do not convert coal into a liquid phase directly. ICL process involves two main steps: (1) gasification to produce syngas (synthesis gas), and (2) converting the hydrogen (H_2) and carbon monoxide (CO) in the syngas into different hydrocarbon fuels. Hydrocarbon fuels are like gasoline, methanol, diesel and various chemicals. Most used process is Fischer-Tropsch (FT) synthesis followed by refining of fuels. Methanol from syngas can be converted into gasoline utilizing MTG process invented by ExxonMobil. In comparison to DCL process the ICL process do not need external hydrogen source. DCL process can receive the additional hydrogen from gasifying additional coal or as the residue from the DCL process. ICL process has been in commercial use by Sasol since the 1950s.

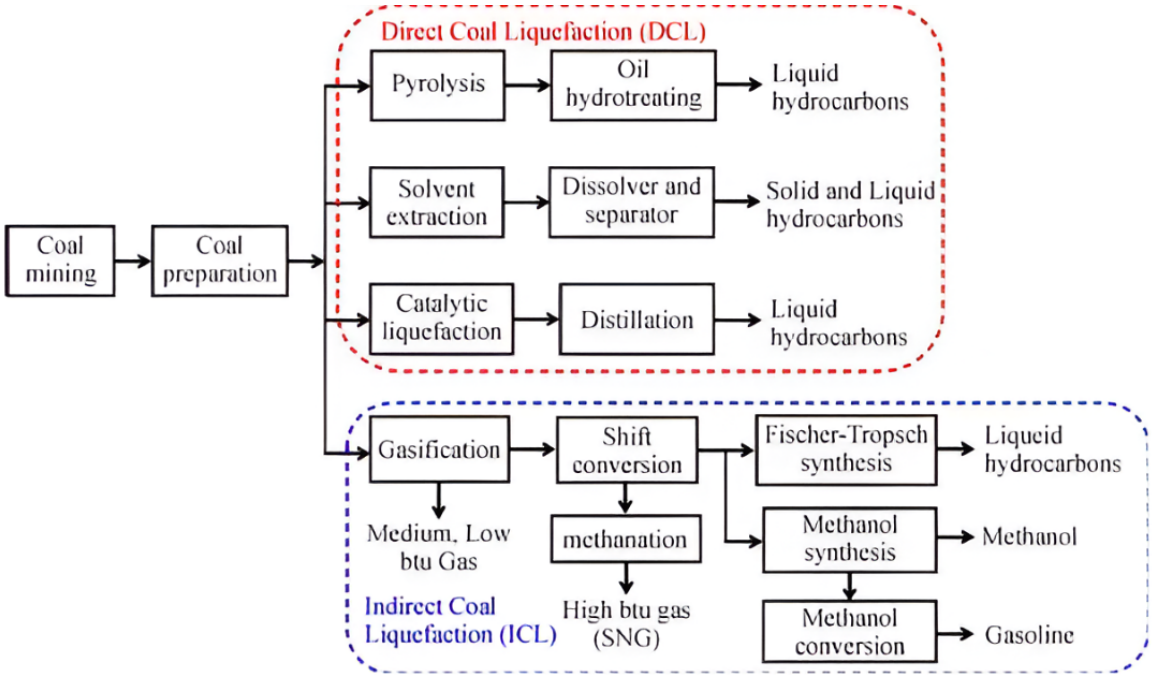


Figure 4. ICL and DCL process chart (Basha et al., 2016)

3 Types of direct coal liquefaction

Several different DCL processes have been developed over the decades of 20st century (Wang et al., 2024). Presented in the Figure 3. SRC-II was developed in USA, H-Coal in USA, IGOR+ in Germany and NEDOL in Japan. However, most of the technologies and plants were never industrialized in full scale. Since 2008 China Shenhua DCL plant has been in operation being world´s first million-ton DCL plant. Shenhua plant has been commercially active around 15 years, and it is the only large-scale industrial plant operating DCL process.

Process	Country	Catalyst	Reaction condition	Process scale	Coal
IGOR	Germany	Fe sulfide	460-470 °C, 30-31 MPa	200 t/d	brown coal, bituminous
NEDOL	Japan	Pyrite	460 °C, 17-19 MPa	150 t/d	bituminous, sub-bituminous
H-Coal	H-Coal	Co-Mo	455 °C, 20-21 MPa	200-600 t/d	bituminous, sub-bituminous
EDS	USA	none	425-250 °C, 14-18 MPa	250 t/d	brown coal, bituminous
SRC-II	USA	none	460 °C, 14 MPa	30 t/d	bituminous, sub-bituminous
CC-ITSL (HTI)	USA	Co-Mo, Ni-Mo	450/400 °C, 17 MPa	6 t/d	bituminous, sub-bituminous
BCL	Japan	limonite	450 °C, 15 MPa	50 t/d	brown coal, sub-bituminous
CT-5	Russian	emulsified Mo	425 °C, 10 MPa	10 t/d	brown coal
Shenhua	China	Synthetic Fe sulfide	450 °C, 18 MPa	6000 t/d	bituminous

Table 1. Major international CTL process plants (Song et al., 2025)

3.1 Single stage DCL processes

In single stage process the coal is heated and treated directly in one reaction temperature producing a liquid fuels (Song et al., 2025). During the single stage process the structure of coal is fractured and reorganized. These substances are typically hard to produce using conventional processes and the substances easily aggregate. Single stage process delivers the liquid product directly from the operating reactor or reactors operating in a similar conditions in a row. Anyway, single-stage process might be a problematic process for lower-rank coals. At elevated reaction temperatures types of bonds might be disrupted. Some bonds hydrogenate too slowly to capture free radicals formed. In addition, their coupling reactions might lead to lower liquid yields. Connected with high capital and operational costs the single-stage process competitiveness is not the best one. Research proposes that further study for catalysts efficiency, more efficient reactors and slurry preheaters as well as effective carbon dioxide management systems should be done.

SRC-II, EDS (Exxon donor solvent) and H-Coal processes have been single stage DCL processes (IIT Roorkee, 2018). SCR-II is a solvent refined coal liquefaction process. The process uses solvent only and no catalyst is used. Coal is first crushed and grinded to small particles. Coal slurry is produced mixing the crushed coal particles and recycle solvent from the process. Recycled solvent is produced in the heating and vacuum distillation step and recycled back to early stage of the process as solvent. After feed slurry mixer the hydrogen is added. Hydrogen can be added from two sources. First is the gasifier step that uses steam, oxygen and coal to produce hydrogen. Second source is recycled hydrogen from the gas separation and purification step from where the hydrogen is recycled back to the early stage of the process. At this point the slurry is formed and hydrogen added. Next step is to preheat the slurry up to 425ºC before pressurizing the slurry in dissolver. 12 MPa pressure is used in dissolver with around 440ºC of temperature. Conversion is reached within time of 15-45 minutes in dissolver. Vapor and liquid separation take place after the dissolver step. Gas product from the vapor-liquid separation continues to purification process from where the recycled hydrogen is led back to process while fuel gas is led to fuel gas plant. Product slurry is led to light liquid separator where light distillates are produced. Separator operates at 1 MPa pressure with 290ºC of temperature. Separator produces naphtha and filter residue.

Filter residue is led back to the process for gasifier step to produce makeup hydrogen for the process. Heavy liquid after the filtration step is led to heating and vacuum distillation. The distillation process step produces heavy distillate back to the early stage of the process, heavy distillate fuel oil, solvent refined coal as well as wash solvent in vapor form back to filter.

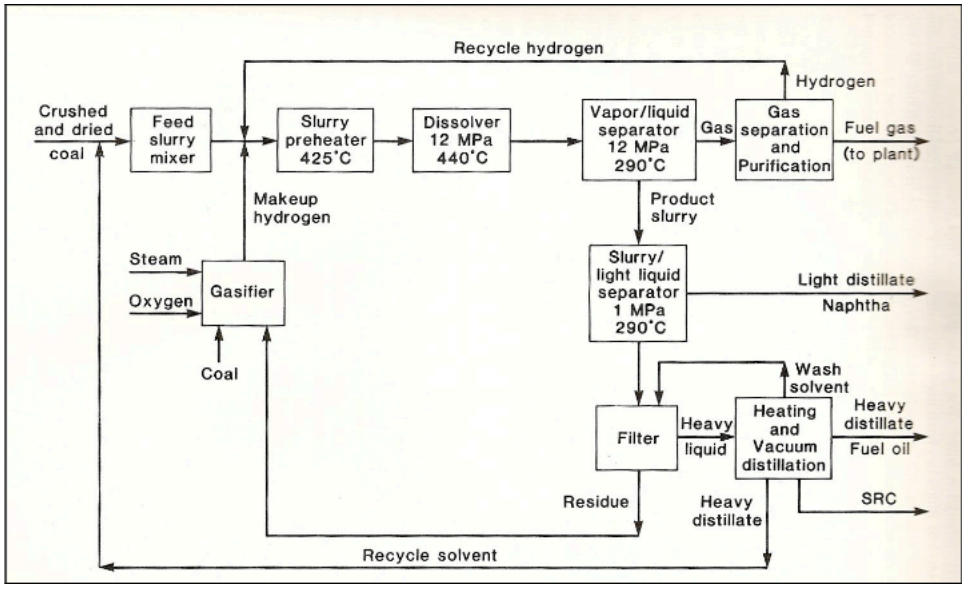


Figure 5. Flowsheet of SRC-II process (Robinson, 2009).

Process yields for SRC-II process are presented in Table 2. Data is from Kentucky plant using bituminous coal as feedstock. Comparison to other SCR-II plants is not done. Comparing data to Illinois bituminous plant utilizing EDS process shows that in SRC-II process the fuel gas production is much higher than in EDS process. SRC-II process produces fuel gas C_1–C_4 of 17,6 wt% while EDS process produces only 7,3 wt% of C_1-C_3 gas. Quite high variation is also for naphtha production. SRC-II produces 13,0 wt% of C_5-350F naphtha while EDS process produces almost 39 wt% of C_4-1000F liquid.

	Wt%
Consumed	
Hydrogen	4.8
Total	4.8
Produced	
Fuel Gas (C_1–C_4)	17.6
Naphtha (C_5–350 °F)	13
Fuel Oil (350–850 °F)	25.8
Solvent Refined Coal	26.5
Unconverted Carbon	6.3
Hydrogen Sulfide	2.5
Carbon Oxides	2
Water	5.7
Ammonia	0.6
Total	100

Table 2. Process yields of SRC-II process (Robinson, 2009).

EDS is Exxon donor solvent process and is one of the single-stage DCL processes (IIT Roorkee, 2018). Early stage of the process is like SCR-II process. Coal is prepared and injected into slurry mixing tank with recycle solvent injection before the mixing tank. Recycle solvent is led to a coal slurry from the fixed bed hydrotreater instead of vacuum distillation (SCR-II). Fixed bed hydrotreater receives the liquid coal from the vapor-liquid separator and produces solvent to be recycled back to process. Tubular reactor in EDS process operates in pretty much the same temperatures versus SCR-II process but with higher pressure. Pressure in EDS process is around 13,5 MPa instead of 12 MPa (SCR-II). Tubular reactor operates without catalyst. Gas-liquid separation is just after the tubular reactor. Vacuum distillation after the gas separation step produces naphtha, distillate fuel oil and solid components to flexi coker. Flexi coker produces distillate fuel oils and coke to gasification.

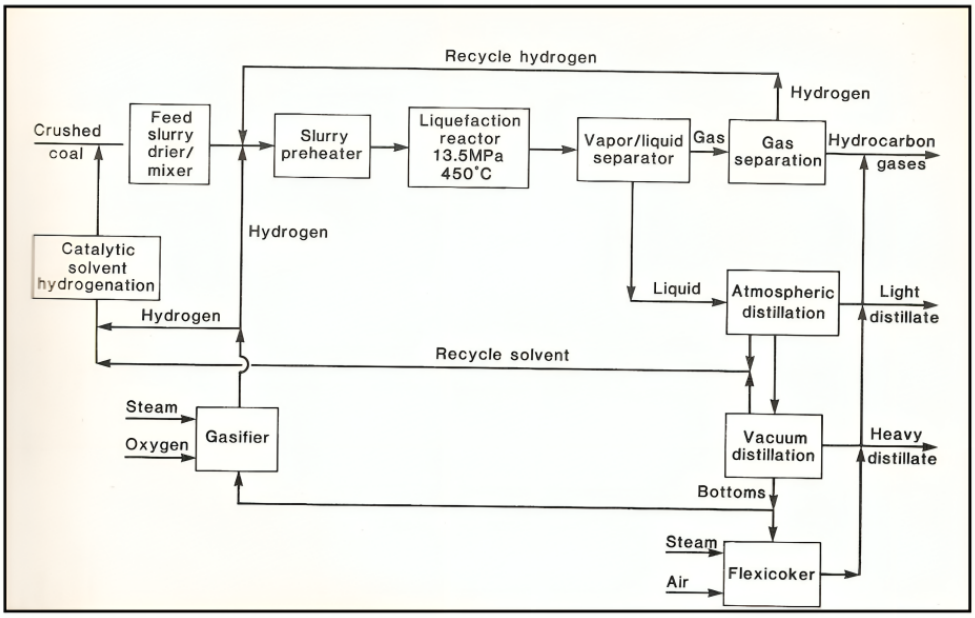


Figure 6. Flowsheet of Exxon donor solvent process (Robinson, 2009).

Process yields for EDS process are presented in Figure 9. for EDS process in Illinois, Wyoming and Texas plants (Robinson, 2009). Residence time varies from 25 minutes to 60 minutes. H_2O+CO_x yield varies much depending the plant and type of coal used. Illinois bituminous coal yield for H_2O+CO_x is only 12,2 wt%. Wyoming and Texas subbituminous and lignite coals produces around 21-22 wt% yield. C4–1000F liquid yield stays between 33-39 wt% for all plants.

	Illinois No. 6 bituminous (Monterey)	Wyoming subbituminous (Wyodak)	Texas lignite (Big Brown)
Residence time (min)	40	60	25-40
Yields (wt% maf coal)			
H_2	-4.3	-4.6	-3.9
$H_2O+CO_{(x)}$	12.2	22.3	21.7
H_2S+NH_3	4.2	0.9	1.7
C_1-C_3 gas	7.3	9.3	9.1
C_4-1000 °F liquid	38.8	33.3	33.3
1000 °F + bottoms	41.8	38.8	38.1

Table 3. Product yields of EDS process (Robinson, 2009).

3.2 Two stage DCL processes

The process consists of two stages both with unique operating conditions (Song et al., 2025). Both reaction conditions use independent reaction temperature. In the first stage the coal is dissolved in high reaction temperatures breaking the coal into highly functional fragments. Second stage include refining of the coal within lower temperatures and with help of catalysts. The idea of using catalysts is to reduce functional groups, promote hydrogenation and lower molecular weight of coal. The approach improves product selectivity and decrease the number of impurities like sulfur and nitrogen. Purer liquid product after coal liquefaction eases the process during fuel refining. In two stages process the carbon utilization is more optimized resulting in higher conversion rates and minimized waste.

Two stage concept was first developed by Chevron Research in the late 1970s in the United States (Robinson, 2009). Pilot plant tests conducted by HRI, Inc. in mid 80´s demonstrated exceptional results. Distillate yields were high while residuals production was minimal. In 1986-1987 Wilsonville SRC plant was upgraded to tightly couple the first dissolver and second stage hydrotreater. This configuration was named as RITSL (Reconfigured Integrated Two-Stage Liquefaction). There was a discussion about sequencing the temperature stages since differences between high/low and low/high combinations were minimal. HRI preferred to have low/high combination in use while Amoco preferred to have high/low combination to achieve more aromatic saturation in the second stage.

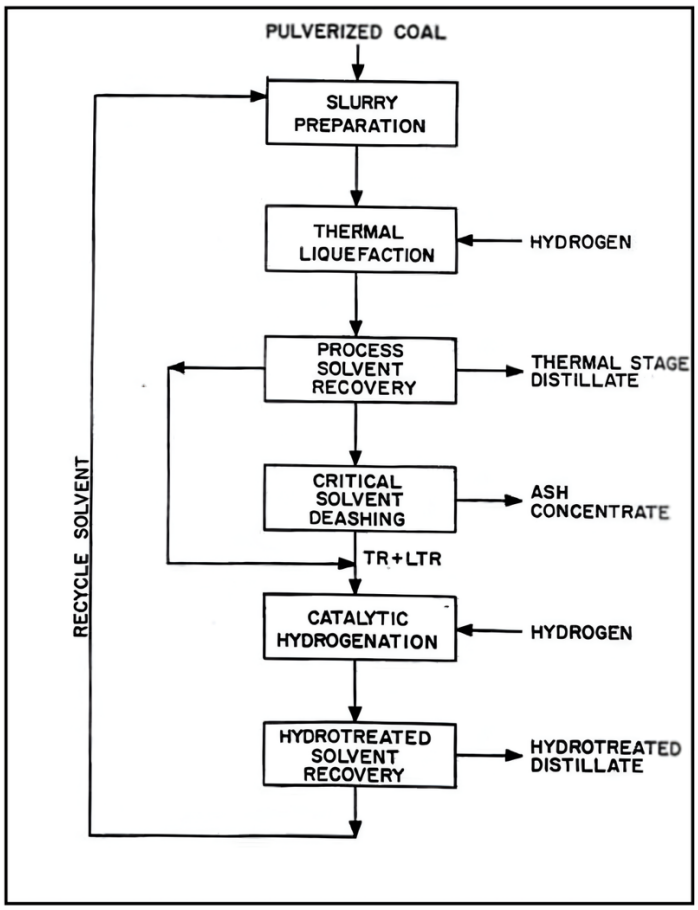


Figure 7. Flowsheet of ITSL two stage process at Wilsonville (Robinson, 2009).

Largest two stage direct liquefaction reference in China is Shenhua DCL plant (Nolan et al., 2004). Shenhua is one of the world's largest coalfields. The process development is based on HTI process in the United States and was firstly discussed between the top leadership of both countries. Besides China had to consider its energy security in early 2000 it had three other features to also tackle regarding energy infrastructure. First aspect was to enhance environmental protection since direct coal liquefaction can reduce the direct combustion of coal therefore environmental pollutions. Second aspect was to find new growth area and offer more work to people who lose their jobs due to industry automation. Third aspect was to assist with the formation of a common strategy for industrial growth and its energy needs.

Shenhua direct coal liquefaction process is presented in Figure 8. The process utilizes two backmixed reactors as well as proprietary dispersed iron catalyst. Additionally fixed-bed in-

line hydrotreater is used. Unconverted residuals are recycled and used for hydrogen production. Process is using low/high temperature combination for reactors. The process operates at 17 MPa and with temperatures in range of 400-460C. The effluent after the first stage goes to separation to remove the gases. Slurry stream is sent to the higher temperature reactor as second step. Effluent after second stage is processed in hot separator. The vapor stream from the hot separator is led to the fixed-bed in-line hydrotreater. Hydrotreater upgrade the stream into very clean fuels. The overhead stream from the atmospheric flash is firstly condensed and then led to hydrotreater also. Major liquefaction product comes from the hydrotreater in form of naphtha and diesel fuel fractions. Bottom stream of atmospheric flash containing mainly solids is recycled to vacuum still and then to solids separation. Solids are partly used again as hydrogen production and partly recycled.

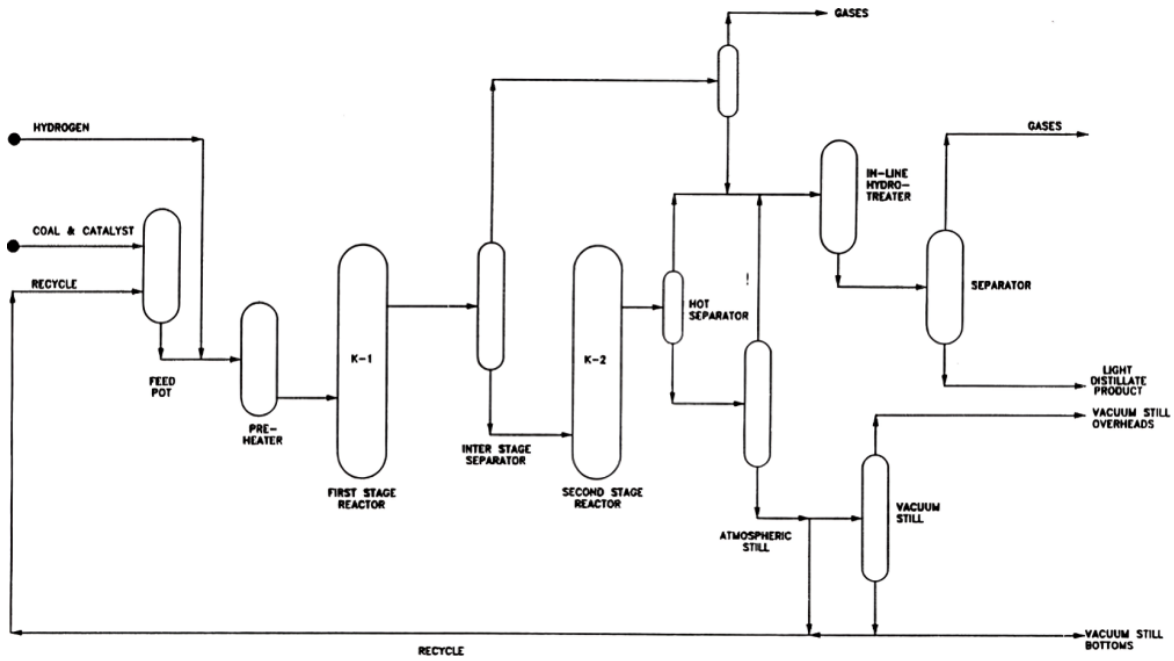


Figure 8. Flowsheet of Shenhua two stage DCL process (Comolli et al., 1999)

3.3 Catalyst options for direct coal liquefaction processes

Both processes are limited by the availability of appropriate catalysts. Table 4 explains the advantages and disadvantages of each process and their catalysts options. For single stage process the suitable catalysts are synthetic high-efficiency iron-based options like FeOOH

and molybdenum-based catalysts such as nanoscale molybdenum trioxide which offer high activity and ease of use. For two stage processes more suitable catalysts are solid acid catalysts and composite catalysts including combinations like Ni-Mo, Co-Mo and Fe-Mo. Two stage process catalysts are more complex, but they offer higher conversion rates and improved product quality. Typically, single stage process has been prioritizing highly active catalysts while two stage process has favoured more complex combinations to achieve superior end product quality and improved conversion rate. Selection of catalysts is important to ensure efficiency, suitability with different coal types, regeneration abilities, minimize catalyst consumption and lower the operational costs.

Advantages	Simple and easy to operate	Higher conversion rates and conversion of coal
	Lower equipment requirements	Equipment adaptability is strong, can handle a variety of coal types
	Relatively low energy consumption	Relatively low energy consumption due to optimized hydrogen utilization
	Average fuel quality	Higher fuel quality with lower sulfur and nitrogen content
	Low investment cost	Lower long-term operating costs and high conversion efficiency
	Single stage process	**Two-stage process**
Disadvantages	Lower conversion rates and incomplete conversion of coal	Complex operation and high investment cost
	Poor adaptability to coal types, difficult to deal with hugh sulfur, high nitrogen coal	Higher equipment investment and maintenance costs
	Higher energy consumption because more hydrogen is needed for conversion	Need for more hydrogen resources
	Higher levels of impurities such as sulfur and nitrogen in the fuel	Complex productio process, difficult to operate
	Higher long-term running costs and lower conversion rates	Higher initial investment costs

Table 4. Comparison of single and two stage approaches (Song et al., 2025).

4 Bergius Pier DCL process

4.1 Friedrich Bergius

Friedrich Bergius, born on October 11th 1884, in Goldschmieden, Silesia, came from a family of scientists and businesspeople (*The Nobel Prize in Chemistry 1931*). Influenced by his father´s chemical factory, he developed and early interest in industrial and scientific methods. He completed his PhD in 1907 after studying chemistry at universities in Breslau and Leipzig. He is famous for the work in high-pressure chemistry, especially the hydrogenation of coal and heavy oils.

Bergius focused on coal liquefaction during World War I. He ended up to large-scale industrial process by the 1920s. He got an honour to receive the Nobel Prize shared with Carl Bosch in Chemistry in 1931 for developing high-pressure chemical processes. At later stage Bergius focused on process converting cellulose to sugar. He moved to Argentina after World War II and died in Buenos Aires on March 30th, 1949.

4.2 Background and early stage of Bergius-Pier process

The Bergius-Pier process, which played an important role in the early development of hydrotreating, originated as an effort to hydrogenate coal into liquid fuels (van Veen, 2017). This method of converting coal into synthetic liquid fuels was driven by increasing demand for alternative sources of energy and fuel in the early 20th century. Friedrich Bergius, a German chemist, first developed the concept in the 1910s, but it didn't reach it´s full potential until the involment of the German chemical company BASF (Badische Anilin- & Sodafabrik) and Matthias Pier.

At early-stage Bergius´ experiments focused on the single stage and high-pressure hydrogenation of the coal. His target was to break down the molecular structure of coal and convert it into a more valuable liquid hydrocarbons. By converting coal to hydrogen gas at

high pressures and temperatures, Bergius achieved partial success, laying the foundation for the future development. His project was considered as high potential but he faced many technical and economic challenges regarding catalysts and corrosion problems.

4.3 Challenges and collaboration

The early commercial realization of the Bergius process proved to be a huge financial burden (van Veen, 2017). Large-scale coal hydrogenation to liquid fuels required substantial capital investments making the process costlier than expected. BASF, later part of the IG Farben, was one of the companies that took on the challenge of scaling up this technology into industrial level. While BASF´s efforts led to substantial progress, international collaborations were mandatory to keep the project alive. To secure additional funding the collaboration with international companies like Shell were pursued.

Despite the challenges, increasing demand for German fuel autonomy provided a strong base to continue the development project for the Bergius´s process. Especially the time before World War II boosted the demand for fuel production in Germany. Under the time of Nazi government, the Bergius-Pier process received substantial support due to its potential to minimize the German´s dependence on imported oil. This state-backed initiative was part of wider strategy of Germany to target energy self-sufficiency especially as nation prepared for war.

4.4 The role of Matthias Pier and catalyst development

Matthis Pier was extremely important person in addressing the technical challenges that hindered the Bergius process in the early stage of development project (van Veen, 2017). His main contribution came through the development of sulfur-tolerant catalyst and dividing the process into two stages that were part of the solution for large-scale coal hydrogenation. First stage produced “middle oil” and second stage converted the “middle oil” into desired fuels. As a result, the throughput was considerably increased as well as product properties were better. Bergius process used, in early stage of the development, iron-based catalyst

similar to those in ammonia synthesis. The catalyst was found as quickly poisoned by sulfur which is typical impurity in coal. Poisoning of catalysts led to ineffective and limited process. In addition the sulfur couldn't be removed from coal prior to hydrogenation.

Pier came up with an idea to test iron sulphide in December 1924. His group of scientists grew up to seventy chemists and engineers while thousands of laboratory workers were also included in project to systematically test elements and compounds from periodic table. At same time he noticed that sulphides of Cobalt (Co), Molybdenum (Mo) and Tungsten (W) were very active catalysts. Next year in 1925 the brown-coal tar was successfully hydrogenated at 200atm pressure with Mo catalyst into pure gasoline using one stage process without forming a coke.

4.5 The two-stage approach

In the beginning the process was single-stage process using brown coal as feedstock (van Veen, 2017). Problem of the single-stage process was stability and performance of catalyst. Complexity of the reaction proved to be too much for a single stage system to handle. First-stage catalyst was injected as powder solution into a paste of crushed coal. The catalyst was impossible to recover from the solution that was a problem also for the future of the single-stage process. Two main candidates were tested as Mo-based catalysts. MoO3/kaolin/CrO3 and MoO3/ZnO/MgO were both selected formulations that should be firstly stable while also active and selective and secondly strong especially in the case of the second stage fixed bed catalyst. In addition, Pier´s group was still unexperienced with the catalyst preparation methods in order to optimize the preparation process.

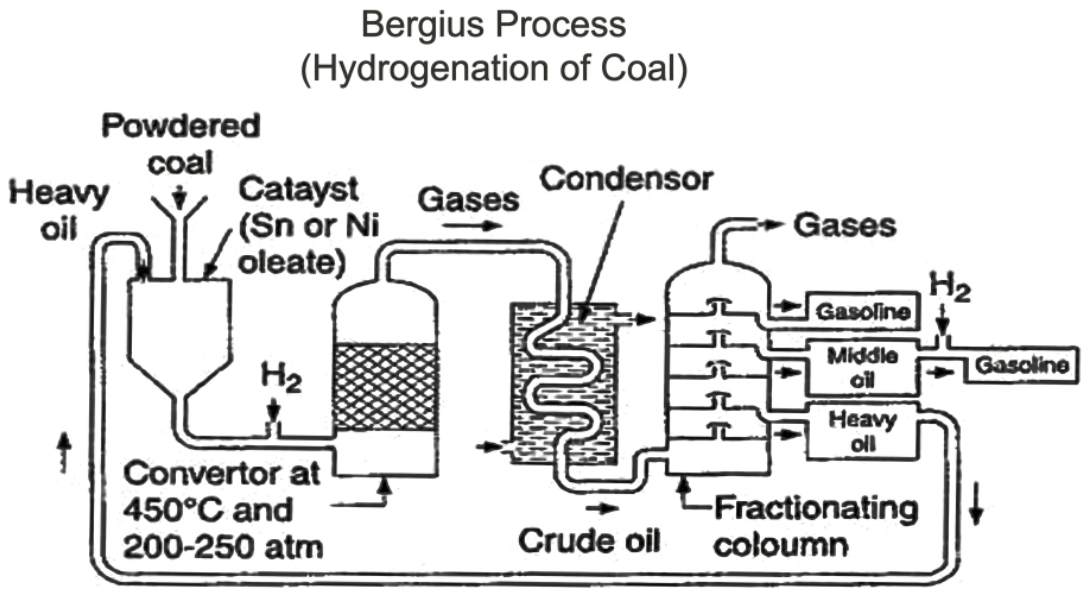


Figure 9. Bergius process – hydrogenation of coal (SynSel, no date).

In the pilot plant test phase the MoO3/ZnO/MgO formulation proved to be more active catalyst and was selected for the first commercial size plant. It was found in catalyst development that replacing the molybdic anhydride to molybdic acid increased the catalyst activity. This innovation led to new catalyst 53.5%w MoO3/30%w ZnO/16.5%w MgO. The new catalyst were taken into use in Leuna plant but at the end it was proven that economics of Mo based catalyst looked unfavourable. The whole research for a cheaper catalyst ended back to iron.

At first all kind of Fe based formulations did not work well. Progress was achieved using low-density carrier like 65%w carbon-rich fly ash from a Winkler coal-gasification plant. The most active catalyst was the one with ammonium molybdate impregnation. Also, Fe based catalyst with impregnation of $FeSO_4$ and NaOH. This formulation remained in service in Leuna plant from 1928 until 1959 when it was replaced by a modern hydrocracking process. In other countries like France and England various catalysts were researched at the same time. Dehydration catalysts in France and Sn formulations with HCl in England proved to be an excellent choice.

The second stage of catalyst development included the target to find formulation with strong physical durability and stability while achieving a lifespan minimum 11 months. Many promising formulations were found but many were discarded since they didn't pass the lifespan requirement. 53.5%w MoO_3/30%w ZnO/16.5%w MgO is a catalyst that was used in the first stage. It was chose to industrial use due to it´s good activity, selectivity and mechanical strength. It was used for liquid and vapor phase hydrogenation at around 450°C. It was working well for three years, and it was primarily a refining catalyst.

The originally used Mo and W oxides were sulphided. Pre-sulphiding the catalyst with H2S lowered the reaction temperature of hydrocracking but led to corrosion issues preventing scaling into large scale application. This led to development of "famous Pierean tungsten sulphide". This catalyst tripled the throughput and even with 50°C less temperature compared to 53.5%w MoO_3/30%w ZnO/16.5%w MgO catalyst. Despite the success the high tungsten content of the catalyst turned out to be a problem. Demand in the steel industry was high at that time. Another problem was its strong hydrogenation power that led to a gasoline with too low octane number.

Higher octane number was achieved developing again a new catalyst but the sensitivity to organic nitrogen in the feed increased again too much. Commercial application for this catalyst looked uncertain. This led to divide the vapor phase process into two steps. A first refining step to remove oxygen and nitrogen compounds and second hydrocracking step to produce gasoline.

The Pierean tungsten sulphide catalyst was successfully applied in refining gasolines, but its high hydrocracking activity required some modifications. This led to a development of 4.5%w NiS/25%w WS_2/70.5%w alumina catalyst with improved performance. This catalyst was main catalyst for German coal-hydrogenation plants.

4.6 Impact and later stage for Bergius-Pier process

By the end of 1930s and early 1940s several industrial plants operating Bergius-Pier process were primarily in Germany (van Veen, 2017). These plants were in a significant role in producing synthetic fuels for German army during the war. Over 11 plants started between 1936 and 1943. Even the political and economic circumstances kept the Bergius-Pier process alive the technical breakthroughs by Pier and his team laid the groundwork for the modern hydrotreating and hydrocracking processes used still at nowadays.

Main innovations by Pier were sulfur tolerant catalyst and two stage process. The Bergius-Pier process became obsolete in the post-war era, but it influenced to processes conditions still applied in today's plants. Pier´s team´s development work around researching better catalyst help the research even today especially in industries requiring conversion of heavy feedstock into lighter products.

5 Reasons why DCL have been closed and what have been the challenges in process?

For many years, cost control has been a key concern for developed countries and a significant factor preventing coal liquefaction from achieving large-scale production. (Nolan et al., 2004). The United States prioritized coal liquefaction in 1978 aiming to produce 50 million tons of oil products by 1990 and 150 million tons by 2010. Between 1978 and 1999 the United States invested $3.5 billion in research and development of coal liquefaction. As a result, the investing and research was stopped due to high costs and inability to compete with the oil prices on global market. The import of oil products has significantly increased in countries where oil is not a common feedstock but who have more coal reserves like China and South-Africa (Kong et al., 2022). As Germany is country with abundant coal reserves but almost no petroleum deposits the natural interest towards utilizing coal as feedstock for gasoline engines fuel was born around World War II. (Miller, B. G. 2017) Also South-Africa studied how to convert coal reserves into liquid fuels and chose Fischer-Tropsch process that German had successfully used. Also, in United States the interest in converting coal to liquid

fuels has been cyclic. It has been affected by cost and availability of petroleum. Petroleum after its discovery quickly became dominant over the coal but expanding automotive industry again rose the concern of depleting petroleum reserves. After up and down movements for interest for coal liquefaction the major discovery of petroleum in the Middle East in the mid-1940s finally resulted coal liquefaction to be uneconomical. All in all, the coal liquefaction process development globally has been suffering the cyclical interest and investing on it mainly due to petroleum supplies. Coal liquefaction with sustainable energy sources is an opportunity for the mentioned type of countries. These countries could also benefit to decrease the dependence on international oil import and countries supplying crude oil. Mentioned countries could also improve national energy security and adjust the energy structure of the coal industry.

The COVID-19 pandemic and global climate change have significantly impacted on energy markets and primary energy use (Li et al., 2023). China is still increasing the primary energy consumption by 2.1% mainly driven by coal. Low-rank coals like lignite and subbituminous coals are inexpensive and abundant. Since the oil crisis in 1970s, wide interest in alternative oil producing technologies has been surging again. Step by step DCL technology has been researched more and various efforts around process parameters have been done in countries like Germany, the United States, Japan, Russia and China. Nevertheless, DCL technology research of low-rank coals has faced several problems. High moisture content (25% - 70%) requires drying pretreatment. The oxygen content (15 % - 30%) is high. In addition, process requires large amount of expensive hydrogen.

During the early steps of DCL development, the EDS and H-Coal processes demonstrated the feasibility of technical and engineering approaches (Winslow, accessed 29.12.2024). Department of Energy (DOE) in United States supported DCL research by $3.6 billion during 1976-2000. Program included two phases; Phase I to develop the technology as a short-term solution for 1970s energy crisis and Phase II to solve the technology problems occurred in Phase I. SRC, EDS and H-Coal processes were included in the program. Many details around process yield, selectivity of catalyst, product quality and economic potential were not solved during Phase I. Later on, the Lummus-Crest and Wilsonville projects

pointed out several options to improve and further develop the process conditions. Improvements achieved in distillate yields and product quality. Naphtha and mid-distillates yields increased from 50% to 70%. Especially increased distillate yields were breakthrough due to nature of high capital-intensive process. Liquids quality improved from Phase I to Phase II. In Phase I liquids were unstable, highly aromatic and had high heteroatom contents (sulfur, nitrogen, oxygen). Phase II introduces liquids with no metals and low levels of heteroatoms. These improvements helped the further processing of liquids into high-quality gasoline in conventional refineries.

According to Winslow, long list of different approaches and tasks should be done to achieve sustainable and well performing DCL system. The topics include hybrid processes targeting better carbon management systems, producing hydrogen from non-carbon sources including biomass, operation issues versus indirect technologies, waste product use or disposal, combination of coal and renewable energy concepts, less severe processing meaning lower capital and process costs, product integration with refinery as well as component material and reliability studies. In addition, LCA (life-cycle assessment) studies should be done and for comparison of two and single-stage processes.

5.1 How slurry cracking differs versus direct coal liquefaction process?

Slurry cracking is a potential technology and process to convert heavy crude oils into a liquid fuel (Prajapati et al., 2021). Main difference between slurry cracking and direct coal liquefaction is the feedstock. Slurry cracking is to some point tested for feedstocks like heavy distillate of crude oil and some residues. Literature is not saying the feedstock couldn't be coal but obviously process conditions are economically more attractive if the feedstock is liquid oil versus coal powder.

The selectivity of the products can be adjusted similar way to direct coal liquefaction utilizing different catalyst compositions. Catalysts used in the slurry cracking process are mainly oil-soluble, water-soluble or powder catalysts. End products for the slurry cracking

process varies between light distillates, gases and pitch. Pitch is the heaviest fraction and can be utilized as an alternative fuel in cement kilns with high calorific value. End products are shown in Figure 10.

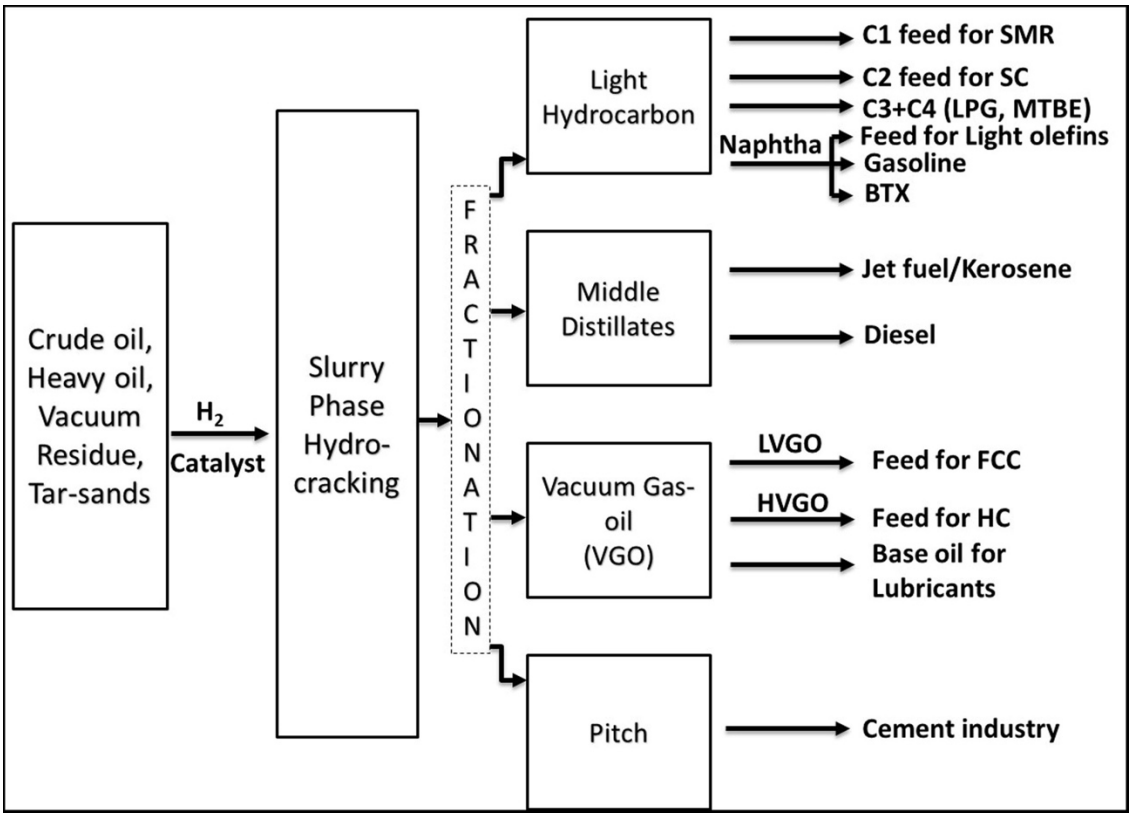


Figure 10. End products of slurry cracking process (Prajapati et al., 2021).

5.2 Has hydrogen been bottleneck for the history development and could cost competitive green hydrogen change the situation? What are the alternative processes scaled up?

Hydrogen consumed in direct coal liquefaction process mainly comes from gasification of coal (Kong et al., 2022). Coal gasification consumes roughly 30% of total coal consumed in the process. Additionally significant amount of water vapor is consumed in the direct coal liquefaction process. Water vapor is mainly produced by boiler units that utilizes coal as

combustion feedstock. Producing water vapor based on coal combustion accounts roughly 10% of total coal combustion in the process. Coal as feedstock also leads to harmful emissions like sulfur, nitrogen and carbon dioxide compounds. Roughly 65% of the steam produced is addressed to coal gasification unit, separation unit and air separation unit. Decreasing the amount of steam consumed directly decrease the amount of coal consumed. Key task is to find out how to decrease the steam consumption demand and improve the hydrogen production methods with more sustainable energy mass balance.

Green hydrogen could help to minimize the need of coal gasification. Photovoltaic and photothermal hydrogen production methods have been proposed to replace fossil-based hydrogen production. Especially United States, European Union and Japan have launched many hydrogen production projects addressing the green hydrogen demand. According to Kong et al., (2022) the solar thermochemical cycle-based direct coal liquefaction system could address the problems of hydrogen production but also steam and electricity production. Solar thermochemical cycle-based direct coal liquefaction could reduce the fuel coal consumption for steam and electricity production by utilizing waste heat recovery from the solar thermochemical cycle. The fossil energy conversion efficiency of the solar thermochemical cycle-based process improved from 59% to 90% compared to conventional direct coal liquefaction process. Solar thermochemical or photovoltaic water electrolysis are expensive processes versus conventional coal-based hydrogen processes. In order to achieve an advantage, the carbon neutrality, carbon taxes and carbon constraints are elements to help investments get green light.

Direct coal liquefaction in the system containing water and syngas or CO (DCLS) is one alternative option for direct coal liquefaction of low-rank coals (Li et al., 2023). DCLS is a solution for low-rank coals that contains large amounts of moisture (25%-70%) and oxygen (15%-30%) as well as have low calorific value and are chemically instable. Low-rank coals are often limited due to these features but its favourable raw material for coal liquefaction due to its high reactivity. DCLS can reduce costs since drying is not needed as well as possibly expensive hydrogen is not needed. DCLS process uses low-cost syngas instead of hydrogen. Planned hydrogen utilization allows coal-bound water to act as a liquefaction

solvent. Hydrogen can be formed having WGSR (water-gas shift reaction) reaction where carbon monoxide and water vapor form carbon dioxide as well as hydrogen. DCLS could be a solution for processes using high-water-content feedstocks and as result cut the operating costs due to absence of drying process and using syngas or WGSR instead of fossil hydrogen produced separately. Key research milestones are presented in Table 5 including study of liquefaction performance, optimization of liquefaction process as well as applications studies.

1921-1990	1990-2010	2010-now
• Key research points • Evaluation of liquefaction effect of liqefied coals, catalysts and solvents • Proposal of liquefaction process • Mainly based on formate mechanism	• Key research points • Evaluation of catalytic effect of multi-component catalysts • Research on liquefaction mechanism of model compounds • Mainly based on reduction mechanism, reaction mechanism, formate mechanism, and others	• Key research points • Application and promotion to biomass liquefaction, heavy oil hydrogenation, coking pretreatment, and other fields • Optimization of liquefaction process • Study on liquefaction mechanism with DFT

Table 5. Development of DCLS process (Li et al., 2023)

5.3 Can carbon monoxide serve as a carbon dioxide sink when implementing green hydrogen and industrial emissions for carbon monoxide?

The transformation of CO^2 into valuable chemical products and fuel sources is widely noticed opportunity and an option to reduce greenhouse gas emissions (Keller and Otomo, 2020). CO^2 can be extracted from industrial plants like power stations and cement

manufacturing. CO^2 can be also extracted from the air but at the moment it is more cost-effective to catch from industrial plants due to higher concentration levels. Atmospheric capture may become mandatory in the future despite the higher costs if the global CO^2 levels won't match with the targets. RWGS (reverse water-gas shift) reaction is an option for converting CO^2 into CO utilizing green hydrogen. Even this is theoretically viable option the process is not fully tested and remains in its early stages. Major development steps are still needed to improve efficiency, product selectivity as well as conversion rates of CO^2 into CO.

Literature didn't reveal if the CO could act as CO^2 sink since CO is again oxidized to CO^2 if burnt after conversion into liquid fuels. Possibly chemicals or materials with longer lifespans could be an option to increase time period when CO^2 is not emitted back to atmosphere.

6 Biochar process

Biochar, a solid biomass-based material produced through the different thermochemical techniques, has been used for thousands of years (Weber and Quicker, 2018). Most of us recognizes charcoal that is derived from wood. Charcoal or biocoal can be used in variety of fields, including energy production, flue gas cleaning, metallurgy, agriculture, livestock management, construction and medicines. Recently, biochar has gained prominence as a sustainable alternative to fossil carbon contributing to efforts to slow down climate change. The carbonization process breaks down the parts of the biomass while maintaining most of its carbon content. This transformation impacts the material characteristics by increasing carbon content and making it more suitable for industry use. carbon content increases due to evaporation of water and the release of volatile components. The choice of feedstock and carbonization conditions is crucial to producing biochar with specific properties. For instance, torrefaction meaning slow pyrolysis process conducted at up to 300ºC and without the addition of external oxygen, improves many challenging properties of raw biomass such as mechanical strength and grindability. These improvements are needed for co-firing and co-gasification processes where original equipment are designed for coal. Higher process temperatures are needed to achieve properties like an exceptionally high carbon content.

Although fast pyrolysis and gasification processes also produce a solid material with higher carbon content, it is typically a by-product, and its quality if often poor for many applications. Chars derived from flash pyrolysis often have low carbon content. In addition, solid product from gasification may contain harmful substances like polycyclic aromatic hydrocarbons.

Main markets for biochar are in North-America and in Europe (Hadiya et al., 2022). North-America market share as biggest market is 36%. Market share for Europe is 24% while less market is in Latin America as 8%. Forecasted growth annually for the biochar market is 15.35% between period of 2021-2031. Dominant technologies are pyrolysis and gasification. Major application areas are agriculture, forestry and electricity generation.

The process begins with the drying of biomass. As the material is further heated, volatile components are released from the solid material. These volatile components may form gases such as CO_2, CO, CH_4 and H_2 or condensable organic compounds like methanol and acetic acid. In the gas phase, subsequent reactions such as cracking and polymerization can occur, altering the overall product composition. The process results in three distinct products like permanent gases, one or more liquid phases and solid residue. The pathways leading to these products often compete and the distribution of products can be influenced by key process conditions. Desired outcome and product yields are optimized to control the process parameters like operating temperature and residence time.

In addition to process conditions, the characteristics of the feedstock significantly influence the conversion process and the properties of the end products. Biomass mainly consists of three organic compounds like cellulose, ligning and hemicellulose. Each of these behaves differently during heat treatment affecting to a product yield and properties. Hemicellulose refers to a groups of branched polysaccharides that are the most reactive of the three components. It decomposes at temperature range approximately 220ºC – 315ºC. Its breakdown process is primarily the torrefaction. The most notable changes in biomass properties occur in this operating temperature window making torrefaction a highly sensitive and challenging process to control. Cellulose, another polysaccharide, differs from

hemicellulose in ints unbranched structure and greater thermal stability. It decomposes at higher temperatures, typically between 280ºC – 400ºC. Despite being the most abundant organic compound on earth and widely studied, the thermal decomposition of cellulose remains only partially understood. Lignin, a complex three-dimensional macromolecule with diverse chemical bonds, decomposes over a much wider temperature range compared to cellulose and hemicellulose. Thermal degradation of lignin begins at 200ºC and can extend to as high as 900ºC. The wide range is due to the presence of multiple functional groups with varying thermal stabilities. The differing thermal stabilities of cellulose, hemicellulose and lignin directly affect the required treatment temperature and the mass yield under specific conditions. In contrast, waste biomasses such as animal manure and sewage sludge lack significant amounts of these components due to their distinct origins and must be characterized using alternative criteria. Typical biochar production process is shown in figure 11.

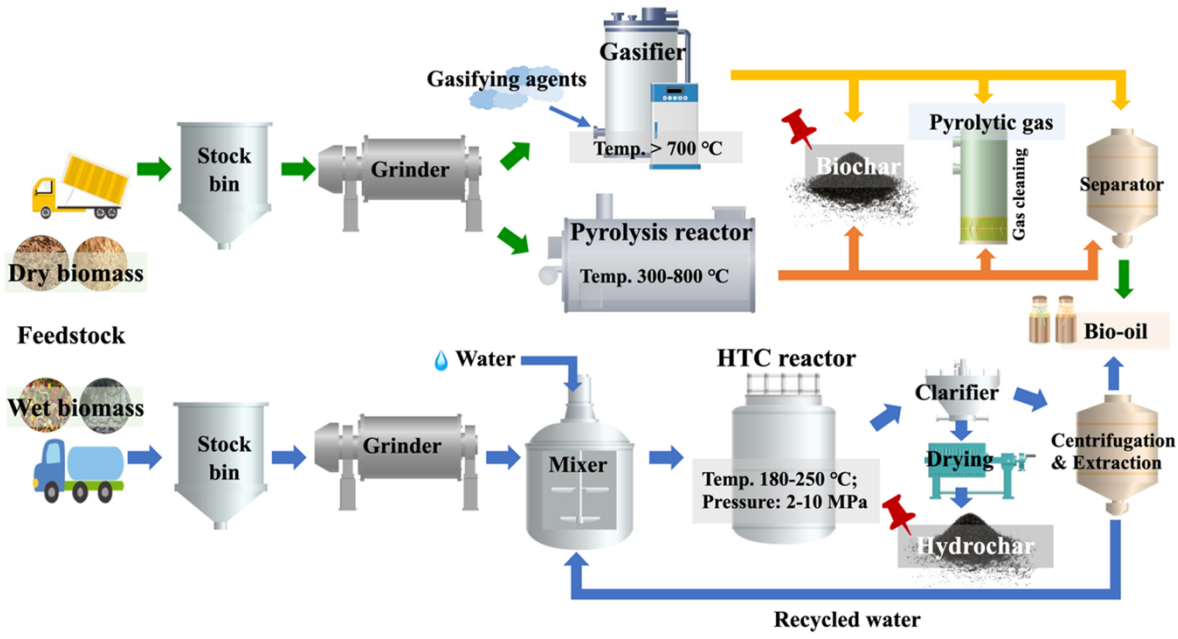


Figure 11. Production process flowchart for biochar production (Zhang et al., 2022)

6.1 Hydrothermal processes and slow pyrolysis for biomass

Hydrothermal processes can be divided into hydrothermal liquefaction, hydrothermal carbonation and hydrothermal gasification. (Ponnusamy et al., 2020) The main difference on these processes is the operating temperature. Pyrolysis in the operating temperature between 200ºC and 300ºC is referred to as torrefaction. According to Akbari et al., the torrefaction is dry torrefaction when feedstock is dry. When feedstock is wet the torrefaction is wet torrefaction or hydrothermal carbonization. Comparison of biocoal production parameters in different production methods are shown in table 5.

Char Type	Slow Pyrolysis biochar	Hydrothermal Carbonization Hydrochar	Hydrothermal Liquefaction Hydrochar
Operating temperature (°C)	300-650	180-260	260-400
Residence time (min)	5-720	5-720	1-120
Heating rate (°C per min)	10-30	5-10	10-100
Pressure (bar)	Atmospheric pressure	20-40	40-200
Feedstock pre-treatment	Dry biomass	Wet biomass	Wet biomass
Process economics			
Net energy ratio	2.3-5.3	1.4 to 5.2	0.81-1.23

Table 5. Process parameters for different biochar production methods (Ponnusamy et al., 2020)

Hydrothermal carbonization (HTC) converts the biomass to biocoal (Akbari et al., 2019). This type of coal is called hydrochar instead of biochar. Temperature window for hydrothermal carbonization is 180ºC – 200ºC. HTC process can receive wet feedstock and do not require the drying process to reach moisture content below 15% like in dry HTC process. This enables HTC process to convert wet feedstocks like municipal solid waste, animal manures and acquaculture residues into hydrochar. Simplified process chart for HTC plant shown in figure 12.

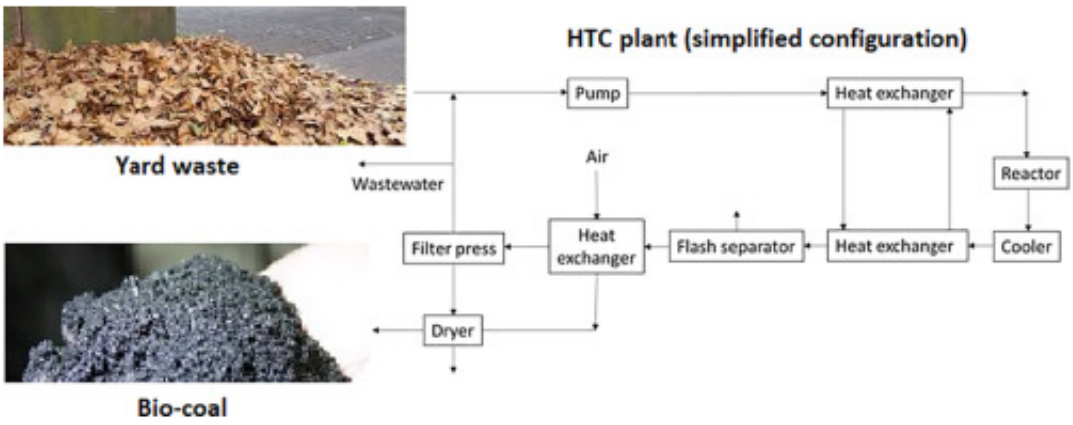


Figure 12. Process chart for HTC plant (Akbari et al., 2019).

Pyrolysis process means the thermal decomposition of organic materials in an oxygen-free environment within a temperature range of 250ºC – 900ºC. This process offers a solution to transform biomass into products such as biochar, syngas and bio-oil. Biochar production via pyrolysis employs various types of reactors, including paddle kilns, bubbling fluidized beds, wagon reactors and agitated sand rotating kilns. As the pyrolysis temperature increases, biochar yield typically decreases while syngas production rises. Fast pyrolysis is a direct thermochemical method designed to convert solid biomass into liquid bio-oil. Fast pyrolysis operating temperature window is 400ºC – 600ºC. In contrast, slow pyrolysis involves much slower heating rates and longer residence times. This method yields a higher proportion of biochar compared to other pyrolysis and carbonization techniques. Process yields comparison for different biochar production methods are shown in table 6.

Technique	Temperature (°C)	Residence time	Yield of biochar (%)	Yield of bio-oil (%)	Syngas production (%)
Pyrolysis	300-700 (slow)	< 2 s (slow)	35 (slow)	30 (slow)	35 (slow)
	500-1000 (fast)	Hour-day (fast)	12 (fast)	75 (fast)	13 (fast)
Hydrothermal carbonization	180-300	1-16 h	50-80	5-20	2-5
Gasification	750-900	10-20 s	10.0	5.0	85.00
Torrefaction	290.0	10-60 min	80.0	0.0	20.00
Flash carbonization	300-600	< 30 min	37.0	--	--

Table 6. Product yields comparison for different biochar production methods (Yaashikaa et al., 2020).

6.2 Has biochar been converted using DCL process?

Biochar has been studied for successfully converting biochar into liquid fuels (Feiner et al., 2013). Feiner et al., did an experiment using biochar derived from pilot scale pyrolysis plant from Vienna. Analysis examined the feasibility of using biochar as a feedstock for direct coal liquefaction process achieving significant conversion rates and oil yields under controlled conditions. While DCL processes are well-documented the application for biochar liquefaction represents a new approach to renewable fuel production.

The process was conducted in a 450ml batch reactor with operating temperature at 425ºC. Initial hydrogen pressure was 50 bar. Tetralin was used and served as the hydrogen donor solvent. The reaction proceeded for 30 minutes under continuous stirring at 500 rpm to ensure proper mixing and heat distribution. The study investigated two pressure configurations. The first configuration maintained an initial pressure of 50 bar without hydrogen metering. Second configuration used same initial pressure but included hydrogen metering at 180 bar allowing for comparison of hydrogen availability effect on conversion rates.

The process achieved results with a maximum biochar conversion of 84.1% and oil yield of 72.4%. The reaction initiated at approximately 370ºC while initial heating period contributed significantly to the overall conversion. During the initial heating stage between 370ºC to 425ºC the process succeeded rapid conversion rates of approximately 6wt% per minute. The subsequent isothermal stage at 425ºC showed steady conversion increases while progressive decreases in the molecular size of products. The optimal duration was 30 minutes for maximum conversion. The study also shows that conversion rate didn’t change much between two scenarios of hydrogen pressure levels. Conversion rates are shown in Figures 13 and 14 below.

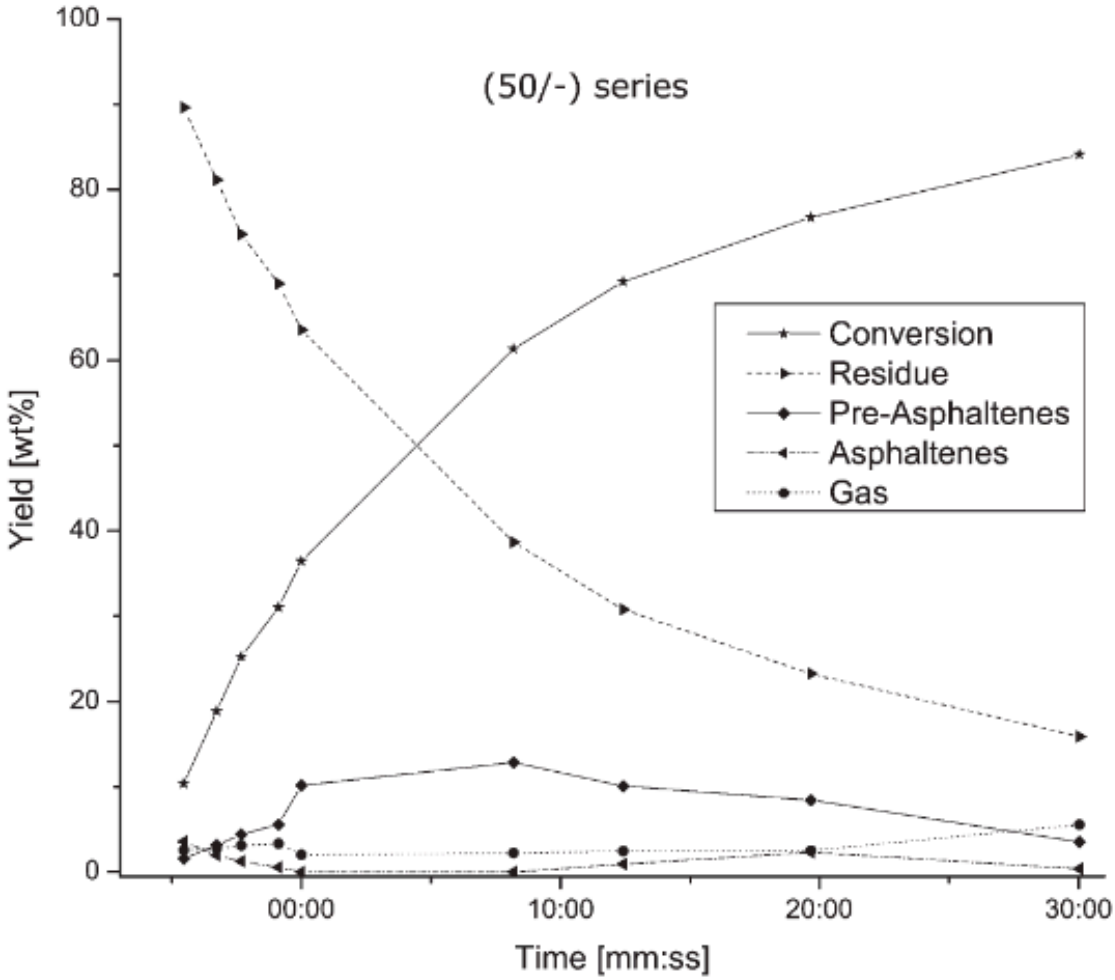


Figure 13. Liquid fuel conversion without additional hydrogen metering (Feiner et al., 2013).

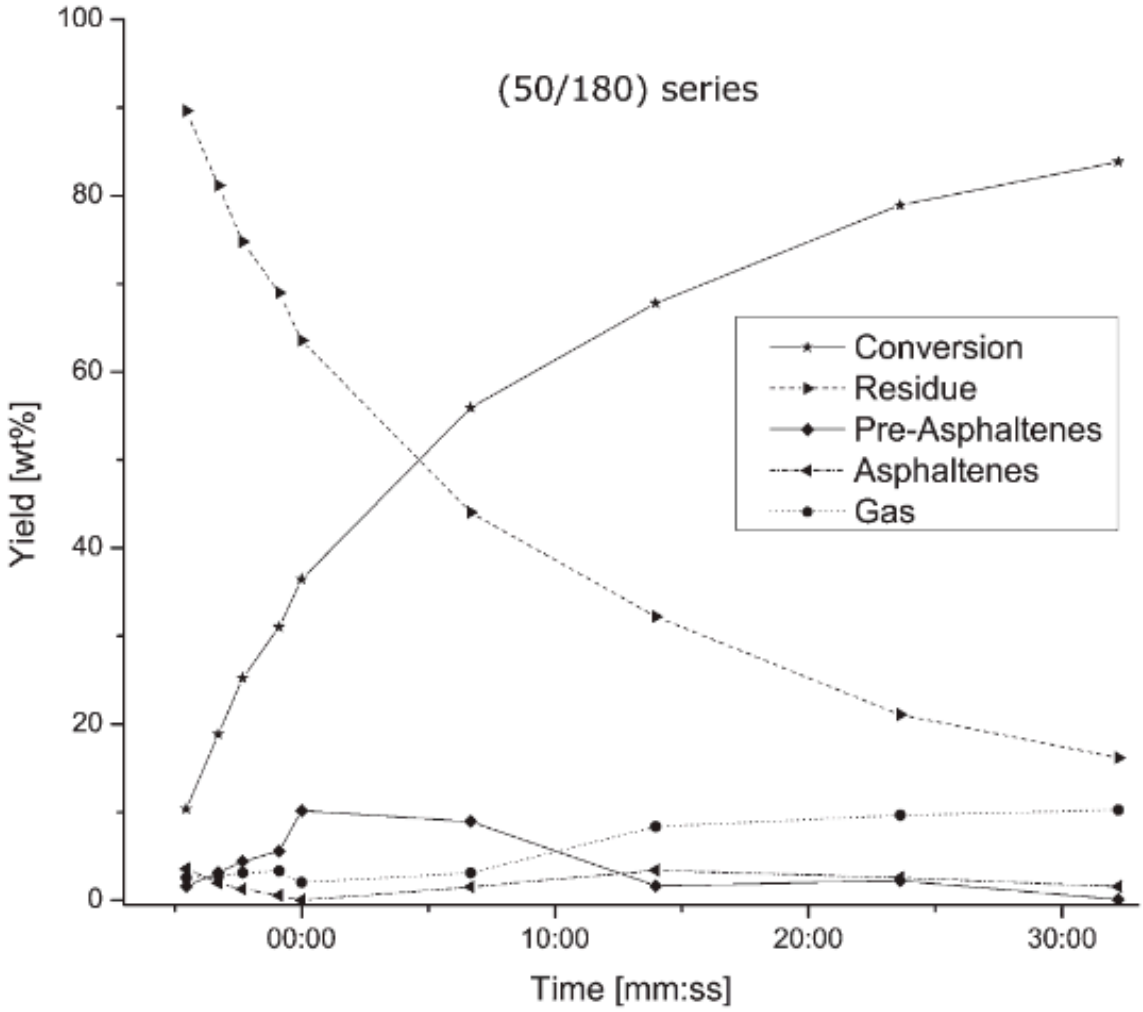


Figure 14. Liquid fuel conversion with hydrogen metering (Feiner et al., 2013).

Tetralin demonstrated good performance as a hydrogen donor solvent. The process also showed remarkable efficiency with tetralin degradation ranging from 2% to 5% of the initial feed. This result combining with limited impact of additional hydrogen pressure shows that more tetralin played more crucial role behind the results compared to system pressure changes. The overall results with conversion rates are shown in a table 7 below.

Time (min)	T (°C)	Gas (wt%)	Oil (wt%)	Asphaltenes (wt%)	Pre-Asphaltenes (wt%)	Conversion (wt%)
Initial Heating Stage						
-4.5	370	2.8	2.3	3.6	1.6	10.3
-3.5	390	3.0	10.7	2.0	3.1	18.8
-2.5	405	4.5	15.1	1.3	4.4	25.3
-1.0	415	5.1	19.8	0.5	5.6	31.0
0.0	425	2.5	23.8	0.0	10.2	36.5
Isothermal Stage (50/-0) series						
0.0	425	2.5	23.8	0.0	10.2	36.5
5.0	425	3.4	45.1	0.0	12.8	61.3
10.0	425	3.8	54.4	0.9	10.1	69.2
20.0	425	3.9	62.1	2.3	8.4	76.7
30.0	425	7.8	72.4	0.4	3.5	84.1
Isothermal Stage (50/180) series						
0.0	425	2.5	23.8	0.0	10.2	36.5
5.0	425	3.7	41.8	1.5	9.0	56.0
10.0	425	8.4	54.4	3.4	1.7	67.9
20.0	425	9.6	64.5	2.5	2.2	78.8
30.0	425	10.3	72.0	1.5	0.1	83.9

Table 7. Overall results for biochar liquefaction (Feiner et al., 2013).

6.3 Rice straw as feedstock for biochar liquefaction

Rice straw can be used as feedstock for first pyrolyzing the biochar and secondly converting it into liquid biofuel (Zhou et al., 2024). Zhou et al., found an eco-friendly way to produce liquid fuel where HHV, O and S content as well as mass and energy yield were remarkably improved compared to direct biomass-to-liquids. At the same time the end product met with the biodiesel standard. Zhou et al., also found out that economic feasibility was also proved since the total production cost was only around 171$/bbl. This price level was expected to be achieved using concentrating the biofuel by a 10-fold multiple and maximizing tetralin recovery. This optimization minimizes the use of pumps and auxiliaries to reduce power consumption.

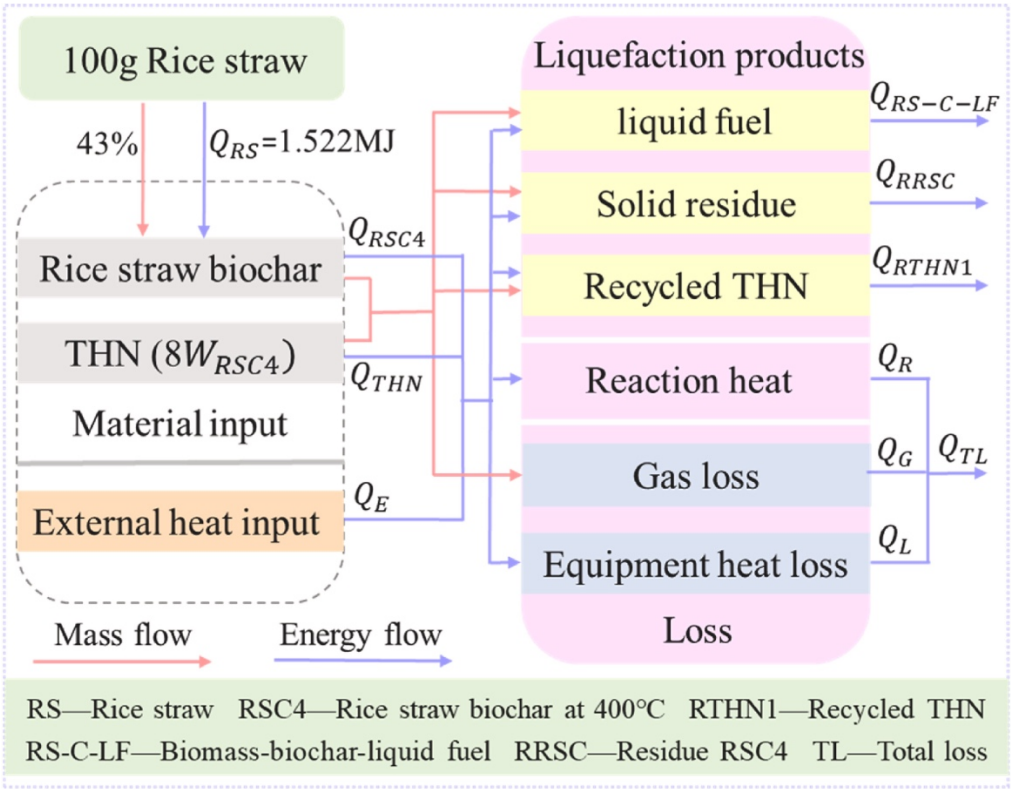
Rice straw samples were collected from Huazhong Agricultural University´s experimental fields in Wuhan, China. The preparation process included air-drying the straw, grinding, filtering with mesh sieve and at the end drying at 105ºC for 24 hours. End product at this phase was straw powder. After this, powder was pyrolyzed in a furnace under nitrogen gas flow at three different temperature levels for 20 minutes. The original rice straw and end products as biochar were analysed for their elemental composition and heating value. The analysis showed that pyrolysis increased carbon content and heating value while reduced oxygen content. Higher pyrolysis temperatures led to decreased hydrophilic properties and

increased aromatic character. In addition, nitrogen and sulfur contents in biochar were notably lower versus lignite. The process used tetralin based hydrogen donor solvent to enhance the process reactions. The biochar produced at 400ºC had properties like lignite for oxygen value, aromatic characteristics and heating value. This resulted that rice straw-based biochar pyrolyzed at 400ºC was superior choice for developing environmentally friendly liquid fuel and replacing lignite as feedstock. Element analysis and key values of rice straw biochars are shown in table 8 below.

Liquefied raw materials	C (wt.%)	H (wt.%)	O (wt.%)	N (wt.%)	S (wt.%)	H/C	O/C	HHV (MJ/kg)
RS	39.34	5.84	41.47	0.75	0.44	1.80	0.79	15.22
RSC_3	44.17	4.87	33.96	0.78	0.48	1.32	0.58	18.40
RSC_4	54.92	3.60	17.19	1.00	0.47	0.72	0.24	19.51
RSC_6	56.83	1.55	8.00	0.88	0.45	0.36	0.11	20.24
Tunçbilek Lignite	67.13	4.68	21.93	2.52	3.64	0.84	0.29	21.80
Inner Mongolia Lignite	66.15	4.30	25.88	1.21	2.46	0.78	0.29	19.18
Huolinhe Lignite	67.67	4.62	25.44	1.05	1.21	0.81	0.28	24.95

Table 8. Rice straw based biochars and their element analysis (Zhou et al., 2024).

Energy and mass balances of the study were key part of the findings. Zhou et al., study shows that rice straw-based biochar and processing it into liquid fuels gave roughly 65 – 68% yield as liquid fuel ($Q_{RS\text{-}C\text{-}LF}$). Solid residues yield was around 5-6% (Q_{RRSC}). Recycled tetralin based hydrogen donor solvent yield was around 22% (Q_{RTHN1}). Losses as reaction heat, gas and equipment heat loss were roughly 2-7% (Q_{TL}). Mass and energy balance is shown in a Figure 15 below.



Liquefaction temperatures			100°C	150°C	200°C	280°C
Energy	Input (%)	Q_{RSC4}	5.33	5.30	5.28	5.23
		Q_{THN}	91.65	91.17	90.69	89.92
		Q_E	3.02	3.53	4.03	4.85
	Output (%)	$Q_{RS-C-LF}$	68.31	69.42	67.57	65.23
		Q_{RRSC}	6.23	6.13	5.99	5.86
		Q_{RTHN1}	22.59	22.40	22.24	22.02
		Q_{TL}	2.87	2.05	4.20	6.89

Figure 15. Mass and energy balance of rice straw based biochar liquefaction (Zhou et al., 2024).

The economic assessment was done for rice straw-based biochar liquefaction where process temperature point as 150ºC was selected. Zhou et al., explain that total production costs of 264,89$/bbl were achieved. This is roughly 1803€ per ton of liquid fuel. Study also explains that cost level of 1803€ per ton of liquid fuel exceed the diesel and biodiesel marker prices mainly due to high tetralin based hydrogen donor solvent cost and increased power consumption from solvent recovery. Zhou et al., propose that implementing more efficient pipeline design and minimizing pump usage could enhance the total process

efficiency and make total production costs more attractive. Total production costs and costs breakdown are shown in a table 9 below.

Items	Detail	Unit price
Raw material expenditure (RMEx)	RSC_4[a]	€130.38/t
	Tetralin[a]	€568.56/t
	H_2[a]	€0.73/L
Capital Expenditure (CapEx)	Equipment	60 % of CapEx
	Maintenance and repair	20 % of CapEx
	Depreciation	Life period 20 years, salvage value 4 %
	Emergency	15 % of CapEx
Operating Expenditure (OpEx)	Operating labor	10 % of ToEx
	Electricity[b]	€0.08/(kWh), 20 % of TPE
	Management	10 % of ToEx
	Marketing	5 % of ToEx
The total production expenditure of RS-C-LF150 (TPE)		€246.35/bbl[d]

Table 9. Economic assessment of rice straw derived liquid fuel with dollar currency exchanged to euros (Zhou et al., 2024).

6.4 Black liquor as raw material for biocoal

Black liquor is key element in papermaking industry and has large amount of organic matter like alkalis and salts that has high value in industrial use (Bao and Liu, 2024). According to Bao and Liu, black liquor could be the feedstock for hydrochar production. Analysis shows that hydrochar from black liquor has similar combustion properties and behaviour versus bituminous coal. Optimum production parameters for hydrochar production utilizing black liquor was 260ºC and 30 minutes residence time. Carbon content was 59.90% with calorific value of 24.14MJ/kg. According to Zhao et al., only few studies have been conducted for investigation of HTC of black liquor into hydrochar. In addition, properties and applications of hydrochar derived from black liquor using HTC process are still lacking.

7 Biochar properties in comparison to fossil coal properties

Biochar has a characteristic that differentiate it from coal and mainly differences are in composition and physical properties (Safarian, 2023). Processed biochar has similar properties compared to low-volatile coal. It also depends on if the biochar is produced using

wood-based feedstock or crops-based feedstock. According to the proximate analysis biochar can contain similar amount of volatile matter versus coal. Range for volatile matter was 9wt% - 41wt% while coal is having 41,5wt%. Proximate analysis also reveals that biochar has lower ash content versus coal that might have different benefits regarding the application. Wood-based biomass also has promising fixed carbon content versus coal. Biomass fixed carbon content around 85wt% is much higher versus 40wt% in coal.

The ultimate analysis shows carbon content varies depending which feedstock is used. Wood-based biochars contain approximately 87wt% carbon while coal contain only 80wt%. What is important that biochar is having zero sulfur content while coal is having 3,6wt% sulfur. This is great environmental benefit for biochar. Hydrogen and nitrogen contents are in general lower in biochar versus coal. This contributes to cleaner combustion characteristics if biochar is burnt. Another difference is oxygen level that is in general higher in biochar versus coal. This might also have an effect to combustion applications. The gross calorific value of wood-based biochar is much higher having around 30 MJ/Kg for biochar while only 20,6MJ/Kg for coal. Another aspect for certain applications is the surface area and porosity of biochar and coal. Biochar has much higher surface area and porosity. The bulk density figures also reveal interesting differences having values around 5g/cm^3 for biochar while around 1,7g/cm^3 for coal. Higher density with higher porosity can be better in various applications like sintering process in steel making where material interaction and heat transfer are essential factors. These figures collectively suggests that biochar derived from wood-based biomass can be highly interesting option for coal for many applications.

Properties	Biochar	Biochar2	Biochar3	Biochar4	Biochar5	Fossil	Fossil6
	Rubber wood char [37]	Wood pellets [37]	Corn straw [36]	Lignin [38]	Walnut shell [39]	Coke [40]	Coal [41]
Proximate Analysis (wt.%)							
Moisture content	0.83	1.94	4.7	0.5	5.7	1.34	6
Volatile matter	9.08	11.06	13	41	35.7	10.3	41.5
Fixed carbon	87.49	83.04	72.9	58	56.6	88	39.6
Ash	2.6	3.96	14.1	0.5	1.8	0.4	12.9
Ultimate Analysis (wt.%)							
Carbon	87.17	87.32	91.53	75.3	56.57	87	80.7
Hydrogen	1.23	1.43	1.54	5.14	5.2	3.5	5.8
Nitrogen	0.4	0.33	0.7	0.97	1.5	1.1	1.2
Oxygen	11.2	10.9	6.16	18	36.6	0.5	8.7
Sulphur	-	-	-	-	-	7.9	3.6
Gross Calorific Value(MJ/kg)	30.38	31.07	27.6	30.18	25.54	27.2	20.6
Surface Area (m²/g)	112.6	247.03	25	-	5.89	4.4	4.13
Bulk density (g/cm³)	4.95	5.3	1.4	1.36	1.32	2.01	1.72

Table 10. Comparison of biochar and coal properties (Safarian, 2023).

Biochar properties can vary, and different properties require separate analysis methods (Hadiya et al., 2022). Different analysis methods are shown in Table 11. for studying detail properties of biochar.

Analytical methods	Description	Purpose/Type of analysis
GC/MS	Analysis of pyrolysis products that are specific to each chemical group	Molecular
(Brunauer-Emmett-Teller) method	BET surface area	Surface
Electrospray ionization Fourier transform ion	Ionizing polar functional groups prior to mass spectrometric analysis	Molecular
Thermogravimetric analysis	Calculation of fixed carbon	Proximate
Raman spectroscopy	Carbon structure stability	Structural
Liquid chromatography	Detecting organic carbon	Ultimate
SEM-EDX	Elemental composition	Ultimate
Polanyi theory	Pore volume	Surface
Zeta potentials analysers	Zeta potentials	Surface
X-ray photoelectron spectroscopy	Types of functional groups and their contents	Ultimate
Fourier transform ion cyclotron resonance mass spectrometry	Determination of mass-to-charge ratio (m/z) of ions based on frequency of cyclotron	Molecular
Contact angle system precision tensiometer	Angle of contact measurements	Surface
FTIR	Functional group	Surface
Brunauer-Emmett-Teller) method	Average pore width	Surface

Table 11. Methods to investigate biochar properties (Hadiya et al., 2022)

8 Fischer-Tropsch challenges and benefits

The Fischer-Tropsch (FT) process is a significant industrial method to convert fossil raw materials, biomass or natural gas into liquid fuels (Moreroa et al., 2024). Anyway, process required step to produce syngas before converting the gas into liquid fuels. This is major difference compared to direct coal liquefaction process. The FT process operates normally at temperatures between 120ºC -300ºC using iron or cobalt-based catalysts to produce liquid fuels. The process generates by-products and effluents as DCL process do. The main components include water, alcohols, ketones, volatile fatty acids, inorganic components and aldehydes. Environmental impacts of FC process are huge. CO_2 emissions, high fresh water consumption, hazardous catalytic waste disposing, air emissions like nitrogen oxides and volatile organic compounds are typical for FT process. Various treatment methods are of course developed for previously mentioned effluents. These include biological treatments as well as non-biological treatments like electrolysis and membrane filtration. Each method has its limitations and advantages. These challenges can be improved by carbon capture, water recycling and reuse programs, circular economy approaches, biogas generation from effluents, and recovery of valuable by-products. FT process shows potential for future for better sustainability through integration of circular economy methods. Valuable resources

recovery like methane, CO_2 and hydrogen would improve the attractiveness of the process. The key focus remains on optimizing industrial productivity, environmental responsibility as well as recovering as much as possible the effluents and side streams.

Main differences of FT process compared to DCL process is feedstock requirements as FT needs syngas preparation and do not accept raw coal feedstock while end product quality is better and more clean after FT process.

9 Commercial study for DCL process utilizing biochar as raw material – mass balance/energy balance with commercial values (OPEX / CAPEX)

9.1 Introduction to business case calculation

This analysis presents techno-economic assessment of a biochar liquefaction plant design. Plant design is based on similar design used for DCL processes in the past. The investment analysis was performed using Microsoft Excel including calculation for capital expenditures (CAPEX), operational expenditures (OPEX), consumables, utilities as well as annualized capital charges. Used parameters for business case analysis are partly from literature and partly from public internet sources. Key components for CAPEX assessment are equipment costs for major process units like reactors and separation units, installation costs and auxiliary equipment, engineering and construction costs as well as contingency and project management costs. OPEX assessment covers raw material costs, catalyst and hydrogen consumption, utility costs, labor and maintenance costs as well as general and administrative expenses. Project scope is plant with production capacity of 100,000 tonnes of biochar per year with liquid fuel yield of 33,000 tonnes per year. Operating time for plant is 8,000 hours annually.

Main objectives for the analysis are to evaluate techno-economic feasibility of biochar as feedstock for conventional direct coal liquefaction process. In addition, study objectives are to find possible weak points to improve and validate stable parts of the process already on feasible level for possible industry scale operation. Main interest for the study from environmental point of view is to offer alternative and sustainable feedstock source for the process instead of fossil feedstock.

Analysis serves giving answers for markets in need for sustainable liquid fuels, increasing pressure of decarbonization in the transportation sector, development of circular economy and utilizing bioenergy value chains.

9.2 Process description

The analysed direct biochar liquefaction process converts biochar into liquid fuels through a high-pressure catalytic hydrogenation process. The plant is designed with the annual biochar processing capacity of 100,000 tonnes. Plant processing capacity is estimated based on realistic possibilities to acquire selected amount of biochar from local market. Process yield is calculated to be as 33,000 tonnes of liquid fuels annually.

Key unit operations for the process are coal preparation, liquefaction in high-pressure catalytic reactors, distillation, gas treatment and hydrogen production. Hydrogen production was not studied separately but assumed that green hydrogen is used as produced externally.

Catalytic reactors operate at 400ºC -450ºC temperatures and in pressures of 150-200 bar. Iron based catalyst is used. The product separation process includes a series of high-pressure separators by distillations units. Distillation unit can be designed to allow outputs of gas, naphtha, jet fuel and diesel for refinery. Process flowchart is shown in Figure 29 with mass balance figures.

9.3 Economic parameters and assumptions

Calculation includes scaling from bigger plants to smaller size. Scaling method for used was "Rule of Six-tenths" (Process Equipment Cost Estimating By Ratio And Proportion, accessed 8.2.2025). Mentioned rule is used for chemical plants to scale costs from size to another. It gives approximate cost within plus or minus 20% range. Liquid fuel yield of roughly 33% of biochar consumed is from Lepinski´s report and is based on U.S location and sub-bituminous coal as feedstock (Lepinski, n.d.). CAPEX costs are based on Lepinski's report. Catalyst ratio per consumed biochar is based on Shui's report (Shui et al., 2010). Hydrogen ratio per consumed biochar is based on Robinson's report (Robinson, 2009). Electricity consumption is based on H-coal process and Burke's report (Burke et al., 2001). Electricity consumption is evaluated studying the relation of coal used, electricity consumed, and liquid fuel produced in H-coal process.

Costs and prices for liquid fuel, biochar feedstock, catalyst, green hydrogen and catalyst disposal are assumptions. Assumptions are based on public sources regarding market prices and include some amount of variation depending on the source. Liquid fuel price of 2,000 - 2,500€ per ton is estimated based on Scandinavian market price for diesel but exact selling price for the plant is hard to estimate since plant is not directly selling to consumers. Biochar price of 200€ per ton is based on public sources of different market prices. Catalyst price can vary a lot and price of 1,500€ per ton include some amount of uncertainty. Green hydrogen cost of 3,000€ per ton is based on public sources of different market prices. Catalyst disposal cost of 7€ per ton is assumed value based on various public sources. Utilities like electricity, steam and water are assumptions based on public data for chemical plants. Cost for electricity in MWh is estimated to be 50€ per MWh. Costs for MP and LP steam per ton are estimated to be 40€ for MP and 35€ for LP steam. Cooling water cost per ton is estimated to be 1€ per ton.

9.4 Mass balance calculation

The mass balance calculation for the biochar-based DCL plant is based on tracking material flows from feedstock input until the final products and utilities stream. The analysis accounts

for main feedstock with consumables, product yields as well as utilities. On more detailed economic calculations also OPEX and annual capital charges are considered. Economical calculation also considers catalyst disposal as waste stream, but impact is minor.

Plant is consuming annually 100,000 tons of biochar. Later in this study the evaluation for up scaled capacity is done and impact for total profitability calculated. Hydrogen input is 4,800 tons per year. Catalyst input is 500 tons per year. Process consumes electricity of 0.64MWh/ton liquid fuel. Medium pressure steam is consumed 1.2t/ton liquid fuel. Low pressure steam is consumed 1.2t/ton liquid fuel. Cooling water is consumed 75t/ton liquid fuel.

Total useful products are roughly 56.000 tons as output including also gaseous products. Liquid fuel output is 33,000 tons per year. Liquid fuels are expected to be diesel, naphtha and jet fuels. Process also produces waste streams like catalyst disposal, solvent refined coal and other wastes like carbon oxides, hydrogen sulfides, waste water and ammonia. Total waste stream is expected to be roughly 40,000 – 45,000 tons.

The mass balance indicates significant mass loss in the process, which is result of formation of gaseous products, water formation from oxygen removal as well as process losses. Mass balance analysis is based on steady-state operation, perfect mixing in reactors, constant feedstock composition as well as no disturbances in storage and handling processes.

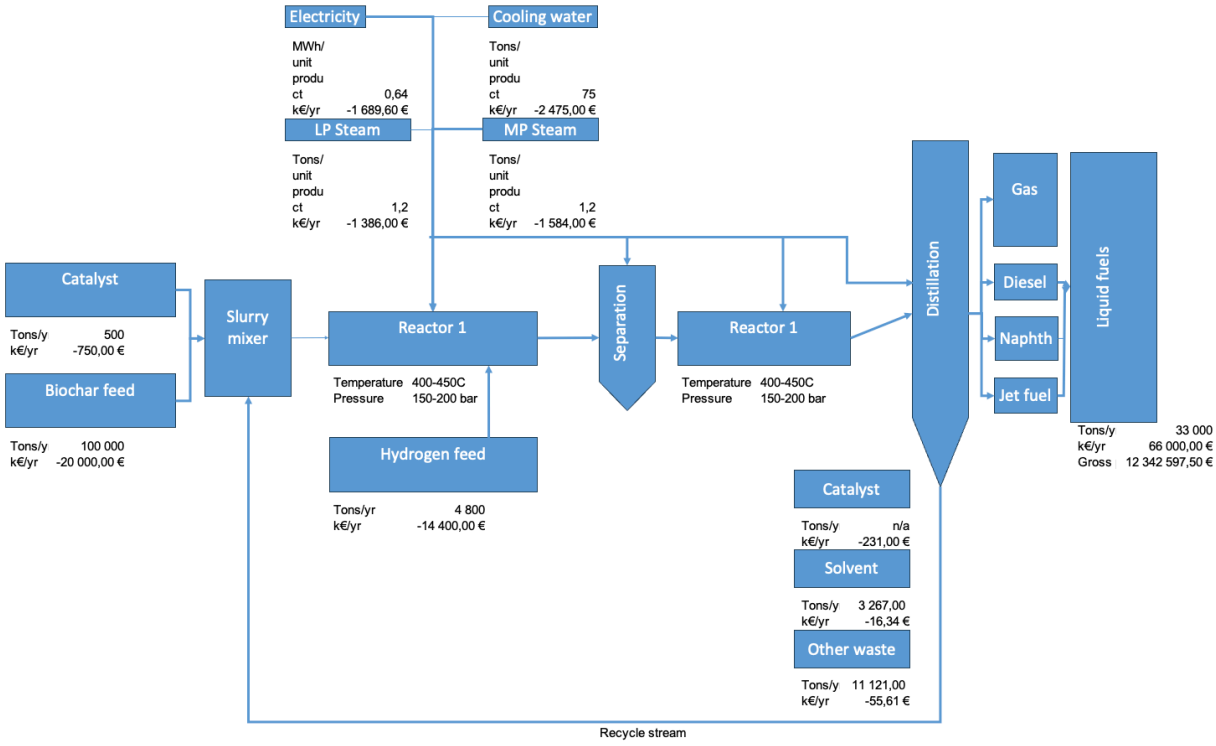


Figure 16. Techno-economic balance calculation with liquid fuel price of 2,000€/ton

9.5 Capacity and scale analysis

The proposed biochar-based liquefaction plant has been analysed for various capacity scenarios to understand the implications of scale for technical and economic performance. The analysis is essential for understanding the optimal plant size and understanding the suitability of the technology for different market conditions. The analysis method used for plant size configuration is Rule of Six-tenths. Selected method is common factor and widely used in chemical process industries for estimating how capital costs change with capacity scaling. This is an empirical rule and it shows ratio of costs between two differently sized plants. Approximation is the ratio of two different plants capacities raised to power of 0.6. The relationship is expressed as in formula below.

$$Cost_2 = Cost_1 \times (Capacity_2/Capacity_1)^{0.6}$$

$Cost_2$ is estimated cost of the new plant. $Cost_1$ is known cost of the reference plant. $Capacity_2$ is the capacity of the new plant. $Capacity_1$ is the know cost of the reference plant. As an example, the scaled plant with 50% capacity increase would be 265.4 million euros using formula $Cost_2 = 209m€ \times (150{,}000/100{,}000)^{0.6}$. This demonstrates that while the capacity increases by 50% the capital costs increases only by 27%. This explains the economy of scale principle where larger plants typically have lower unit capital costs. The 0.6 exponents reflect the average relationship between scaled plants. Actually, the used exponent can vary depending on the type of process and equipment used. The rule is useful for study phase exercises and in early-stage cost estimations. Feasibility and magnitude of impacts can be review without detailed engineering designs.

9.5.1 Base case configuration

The current design basis represents a medium-scale facility with an annual biochar processing input of 100,000 tonnes. The plant operates for 8,000 hours meaning daily processing rate of 300 tonnes of biochar. With the established yield of 33% the annual liquid fuel production is 33.000 tonnes. The base case configuration has capital charge of 548 € per ton, fixed costs of 346€ per ton and variable costs of 1269€ per ton. This base configuration serves as a reference point for scaling estimations.

9.5.2 Capital investment scaling

The rule of six-tenths indicates that capital costs do not increase linearly with capacity. Analysis for 50% capacity in relation to reference point shows ISBL (inside battery limits) investment reduces to 138 million euros. Doubling the capacity to 200,000 processing capacity will increase the ISBL investment to 316 million euros. Scaling analysis in steps of 50,000 tonnes, starting from capacity of 50,000 tonnes until capacity of 250,000 tonnes, shows how total investment cost in millions of euros as well as how total investment cost in euros per ton changes when capacity is adjusted. Investment cost per tonne will reach below 1500€ point just when the plant capacity is increased to process input capacity of 250,00

tonnes. It is significantly lower than 2757 euros per ton when process input capacity is 50,000 tonnes. Fixed operating costs are not analysed in this graph since it´s expected to be more stable while capacity is adjusted. Of course, operating costs will be increased while capacity is increased but Rule of Six-tenths only covers the investment costs. Scaling analysis results are shown in table 12 below.

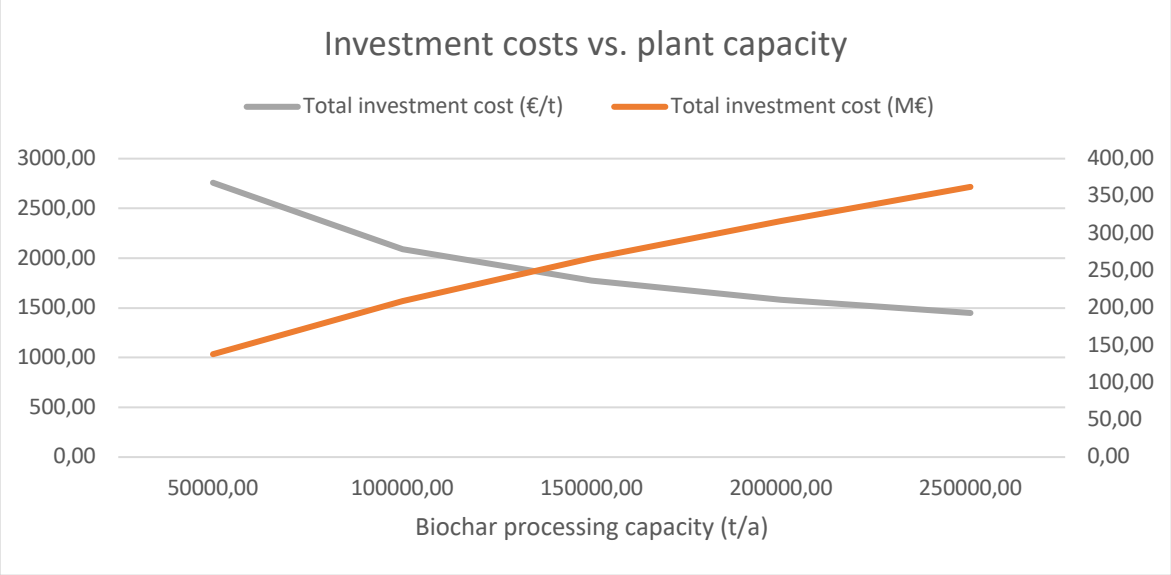


Table 12. Scaling analysis using “Rule of Six-tenths”

Variable costs including raw materials, utilities and consumables typically scale almost linearly with capacity. However, larger operations allow more purchasing power for instance for raw materials potentially allowing improvements on unit costs level.

Capacities below 50,000 tons seems to end up with a result of high fixed costs as unit basis. 50,000 tons capacity can be understood as somehow minimum feasible level for investment from that perspective. Maximum feasible point can be tricky to estimate due to technical aspects like possible equipment size constraints, process efficiency considerations, regional biochar availability, product market absorption capacity and logistical aspects. Without further studies for those the 250,000 tons capacity offers attractive investment cost per ton figure. Ideal capacity range might be between 100,000 tons and 150,000 tons offering already almost as attractive investment cost per ton figure

as bigger plants like 250,000 tons capacity plant. Investment cost per ton lies between 1780€ to 2090€ for 100,000 tons and 150,000 tons plants.

9.6 Sensitive analysis for hydrogen cost variation

The sensitivity analysis shows that the biochar liquefaction plant´s profitability is significantly influenced by hydrogen cost which currently plays major role in operational costs at 3000€ per ton. At current hydrogen prices the plant maintains a healthy gross profit only if price for liquid fuel stays above 2,200€/t. Prices above 2,200€/t obviously are too optimistic for Scandinavian market where price is heavily correlated with fossil based liquid fuels. Estimated price for the plant for liquid fuel is challenging to set up but the analysis shows that using 2,000€/t brings negative profitability for the plant. Profitability declines linearly with increasing hydrogen costs. Sensitive analysis results for hydrogen cost variation are shown in tables 13 and 14 below.

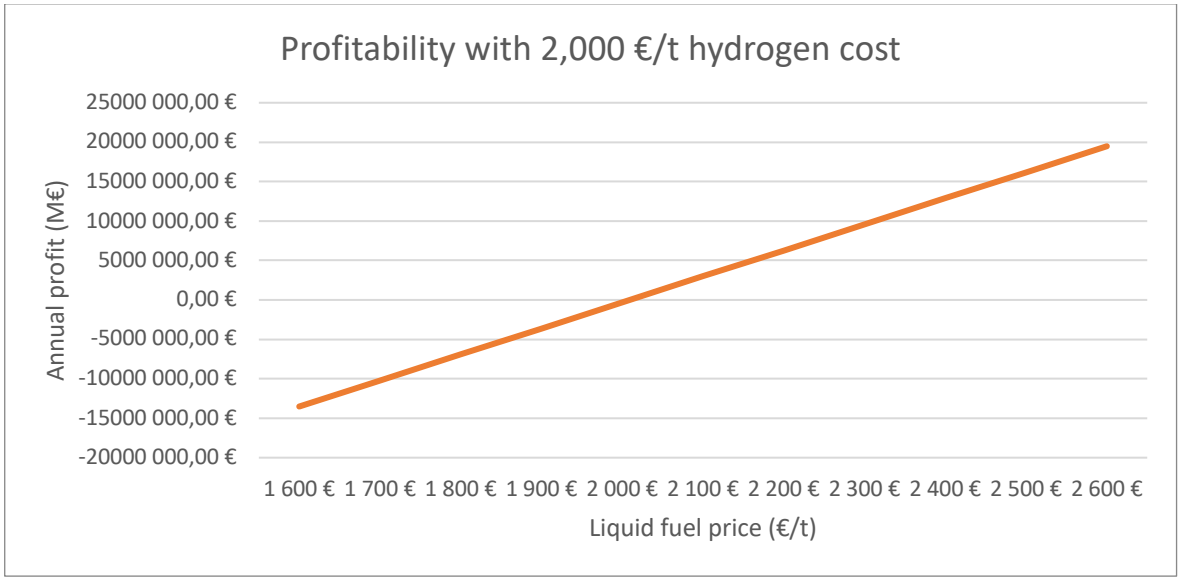


Table 13. Sensitive analysis for liquid fuel price if hydrogen cost is 2,000€/t

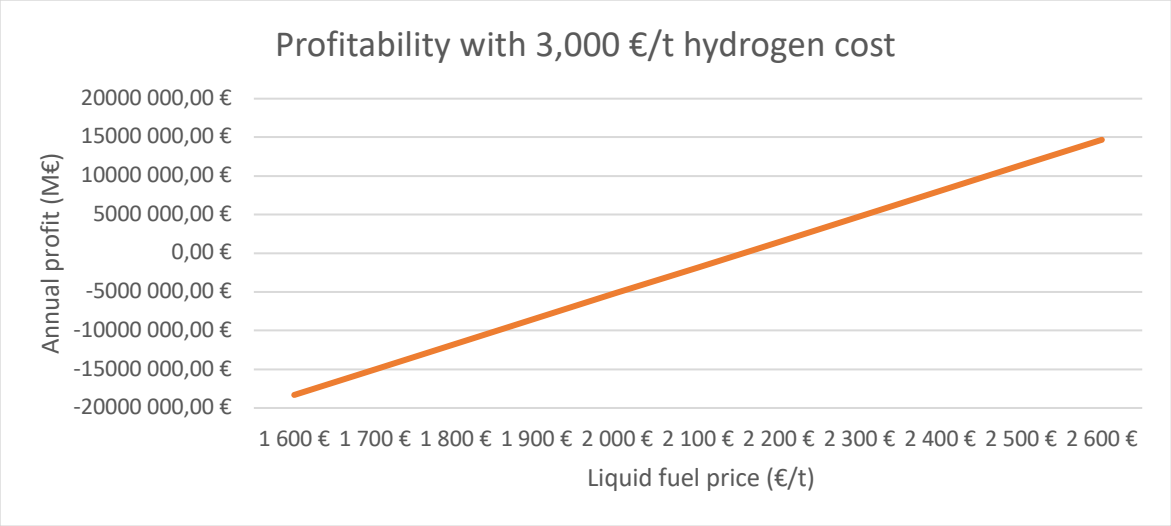


Table 14. Sensitive analysis for liquid fuel price if hydrogen cost is 3,000€/t

9.7 Sensitive analysis for electricity cost variation

The sensitivity analysis reveals that electricity variation is not a huge risk for project scenarios. Analysis shows that when comparing 50€/MWh and 80€/MWh the result for profitability won´t change with same correlation. Estimation is based that hydrogen cost is fixed 3,000€/ton and the sensitivity analysis for electricity do not expect that plant produces hydrogen on its own facility. This shows that already high peaks meaning doubling or tripling the current electricity costs of 50€ per megawatt hour of course reduces the annual gross profit but only with minor impact. That kind of electricity cost increase is already extra ordinary. The impact of electricity cost increases is notably less severe comparing to hydrogen cost variation since electricity represents smaller portion of overall operating costs. Sensitive analysis results for electricity cost variation are shown in tables 15 and 16 below.

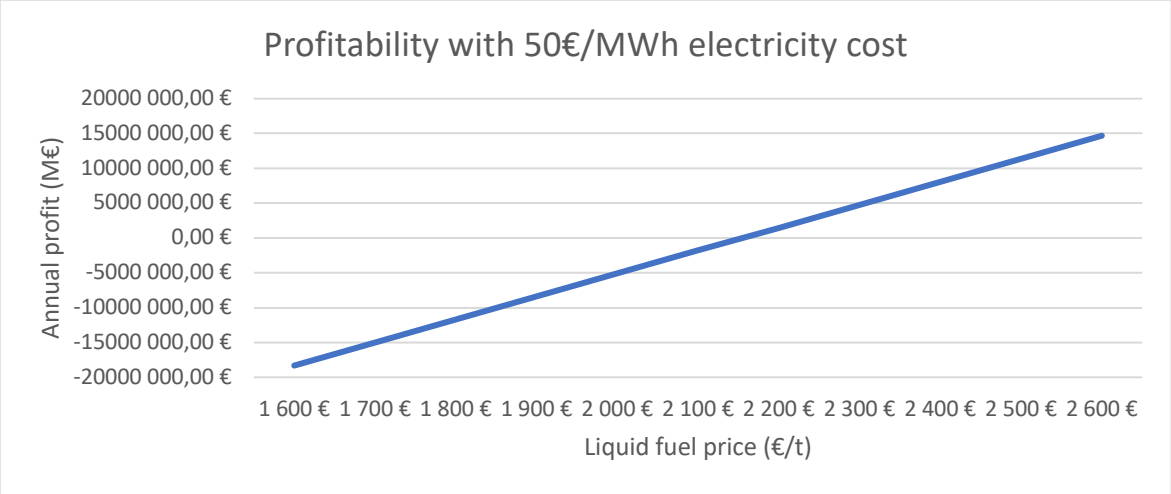


Table 15. Sensitive analysis for electricity cost of 50€/MWh

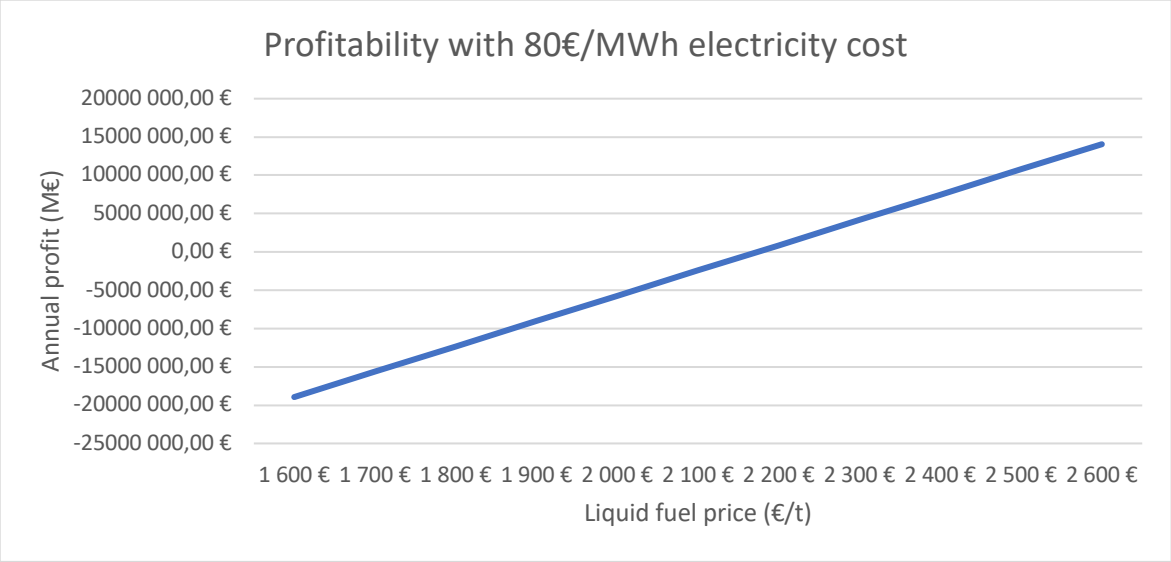


Table 16. Sensitive analysis for electricity cost of 80€/MWh

9.8 Sensitive analysis for total fixed capital cost variation

The sensitivity analysis shows a clear relationship between capital investment and project profitability. At the base case of roughly 255 million euros of capital investment the plant provides positive profit with liquid fuel price of roughly 2,150 – 2,200€/t. Another case with 200 million euros of capital investment brings positive profit with liquid fuel price of roughly

2,000 – 2,100€/t. This is a result of total product revenues and total cost of production. Total cost of production also includes total annual capital charge that is a result of 5% interest rate level for the investment. Plant estimated lifetime is 25 years. The sensitivity is driven by three capital-dependent cost factors. The factors are annual capital charges, maintenance costs as well as tax and insurance expenses. The analysis reveals that plant will stay with positive profitability with estimated CAPEX figures but price for liquid fuel must be such high. Sensitive analysis results for total fixed capital cost variation are shown in tables 17 and 18 below.

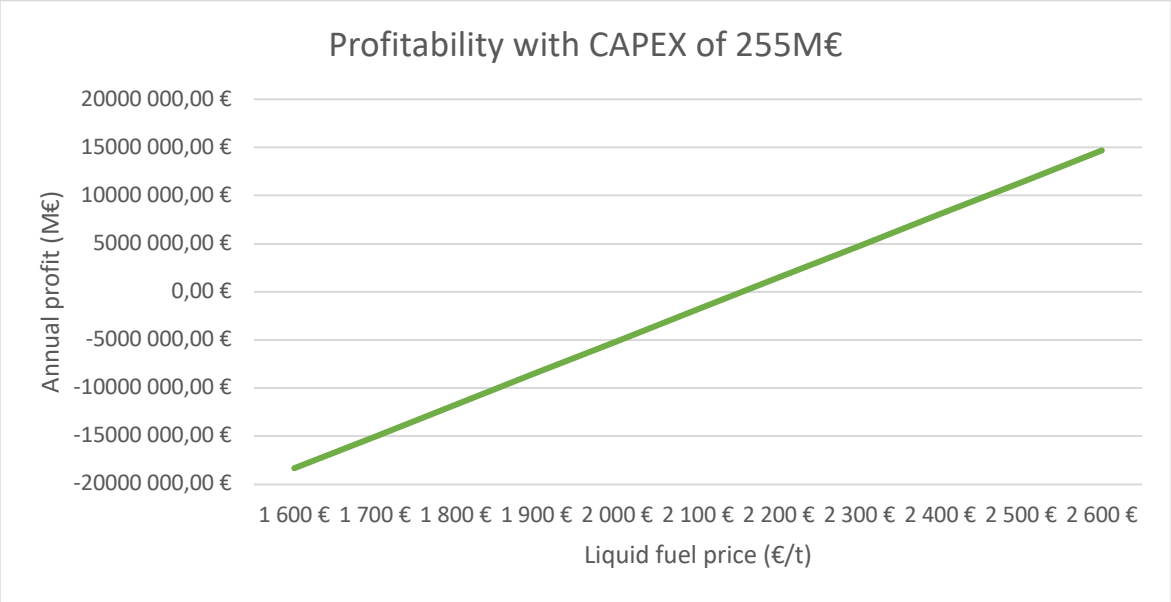


Table 17. Sensitive analysis for total fixed capital cost of 255M€

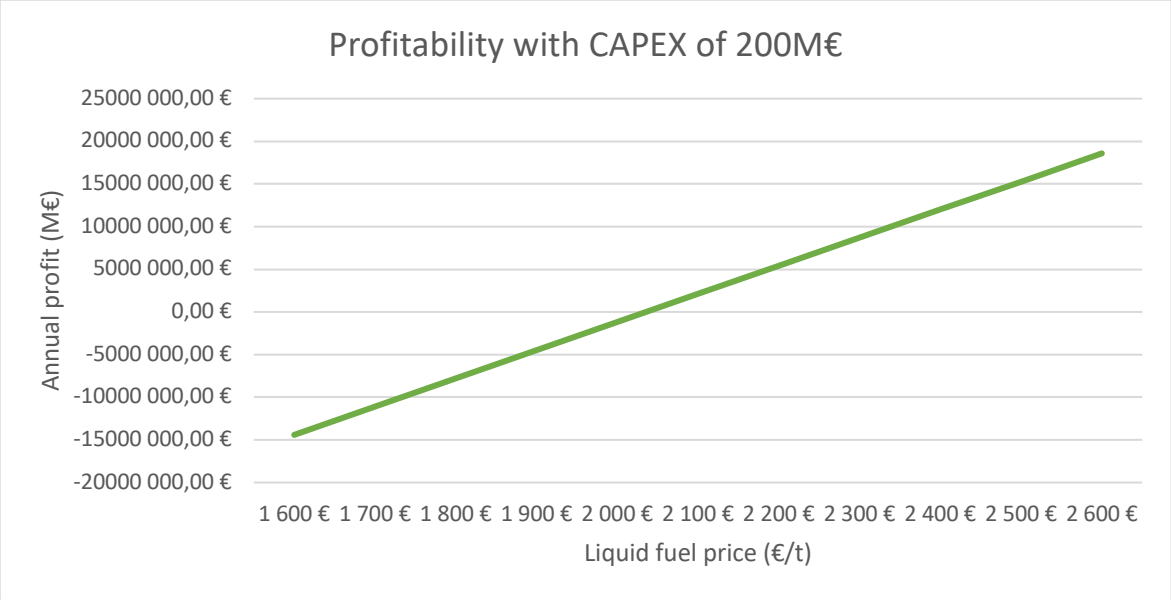


Table 18. Sensitive analysis for total fixed capital cost of 200M€

9.9 Feedstock analysis comparing lower heating values of biochar and lignite

This visualisation reveals that the project would need notably more lignite as feedstock comparing to biochar if plant would run using lignite raw material. Lignite has lower heating value (LHV) of roughly 14MJ/Kg ("Higher Calorific Values of Common Fuels," n.d.). Biochar has lower heating value of roughly 23.55MJ/Kg (Rathod et al., 2023). Since lignite has lower LHV the plant would require almost 70% more feedstock mass in order to achieve same LHV amount in end product. The findings are that using lignite would require higher material handling capacity, larger storage requirements, more transportation as well as potentially larger reactor volumes. Bar chart analysis for feedstock LHV comparison is shown in table 19 below.

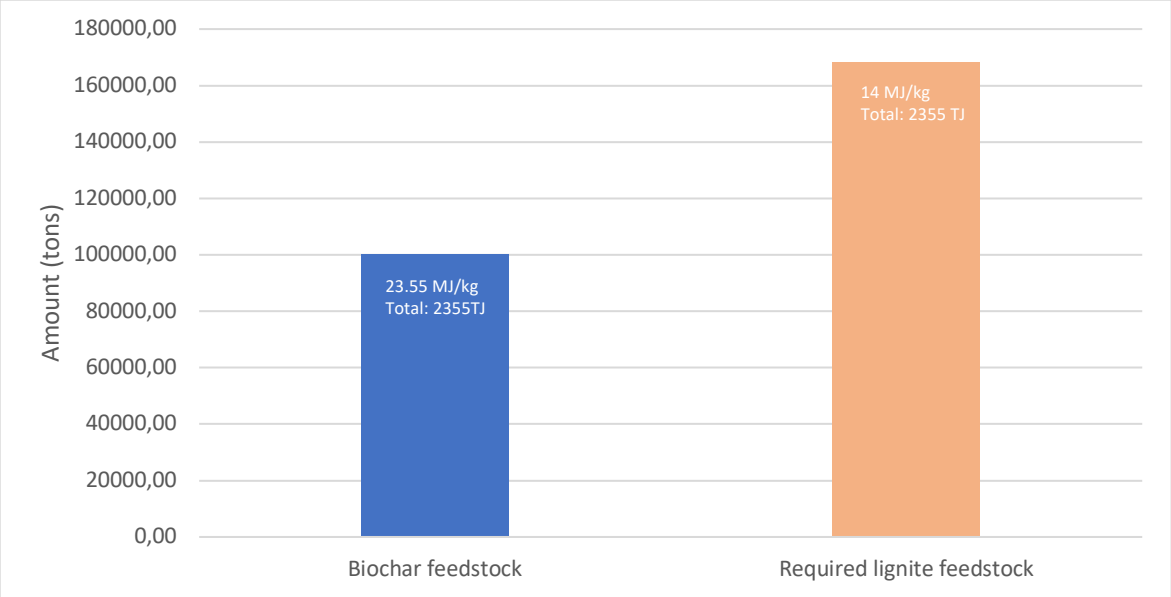


Table 19. Feedstock amount comparison for equal energy content

9.10 Operating cost analysis

The economic feasibility of biochar liquefaction plant depends on their operating cost's structure. Labor and maintenance costs are a significant part of fixed operating expenses in production facilities. The plant is estimated to use four shifts with five operators per shift position. Total cost per operator is estimated to be 75,000€ annually. These operators typically consist of operational staff handling the production processes and are supported by technical and administrative personnel for maintaining and developing the processes. Also, quality control is estimated to be part of administrative work. Total Fixed Cost of Production (FCOP) are estimated to be roughly 333€ per unit main product. Labor costs including shift operators, supervision and direct overheads are roughly 50€ per unit main product. Therefore, labor costs are roughly 15% of total FCOP.

Maintenance, taxes and insurances represent another significant cost components. Annual maintenance expenditure is estimated to be 2% of ISBL investment. It is roughly 126€ per unit main product meaning almost 38% of all FCOP costs. Tax & insurances are estimated to be 2% of total fixed capital costs. Total fixed capital costs are 254,980,000€ including

ISBL and OSBL capital costs, engineering costs as well as contingency of 2%. Therefore tax & insurances are high part of FCOP costs ending up to 46%. Plant overhead costs are estimated to be 2% of labour and maintenance costs and ends up being in minor part in total FCOP, only 1%. Interest on debt financing is also in minor role, only 0,5% and is based on 6% of working capital. Working capital is estimated to be 1,000,000€

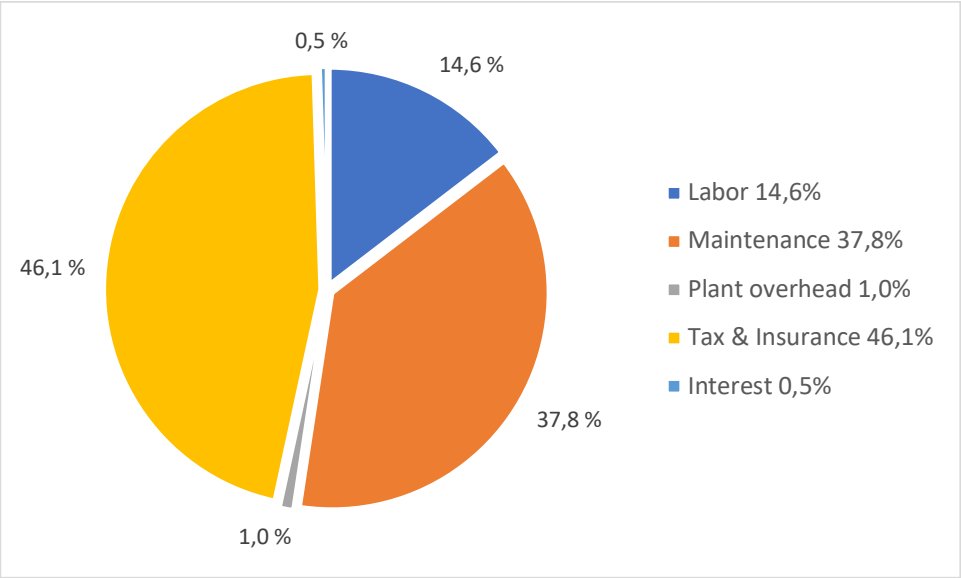


Figure 17. Fixed Cost of Production (FCOP) share

Utilities also represent notable amount of cash cost of production. Utilities including electricity, MP and LP steam as well as cooling water. Utilities costs together are 197€ per unit main product. Therefore, utilities represent roughly 15% of total variable cost of production. Cash cost of production consists of variable and fixed cost of production. Variable cost of production is estimated to be 1,271€ per unit main product while fixed cost of production is estimated to be 335€ per unit main product. These costs sum up to 1,606€ per unit main product. Total variable cost of production (VCOP) is calculated adding raw materials, byproducts and wastes, consumables and utilities together.

9.11 Revenue projections and estimation of business impacts

Primary revenue stream for the plant is sale of liquid fuel with the following key parameters. Annual production capacity of 33,000 tonnes of liquid fuel with base selling price of 2.000€ per ton. Annual revenue potential is therefore 66,000,000€ and is based on plant availability of 91.3%. Plant availability can be raised up to 95% or even a bit more increasing potential for more revenue. Liquid fuel pricing is influenced by various factors. Global oil market dynamics, regional fuel market conditions, product quality specifications, compliance requirements as well as competition of alternative fuels.

Revenue stability could be supported focusing on long-term supply contracts but also forecasting the transportation sector demand and possible changes. Industrial fuel applications development would be part of stabilizing revenue stream for the plant. Renewable fuel regulation and possible credits should be taken into consideration for planning the revenue streams. Possible risks for revenue streams to be considered are global crude oil price volatility, alternative fuel competition including electrification, regulatory changes globally and regionally, carbon pricing development, transportation sector development as well as general industrial sector demand and volatility.

Biochar liquefaction plant should consider strategic revenue opportunities since it´s operation profile would be almost unique, and it differs in sector of fuel industry. Plant should consider how to charge some premium for biochar derived fuels. In addition, environmental compliance credits as well as carbon credits could offer opportunities for strategic sales. From technology perspective the licensing of technology could offer interesting approach to expand the business with strategic industrial partners. Plant could also consider how to succeed with superior product quality consistency or supply reliability over the competition. In addition, some competition advantage could be found from strategic location of the plant.

9.12 Financial analysis in Excel format

Item	k€	%	€/unit main product
ISBL Capital Cost	209000,0		
OSBL Capital Cost	20900,0	10 %	
Engineering Costs	20900,0	10 %	
Contingency	4180,0	2 %	
Total Fixed Capital Cost	254980,0		
Working Capital	1000,0		
	k€/yr		€/unit main product
Key Products	66 000,00		2000,00
	k€/yr		€/unit main product
Fixed Capital Investment	-18 091,46		-548,23
	k€/yr		
Total Annual Capital Charge	-18 091,46		-548,23
	k€/yr		€/unit main product
Variable Cost of Production	-41954		-1271,33
Fixed Cost of Production	-11070		-335,45
Cash Cost of Production	-53024		-1606,78
Total Cost of Production	-71115		-2155,01
	k€/yr		
Gross Profit	12 976,20 €		393,22 €

Table 19. Financial analysis in Excel format

10 Conclusions

Based on the investment calculation and process analysis done in this thesis, several key conclusions can be presented regarding the feasibility of biochar-based direct coal liquefaction technology. The investment calculation demonstrates that the minimum selling price required for the liquid fuel is approximately 1,606€ per ton since cash cost of production is on this level. However, including capital charges, the full cost of production rises to 2,155€ per ton. With the current assumption of 2,000€ - 2,500€ per ton for the liquid fuel product and 200€/ton cost for biochar the process generates positive gross profit. Main aspect is the price assumption for the liquid fuel and the relation of end users price sensitiveness for the liquid fuel. Price level of 2,500€ per ton might be high for liquid fuel as output from DCL plant but further studies should be done to understand the market position for the end product and what could be the acceptable price level. Fischer-Tropsch process, as competitor for DCL process, might end up to higher capital investment costs as well as lower carbon conversion efficiencies giving room for DCL process to gain market share.

The economic viability of the biochar-based DCL process is highly depending in several critical parameters. Hydrogen costs, biochar costs, production scale as well as product yield are the main areas to focus and to improve to make a process attractive. At the current assumption of 3,000€ per ton hydrogen represents one of the most significant variable cost components. The process remains profitable even hydrogen price will increase but assumption requires that other cost components won't increase, or the liquid fuel price won' t decrease. Hydrogen production methods are key areas to focus to achieve sustainable feedstock for hydrogen as well as keep cost control for the cost element. As the green hydrogen economy develops the biochar-based DCL process might benefit. The process is hydrogen-sensitive requiring approximately 0.048 tons of hydrogen per ton of biochar processed. Current assumption is relatively expensive hydrogen and as green hydrogen production scales in countries like Finland with low electricity costs the economics of the process might improve. Potential integration with electrolysis facilities could provide synergies in both operations and capital investment.

The expanding biochar market represents opportunities and challenges for the biochar-based DCL process. Competition for this feedstock obviously will increase as soil amendment and other use cases gain more attraction. More users for the biochar might increase the prices but on the other hand more production obviously will follow to stabilize the prices. The biochar-based DCL process might be able to use lower-grade biochars unsuitable for soil applications. This would lead into tiered market where premium biochar serves agriculture and industrial-grade material suppliers liquefaction processes. These different levels of biochar quality could serve all end users and allocate correct quality per applications requirements.

One aspect is to test more biochar-based process even rice straw-based biochar process was already successfully tested. In addition, slurry cracking could be potential process to study more with biochar as feedstock. Kraft lignin conversion to liquid fuels has also been studied and interesting results like around 50% yield with similar catalysts to DCL process would be worth to study more.

Production scale of 100,000 tons of biochar as input appears to be near the minimum economically viable size. Scaling up would improve the economics. Product yield is assumed to be 33% while some studies have shown up to 60-70% yield for biochar-based liquefaction process. Increase on product yield would be major element to support feasibility of the plant.

Biochar-based DCL plant could potentially serve also other applications than liquid transportation fuel. Chemical feedstocks would be obviously interesting application area and could be potentially more profitable market versus transportation fuel market. Specialty lubricants could be another area to study for further processing of biochar-based DCL process output. Heavier fractions of DCL process output might be interesting for carbon fiber producers to study alternative feedstock for their process. Areas like aviation where electrification is challenging might be also interested in to use biochar-based liquid feedstock to develop aviation fuels. Biochar-based DCL plant might not end up to as good quality end products as Fischer-Tropsch process but process simplicity, carbon efficiency, product selectivity as well as CAPEX requirements might serve interesting further studies to prove DCL process to be more attractive for certain investment compared to Fischer-Tropsch.

However, Fischer-Tropsch benefits from decades of commercial implementation and optimization while modern DCL processes are relatively less developed. Local conditions, feedstock availability, hydrogen costs and end product specifications should be closely studied to form a recommendation for certain plant to select the process technology.

As final assessment the investment and process analysis support the technical and economic feasibility of the biochar-based DCL process under current market conditions. One main technical and economical aspect to support this conclusion was study by Zhou et al., 2024 for rice straw biochar-based liquid fuel. They provided interesting results to study the process in larger scale. Pilot-scale validation of the process with various biochar types would be essential next step. As conclusion biochar-based DCL process represents promising route to sustainable liquid fuel production while keeping path open for also other end applications.

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Appendix 1. Use of AI and language translators

For linguistic accuracy, ChatGPT and DeepL were used to ensure the correctness of the linguistic expression. Use is sporadic throughout the work.