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TECHNO-ECONOMICS OF LONG AND SHORT TERM TECHNOLOGY PATHWAYS FOR RENEWABLE TRANSPORTATION FUEL PRODUCTION

- DETAILED REPORT

Report from a project within the collaborative research program *Renewable transportation fuels and systems*

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PREFACE

This project has been carried out within the collaborative research program *Renewable transportation fuels and systems* (Förnybara drivmedel och system), Project no. 42406-1. The project has been financed by the Swedish Energy Agency and f3 – Swedish Knowledge Centre for Renewable Transportation Fuels.

f3 Swedish Knowledge Centre for Renewable Transportation Fuels is a networking organization which focuses on development of environmentally, economically and socially sustainable renewable fuels, and

- Provides a broad, scientifically based and trustworthy source of knowledge for industry, governments and public authorities
- Carries through system oriented research related to the entire renewable fuels value chain
- Acts as national platform stimulating interaction nationally and internationally.

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The work has been carried out in a collaborative effort with Project no. 39587-2 *Advances in Bio-oil Production*, within the Renewable transportation fuels and systems research program, and shares selected data with this project.

SUMMARY

This project has evaluated different transportation biofuel production technologies that have been identified as strategically interesting in the short and long term by the Swedish Energy Agency. The use of forest-based feedstocks for the production of transportation biofuels is a general priority in Sweden because of their good greenhouse gas performance and plentiful local supply. In the short term, drop-in hydrocarbon-based fuels have been pinpointed as a priority. The longer-term focus is on high-blend or pure biofuels, such as gasification-based methane, methanol and dimethyl ether (DME) as well as second generation ethanol, for reasons of energy and resource efficiency.

Four technology tracks spanning a range of feasible biofuel production pathways were studied. For each track, the selection of specific cases for more detailed evaluation was based on an initial screening for technical and commercial relevance in 2030. The chosen cases represent two types of forest-based feedstocks and two types of production technologies, as shown in the table below. For one of the tracks, based on the upgrading of fast pyrolysis oil, three separate cases were investigated (cases 3a-c). For cases 1-3, biofuel production was integrated with a market pulp mill, while for cases 1 and 3, it was also integrated with an oil refinery. Case 4 denotes a stand-alone biofuels plant, since integration was not found to give substantial benefits for this case with the assumptions used in this work.

		Technology/product	
		Liquefaction - hydrotreatment	**Gasification – catalytic synthesis**
Feedstock	**Kraft lignin**	Case 1. Membrane-based separation and depolymerisation of kraft lignin and hydrodeoxygenation-based upgrading to diesel and petrol (**1 MSL-HDO**)	Case 2. Black liquor gasification-based methanol production (**2 BLG**)
	Forest residues	Case 3. Pyrolysis of forest residues and upgrading to diesel and petrol a) Fast pyrolysis and hydrodeoxygenation-based upgrading (**3a Pyr-HDO**) b) Fast pyrolysis and upgrading by co-processing in a fluid catalytic cracker (**3b Pyr-FCC**) c) Integrated hydropyrolysis and hydrotreatment (**3c Hydropyr**)	Case 4. Forest residues gasification-based methanol production (**4 BMG**)

The cases were evaluated with respect to energy efficiency, profitability, investment requirement, production potential, greenhouse gas (GHG) performance, and technology maturity. The analysis was based on information that is either available in the open literature or that technology developers elected to share with the project group. For each case studied, mass and energy balances were performed and used as the basis for the estimation of selected performance indicators.

An aggregated overview of the evaluation results is presented in the table below. With one exception, the scale used is relative and shows the merits of each case in comparison to those of the others. The exception concerns GHG performance, which is measured against future Renewable Energy Directive sustainability criteria.

	Energy efficiency	Profitability	Investment requirement[a]	Production potential	GHG perform.	Technology maturity
1 MSL-HDO	+	+	+	0	0	-
2 BLG	+	+	0/-	+	+	+
3a Pyr-HDO	0	-	-	+	-	0
3b Pyr-FCC	-	-	+	-	+	+
3c Hydropyr	0	+	+	+	+	0
4 BMG	0	+	+	+	+	+

[a] + means small investment requirement, - large investment requirement

Various measures of energy efficiency were calculated for each case to evaluate performance from different viewpoints. From a systems perspective, the most relevant mode of assessment is based on system expansion that includes the effect of refinery and mill integration. It was found that integration greatly improved the system efficiency in kraft lignin-based cases (1 MSL-HDO and 2 BLG), giving efficiencies >100 %. This is explained mainly by the presence of an energy surplus at the considered modern market pulp mill. The energy surplus can be converted to biofuels with high efficiency, instead of being converted to electricity with lower efficiency. For cases 3a Pyr-HDO, 3c Hydropyr and 4 BMG, the system efficiencies were considerably lower (around 60 %). Owing to uncertainties in fossil yields from the co-processing of fast pyrolysis oil in a fluid catalytic cracker, the efficiency of 30 % for case 3b Pyr-FCC is difficult to interpret.

Profitability was assessed by using mass and energy balances in combination with commodity prices to calculate an “investment margin” for each case. This margin was then compared to specific investment costs from the literature, computed with a discount rate of both 10 % and 15 %. Three different energy market scenarios were applied: a 2018 scenario based on current prices and two 2030 scenarios, where one represented current policy and the other a more progressive climate policy (targeting 450 ppm atmospheric CO_2-eq). Practically all technology tracks were found to benefit from the projected increase in energy prices, which meant that compared to the 2018 scenario, biofuel product prices were nearly 50 % higher in the 2030 scenarios. Investment margins were found to be the highest in cases 2 and 4. Nevertheless, the lower investment cost estimates for cases 1 and 3c in the literature meant that, if a high level of technology maturity is assumed also for currently immature technologies, a good return on investment was achievable for cases 1 MSL-HDO, 2 BLG, 3c Hydropyr and 4 BMG. For a first-of-a-kind plant, it is likely that support will be required for at least cases 2 BLG and 4 BMG. When interpreting these results, it must be noted that the uncertainty in investment cost estimates is relatively high, in particular for cases 1, 3a and 3c, which represent technologies that are less mature.

For biofuel plants, in addition to profitability, the absolute size of investment is also an important factor in the investment decision, especially in the case of first-of-a-kind facilities for which the technical risk is higher. The investment requirement for each case was evaluated by scaling specific investment costs from the literature to identify the minimum plant size above which the available investment margin was enough to cover capital and O&M costs. For nth-of-a-kind plants, all economically feasible cases were found to be profitable already at relatively small scales, thus reducing the required investment. For case 2 BLG the size of the pulp mill will, however, in practice determine biofuel plant size, with the process design assumed here. The common perception that gasification is only economically feasible in large scale was not supported by this study, which indicates that the production of gasification-based biofuels can be feasible also in relatively small scales. More research is required to confirm this.

The biofuel production potentials for all the cases that use forest residues as feedstock or as feedstock replacement (case 2 BLG) were around the range 10-25 TWh/y. Compared with the rest, case 3a Pyr-HDO had a slightly lower potential because of a lower conversion efficiency. For case 1 MSL-HDO, the potential is around 4-8 TWh/y and limited by the potential lignin supply, which in turn is limited by technical restrictions related to reliable recovery boiler operation. For case 3b Pyr-FCC, the current potential is constrained by a combination of restrictions on allowed bio oil co-processing fraction and installed cracker capacity.

Greenhouse gas performance was assessed in a simplified manner. Only the major contributing factors were considered, namely, feedstock supply, electricity use and hydrogen production. For refinery-integrated production pathways (1a MSL-HDO and 3a Pyr-HDO), there is a substantial positive effect on performance if heat integration is also taken into account. The hydrodeoxygenation processes produce both heat and fuel gases that can reduce fuel oil consumption in the oil refinery. However, the accounting methodology set out in the Renewable Energy Directive permits the inclusion of the effect from fuel gases only. Compared to a fossil reference, results obtained using this methodology showed a reduction in greenhouse gas emissions of 83-95% for cases 2 BLG, 3b Pyr-FCC, 3c Hydropyr and 4 BMG. On the other hand, cases 1 MSL-HDO and 3a Pyr-HDO, both of which use a substantial amount of hydrogen that is assumed to be produced from natural gas, only registered reductions of 70% and 66%, respectively. Consequently, these technology tracks, as defined in the present work, clearly run the risk of not meeting future Renewable Energy Directive sustainability criteria. This in turn provides motivation for the use of carbon capture technologies or alternative means of hydrogen production.

For several of the studied cases, technology maturity is very heterogeneous within a specific production chain. To account for this, two methodologies were used in this study: a weighted average approach and a weakest link approach. The weighted average approach is applied over the full production chain and it gave TRLs of 6-7 for cases 2 BLG, 3b Pyr-FCC and 4 BMG, indicating that the next development step for these tracks is a commercial-scale demonstration. Consideration of the weakest link resulted in slightly lower TRLs that were nonetheless still above 5 for cases 2 and 4, and above 4 for case 3b. However, for cases 1 MSL-HDO, 3a Pyr-HDO and 3c Hydropyr, the TRLs were below 4, indicating that substantial R&D effort is still needed to bring these technology tracks to the point at which commercial deployment is feasible. It is a paradox that short-term priority is being given to the drop-in biofuel producing cases 1 and 3a-c that are based on technologies with the lowest maturity and corresponding high technical uncertainty.

In summary, the results from the assessment indicate that all four technology tracks that were studied in this project can provide good performance with respect to most of the aspects evaluated. This is subject to the provision that ongoing technology development, which is in particular needed for the refinery-integrated cases 1 and 3, is successful and leads to technological demonstration at a level that meets the assumptions made in this study about yields and costs. For case 1 MSL-HDO, an alternative hydrogen production technology will most likely be required to satisfy future emission reduction targets. The production technologies based on introduction of fast pyrolysis oil into an oil refinery (cases 3a Pyr-HDO and 3b Pyr-FCC) seem to be more problematic, both with respect to economy and other aspects.

SAMMANFATTNING

Detta projekt har utvärderat olika tekniker för produktion av skogsbaserade biodrivmedel, som har identifierats som strategiskt intressanta på kort respektive lång sikt av Energimyndigheten. Skogsbaserade råvaror för biodrivmedelsproduktion är en generell prioritet i Sverige på grund av god växthusgasprestanda och ofta relativt rikliga lokala tillgångar. På kort sikt prioriteras kolvätebaserade *drop in*-bränslen. I ett längre perspektiv är fokus av energi- och resurseffektivitetsskäl på höginblandade eller rena lågmolekylära biodrivmedel, som exempelvis cellulosabaserad (andra generationens) etanol likväl som förgasningsbaserade bränslen som metan, metanol och dimetyleter (DME).

Fyra teknikspår som spänner över en rad möjliga produktionsvägar har studerats. För varje teknikspår valdes en eller flera specifika tekniker ut för analysen, utifrån vad som antogs vara kommersiellt och tekniskt relevant år 2030. De utvalda fallen representerar två typer av skogsbaserade råvaror och två typer av produktionsteknik, som visas i tabellen nedan. För de fall där det var relevant antogs processerna vara integrerade med ett avsalumassabruk (fall 1, 2, 3a-c) och/eller oljeraffinaderi (fall 1, 3a-c). Fall 4 representeras av en stand-alone-anläggning, eftersom integrering i bedömdes ge några substantiella fördelar med de antaganden som gjorts.

		Teknik/produkt	
		Förvätskning-vätebehandling	**Förgasning-katalytisk syntes**
Råvara	**Kraftlignin**	Fall 1. Membranbaserad ligninseparation följt av depolymerisering och hydrodeoxygenering för att producera bensin och diesel (**1 MSL-HDO**)	Fall 2. Svartlutsförgasning samt katalytisk uppgradering av syntesgas till metanol (**2 BLG**)
	Skogsrester	Fall 3. Pyrolys av skogsrester samt uppgradering till bensin och diesel genom tre olika vägar: • hydrodeoxygenering (**3a Pyr-HDO**) • samprocessning i kracker (**3b Pyr-FCC**) • hydropyrolys (**3c Hydropyr**)	Fall 4. Förgasning av skogsrester samt katalytisk uppgradering av syntesgas till metanol (**4 BMG**)

De studerade fallen utvärderades med avseende på energieffektivitet, lönsamhet, investeringsbehov, produktionspotential, växthusgasprestanda och teknikmognad. Analysen baserades på öppen information samt på information som teknikutvecklare valde att dela med projektgruppen. För varje fall sammanställdes mass- och energibalanser som användes som bas för den vidare utvärderingen. En övergripande sammanfattning av utvärderingen visas i tabellen nedan. Bedömningen är gjord på en relativ skala, utom för växthusgasprestanda som bedöms gentemot hållbarhetskriterier i Förnybarhetsdirektivet.

	Energi-effektivitet	**Lönsamhet**	**Investeringsbehov [a]**	**Produktions-potential**	**Växthusgas-prestanda**	**Teknik-mognad**
1 MSL-HDO	+	+	+	0	0	-
2 BLG	+	+	0/-	+	+	+
3a Pyr-HDO	0	-	-	+	-	0
3b Pyr-FCC	-	-	+	-	+	+
3c Hydropyr	0	+	+	+	+	0
4 BMG	0	+	+	+	+	+

[a] + betyder lågt investeringsbehov, - betyder stort investeringsbehov

Bedömningen av energieffektivitet gjordes utifrån flera olika mått. Från ett systemperspektiv är ett angreppssätt som inkluderar systemexpansion, för att innefatta även effekter relaterat till de integrerade industrierna, mest relevant. Med denna metodik uppvisade tekniker som integreras med massabruk och använder kraftlignin som råvara (1 MSL-HDO och 2 BLG) mycket hög effektivitet (>100 %). Detta beror huvudsakligen på att det studerade moderna avsalumassabruket har ett värmeöverskott som utan biodrivmedelsproduktion konverteras till elektricitet med relativt sett låg verkningsgrad. Genom att istället konvertera denna överskottsenergi till biodrivmedel med bättre verkningsgrad erhålls en mycket hög beräknad systemeffektivitet på energibasis. För fallen 3a Pyr-HDO, 3c Hydropyr och 4 BMG var systemeffektiviteten betydlig lägre (runt 60 %). På grund av komplicerade och osäkra effekter på fossila utbyten i raffinaderiet vid samprocessning av pyrolysolja med fossil råvara i katalytisk kracker är den låga energieffektiviteten för fall 3b Pyr-FCC (30 %) svår att tolka.

Lönsamheten utvärderades genom att beräkna en ”investeringsmarginal” för varje fall, baserat på de framtagna mass- och energibalanserna och olika råvarupriser. Denna marginal jämfördes sedan med specifika investeringskostnader från litteraturen, beräknat med en kapitalavkastning på 10 % respektive 15 %, över 20 år. Tre olika energimarknadsscenarier tillämpades, med två olika tidsperspektiv: ett 2018-scenario baserat på dagens priser, och två 2030-scenarier, där ett representerade nuvarande policy (*”WEO-CP”, Current Policies*) och ett representerade en mer ambitiös klimatpolicy (*”WEO-450”, 450 ppm* CO_2). I princip uppvisade samtliga studerade fall en fördel av de högre energipriserna i 2030-scenarierna, i vilka biodrivmedelspriserna var nästan 50 % högre än i 2018-scenariot, vilket resulterade i motsvarande ökning i produktpriserna för producerade biodrivmedel. Lönsamhetsbedömningen uppvisade högst investeringsmarginaler för fall 2 BLG och 4 BMG men med de lägre investeringskostnaderna för fall 1 MSL-HDO och 3c Hydropyr, visade resultaten god lönsamhet för fallen 1, 2, 3c och 4. Detta gäller om bedömda investeringskostnader för mogen teknik (*nth-of-a-kind, NOAK*) används även för teknik som i dagsläget har låg teknikmognad. För omogen teknik (*first-of-a-kind, FOAK*), kommer sannolikt investeringsstöd behövas åtminstone för fall 2 BLG och 4 BMG. Noteras skall att eftersom osäkerheten i investeringskostnadsbedömningarna är relativt hög, i synnerhet för fallen 1, 3a och 3c vilka representerar tekniker av lägre teknikmognad, bör resultaten tolkas med försiktighet.

Utöver lönsamheten är även det totala investeringsbehovet en relevant faktor i investeringsbeslut rörande biodrivmedelsproduktion, i synnerhet vad gäller omogen teknik (FOAK-anläggningar) där den tekniska risken är högre. Investeringsbehovet för respektive fall utvärderades genom att skala specifika investeringskostnader från litteraturen för att för respektive teknik kunna identifiera den minsta anläggningsstorleken där den beräknade tillgängliga investeringsmarginalen var tillräcklig för att täcka kostnaderna för både kapital och drift och underhåll. Vad gäller mogen teknik (NOAK-anläggningar) så uppvisade samtliga fall som identifierats som ekonomiskt genomförbara lönsamhet redan vid relativt liten skala. För fall 2 BLG är det dock i praktiken massabrukets storlek som avgör storleken på biodrivmedelsanläggningen, med processdesignen som beaktats här. Det faktum att resultaten indikerar att förgasningsbaserad biobränsleproduktion, för mogen teknik, kan vara ekonomiskt möjlig även i relativt liten skala är intressant och går emot en allmän uppfattning att förgasningsteknik endast är ekonomiskt möjlig i stor skala. Mer forskning behövs för att bekräfta dessa resultat. För svartlutsförgasning kan detta motivera ytterligare studier specifikt avseende anläggningar baserade på endast en andel av ett massabruks svartlutsflöde, vilket kan bidra till avlastning av sodapannan på samma sätt som ligninseparation enligt fall 1 MSL-HDO.

Produktionspotentialen är av samma storleksordning, ca 10-25 TWh/år, för alla fall som använder skogsrester som råvara eller som ersättning för råvara, som för fall 2 BLG. Den lägre effektiviteten i fall 3a Pyr-HDO minskar denna potential något. För fall 1 MSL-HDO är potentialen 4-8 TWh/y och begränsas av möjligt ligninuttag, vilke ti sijn tur begränsas av tekniska restriktioner för pålitlig drift av sodapannan. För fall 3b Pyr-FCC är potentialen lägre och begränsad av integrationen genom maximal inblandning vid samprocessning och krackerkapacitet.

För växthusgasprestanda gjordes en förenklad form av utvärdering, där endast faktorerna med störst bidrag till växthusgasavtrycket beaktades (råvaruförsörjning, elanvändning och vätgasproduktion). Utvärderingen gjordes dels enligt Förnybarhetdirektivets metodik, dels genom tillämpning av systemexpansion. För de raffinaderiintegrerade fallen, 1 MSL-HDO och 3a Pyr-HDO, fanns betydande positiv effekt på växthusgasprestandan när även värmeintegration i raffinaderiet beaktades (systemexpansion), eftersom hydrodeoxygeneringsprocesserna ger upphov till både överskottsvärme och bränngaser som har potential att minska användningen av eldningsolja i raffinaderiet. Dock tillåter beräkningsmetodiken i Förnybarhetsdirektivet endast att substitutionseffekter från gaserna beaktas, och inte från överskottsvärmen. Fallen 2 BLG, 3b Pyr-FCC, 3c Hydropyr och 4 BMG uppvisade alla tillfredsställande växthusgasprestanda, med växthusgasbesparingar på 83-95 % jämfört med en fossil referens, när beräkningarna görs enligt Förnybarhetsdirektivets metodik. För de fall som använder stora mängder (naturgasbaserad) vätgas, fall 1 MSL-HDO och fall 3a Pyr-HDO, var reduktionen endast 70 % respektive 66 %. Det innebär att dessa teknikspår, med den utformning som antagits här, tydligt riskerar att inte uppfylla framtida hållbarhetskriterier. Detta motiverar i sin tur användning av koldioxidinfångning, och/eller alternativ vätgasproduktionsteknik, för att kunna uppfylla kraven.

Teknikmognaden är för flera av de studerade fallen mycket heterogen även inom en given produktionskedja. Två olika angreppssätt för att ta hänsyn till detta användes i denna studie: en ”viktat medelvärde”-skattning och en ”svagaste länken”-skattning. Bedömningen av teknikmognad baserat på hela produktionskedjan (viktat medelvärde) visade TRL 6-7 för fall 2 BLG, 3b Pyr-FCC och 4 BMG, vilket indikerar att nästa steg för dessa spår är en kommersiell demonstration. Att istället beakta den svagaste länken resulterade i något lägre TRL, men fortfarande över 5 för fall 2 och 4, och över 4 för fall 3b. För fall 1 MSL-HDO, fall 3a Pyr-HDO och fall 3c Hydropyr, visade sig TRL från ett svagaste-länken-perspektiv däremot vara under 4, vilket indikerar att stora FoU-insatser krävs för att få dem till ett tillstånd där kommersiell implementering är möjlig. Det är en paradox att de kortsiktigt prioriterade teknikspåren i fall 1 och 3a-c, vilka producerar *drop in*-bränslen, i allmänhet baseras på tekniker med lägre teknisk mognad.

Sammanfattningsvis indikerar resultaten från utvärderingen att produktionsteknik från alla de fyra studerade teknikspåren kan ge god prestanda med hänsyn till de flesta av de utvärderade aspekterna. Noteras skall dock att detta förutsätter att den fortsatta teknikutveckling som fortfarande krävs, i synnerhet för de raffinaderiintegrerade fallen (1, 3a-c), är framgångsrik och leder till demonstration av teknik som uppfyller de antaganden som gjorts i denna studie om utbyten och kostnader. För fall 1 MSL-HDO kommer alternativ vätgasproduktionsteknik sannolikt behövas för att kunna uppfylla framtida hållbarhetskriterier. Produktionsteknik baserat på uppgradering av pyrolysolja i ett oljeraffinaderi (fall 3a och 3b) verkar, utifrån resultaten i denna rapport, ha större utmaningar än de andra alternativen, både ekonomiskt och vad avser andra aspekter.

CONTENTS

1 INTRODUCTION

1.1 BACKGROUND

This project evaluates different transportation biofuel production technologies that have been identified as strategically interesting in the short and long term by the Swedish Energy Agency. Forest biomass-based biofuels are a strategic priority because of their, generally, good greenhouse gas performance and plentiful supply in Sweden [1]. SOU 2013:84 (”Fossilfrihet på väg”) [2] estimates the potential for biofuels from domestic feedstock to be 25-30 TWh/y. A large share of this is expected to come from forest based feedstock, especially harvesting residues (12-18 TWh/y). Additional potential can be realised by utilising industrial by-products such as saw dust, bark and lignin.

The Swedish Energy Agency pinpoints drop-in fuels as a short term priority [3] and currently gives significant support to development of technology for forest based drop-in fuels, primarily the upgrading of kraft lignin (e.g. the RenFuel pilot plant development in Bäckhammar, Sweden). The long-term focus is on high-blend or pure biofuels, such as gasification-based methane, methanol and dimethyl ether (DME) as well as second generation ethanol, for reasons of energy and resource efficiency [3].

Economically competitive biofuels that meet sustainability criteria are a prerequisite for a transition to a fossil independent transportation system. So far, the authors of this report are not aware of any publically available study that evaluates and compares techno-economic performance of short and long-term forest-based biofuel alternatives. This study aims to fill this gap by comparing four different strategically interesting technology tracks for production of biofuels from forest biomass.

The integration of biofuels production with existing industries can be beneficial from several perspectives. For example, there may be opportunities for more efficient heat integration as well as utilization of surplus energy and by-product flows. The size of required investment can also be reduced through the use of existing processing equipment, avoided alternative investments, logistic synergies and, last but not least, the utilization of important know-how. For this reason, the present work focuses on the integration of biofuel production technologies with pulp mills and refineries as they are considered the most promising alternatives for achieving the above-mentioned benefits.

There are a range of different options available for pulp mill-integrated biofuels production. Black liquor gasification is a well-studied track that has been demonstrated in pilot scale [4–8] and evaluated techno-economically [9–11]. A different approach for the use of kraft lignin as feedstock involves its separation from black liquor. Several R&D projects investigating this pathway are underway in Sweden, including at companies such as RenFuel AB, SunCarbon AB and SCA. When forest residue is used as the feedstock, the most relevant technologies at present are pyrolysis and gasification. In these cases, pulp mill integration relates primarily to logistics and heat integration.

The four technology tracks that have been selected for evaluation in this study represent two main types of production technologies: (i) liquefaction (depolymerisation through pyrolysis or hydrothermal treatment) followed by hydrotreatment upgrading, and (ii) gasification followed by catalytic synthesis. Two main types of feedstocks are used (a) kraft (black liquor) lignin, and (ii) forest residue (tops and branches) as shown in Table 1. More details on the technology tracks are provided in Chapter 3.

Table 1. Technology tracks studied.

	Liquefaction - hydrotreatment	**Gasification – catalytic synthesis**
Kraft lignin	Case 1. Depolymerisation and upgrading of kraft lignin to diesel and petrol	Case 2. Black liquor gasification-based methanol production
Forest residues	Case 3. Pyrolysis of forest residues and upgrading to diesel and petrol	Case 4. Forest residue gasification-based methanol production

The descriptions of the refinery integrated, liquefaction – hydrotreatment, processes make quite extensive use of refinery specific nomenclature. For readers unfamiliar with the terminology, Appendix A provides brief descriptions of the process types discussed.

1.2 PROJECT OBJECTIVES

The primary aim of this project is to deliver a comparative techno-economic evaluation of four different technology tracks for the production of biofuels from forest-based feedstocks. The chosen tracks are listed in Table 1 and their selection follows the short and long-term priorities of the Swedish Energy Agency. The evaluation is carried out with both current energy prices and scenarios for future energy market conditions.

Specific aims are to present:

- material and energy balances for the integration of biofuel production with a pulp mill and an oil refinery
- acceptable specific investment cost for selected technology pathways and to compare these with investment cost estimates from the literature. Resulting acceptable specific investment cost can be used as targets for future technology development.
- estimates of required scale for a commercial implementation of each technology, based on economies of scale.
- estimates of technology maturity for each technology pathway, including for individual process steps
- estimates of national production potential for each technology track

1.3 DISPOSITION OF THIS REPORT

This report is structured as follows:

- Chapter 2 provides an overview of the methodologies used to perform the following assessments: (a) techno-economic analysis (2.1), (b) greenhouse gas (GHG) accounting (0), (c) technology maturity determination (2.3) and (d) production potential estimation (2.4).
- Chapter 3 contains detailed descriptions of the evaluated technology tracks and the assumption used to model their technical performance for carrying out mass and energy balances calculations. It also describes the procedure used to select the specific technologies analysed in cases where several alternatives are possible.
- Chapter 4 discusses the results. First (section 4.1), the energy balances are presented and the need for additional commodities like hydrogen and electricity is addressed. Various measures of efficiency are also discussed. After that (section 4.2), production costs (excluding capital costs) are estimated based on the balances. These are used together with expected revenues from biofuel sales to identify specific investment margins for each case, which represent the sums available for covering investment and O& M costs. Sections 4.3,

4.4 and 4.5 discuss GHG performance, technology maturity and domestic production potential for each of the tracks. Finally, section 4.6 gives a comparison of the available investment margins (from 4.2) with investment cost estimates from the literature in order to assess the economic feasibility of each case. The influence of economies-of-scale and technology learning (first plant versus n^{th} plant) on economic viability is also discussed in this section.

- Chapter 5 provides a set of conclusions based on the analysis of the results.

2 METHODS

This report provides a comparative techno-economic assessment of different technological pathways for producing forest-based biofuels. As noted in section 1.1, the selection of pathways was guided principally by the Swedish Energy Agency’s roadmap for the establishment of a fossil-independent transportation fleet in Sweden. Drop-in fuels are to be prioritized in the short run, while both pure biofuels and high blend variants are seen as viable options in the medium to long run. Based on this and on the assumption that forest residue and lignin are promising feedstocks from a Swedish perspective, the selection of tracks is as shown in Table 1 above.

The first part of the project consisted of a preliminary screening of each of the four studied technology tracks, in order to select the specific cases to include in the analysis. This was done partly as a literature review, partly by contacts with technology developers for each of the tracks. The final selection of cases to include was primarily based on data availability and reliability and is discussed separately for each case in chapter 3, in combination with the assumptions made and the sources of those.

For each selected case, mass and energy balances were performed and used as basis for the assessment evaluation. The selected cases were evaluated with respect to energy efficiency, profitability, investment requirement, greenhouse gas (GHG) emission performance, technology maturity, and production potential.

This chapter discusses the methodology used in the analysis carried out for the selected technology tracks from a generic point of view.

2.1 TECHNO-ECONOMIC ANALYSIS

Normally, in a techno-economic analysis aimed at cross-technology comparisons, investment costs would be used in conjunction with income projections and operating expenditure to perform a discounted cash flow analysis, which can be set up to generate handy specific production cost estimates. However, the use of this approach for the cases under focus in this study was rendered problematic by the findings of the initial assessment, which indicated that the selected technology tracks are at substantially different stages of technical maturity in their development curve. This has two important consequences. First, there is a paucity of key technical data such as product yields and conversion efficiencies in published literature for those cases that are at a comparably early stage of R&D. Second, and more problematically, component costs and other assorted capital expenditure (CAPEX) costs figures are even harder to source, especially from the public domain. A high level of uncertainty thus exists regarding the very possibility of establishing reasonably accurate investment estimates.

The first obstacle was mitigated against by liaising with technology developers to source pertinent data that was not available in scientific literature. This enabled the determination of mass and energy balances with the most up-to-date information available, as discussed in chapter 3 for each of the chosen cases. The second obstacle was harder to overcome, as the numbers concerned may be unavailable or withheld for reasons of commercial confidentiality. However, the issue was bypassed by turning the standard approach on its head to instead calculate a specific margin for investment in each case, which was compared with existing investment estimates from multiple sources to assess and compare the viability of the selected cases. This approach uses uncertain CAPEX estimates in a more qualitative manner, which was considered an advantage.

2.1.1 Mass and energy balances, energy efficiency assessment and integration

Detailed mass and energy balances were performed to calculate product yields and energy efficiencies for all cases, which formed the basis of the subsequent economic analysis. Data was extracted from existing studies in the first instance, while the missing gaps were filled by consulting relevant technology developers and carrying out process modelling in Aspen Plus, where required. Given the differences in feedstocks, the exact system boundaries differed somewhat between the cases, but in general the balances took into account all process steps from the acquisition of as-received feedstock to the final production of biofuels, including any upgrading or refining. Chapter 3 provides a more detailed account of the data used and the underlying assumptions for each case.

Integration of biofuels production with existing industries is often beneficial, see e.g. [12,13]. With one exception, that of the gasification of forest residues case, the production of biofuels was assumed to be integrated with an existing pulp mill. Similarly, the final upgrading of the products from all liquefaction-hydrotreatment cases was assumed integrated with existing operations at a crude-oil refinery. The effects of the integration of biofuel production on the energy balance of existing facilities were included in the calculations. Thus, the cases evaluated in this report represent integrated energy systems with a multiplicity of energy carriers and conversion steps.

Depending on where the system boundary is drawn, the efficiency of energy use can be calculated differently. Accordingly, three different measures of efficiency, as defined below, were calculated in order to facilitate performance comparisons across the cases in a consistent manner.

$\eta_{conversion}$ is a measure of the thermal efficiency of biomass conversion to transportation fuels. It is defined in Eq. 1 as the ratio between energy in the finished biofuel product and energy in the biomass feedstock, both in units of MW HHV (higher heating value). The feedstock is black liquor for the case 2 BLG, lignin retentate for case 1 MSL-HDO, and forest residues in all the other cases. The finished biofuel products are either methanol (cases 2 and 4, see chapter 3) or refined petrol and diesel (cases 1 and 3, see chapter 3).

$\eta_{conversion} = E_{Product} / E_{feedstock}$ (Eq. 1)

Besides the thermal energy in the feedstock itself, other energy sources such as electricity and hydrogen may also be needed for the production of the finished biofuel product. This calls for the estimation of an efficiency figure that takes into account all the energy inputs into the biofuel production process. Thus, an overall system efficiency η_{system} for the process is defined in Eq. 2 as the ratio, in units of MW, of the energetic value of biofuel products to the energetic value of all primary energy inputs. Only net flows are considered and the primary inputs for each case are presented in Table 2.

$\eta_{system} = E_{Product} / E_{Primary\text{-}Inputs}$ (Eq. 2)

In the definition for η_{system} given above, the system boundary is drawn around the biofuel production process. With the exception of one case (case 4), all other cases include the integration of different stages in the biofuel production chain with existing refinery and pulp mill operations. In most cases, the integration also causes a change in existing heat and electricity balances. The effect of integration can be quantitatively assessed across the cases with the help of a measure such as an efficiency $\eta_{system\text{-}exp}$ that takes into account effects such as reduction in heating oil usage or electricity generation. This requires that the system boundary be expanded to include the integrated mill/refinery. For the MSL-HDO (case 1) and BLG (case 2) cases, the feedstocks, black liquor and lignin retentate, respectively, are actually inside the system boundary on account of being by-products of the pulping process. Thus, a set of "Integrated inputs/products" are identified for each case and these are presented in Table 2. Accordingly, $\eta_{system\text{-}exp}$ is defined in Eq. 3 as the ratio, in units of MW, of the energetic value of the biofuel product to the net energetic value of "expanded inputs".

$$\eta_{system\text{-}exp} = E_{Product} / E_{Expanded\text{-}Energy\text{-}Inputs} \quad \text{(Eq. 3)}$$

Table 2. Energy inputs for efficiency calculations. Cases described in chapter 3.

	Primary Energy Inputs [η_{system}]	**Integrated Energy Inputs/Products [$\eta_{system\text{-}exp}$]**
1 MSL-HDO	Net BL flow, hydrogen, electricity	Reduction in electricity export from pulp mill to grid, hydrogen, reduction in heating oil use in the refinery.
2 BLG	Black liquor, electricity	Additional biomass (forest residues), Increase in electricity import from grid to pulp mill.
3a Pyr-HDO	Biomass (forest residues), hydrogen[A]	Biomass (forest residues), hydrogen, reduction in heating oil use in the refinery.
3b Pyr-FCC	Biomass (forest residues), hydrogen[A]	Biomass (forest residues), hydrogen, Increase in heating oil in the refinery.
3c Hydropyr	Biomass (forest residues), hydrogen[A]	Biomass (forest residues), hydrogen
4 BMG	Forest residues, electricity	Forest residues, electricity

[A] Electricity consumption in cases 3a, 3b and 3c has been neglected.

On account of the difference in energy quality between electricity and thermal energy carrier, two additional measures $\eta_{system\text{-}el}$ and $\eta_{system\text{-}exp\text{-}el}$ were calculated by converting all the relevant energy flows into their electricity equivalents. The conversion efficiencies used in the calculations are listed in Table 3.

Table 3. Conversion factors for the calculation of energy inputs in electricity equivalents.

Energy carrier	**Power generation efficiency [%]**	**Comments**	**Reference**
Methanol	55.9		[14]
Hydrogen	60.0	250 kW molten carbonate fuel cells	[15]
Biomass	46.2		[14]
Heating Oil	44.0	Steam turbine fuel-oil power plant	[16]
Diesel/Petrol	44.0	Assumed to be same as for heating oil	[16]
Black Liquor	22.0	Black liquor gasification combined cycle	[17]
Lignin retentate	21.34	Using an electricity to BLG efficiency of 22%	Estimated Internally

Pulp mill integration

Pulp mill integration considers material integration for cases 1 and 2, since the feedstock in these cases is black liquor lignin, and heat integration for cases 1, 2 and 3. A generic state-of-the-art pulp mill producing 700 kADt of kraft softwood pulp per year [18] was selected to represent the integration. In normal operation, all BL (557 MW) is combusted in the recovery boiler and the falling bark is combusted in the lime kiln (34 MW) and a power boiler (82 MW). Both boilers are used to produce process steam (455 MW) and the surplus is used to generate electricity for internal use and a surplus of 82 MW_{el}, which is sold to the grid as shown in Table 4. The efficiency of the low-pressure turbine in the model was 20.2 %.

Table 4. Power balance for model pulp mill.

	kWh/ADt	MW
Generation back-pressure part of the turbine	840	70
Generation condensing part of the turbine	871	73
Process consumption	726	61
Sold	984	82

The effects of integration on mass and energy balances are specific to the biofuel production process, i.e. different for different cases. For this reason, the details of the material integration are discussed for each case in chapter 3. The effect of integration of a specific biofuels process is dependent on the type of pulp mill considered. Hence, the use of a mill that is energy deficient would give other results as discussed further in chapter 3 for relevant cases. There is also a potential synergy concerning feedstock logistics, but the value of this has not been quantified.

Refinery integration

The upgrading of intermediate products from all liquefaction-hydrotreatment cases was considered to be integrated with existing operations at a crude-oil refinery. It should be noted that the refinery in question for case 3b involving fluid catalytic cracker (FCC) integration can only be realized in a refinery that has FCC equipment in operation.

The considered integrated refinery has sufficient hydrogen production capacity for the hydrogen needed for upgrading of liquefied biomass to finished fuel product in cases 1 and 3. We have assumed natural gas based hydrogen production as the base case but alternatives, including electrolysis and reforming of biogas are possible and interesting alternatives from a sustainability perspective (see discussion in 4.3). It was further assumed that a hydrocracker is available and can be used to crack the heavy fractions from both hydrodeoxygenation (HDO) and FCC processing of bio-oils. In order to make Swedish diesel (“MK1”), a final upgrading unit to adjust for example aromatics content is required.

For refinery internal heat integration, it was assumed that heat for the (fossil) refinery processes is supplied through a furnace firing fuel oil. It was further assumed that NCGs formed in the biofuel production processes would partly replace the fuel oil and that process heat recovered from the biofuel production processes would be used to decrease fuel oil consumption in the furnace.

As discussed further in 2.1.2, it was assumed that the refinery is operating at capacity and that biofuel production in existing refinery equipment limits the fossil fuel production capacity.

Heating values

All energy balances for the production pathways were made and reported on higher heating value basis unless otherwise specifically noted. GHG emissions were reported on lower heating value basis since that is the conventional approach. Table 5 lists generic higher heating values used in this work unless otherwise specifically noted. Lower heating values are also presented but are not used in this report except for section 4.3. Heating values for energy carriers that are process internal and case specific are presented for each case in chapter 3.

Table 5. Higher heating values (HHV) and lower heating values (LHV) of energy carriers.

Energy carrier	MJ/kg DS (HHV)	MJ/kg DS (LHV)
Black liquor	13.2	9.5
Pyrolysis oil	22.7	21.6
Forest residues	20.7	19.7
Hydrogen	142	120
Fuel oil (for refinery heat production)	43	40
Methanol	22.9	20.1
Diesel	45	42
Petrol	47	43

2.1.2 Economic evaluation

As the first step in the economic evaluation, *a specific investment margin* is calculated for each case in units of SEK/MWh to estimate the amount available for covering the capital and O&M costs of an investment. This is done by dividing the net annual operating income, which is defined as biofuel product revenues minus direct costs, by the annual biofuel production volumes.

A *specific investment cost* was calculated for each selected investment estimate by dividing the specified total investment cost by the given biofuel production volumes. The yearly capital expenditure of the investment was calculated using a 15 % internal rate of return (IRR) over 20 years, giving an annuity of approximately 0.16. As an alternative 10 % rate over 20 years, giving an annuity of 0.12, was used. An operation and maintenance (O&M) cost that was set at 4 % of the investment cost was added to the resulting capital expenditure.

Although the parameter *expenditure on commodities* is used as a representative measure of OPEX costs in this report, the two terms are not fully synonymous. This is because, in addition to the costs of feedstock, utilities and process consumables such as catalysts and chemicals, the latter is generally also defined to include operations & maintenance (O&M) costs. They are generally calculated on an annual basis as a percentage of CAPEX. For this reason, O&M costs were added to the investment estimates and thus form part of the annualized specific investment costs.

A systematic description of how the concept of specific investment margin was used to assess the economic performance of the chosen cases, in conjunction with investment estimates from the literature, is provided below[1]:

1) Specific investment costs were compared with the estimated specific investment margins to evaluate the economic attractiveness of the cases. Given the significance of the learning curve effect to emerging technologies, a distinction was made between first-of-a-kind (FOAK) and nth-of-a-kind (NOAK[2]) investment estimates in the comparison.
2) The effect of technology readiness on the economic performance of the cases was investigated. This was done in a somewhat simplistic yet illustrative manner by varying the on-

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[1] For details on the studies and the used capital cost estimates, see section 4.6.1.

[2] NOAK is not a well-defined term and the interpretation may differ between the different studies from which investment cost data are sourced (see section 4.6). Generally, NOAK is denotes a state in which construction and operation experience is substantial enough so that the learning curve has flattened.

stream factor used for calculating specific investment costs for different cases in accordance with the results of the technology readiness assessment (see section 2.3).

3) The influence of plant size on specific investment cost was also examined. In conjunction with the capital investment margin estimates, this enabled the estimation of the required scale for commercial viability. A simplified approach for estimating economies of scale was used; the full investment was scaled using an exponent of 0.65.
4) An appraisal of the contribution of each commodity to the specific investment margin was used to select the most relevant parameters for a one-at-a-time *sensitivity analysis*. Specific investment costs were also computed at a lower IRR (10 %) to examine the impact of expected financial returns on the attractiveness of the selected technologies.

It is apparent that the integration of biofuel production facilities with a pulp mill or a crude oil refinery would also affect financial outcomes such as electricity exports and fuel imports; for example in that the replacement of fossil crude with an equal amount of renewable bio-oil changes the final product mix. A central starting point in this work was the use of existing refinery equipment in order to leverage existing infrastructure and to support a transformation of crude oil refineries to biofuel production. It was assumed that refining units (such as the SynSat reactor) are being operated at full capacity prior to the introduction of bio-oil[3]. Modifications would be needed, if these units were to handle the increased volumes of petrol and diesel coming from the biogenic feedstock. The possibility and cost of such debottlenecking is very case specific and it is impossible to accomplish a generic analysis. Our approach assumes a reduction in fossil diesel and petrol throughput by an amount equal in volume to the extra renewable diesel and petrol produced from the bio-oil[4]. In conjunction with the latest data on gross margins for Swedish diesel and petrol, this enabled the estimation of a "reduced fossil production" cost. This can be viewed as that when biofuel production processes use existing equipment that was previously used for fossil fuel production, they avoid investments but instead the production is associated with a cost for the used equipment capacity equal to the profit that the corresponding fossil production would have given.

All prices and cost estimates used in this study are denominated in 2017 Swedish crowns and they were adjusted for inflation, where applicable, before being converted using the exchange rates 9.5 SEK/EUR and 8.5 SEK/USD.

2.1.3 Energy market scenarios and cost of commodities

The evaluation was made using two different time-frames for energy related costs and prices: near-future (2018) and medium-future (2030). For other costs, no difference was assumed between the two time-frames.

For 2018, energy market parameters based on the current situation were used for the evaluation, while for 2030, energy market scenarios created using the ENPAC tool (Energy price and Carbon Balances Scenarios tool) were applied. The ENPAC tool has been developed in order to generate

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[3] Such an assumption would of course not be applicable to new-built facilities. However, in this study only integration with an existing refinery was considered.

[4] This generic approach corresponds to assuming an ideal refinery in which every equipment/pipe/tank is used at its capacity. Hence, any increase in production of biofuel will replace production of fossil fuels.

consistent energy market scenarios (e.g. that consider inter-parameter relationships between different energy market parameters) for different time horizons. The tool has been updated regularly in order to reflect changes in the energy market and energy system, and has been described by Axelsson et al. [19–22]. The ENPAC tool calculates energy prices for large-volume customers based on input in the form of forecasted global fossil fuel prices, and projected policy instruments (e.g. costs for emitting CO_2, green electricity certificates, and subsidies for renewable transportation fuel production). For this study, two different scenarios generated using ENPAC 2.0 were considered: the "2030-450" scenario (450 ppm; scenario for limiting the concentration of greenhouse gases in the atmosphere to around 450 parts per million of CO_2-equivalents) and the "2030-CP" scenario (Current Policies scenario) [22]. The ENPAC scenarios were generated using estimations from the World Energy Outlook (WEO) 2016 as basis for fossil fuel prices and costs for emitting CO_2.

Central aspects for the analysis in this study were in particular the assumed biomass price and the assumed revenue for produced biofuels. For biomass, the market price in ENPAC has been defined as potential marginal (price-setting) users' estimated willingness to pay for biomass. Potential marginal users currently considered in the tool are coal power plants for co-firing, and biofuel production (transport fuels). This reasoning builds on the assumption that biomass is subject to competition and traded on a functioning (international) market, which is not the case today.

For renewable transportation fuels, ENPAC distinguishes between diesel and petrol substituting fuels, in order to reflect real world dissimilarities between fossil products, as well as differentiated taxes adhering to the different fossil fuels. Renewable transport fuels have been assumed to be subject to energy taxes similar to their fossil counterparts, but not to any CO_2 charge. In order to create a willingness to pay for biofuels, and to counteract the higher costs of producing biofuels compared to fossil fuels, renewable transport fuels have instead been assumed to be entitled to support in the form of e.g. a tax reduction or green certificates (compare green electricity certificates). This has in ENPAC 2.0 been set to the EU average tax reductions in 2009 [23]. The willingness to pay for renewable transport fuels (at the gate) is then calculated as the product price at the pump for conventional transport fuels (diesel or petrol, respectively) minus the distribution cost for the renewable fuel, plus the possible support for renewable transport fuels. A fundamental assumption in ENPAC is that of a harmonised CO_2 market, with the same CO_2 charge in the transport sector as in other sectors. In the 2030-450 scenario, this thus results in an increase of almost 50 % of the petrol price when the CO_2 charge is added. In addition to this, biofuels have been assumed to be entitled to tax reductions or similar support, which in combination gives relatively high biofuel selling prices.

ENPAC will likely undergo substantial revision, in particular regarding the biomass and transport fuel modules, which have been highlighted as in need of improvement [22]. For this study, the ENPAC scenarios in their existing form were deemed acceptable for the analysis, in combination with the near-future scenario based on current energy market conditions.

Table 6 shows the energy market parameters for the three scenarios, as used in the techno-economic evaluation. Table 7 summarises the catalyst costs. As noted in the table, prices for HDO catalysts are projected to fall sharply in the coming years, though in the absence of data from the last two years, the extent to which this is likely to improve the economics is as yet unclear. Table 8 shows other chemical costs. In Table 9, other cost parameters are finally summarised.

Table 6. Energy market parameters for the three studied scenarios. General sales tax is not included.

	Scenario 2018			Scenario 2030-450 [ENPAC] [i]		Scenario 2030-CP [ENPAC] [j]	
Commodity	**[SEK/kg]**	**[SEK/MWh]**	**Note**	**[SEK/kg]**	**[SEK/MWh]**	**[SEK/kg]**	**[SEK/MWh]**
Forestry residues	0.52	187	a	1.09	390	1.09	390
Heating oil [E10/EO1]	4.46	375	b	6.32	532	9.37	789
Natural gas	4.46	335	c	7.99	551	6.75	466
Ethanol [T2]	6.72	810	d	9.06	1093 [k]	9.61	1159 [k]
Methanol	5.15	810	d	6.94	1093 [k]	7.37	1159 [k]
HVO diesel	8.72	713	d	13.35	1093 [k]	14.28	1169 [k]
Fossil diesel	4.53	356	e	10.27	808	11.48	903
Fossil petrol	5.91	467	e	9.49	751	10.21	808
Hydrogen	15.88	402	f	28.49	721	24.07	609
Electricity (purchased)	N/A	313	g	N/A	485	N/A	485
Electricity (sold)	N/A	289	h	N/A	485	N/A	485

[a] Average Swedish price in 2016 for wet woodchips from both coniferous and deciduous forest residues [24]. Including transportation costs.

[b] Average Swedish price in week 37 of year 2016 for Preem AB (personal communication Åsa Håkansson, Preem AB). Excluding energy taxes and transportation costs.

[c] Average Swedish price in 2016 for industrial customers in the I4 category [3000 – 30000 MWh] [25]. Including distribution costs and taxes.

[d] Average European price of ethanol [T2] and FAME, respectively, in 2016 [26]. FOB ARA: excluding insurance and transportation costs. For methanol, the price is assumed to be the same as for ethanol on an energy basis, given the current absence of a market for methanol as transportation fuel.

[e] Reference fossil fuel price in 2016 (Sweden) [26]. Includes production cost and gross margin but excluding all taxes.

[f] Calculate as 3.564 times the price of natural gas from steam-methane reforming, on a mass basis. The multiple represents the average of a low and a high estimate in the literature.

[g] Average Swedish price in the first half of 2017 for industry customers [70000 – 150000 MWh] [27]. Including transmission costs and tax but excluding electricity certificates.

[h] Average hourly spot price in the first half of 2017 for South Central Sweden [28]. Excluding energy tax, electricity certificates mark up and transmission.

[i] Input data to ENPAC 2030-450: Crude oil (Brent): 85 USD/bbl.; Natural gas (EU import): 9 USD/Mbtu; Coal (OECD import): 57 USD/tonne; CO_2 charge (EU ETS): 90 EUR/tonne CO_2; support “diesel biofuels”: 28 EUR/MWh; support “petrol biofuels”: 37 EUR/MWh [22,29].

[j] Input data to ENPAC 2030-CP: Crude oil (Brent): 127 USD/bbl.; Natural gas (EU import): 11 USD/Mbtu; Coal (OECD import): 80 USD/tonne; CO_2 charge (EU ETS): 27 EUR/tonne CO_2; support “diesel biofuels”: 28 EUR/MWh; support “petrol biofuels”: 37 EUR/MWh [22,29].

[k] Represents willingness to pay for biofuels, including support for biofuels, as defined in the scenarios. The energy tax for biofuels has been assumed to be the same as for fossil fuels, while biofuels have been assumed to not be subject to CO_2 charge.

Table 7. Catalyst costs as used in the estimations of the specific investment margins. The specific cases are further described in chapter 3.

	Case	Catalyst price [SEK/kg biofuel]	Comments	Source
1	MSL-HDO	1.74	The combined cost of catalyst for the hydro-deoxygenation step. Common to both Case 1 and Case 3a. A 2017 projection by PNNL has the annual cost falling to only a quarter of that seen in 2015 (see Table 4 in [30]). Given that this rapid drop does not mirror recent historical trends, we have opted to use a figure that is approximately half that of the most recent available numbers (2015) and very similar to the 2016 projection.	[30]
2	BLG	0.01	The combined cost of methanol and water-gas shift catalysts. Common to both Case 2 and Case 4. The estimate is based on data supplied in confidence by the industry.	Internal/ industrial estimate
3a	Pyr-HDO	1.74	The combined cost of catalyst for the hydro-deoxygenation step. Common to both Case 1 and Case 3a.	[30]
3b	Pyr-FCC	0.12	An estimate of FCC catalyst costs that may be attributed to the upgrading of additional renewable diesel and petrol.	Internal estimate
3c	Hydropyr	0.61	The cost of make-up catalyst for the hydro-pyrolysis step. It may be wise to treat the figure with circumspection since it is based on an estimate that goes back a decade, while the technology has undergone significant development in the interim.	[31]
4	BMG	0.01	The combined cost of methanol and water-gas shift catalyst.	Internal/ industrial estimate

Table 8. Cost of chemicals [32].

Commodity	Price [SEK/kg]
CO_2	0.81
NaOH	18
H_2SO_4	0.72
KOH	9.0
NaOH	3.6
ZrO_2	300
C_6H_5OH	14.4

Table 9. Misc. parameters and costs used in the estimations of the specific investment margins.

Parameter	Value	Unit	Comments	Source
Gross margin for petrol	1.91	[SEK/kg]	The average gross margin for Swedish petrol in 2016. It is used to estimate the cost of employing at-capacity infrastructure for the final refining of additional renewable diesel and petrol.	[26]
Gross margin for diesel	1.28	[SEK/kg]	The average gross margin for Swedish diesel in 2016. Used for the same purpose as the gross margin for petrol.	[26]
Ash disposal	0.36	[% expenditure]	An estimate of the cost of ash disposal in Case 4	[33]
Bed material	0.81	[% expenditure]	An estimate of the cost of bed material in Case 4	[33]

2.2 GREENHOUSE GAS EMISSION PERFORMANCE

A full life cycle analysis based greenhouse gas analysis has been outside the scope of this project, which has primarily focused on production costs and technology maturity. An estimation of greenhouse gas performance of the production pathways analysed is, however, an important component for determining their competitiveness. A simplified greenhouse gas analysis has been included for this reason.

The mass and energy balances developed for each production pathway provide quantification of feedstocks and inputs. Previous work, for example in the f3 project "Value chains with intermediate biofuels" [34], and published data for gasification [35–37] show that including forest residue feedstock supply, electricity and hydrogen covers the absolute largest part of the climate impact of the processes, except when biomass is studied from a marginal perspective with coal-fired coal production as alternative use [37], which is not the case in this work.

We have not included any direct or indirect land use change effects, since the feedstock used is forestry residues. Hence, the footprint associated with the feedstock is only related to collection and transportation. A value of 2.2 g CO_2eq/MJ has been used, corresponding to "typical" collection technology and transport distances [38]. Hydrogen was assumed to be manufactured through steam reforming of natural gas, which is the dominating technology today and also in agreement with planned production at the Preem refinery [39]. A GHG footprint of 91.4 g CO_2eq/MJ LHV (lower heating value) has been used for refinery internal hydrogen production [40]. For electricity, the GHG footprint of Swedish electricity mix of 13.1 g CO_2eq/MJ was used, which is in accordance with recommendations from the Swedish Energy Agency[5] [41]. Fuel oil replacement by refinery integration was accounted for in some GHG evaluation scenarios, see results in 4.3. In those cases, a GHG footprint of 80 g CO_2eq/MJ LHV was used based on the reference value for heat production in the European Commission proposal for a new Renewable Energy Directive (RED) [42].

In the cases involving HDO treatment of lignin oil or pyrolysis oil (cases 1 and 3a), a substantial amount of heat is produced that can be used in the fossil processes on the refinery, as detailed in chapters 3 and 4.1. While RED does not allow allocation of GHG emissions to heat, we have nevertheless also included a GHG accounting methodology using system expansion, in order to illustrate the effects from using this heat. In the Pyr-FCC case, as discussed in 3.3.2 below, there is an impact from the co-processing on yields of fossil products. In the discussions of the GHG effects of this we have applied system expansion and included the use of these fossil fuels with GHG footprints of 93.5 g CO_2eq/MJ and 95.5 g CO_2eq/MJ for petrol and diesel, respectively.

2.3 TECHNOLOGY MATURITY

Technology maturity for the technology tracks studied was estimated, as far as possible, on the Technology Readiness Level (TRL) scale used by the European Commission for the Horizon 2020 program [43]. On this scale, TRL 1 is a technology for which only basic principles are known,

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[5] This value for electricity GHG accounting will most likely be the recommended approach in a near future [41] even if the current recommendation is to use Nordic electricity mix with 34.91 g CO_2eq/MJ, see http://www.energimyndigheten.se/fornybart/hallbarhetskriterier/hallbarhetslagen/fragor-och-svar/vaxthusgasberakning/

while TRL 9 is a technology proven to work in commercial operation. The European Commission's definitions of the TRL levels are sometimes quite vague, which is explained by the fact that the definitions are kept general in order to be applicable for a very wide range of technologies. This can make TRL estimation subjective and difficult. In order to increase precision and objectivity, TRL definitions developed by US Department of Energy (DOE) for clean coal technologies were used as a support [44]. These have the advantage that the definitions are specific for energy technologies, and that they include specific descriptions of the level of integration required. They also contain quantitative thresholds for scale, which is not present in the European Commission's definitions, typically at 5 % and 25 % of expected commercial scale for TRL levels 4-7, which are the most relevant levels for the technologies assessed in this work. TRL definitions for both European Commission and US DOE are shown in Appendix D.

The definitions described in Appendix D are in principle valid for a full system, i.e. a number of integrated technologies. However, for a system that includes a mixture of technologies at different stage of development, it can be useful to apply a modification of the definitions added above. All the technology tracks investigated are value chains consisting of several linked technologies with different levels of maturity, sometimes very different. We have used two approaches: a "weighted average" approach and a "weakest link" approach. For both approaches, each technology track was divided into 3-6 individual process steps. Integration with a refinery/pulp mill was considered as a step of its own.

In the *weighted average* approach, each step was given a weight based on its complexity and its importance for the overall chain. Care was taken to assign weights as objectively as possible, but since terms like "complexity" and "importance" are not well defined, it is nevertheless difficult to completely avoid a subjective aspect.

In the *weakest link* approach, the TRL of the "main" process step with the lowest TRL was considered relevant for the maturity assessment. A "main" process step was here defined as a process step given a weight of at least 20 % in the weighted average approach.

In order to fully account for the integration aspects, which are often important and difficult, the TRL estimate of a sub-process was set considering if the feed to the process is "conventional" or new. If a sub-process is a commercially deployed process (TRL 9) but the feed is of a different quality or coming from a process that is not the one used in commercial deployment, the TRL was corrected downwards to account for this. Typical examples include the use of forest residues in pyrolysis processes, which are commercial for stemwood feedstock, and feeding biosyngas to gas cleaning processes, which are commercial for syngas for gasification of fossil feedstocks. A TRL adjustment of 0.5-2 units was used in these cases.

2.4 PRODUCTION POTENTIALS

The resulting mass and energy balances were used to calculate the biomass-to-biofuel conversion efficiency for each production pathway. In combination with estimations from the literature regarding domestic feedstock potentials for forest residues and kraft lignin, respectively, the conversion efficiencies were used to assess the technical production potentials for the investigated technology tracks. For each type of feedstock, an upper and a lower estimation were made regarding the potential, in order to give a range for the overall potentials.

In addition to the feedstock potentials, some key technology constraints were also considered, regarding maximum possible lignin extraction rates, and maximum blend-in potential in the refinery’s FCC. Also for the technology constraints, an upper and a lower estimation were made.

3 TECHNOLOGY TRACKS – DESCRIPTION AND ASSUMPTIONS

As noted in 1.1, the cases investigated represent two main types of production technologies: (i) liquefaction (depolymerisation through pyrolysis or hydrothermal treatment) followed by hydrotreatment upgrading, and (ii) gasification followed by catalytic synthesis. Two types of feedstocks are used (a) Kraft (black liquor) lignin, and (b) forest residue (tops and branches), see Table 1.

As noted in chapter 2, the work in this project started with a broader investigation of technologies that are currently under development, in order to make the final selection of technology tracks for the continued analysis. Our chosen set of production pathways/cases contains one example each of every technology track. This only exception to this is pyrolytic hydro-deoxygenation of forest residues (case 3), which is represented by three distinct pathways as summarized in Table 10. The screening and selection are discussed in detail for each of the cases below in the relevant section. As we wish to capture the state-of-the-art in forest-based biofuel production, each case chosen is either undergoing active development at the time of writing, or has undergone such development in the recent past.

With the exception of one of the two gasification-based cases, all other cases include the integration of production facilities with an existing pulp mill. Similarly, the upgrading to fuel products in all liquefaction-hydrotreatment cases is integrated with existing operations at a crude-oil refinery. The effects of these integrations on material and energy balances are accounted for in the estimations.

Table 10. General overview of the case set.

	Case abbreviation	Description	Feedstock	Production technology	Pulp mill integration	Refinery Integration
1	MSL-HDO	Hydrodeoxygenation of membrane separated lignin	Kraft lignin	Liquefaction-hydrotreatment	✓	✓
2	BLG	Entrained-flow gasification of black liquor	Kraft lignin [black liquor]	Gasification-catalytic synthesis	✓	×
3a	Pyr-HDO	Fast pyrolysis and hydrodeoxygenation	Forest residues	Liquefaction-hydrotreatment	✓	✓
3b	Pyr-FCC	Fast pyrolysis and fluidized catalytic cracking	Forest residues	Liquefaction-hydrotreatment	✓	✓
3c	Hydropyr	Catalytic hydropyrolysis	Forest residues	Liquefaction-hydrotreatment	✓	✓
4	BMG	Fluidized-bed gasification of biomass	Forest residues	Gasification-catalytic synthesis	×	×

3.1 CASE 1 – DEPOLYMERISATION AND UPGRADING OF KRAFT LIGNIN TO DIESEL AND PETROL

This technology track includes separation of kraft lignin from black liquor followed by catalytic depolymerisation and hydroprocessing to give diesel and petrol products. In the process concept evaluated, lignin separation and depolymerisation steps are integrated with a kraft pulp mill, while the hydroprocessing step is integrated in an oil refinery.

Different technologies are under development for all involved process steps. Initially, two separate sub-tracks were selected with different technology options for lignin separation and depolymerisation:

1. Hydrothermal liquefaction (HTL) of membrane-separated lignin
2. Base catalysed depolymerisation of kraft lignin from the LignoBoost process

Based on results from Anheden et al. [32], the depolymerisation process in option 2 was showing poor economic and environmental performance, for which reason option 1 (hydrothermal liquefaction of membrane-separated lignin) was ultimately selected for further evaluation.

3.1.1 Membrane-separated lignin depolymerisation followed by hydrodeoxygenation (case 1 MSL-HDO)

Lignin separation and HTL process

This section contains overall process descriptions and assumptions about technology performance. mainly based on information from SunCarbon about on ongoing development work [45]. Due to the novelty and commercial nature of this technological development there is, to our knowledge, no relevant open literature.

The separation-depolymerisation process has been designed to be integrated with a pulp mill. Weak black liquor is fed to the process, where it is filtered, treated hydrothermally, and purified into depolymerised lignin, which is sent to a petroleum refinery for final hydrogenation. A simplified flow chart depicting the HTL process is shown in Figure 1.

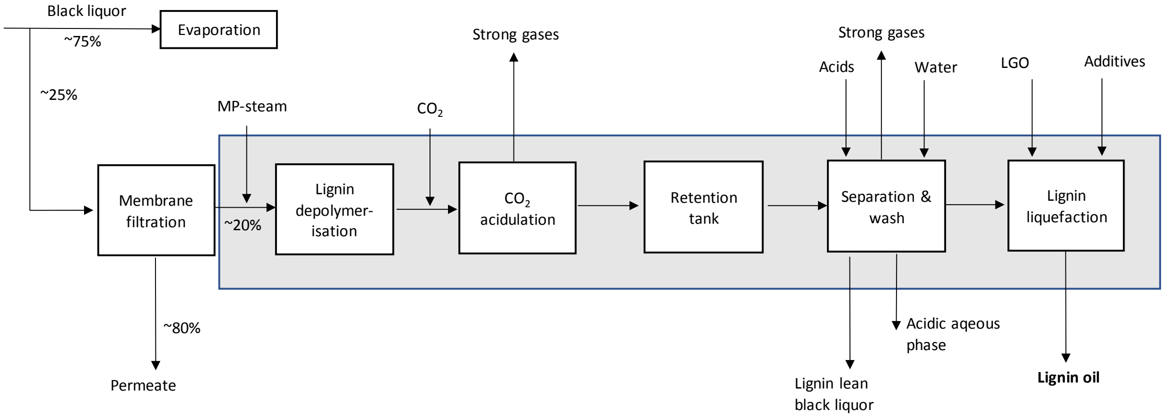

Figure 1. Simplified flow chart depicting the included HTL process, using membrane separated kraft lignin as feedstock. Based on initial design by SunCarbon [32].

In the process, approximately 25 wt% of the mill’s weak black liquor is extracted to a membrane filter. The retentate leaving the membrane filter contains 15-25 % lignin by weight and less than 10 % black liquor alkali. The permeate from the membrane can potentially be recycled to the green liquor preparation stage of the mill However, this has not been verified, and in this study permeate has been assumed to be recycled to the evaporation plant. The lignin rich retentate is mixed with medium pressure steam (23-56 bar) from the mill and fed into a thermal catalytic cracker at sub-critical conditions for depolymerisation. A catalyst internal to the pulp and paper mill is used in the catalytic cracker. The hydrothermally treated lignin mixture is then mixed with low-purity CO_2, which lowers the pH. After spending some time in a retention tank, the lignin mixture is subjected to an acid wash (H_2SO_4) which separates inorganic ash components, residues, lean black liquors and strong gases (mainly H_2S) from the depolymerised lignin fraction. The removed lean black liquor contains around 30 % of the energy content of the initial lignin entering the process, which results in an energy yield of 70 % of the purified depolymerised lignin.

Table 11 below shows heating values of main flows in the thermal liquefaction of membrane separated lignin. The heating values were estimated based on simulations results from the reference mill model and discussions with SunCarbon.

Table 11 Heating values for thermal liquefaction of membrane separated lignin.

	MJ/kg dry, HHV
Weak black liquor	15
Retentate	20
Lignin oil	28

Lastly, the remaining purified depolymerised lignin is suspended in gas oil or tall oil pitch, assisted by additives, which gives a resulting lignin oil product that is ready to transport to a petroleum refinery for hydrogenation. In this project, however, 100 % lignin oil was assumed as feedstock for the hydrogenation process.

Pulp mill integration

For a specific pulp mill, the integration of the HTL process on membrane separated kraft lignin will render different results depending on the mill’s energy situation. If the mill has no energy surplus, energy saving measures or additional biomass may be needed to compensate for the loss of energy due to the lignin removal. If the mill’s recovery boiler constitutes a production bottle neck, the removal of black liquor can support increased pulp production, in a similar fashion as a LignoBoost process. Finally, if the mill has an energy surplus that is used for power production in a condensing turbine, the power production can be reduced to support the energy balance. As noted in 2.1.1, the last situation was adopted as the basis for the calculations. No additional benefit for increased pulp production was considered in the evaluation.

Further, if the mill’s original operation is already in balance with respect to Na/S, the additional sulphur resulting from the addition of H_2SO_4 the membrane separation and HTL processes, might upset the Na/S balance. In this case, Na_2SO_4 will have to be bled off with the electrostatic precipitator ash and compensated by virgin NaOH makeup. There is also an uncertainty in the estimation of how much alkali that is lost with the retentate from the membrane separation, which in turn is connected to the need of NaOH make-up chemicals in the pulping process. We have accounted for the extra NaOH make-up connected to the Na/S balance and a small the loss via the lignin oil. It is assumed that most of the alkali in the retentate is recovered, since any presence of alkali in the lignin oil has a negative impact on the expected catalyst lifetime in the downstream refinery and upgrading process steps.

Upgrading of depolymerised lignin and refinery integration

Figure 2 shows both the upstream lignin separation and depolymerisation step and the integration with the pulp mill and the downstream refining process steps of the depolymerised lignin in the oil refinery. In the refining process, the depolymerised and purified lignin is upgraded to meet final transportation fuel product specifications and standards for diesel and petrol using catalytic upgrading processes, primarily hydrodeoxygenation (HDO).

The need of processing is based on the structure and composition of the bio-oil. Bio-oils containing oxygen are hydrodeoxygenated, normally in two steps, to reach close to zero oxygen content. H_2 is used as reactant together with a solid catalyst to ensure a high oxygen removal rate and efficient utilisation of the hydrogen. Bio-oil hydrotreating means processing the bio-oil with a large excess of H_2, which results in a gas and two liquid fractions. The compounds in the gas product are excess H_2, light hydrocarbons and CO_2. The excess H_2 is assumed to be recovered using pressure swing absorption, and recycled. The liquid products consist of hydrocarbon oil and an aqueous phase, which are easily separated. The aqueous phase carbon content depends upon the degree of bio-oil deoxygenation.

The hydrocarbon oil yield and quality are dependent on the choice of catalyst, the reactor configuration and the hydrotreating conditions. Fixed bed reactors similar to those used in conventional hydroprocessing of petroleum products to final fuels can be used. Chemical reactions during bio-oil hydrotreating are very complex. Overall, the bio-oil is almost completely deoxygenated by a combination of hydrodeoxygenation and decarboxylation, with oxygen removed in the form of water and carbon oxides. A good combination of hydrotreating catalysts and optimum operating conditions can help control reaction selectivity and minimise H_2 consumption.

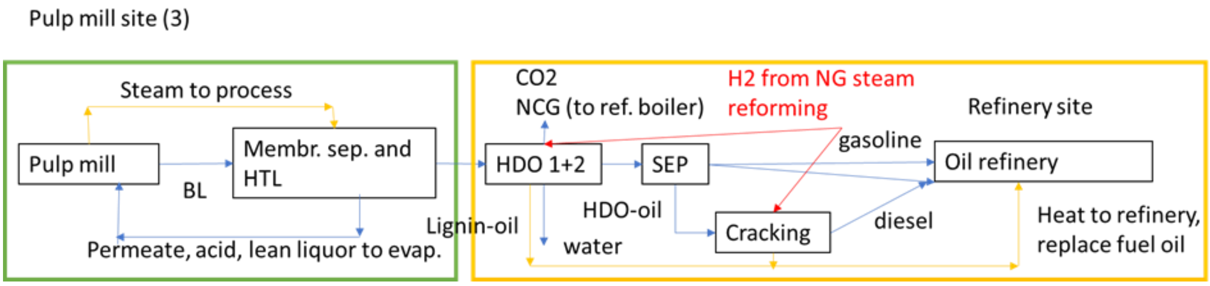

Figure 2 Integration of the value chain with HTL of kraft lignin retentate from membrane separation, followed by deoxygenation (MSL-HDO).

The HDO-product is then distilled in a petrol fraction, a diesel fraction, and in some cases also a heavy fraction. The heavy fraction is further cracked to a diesel fraction, utilising the existing refinery hydrocracker. The goal of the cracking process is to reach hydrocarbon chain-lengths in the diesel range, e.g. about C18, under addition of H_2.

The knowledge base on HDO and upgrading of bio-oils from kraft lignin depolymerisation is so far very limited. Work is currently ongoing in the BioInnovation project BioLi2.0[6], but no results are yet public. The closest experimental data available is the data provided by Löfstedt et al. [46], as well as the associated supporting documentation[7]. We assume that the HTL depolymerised lignin considered in this report, behaves similarly during HDO and upgrading, as the lignin in run 3 in Löfstedt et al. [46]. It should be noted that since a fundamentally different depolymerisation process was considered (based on mild Ni-catalysed transfer hydrogenation using 2-propanol), this is a very uncertain assumption. In addition, the depolymerised lignin in the process described in [46] is esterified with tall oil fatty acid (TOFA) anhydride, in order to make it soluble in light gas oil

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[6] Financed by Vinnova 2016-2019, see http://www.bioinnovation.se/projekt/bioli2-0-fran-lignin-till-biobaserade-drivmedel-och-kemikalier/

[7] This data concerns the process under development by RenFuel AB.

(LGO). The esterified depolymerised lignin was thus co-processed with LGO during hydrotreatment. In the calculations in this study, the no lignin reference run (run 1) in [46] was used to calculate the H_2 consumption related to the tall oil (0.016 g H_2/g TOFA), assuming that the LGO does not consume any H_2. Based on this information it was estimated from run 3 that the H_2 consumption for the depolymerised lignin is 0.05 g H_2/g lignin fragments. Based on the combined information from run 1 and 3, the distribution of products (light gases, fuel components, heavy components, CO_2 and H_2O) was then calculated.

Next, data on the elemental composition of the depolymerised lignin was used as input. The same oxygen reduction as reported in [46] was used as input in the calculations for the first HDO step, and a stoichiometric H_2 consumption assuming direct reduction of oxygen in the oil to H_2O. The H_2 consumption in the second HDO step was calculated assuming the same specific H_2 consumption, resulting in a H_2 consumption of 0.045 kg H_2/kg depolymerised lignin feed.

The yield from the depolymerised lignin to the different fractions from the fractionation column was calculated from run 3 [46]. In this calculation it was assumed that LGO only contributes to the petrol, diesel and heavy fractions, and that the TOFA and LGO contributes to petrol, diesel and heavy fraction to the same extent as in run 1. In addition, it is assumed that the TOFA-contribution to light gases is the same as in run 1. Based on these assumptions, it is possible to calculate the contribution to the different phases from the depolymerised lignin based on the reported overall distribution of phases in run 3 as a difference between the total distribution of a component, minus the distribution from LGO and TOFA. The same normalised percentage of contribution to the different distillation phases from the products from the second HDO step is finally applied to calculate the production rate of CO_2, light gases, petrol, diesel, heavy fraction and water for the MSL-HDO process.

The H/C-ratio of the produced diesel was assumed to follow [46] (Table S4), which means that with an H/C ration of 2.1, no further treatment is required. Regarding the heavy fraction it was assumed that C36 is cracked to C18, requiring 1 mole of H_2 per mole of C36. Finally, the heat from the HDO process that could be used to replace fuel oil in the refinery was calculated based on an energy balance over the two HDO steps. Also non-condensable gases (NCG) formed in the process can be used for energy integration. Table 12 shows the resulting main data, as used for the further evaluations in this study.

Table 12. Main data related to HDO and further upgrading of depolymerised lignin.

Parameter	Value	Unit
H_2 consumption	0.065	kg/kg depolymerised lignin
Excess heat to recovery	5.9	MJ/kg depolymerised lignin
NCG to recovery	5.7	MJ/kg depolymerised lignin
Diesel	0.47	kg/kg depolymerised lignin
Petrol	0.10	kg/kg depolymerised lignin

3.2 CASE 2 – BLACK LIQUOR GASIFICATION BASED METHANOL PRODUCTION

This technology track describes the production of methanol from pressurized oxygen-blown entrained-flow gasification of kraft black liquor. The whole process is fully integrated with the reference kraft pulp mill that is presented in section 2.1.1. However, there is no integration with the refinery as the final product, grade AA methanol, is an automotive fuel that can be used in suitable vehicles.

The gasification of black liquor was first proposed in the 1960s. Several designs have been developed in the intervening decades; however, none of them have seen large-scale commercial implementation. The drivers behind the development have also undergone a shift, from a desire for increased safety in the 60s and 70s to power generation in the 80s and 90s, and more recently, to the production of green transportation fuels [47]. It is for this latter purpose that Chemrec AB developed an oxygen-blown entrained-flow variant, which was demonstrated in a 3 MW_{th} pilot facility for more than 28 000 hours between 2005 and 2016 [6], latterly under the custodianship of Luleå University of Technology, where research and development work on catalytic gasification is ongoing. As this design represents the only advanced and demonstrably technologically viable black liquor gasification alternative for the production of renewable transportation fuels such as methanol, it was chosen to represent the technology track with the label BLG.

Mass and energy balances for the techno-economic analysis were taken from the black liquor case scenario in a study by some of the authors of this report that was recently submitted for publication.[48] The study built on work undertaken in the “Catalytic Gasification” project in the LTU Biosyngas Program 2013-2016. A description of the concept as modelled in the source study is provided below, which is followed by an overview of steam and electricity balances in the methanol plant at the described pulp mill. The major modelling parameters given in the study are summarized in individual sections covering all the key conversion steps. Readers are referred to Chapter 2 in the source study for additional details on system modelling.

3.2.1 Entrained-flow gasification of black liquor followed by methanol synthesis (case 2 BLG)

Process description

Kraft black liquor is an alkaline mixture of pulping chemical, polysaccharides and dissolved lignin in approximately equal amounts (on a mass basis). In modern pulp mills, it is combusted in recovery boilers to produce heat, generate electricity, and most importantly, recover the pulping chemicals for re-use. In comparison, the gasification of black liquor produces a gaseous mixture that is composed primarily of CO, H_2, H_2O and CO_2 as the major components. Small quantities of CH_4, around 1 % on a molar basis, and even smaller amounts of tars (in the ppm range) are also present [4,5,49,50]. A molten slag made up of ash that contains the pulping chemicals is produced as a by-product, and this is returned to the pulp mill in a dissolved form known as green liquor. After exiting the gasifier, the gaseous mixture, known as raw syngas, is cooled to 40 °C in a counter-current condenser that doubles as an efficient particle scrubber. Oxygen for gasification comes from an air separation (ASU) unit.

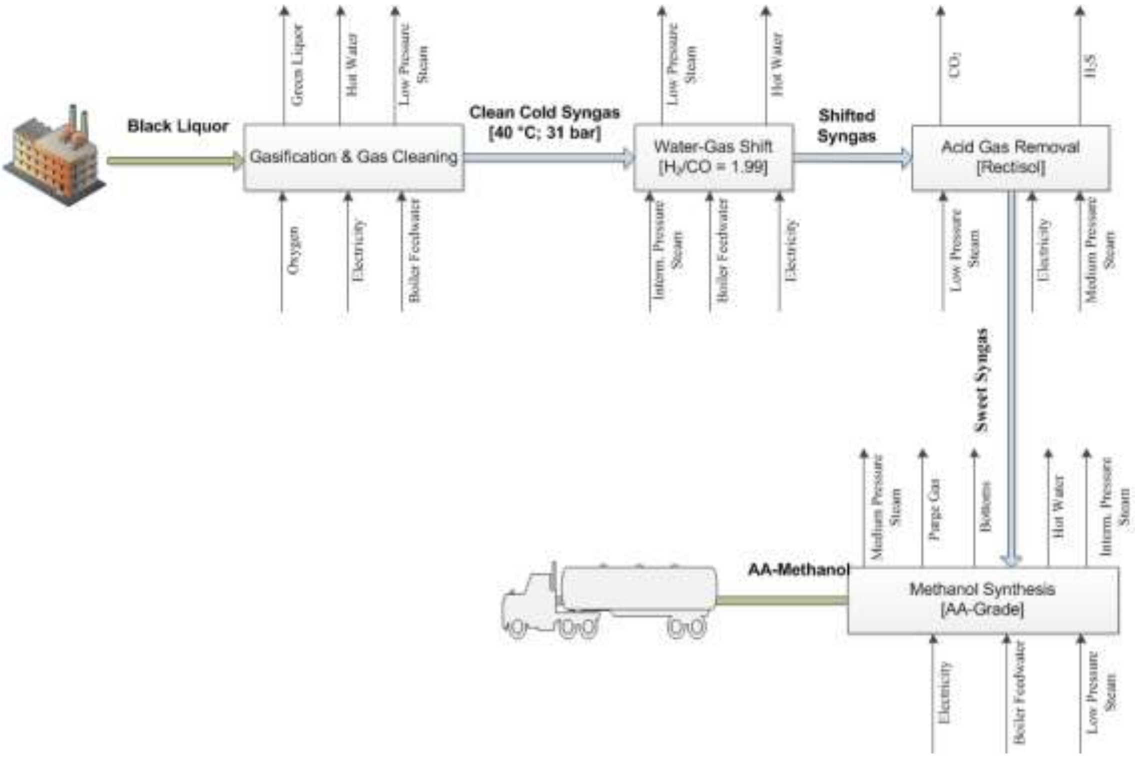

Figure 3 Schematic description of the modeled BLG case.

The cooled raw syngas undergoes conditioning in the water-gas shift (WGS) and acid gas removal (AGR) units prior to methanol synthesis. A schematic presentation of the different processing steps and the media streams crossing them is given on Figure 3. Detailed descriptions of the gasification and syngas upgrading setups are available in the literature.[4,9]

Pulp mill integration

The reference pulp mill was modelled with a capacity of 700 kADt kraft pulp per year. The feedstock used was a fifty-fifty split of spruce and pine softwood. More information is provided in section 2.1.1. The integration between the methanol plant and the pulp mill took the form of materials, namely, black liquor and green liquor, as well as energy exchange. Increased lime reburning demand is a known effect of using BL gasification and a 25 % increase in fuel demand in the lime kiln was included in the energy balances.

Excess heat from different parts of the process was used to provide hot water at 95 °C/1 bar and saturated steam at three levels: low pressure steam (LPS) at 150 °C/4.5 bar, medium pressure steam (MPS) at 200 °C/10 bar and intermediate pressure steam (IPS) at 250 °C and 30 bar. The overview of steam distribution that is presented in Figure 4 shows both the sources and the uses of steam in the methanol plant. Surplus LPS steam was sent over to the pulp mill in amounts that were enough to cover 60 % of its LPS steam requirement. Purge gas from the methanol synthesis unit was also sent to the pulp mill where it replaced the bark combusted in the lime kiln. This bark was instead combusted with additional biomass in an expanded biomass boiler to produce high pressure process steam for the pulp mill. This was expanded into other steam levels (IPS, MPS and LPS) in a back pressure turbine, which was simulated in Aspen Plus and validated with steam and electricity data for the reference pulp mill (see Chapter 2 in [48]). Some of the MPS from the back pressure turbine was also sent to the AGR and the methanol units to cover the shortfall. Isentropic efficiencies of 89 %, 85 % and 71 % were used to model turbines generating IPS, MPS and LPS, respectively for the pulp mill. The generator was assumed to have an efficiency of 98 % and the shortfall in the electricity balance was covered by purchasing electricity from the grid. See Figure 4b in [48] for more details on the integrated electricity and steam balance).

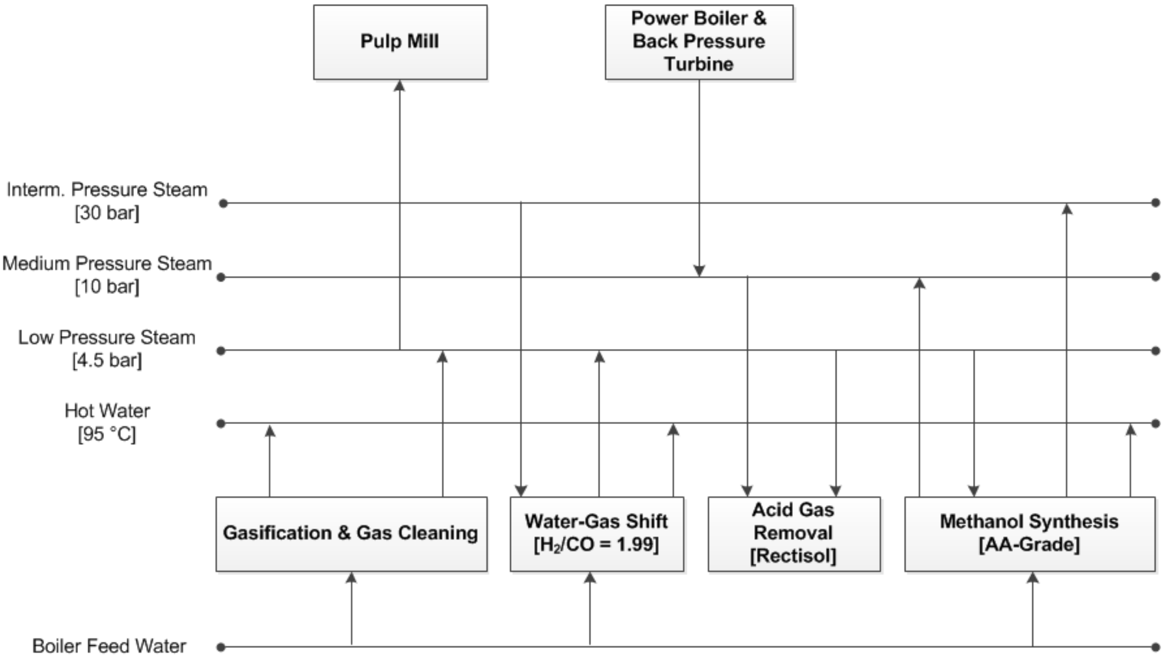

Figure 4 Steam distribution in the methanol plant for the BLG case.

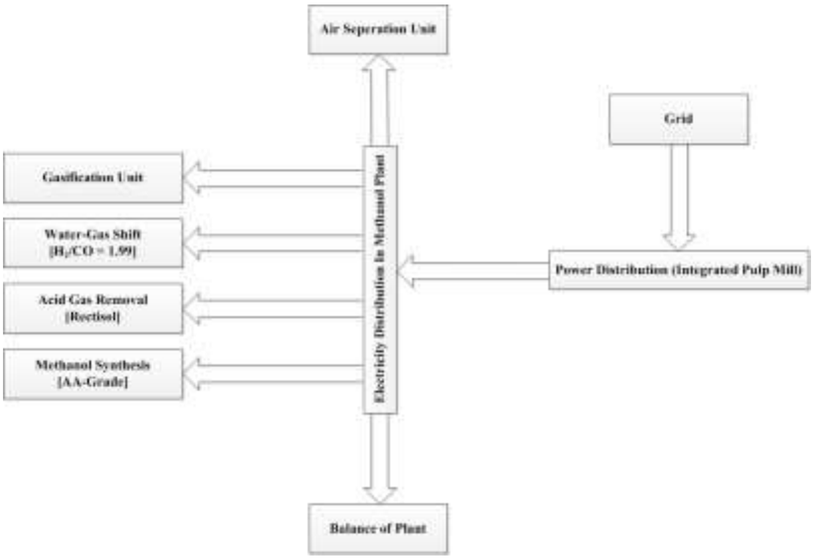

Figure 5 Electricity distribution in the methanol plant for the BLG case.

Gasification and syngas conditioning

The gasifier was modelled as a refractory-lined oxygen-blown entrained-flow reactor in SIMGAS, which is a thermodynamic equilibrium model that was developed to simulate BLG. The model has been validated with pilot experimental data [51]. It uses a global equilibrium-approach to calculate equilibrium composition of the major syngas components and the molten slag, which contains the inorganic pulping chemicals. The concentration of CH_4 was assumed to be 1 mol% in raw syngas, while H_2S was modelled by an empirical modification to the thermodynamic model. Reactor temperature was set at 1050 °C based on data from pilot plant operation. Molar fraction of syngas components in the hot gas at gasifier exit are provided in**Fel! Hittar inte referenskälla.** Table 13, for details see Carvalho et al [48,52]. The simulate scale was 509 MW fuel input on a HHV basis but evaluation of cases (chapter 4) is done using generic balances without considering a specific scale. Hence, balances calculated are assumed to be representative also when scaled up or down.

Syngas upgrading and methanol synthesis

These process steps were simulated in ASPEN Plus. In the first step, hot syngas from the gasifier was cooled to a temperature of 40 °C and the recovered heat was used to generate LPS. The cold syngas was fed to the water-gas shift reactor (WGS). This unit was simulated as an equilibrium reactor and a ratio of 1.95 between hydrogen and carbon dioxide was obtained by adding HPS. The optimal ratio for methanol production is thought to be in the vicinity of 2 [53]. The shifted syngas was sent through a series of heat exchangers to generate IPS, LPS and hot water before entering the water gas shift unit. The purpose of this unit, the modelling of which is based on the widely used Rectisol system, is to remove the acid gases, namely H_2S/COS and CO_2. The outgoing syngas, which is also known as sweet syngas had molar CO_2 and H_2S concentration of 2.9 % and 0.1 ppmv, respectively. The synthesis of grade AA-methanol from the upgrading of sweet gas at 225 °C, which was compressed to a pressure of 90 bar(a) was the final step in the process. Synthesis reactions for methanol are given in Eqs. 1 and 2.

$CO + 2H_2 \leftarrow\rightarrow CH_3OH$ (Eq.1)

$CO_2 + 3H_2 \leftarrow\rightarrow CH_3OH + H_2O$ (Eq.2)

The synthesis reactor was simulated as an equilibrium reactor with a temperature approach of 30 °C. Upon leaving the reactor, the synthesized gas also contains unconverted syngas and inert compounds such as methane and nitrogen besides methanol and water. Thus, it is recycled back to the reactor after being fed through a two-stage separator and purged to avoid the build-up of inert compounds in the loop. The resulting purge gas is sent to the pulp mill lime kiln for use as fuel. The strongly exothermic synthesis reactions release large amounts of heat, which is used to produce IPS and hot water for the facility. Finally, the crude methanol, which contains some water and small amounts of unconverted residues, is purified to a quality of greater than 99.85 % to produce grade AA methanol.

Summary balance data for the biofuels plant in case 2 BLG is presented in Table 14. It does not include mill integration results, which are presented in chapter 4. Energy yield are presented using BL heating values including and excluding sulfur oxidation. The reason for the latter is that sulfur oxidation energy is not available in a BLG process, since sulfur needs to be returned to the pulp mill reduced state to facilitate pulping chemical recovery,

Table 13 Molar fractions of syngas components.

		Raw SG	Shift SG	Sweet SG	AA MeOH	Purge Gas
Temperature	°C	40.0	36.5	26.0	48.0	50.1
Pressure	bar	31.0	24.2	22.0	5.0	4.0
Flow	kg/hr	119023	133447	44539	39975	2285
Energy	MW HHV	319	320	314	252	17.2
CO	mol/mol	0.378	0.195	0.294		0.119
H2	mol/mol	0.350	0.437	0.660		0.264
CH4	mol/mol	0.010	0.009	0.013		0.401
CO2	mol/mol	0.249	0.349	0.029		0.084
N2	mol/mol	0.003	0.002	0.004		0.109
H2S	mol/mol	0.008	0.007			0.000
H2O	mol/mol	0.002	0.002			0.000
Methanol	mol/mol				1.000	0.021

Table 14. Summary balance data for the biofuels plant in case 2 BLG, i.e. without mill integration.

Cold gas efficiency	62.7 %	MJ/MJ (HHV basis)
	71.4 %	MJ/MJ (S free HHV basis)
BL to methanol efficiency	49.4 %	MJ/MJ (HHV basis)
	56.3 %	MJ/MJ (S free HHV basis)
LP steam net	0.073	kg/MJ BL (HHV basis)
MP Steam net	-0.029	kg/MJ BL (HHV basis)
IP Steam net	0.0065	kg/MJ BL (HHV basis)

3.3 CASE 3 – PYROLYSIS OF FOREST RESIDUES AND UPGRADING TO DIESEL AND PETROL

Overview of pyrolysis concepts and technologies

Pyrolysis is a collection of thermochemical processes in which organic compounds are broken down into smaller ones due to heating to high temperatures, in the absence of oxygen. Pyrolysis of biomass results in the products char and vapour. A part of the vapour condensates to liquor, while the rest consists of non-condensable gases (NCG). The distribution of char, NCG and liquor depends on the temperature and residence time.

Slow pyrolysis has a long residence time (minutes to hours) and generally gives low liquid yields. This type of process is typically mainly focused on the production of solids (biochar), and is thus not discussed further here. *Fast pyrolysis* instead aims at high yields of liquid products, and is completed in seconds. This requires very high heating and heat transfer rates (fast heating and cooling) which in turn requires the biomass to be milled to particles on the millimetre scale [54,55]. Fast pyrolysis can be accomplished in different types of reactors of which the most important today are bubbling fluidised beds (BFB), circulating fluidised beds (CFB) and rotating cone reactors.

Thermal fast pyrolysis of biomass is becoming a commercially established process with two plants recently being started in Europe. The Fortum owned 50 kt/y plant in Joensuu, Finland, was designed by Valmet and commissioned in 2013[8]. The Empyro plant, using a BTG-BTL rotating cone design, is marginally smaller and was commissioned in 2015[9]. Both plant runs on low ash stemwood feedstock and have not, to our knowledge, established regular operation on residue based lignocellulosic feedstock such as forest residues.

Fast pyrolysis oils (FPOs) generally contain a large amount of oxygen and acidic components that are not compatible with a hydrocarbon based transportation fuels infrastructure. In order to be able to use FPOs as transportation fuels they thus need upgrading. It is also possible to modify the pyrolysis process, in order to obtain products that contain less oxygen/acids. Examples of modifications are the use of catalysts in the pyrolysis process (in situ), upgrading of the pyrolysis vapours without intermediate condensation (vapour phase upgrading or ex situ catalysis), or the use of H_2 in the pyrolysis process instead of nitrogen, which is called hydropyrolysis. In this report, the liquid product from catalytic pyrolysis is denoted catalytic pyrolysis oil (CPO). RTI International [56] presented technology that integrates catalytic biomass pyrolysis and hydroprocessing to produce drop-in biofuels. The pyrolysis catalyst used is specified as “non-zeolite solid acid”. Catalytic pyrolysis experiments at 525 °C were done in 1 t/d scale with 30 h duration, obtaining a fairly stable operation, which resulted in a yield on mass basis of 9 % of CPO (separated into a heavy and a light fraction) from saw dust. The aqueous liquid product yield was 11 %, while 25 % ended up as char and ash. Higher yields were obtained at 465 °C (17 % organic, 25 % aqueous) but with less stable operation, during only 13 h. The hydroprocessing work done in conjunction with this work has not been found in the open literature.

Research work at Aalto University and VTT [57] investigated catalytic pyrolysis and showed that coke-induced reversible catalyst deactivation is problematic in BFB set-ups, but that it can be effectively handled if a CFB set-up is instead used, as shown by a 96 hour pilot-scale experiment [58]. However, catalyst long-term stability, a key question for catalytic pyrolysis, was found problematic. The combination of biomass-derived inorganic contaminants and severe reaction/regeneration conditions cause irreversible changes in the catalyst structure and properties, which influenced the quality of the bio-oil.

Another catalytic pyrolysis development worth mentioning is the technology of KiOR Inc. KiOR failed to manage to solve technical problems in their demonstration plant, and had to file for reconstruction. A part of the old company is now active under the name Inaeris Technologies. The “rise and fall” of KiOR has recently been described in Biofuels Digest [59].

Researchers at Ghent University [60] compared in situ and ex situ catalytic pyrolysis with special emphasis on catalyst deactivation and effects from biomass ash. They concluded that the use of catalytic pyrolysis (either type) significantly decreased the overall liquid yield compared to thermal pyrolysis, and with significantly changed composition. Biomass ash had several negative effects, including increased gas and char yields, as well as poisoning of the added catalyst. Ex situ catalysis was found to consume less catalyst, give lower gas yields, and have better options for mitigating

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[8] See https://www.fortum.com/en/mediaroom/pages/fortums-bio-oil-plant-commissioned-in-joensuu-first-of-its-kind-in-the-world.aspx

[9] See http://www.empyroproject.eu/ and https://www.btg-btl.com/en/company/projects/empyro

negative effects from ash; and was thus concluded to be a techno-economically more attractive option.

The IH2 (integrated hydropyrolysis and hydroconversion) process is different from other pyrolysis processes, since pyrolysis is carried out in an H_2 environment. It is the currently most technologically mature catalysed pyrolysis process [61–65]. This technology is described in more detail below (section 3.3.3).

Paasikallio [57] gives basic alternatives for pyrolysis based conversion of lignocellulosic biomass to hydrocarbon biofuels. For both FPO and CPO, it is in principle possible to use either catalytic hydrotreating, or FCC co-feeding. A main distinction is that the catalytic hydrotreating option has the potential to provide fuel components that originate only from the biomass, while the FCC option leads to the unavoidable blending of bio-based and fossil components in the FCC products. From a legislative and tax perspective, this distinction can be very important.

In the upgrading of FPO/CPO to hydrocarbon fuels, three individual processes can be said to be involved: 1) oxygen (and sulphur) removal, 2) cracking of heavy compounds into lighter fuel range products, and 3) adjustment of saturation/aromaticity in order to meet petrol and diesel quality requirements. It should be noted, however, that these three steps are typically in practice not separated, but often occur to some extent in parallel in the same processing equipment.

It is widely accepted that removal of particles from FPO is beneficial for FPO upgrading, for example due to particle bound alkali metals that deactivate catalysts. As sufficient separation is typically not obtained in the cyclone separator of the pyrolysis unit, filtration or other measures for removing particles from FPO may be necessary.

A fractionation of the pyrolysis oil prior to catalytic upgrading can be beneficial. The most common fractionation is separation into a water phase and a more hydrophobic phase, of which the former typically contains low molecular weight organic compounds, while the latter typically contains mostly pyrolytic lignin. Hence, processing only the organic/lignin fraction can potentially avoid processing the light acids/ketones/aldehydes of the aqueous phase, that otherwise consumes large amounts of H_2 but do not contribute to the yields of liquid fuel range products. Oil/water phase separation has been shown to be an advantage for upgrading [66]. If the intention is to upgrade both phases to hydrocarbons that is, however, not necessarily true from an overall perspective [67]. Early studies [68] emphasized fractionation before FPO upgrading, while more recent research mostly concerns fractionation of whole FPO or CPO.

Selection of cases for further evaluation

Based on the literature, initial screening, as presented above, two different variants of pyrolysis concepts were selected for this study: thermal pyrolysis, as demonstrated by Valmet/Fortum and BTG-BTL/Empyro (case 3a and 3b), and the Shell/CRI IH2 hydropyrolysis process (case 3c). Due to the uncertainty regarding technological advantages as well as the current development status, catalytic pyrolysis was excluded from the further analysis. Regarding FPO upgrading, two different options for were considered: HDO (hydrodeoxygenation) (case 3a) and FCC co-processing (case 3b). For the hydropyrolysis process (case 3c), only product finishing is required.

3.3.1 *Fast pyrolysis followed by hydrodeoxygenation (case 3a Pyr-HDO)*

In this case, fast pyrolysis is followed by fuel upgrading by deoxygenation, as shown schematically in Figure 6. The pyrolysis process is considered for integration with a pulp mill, while the upgrading is integrated with an oil refinery. This section contains overall process descriptions.

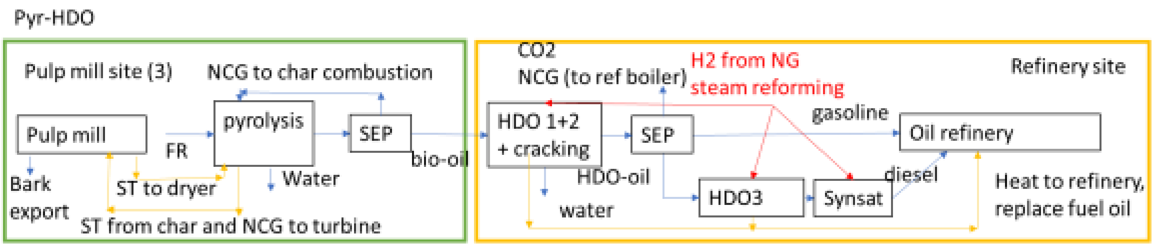

Figure 6. Overview of the process in case 3a – fast pyrolysis of forest residues followed by hydrodeoxygenation (Pyr-HDO).

Pyrolysis process and pulp mill integration

The selected fast pyrolysis process is based on a reactor with rotating cone delivered by BTG-BTL, as described by Benjaminsson et al. [69] and as shown in Figure 7. In the BTG-BTL pyrolysis process, the biomass (forest residues) is ground to a particle size of max 6 mm before being fed into a belt dryer for drying. The forest residues enter the dryer with a moisture content of 50 % and a temperature of 25 ˚C, and are dried to a moisture content 6 %. The dryer uses internally produced steam, with a heat consumption of 1.1 MWh/ton evaporated water. From the dryer, the biomass enters the rotating cone pyrolysis reactor where sand is used as heat carrier. The biomass is disintegrated and forms vapour and solids. The solids (char, 32 MJ/kg) exit the reactor with the sand and are combusted in a char burner. Larger particles are separated from the vapours in two cyclones, before being subjected to condensation. The NCGs (11 MJ/kg) are combusted in a burner for steam production. The condensed vapour (i.e. the liquid which constitutes the desired pyrolysis oil product), is filtered before storage and transport for further upgrading. Steam from char and NCG is produced with an efficiency of 87.5%. Part of the steam is consumed in the dryer, while the rest goes to the pulp mill's steam turbine to generate power.

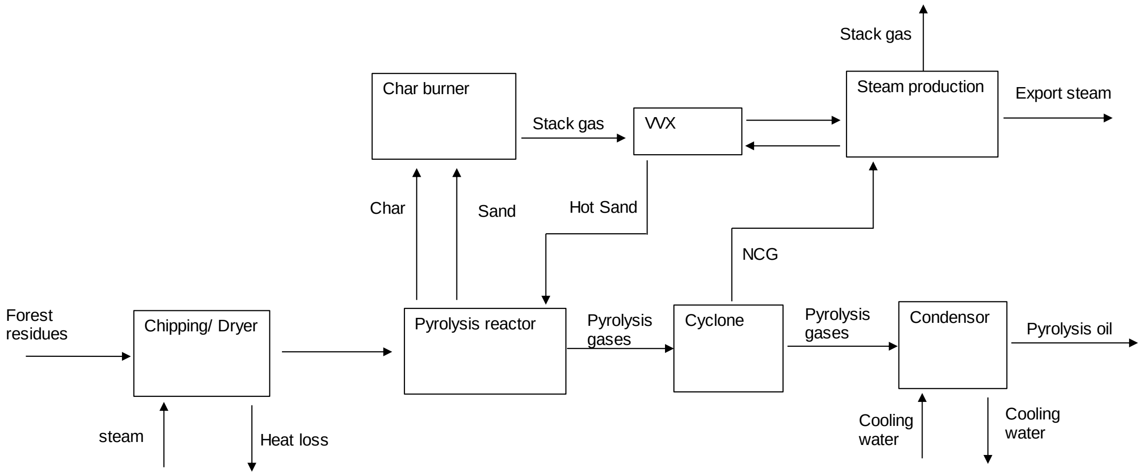

Figure 7. Illustration of the fast pyrolysis process by BTG-BTL [69].

Forest residues contain higher amounts of ash compared to e.g. wood chips or sawmill dust. As ash has been shown to have a negative effect on the pyrolysis oil yield, in favour of more char and NCG, a lower pyrolysis oil yield can be expected when forest residues is used as feedstock. In this study, an energy yield of 54 % from forest residues has been used [69].

Fast pyrolysis oil upgrading by hydrotreatment and refinery integration

Hydrotreatment typically involves catalytic heteroatom removal under high H_2 pressures. For FPO the heteroatom to be removed is mainly oxygen at high concentrations, while fossil feedstocks typically contain nitrogen and sulphur at much lower concentrations. The catalyst manufacturer WR Grace lists the following concerns related to FPO hydroprocessing, of which hydrotreatment is a part, compared to hydroprocessing crude oil derived materials[10]:

- FPO is typically very hydrogen deficient
- The high average molecular weight can lead to condensation and coking on the catalyst surface and pressure drop increases.
- There are often miscibility concerns when co-processing with standard vacuum gas oil (VGO) material. Phase separation could lead to unintended coking, including plugging of feeding equipment.
- FPOs often have extremely high metals levels, and may need to be pre-treated to remove those metals.

It is widely acknowledged that it is advantageous to do FPO hydroprocessing/deoxygenation in two steps. A first, low-temperature, step removes some oxygen but mainly stabilizes the oil by removing the most reactive groups, which suppresses polymerization/char formation in the subsequent step(s) [30,66,70].

Hydrotreating catalysts that have shown the most promise for FPO include traditional hydrotreating catalysts (sulfided NiMo and CoMo) as well as Ru, Pt or Pd based catalysts [66,71]. It has been suggested that using Ru in a first stage to avoid coke formation and NiMo in a second stage to improve deoxygenation is a good solution [72] in agreement with the recent PNNL study for US DOE about the state of technology for fast pyrolysis and hydrotreating [30].

The products from two-stage catalytic FPO upgrading still contain a significant amount of products heavier than diesel. In addition, diesel and petrol fractions typically do not meet specifications and quality demands for petrol and diesel [71]. Hydrocracking is a technology that could crack heavy hydrocarbons to give petrol/diesel range products and at the same time remove phenolics [30,73].

A typical biomass-to-fuel carbon efficiency is 30-50 % for this technology track, with recent technology development leading to numbers in the upper end of this range [30]. The H_2 requirement for hydroconversion of fast pyrolysis oils to transportation fuels is dependent on the route taken and if oxygen is removed only as water, or also as CO_2/CO. Marker et al. [65] states that hydroconversion

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[10] https://grace.com/catalysts-and-fuels/en-us/Documents/112-Predict%20FCCU%20Performance%20with%20Laboratory%20Testing.pdf

of fast pyrolysis oils into petrol and diesel typically requires the addition of 3–5 wt% H_2. The BIO-COUP project states that "deep deoxygenation" requires 800 NL/kg FPO (normal litres per kg), which translates to 7 wt% H_2 [70].

In our calculations regarding HDO of pyrolysis oil, the analysis has included use of data from [74], assuming the exact same composition of the pyrolysis oil as assumed in this report. Based on the decomposed data from the report, we have estimated the H_2 demand in HDO step 1 and 2 (see Figure 6) from the stream data, as well as additional H_2 used in step 3 to finish the heavy fraction. The distribution of components in the stream of non-condensable gases from the HDO and cracking treatment has also been collected from the same report. A rough energy balance over these process steps has been estimated based on stream flow data and heating values of the components.

The reported final diesel fraction in Jones et al [74] does not meet the specified requirements of zero oxygen content and an H/C ratio of 2. An evaluation was made of the C, H, O, N, S content of the reported diesel fraction (stream 326 in the report), which resulted in the conclusion that this stream contains 0.92 wt% oxygen and had an H/C-ratio of 1.6. Based on this, a third HDO step was added (Figure 6) and the theoretical H_2 demand was calculated based on simple stoichiometry for oxygen removal and the reactions in the SynSat treatment. It should be noted that in a real process design, it is very likely that full deoxygenation would be strived for in the initial HDO, making this approach superfluous but given the data source used here [74] it is required.

These estimations resulted in a final estimated H_2 demand, which was used for the further evaluations in this study (Table 15). In addition, 0.25 kg diesel/kg dry FPO was estimated to have to go through the SynSat treatment, and consequently taking away capacity from the fossil production. 3.4 MJ/kg dry FPO of process heat was estimated to be recovered to replace use of fuel oil for heating purposes, as well as 9.0 MJ/kg dry FPO of non-condensable gases.

Table 15. Hydrogen demand for upgrading of pyrolysis oil (own calculations based on [74].

Hydrogen demand	**kg H_2/kg dry FPO**
H_2 for HDO	0.057614
H_2 for cracking	0.008551
H_2 for HDO3, diesel	0.000573
H_2 for SynSat, diesel	0.006758
Total H_2	**0.073496**

3.3.2 Fast pyrolysis followed by fluidised catalytic cracker co-feeding (case 3b Pyr-FCC)

Pyrolysis process

The case with fast pyrolysis followed by fluidised catalytic cracker (FCC) co-processing is based on the same pyrolysis process as presented in section 3.3.1 (Figure 7).

Fast pyrolysis oil upgrading by FCC co-processing and refinery integration

An overview of the fast pyrolysis process with FCC co-processing is seen in Figure 8. The FCC process typically uses a zeolite based catalyst at 500-550 °C to crack the feedstock and it can reduce the oxygen content in FPO without the addition of H_2. Coke formation is reported to be a general problem in FPO FCC [71]. It has been noted in several studies that a stabilisation of FPO before feeding to an FCC unit prevents coke formation by removing the most reactive functional groups in the FPO [70,71,75]. Shell, as a part of the BICOUP project, performed FCC co-processing of up to 20 % (on mass basis) “upgraded” bio-oils (oxygen content in the 10-30 % range, unspecified) with VGO in lab scale. In an FCC riser pilot plant, 20 % co-feeding was found to lead to technical issues, while 5 % co-feeding was technically feasible [70].

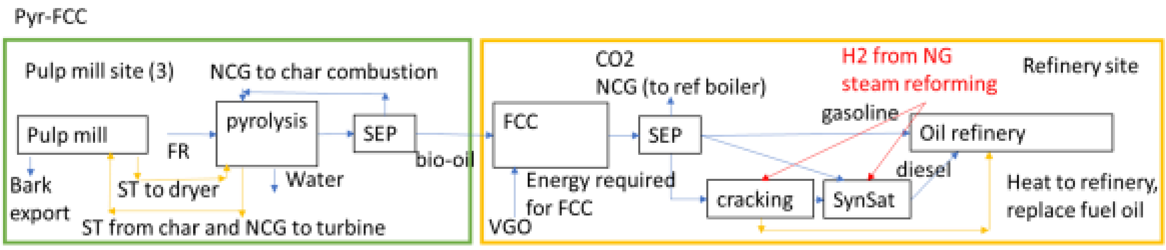

Figure 8 Overview of the process in case 3b – fast pyrolysis of forest residues followed by FCC co-processing (Pyr-FCC).

Pinho [76,77] used a pilot scale unit to co-process 5-20 % FPO with VGO at 150-200 kg/h. Their work showed that coke formation was lower in pilot scale units compared to in lab scale equipment, which indicates that FCC co-processing of “raw” FPO may be more feasible than lab scale studies indicate. There are several reasons behind this, the most important being the presence of multiple feeds, which means that FPO and VGO do not have to be mixed before processing, and the thermal shock effect caused by the non-isothermal profile of the larger equipment.

FCC co-processing yields used in this study were based on the pilot plant results [77]. FCC yields at 5 % PO were recalculated to incremental FPO yields using the VGO FCC yields as a baseline. It is very clear from this calculation that the addition of PO to the FCC has large influence on the behaviour of the fossil (VGO) components, since a lot of negative incremental yields results, in particular for light gases, as shown in Table 16 (total yields). H/C ratios for diesel and petrol are based on earlier results from the same pilot plant [78]. Since no H/C ratio was published for the bottoms fraction, a value of 1.3 was assumed here. Similarly, a value of H/C=0 was assumed for coke, meaning that coke is counted as pure carbon.

Based on the ^{14}C analysis of FCC products, which shows that 30 % of the biogenic (i.e. FPO) carbon is found in the petrol and light cycle oil (LCO) streams [77], these streams were divided in a green and a fossil component. It was further assumed that all CO, CO_2 and H_2O originate from the FPO, which is supported by the fact that these species are the primary means of removing oxygen from the FPO components. In this way, 70 % of the PO FCC products mass yields are explained, leaving 30 % (mass) of the FPO to be assigned to product yields without any useful information in the published results. From other FCC FPO processing results in smaller scale [79,80], it is known that the yields of heavy hydrocarbons and coke are high. Hence, the reminder of the FPO was assigned to yields of bottoms and coke. Based on that the total incremental coke yield when adding 5 % PO is negative, it was assumed that the green contribution to coke is minimal. This means that the full increased bottoms production was assumed to originate from PO and coke, which was used to balance the PO yields. It should be noted that no published experimental results have been found that support the assignment of green and fossil fractions of coke and bottoms. The approach used is optimistic, since it maximises biofuels yields obtained through hydrocracking of bottoms to diesel.

Using the experimental results and assumptions discussed above, the FCC can be divided in a “green” part, which shows the yields of FPO feed, and a “fossil” part, which shows the yield shifts in VGO products resulting from the FPO co-feeding, as shown in Table 16. It is apparent that the fossil yield shifts are favourable, with increased petrol and LCO yields, at the cost of gas and coke.

Table 16. Incremental mass and energy FCC PO yields at 5% FPO in VGO. Based on results from pilot scale experiments [77,78], see text. All yields on wet FPO basis.

	Total		Green		Fossil		H/C
	kg/kg FPO	MJ/MJ FPO	kg/kg FPO	MJ/MJ FPO	kg/kg FPO	MJ/MJ FPO	
H_2	-0.1 %	-0.9 %			-0.1 %	-0.9 %	
CH_4	-6.5 %	-16.8 %			-6.5 %	-16.8 %	4
C_2H_4	-2.8 %	-6.6 %			-2.8 %	-6.6 %	2
C_2H_6	-10.6 %	-25.5 %			-10.6 %	-25.5 %	3
C_3H_6	-5.7 %	-13.0 %			-5.7 %	-13.0 %	2
C_3H_8	-9.8 %	-23.0 %			-9.8 %	-23.0 %	2.67
C_4	6.0 %	13.9 %			6.0 %	13.9 %	2.5
Petrol	40.8 %	89.4 %	10.2 %	22.3 %	30.6 %	67.1 %	1.77
LCO	18.0 %	37.8 %	4.4 %	9.3 %	13.6 %	28.5 %	1.48
Bottoms	17.7 %	35.5 %	17.7 %	35.5 %			1.3
Coke	-3.2 %	-4.5 %	11.5 %	16.2 %	-14.7 %	-20.7 %	0
CO	20.1 %	9.5 %	20.1 %	9.5 %			
CO_2	4.1 %	0.0 %	4.1 %	0.0 %			
Water	32.0 %	0.0 %	32.0 %	0.0 %			
Total	100 %	96 %	100 %	93 %	0 %	3 %	

3.3.3 Catalytic hydropyrolysis (case 3c Hydropyr)

Pyrolysis process and pulp mill integration

As mentioned in the introductory part of section 3.3, the IH2 (integrated hydropyrolysis and hydroconversion) process differs from other pyrolysis processes in that the pyrolysis is carried out in an H_2 environment. The IH2 process has been described in a number of publications [61–65]. In the process, which is shown schematically in Figure 9, hydropyrolysis and hydroconversion are carried out under 20-35 bar H_2 [64]. The hydropyrolysis at 350-450 °C gives low oxygen pyrolysis vapours that are “polished” in the hydroconversion unit, giving unfinished petrol and diesel. The H_2 requirement is approximately 5 % (mass basis) of the dry ash free biomass feed. The H_2 is envisioned to be supplied internally from the NCG produced in the process, but this integration has not yet been demonstrated in pilot scale. Here it was assumed that the H_2 originates from steam reforming of the NCGs. The process gives 20-30 % (on mass basis) of products (26 % average for maple) and 35-40 % of water.

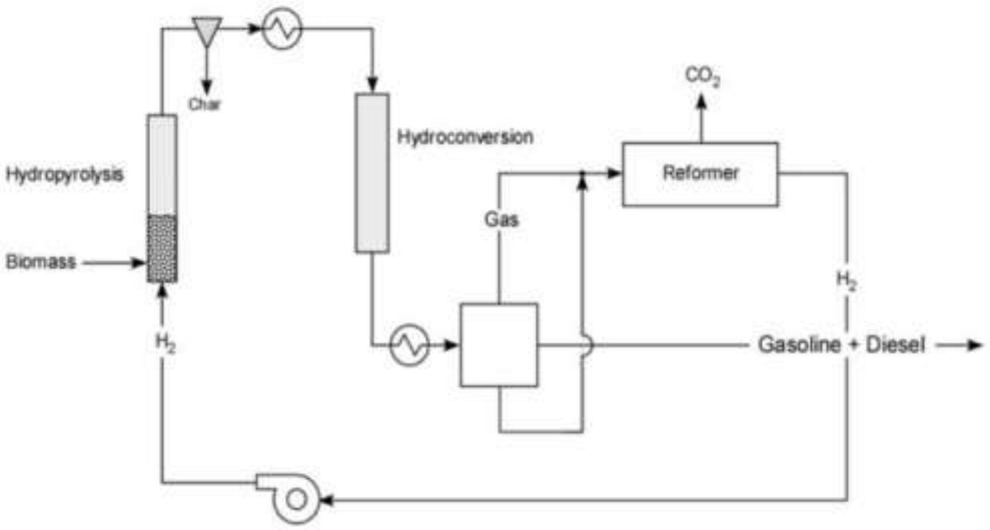

Figure 9. Overview of the IH2 hydropyrolysis process [65].

In this study, the hydropyrolysis is assumed to be integrated with a pulp mill, with the downstream upgrading integrated with a refinery, as shown schematically in Figure 10.

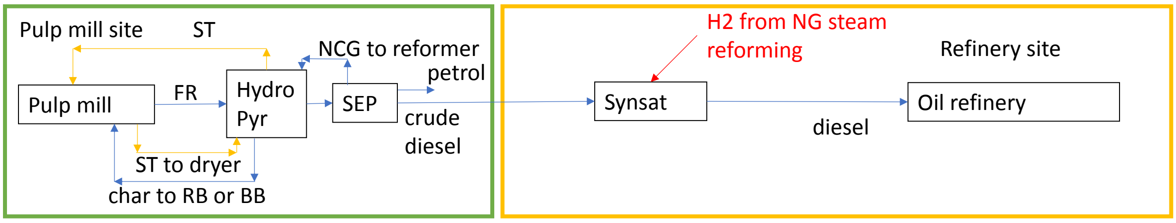

Figure 10. Overview of the process in case 3c – catalytic hydropyrolysis of forest residues.

The forest residues enter a belt dryer with a moisture content of 50 % and a temperature of 25 °C. Low pressure steam (4.5 bar, 150 °C) from the pulp mill is used to dry the wood chips to a moisture content of 6 %. The heat consumption of the belt dryer is 1.1 MWh/ton evaporated water [69]. The use of low pressure steam from the pulp mill results in a loss of power production from the mill. The loss of power was quantified using a model of a softwood pulp mill [18] as described in chapter 2.1.1.

From the dryer, the biomass enters the hydropyrolysis reactor. The mass yield for the liquid products of the IH2 process is 0.27 kg/kg dry biomass and the energy yield is 0.60 MWh/MWh dry biomass [81]. It was assumed that an on-site steam reformer produces enough H_2 to supply the IH2 process, and that no external H_2 will be needed.

The excess heat produced from the exothermic reactions in the hydropyrolysis and the hydroconversion reactors is used to generate steam (40 bar, 370 ˚C) in a boiler internal to the IH2 process, with an assumed efficiency of 80 %. The resulting steam was assumed to be sent to the pulp mill to be used for internal purposes (soot blowing, air preheating, feed water inter-heating etc.) after reducing the pressure in an expansion valve to 13.5 bar (instead of being expanded through a condensing turbine). By adopting this approach, the medium pressure (MP) steam (30 bar) extraction from the turbine could be reduced, thus enabling increased electricity generation. The increase in power production resulting from this integration was also modelled in the above mentioned pulp mill model [18] The efficiency of the medium pressure turbine was 29 %.

All char produced in the pyrolysis reactor was assumed to be combusted in the pulp mill's recovery boiler, thus producing high pressure (HP) steam used for electricity generation in the condensing turbine. The pulp mill's recovery boiler produces HP steam with an efficiency of 72.5 % and the condensing turbine's efficiency is 35.2 %. It was assumed that the condensing turbine can handle an increased condensate flow of 10 %. In fact, due to the large use of LP steam for drying, there was even a decrease of the condensate flow with 20 %.

Proprietary catalysts are used in both the hydropyrolysis and hydroconversion reactors. Both the hydropyrolysis and the hydroconversion processes are exothermal and produce excess heat that is used for steam production (see Figure 10). The diesel and petrol products do not meet current specifications due to high aromatic content [62].

Upgrading of hydropyrolysis products and refinery integration.

The liquid phase leaving the IH2 process consists of one diesel fraction (24 %) and a petrol fraction (76 %). The diesel fraction leaving the IH2 process has an H/C ratio of 1.3 [81], which means that it cannot fulfil the Swedish diesel standard MK1 (SS 155435:2016) that allows max 5 % aromatics. A H/C ratio of approximately 2.0 is required to be able to meet the standard. Therefore, the diesel fraction needs to be treated with H_2 in a SynSat-reactor at the refinery. Based on stoichiometry, it is found that additional 0.053 kg/kg crude diesel of hydrogen is required. It was assumed that the weakly endothermic behaviour of the SynSat process will not require any external heat. The petrol fraction leaving the IH2 process has here been assumed to be used as a gasoline blending component, which is possible without any further processing.

3.4 CASE 4 – FOREST RESIDUES GASIFICATION BASED METHANOL PRODUCTION

The fourth technology track describes production of methanol from gasification of forest residues and subsequent upgrading of produced syngas. The case evaluated represents a standalone facility. It is integrated with neither a pulp mill nor a refinery, which is motivated by the potential size of a gasification based biofuels plant that creates a substantial risk that surplus heat from the plant in the form of steam would not be possible to feed to the existing steam turbine in a pulp mill without exceeding its capacity. Hence, a new steam turbine was considered as a part of the plant investment in this case, giving a very similar electricity production that would have been obtained by feeding steam to an existing turbine with sufficient capacity.

Several BMG based biofuel concepts have been put forward in the recent decades [82]; a few have been demonstrated successfully in pilot-scale, although commercial implementation remains elusive. In nearly all cases that advanced beyond the drawing board, fluidized bed reactors were the gasification technology of choice as the feeding and conversion of solid biomass in entrained-flow gasifier presents challenges that are yet to be fully resolved. In keeping with the approach used for the other tracks, an initial screening of large-scale designs was performed to study process configurations and assess data availability. Designs screened include BioMeet II [83], the Swedish EON

Bio2G gasification project[11] [84], the GTI/Andritz/HTAS/UPM cooperation partly financed by US DOE [85] and the VTT study about large-scale fluidized bed gasification (FBG) [86].

In the BioMeet II and E.On B2G cases, data on mass and energy balances was not publically available. The information found was not considered sufficient enough for the performance of mass and energy balance calculations independently. Hence, VTT FBG and GTI/Andritz/HTAS/UPM concepts were taken forward for more detailed examination. An overview of the two concepts is presented in Table 17. It is followed by a description of the modelled process, the steam and electricity distribution in the plant and a discussion on the parameters used for modelling the different process blocks.

Table 17. Overview of selected gasification concepts.

	GTI/Andritz/HTAS/UPM	VTT FBG
Gasification Type	Oxygen and Steam Blown Circulating Fluidized Bed	Oxygen and Steam Blown Bubbling Fluidized Bed
Oxidising Agent	Oxygen and Steam	Oxygen and Steam
Gasification Temperature	850 °C	850 °C
Gasification Pressure	Pressurized [9 bar]	[Pressurized 5-22 bar]
End Product	Petrol [via methanol/DME]	Methanol
Data Availability	Detailed balances for both pilot and commercial scale [85]	Detailed balances [86]

3.4.1 *Fluidised bed gasification of forest residues followed by methanol synthesis (case 4 BMG)*

Process Description

Fluidized-bed gasification of forest residues, denoted here as biomass gasification (BMG), represents a means of converting solid biomass into clean syngas that can be upgraded to transportation fuels such as methanol and dimethyl-ether. In the gasification process, solid biomass is dried on arrival and fed to a pressurized bubbling fluidized bed gasifier that is operated at 850 °C and 9 bar. Both oxygen and MP super-heated steam are used as gasification agents in the refractory lined pressure vessel. The main product is a syngas that is made up of CO_2, H_2O, H_2 and CO_2. Small quantities, less than 0.5 mol%, of CH_4 and N_2 are also present. Many of the unwanted impurities in biomass form an ash that is removed together with the dolomite bed material periodically. The raw gas exiting the cyclone connected to the gasifier is cooled with water quench to the temperature of the hot gas filter unit, which is followed by a catalytic tar reforming step. The conditioning of syngas in water-gas shift (WGS) reactors and acid removal units is common to the production of methanol from both black liquor (case 2, see 3.2) and solid forest residues. Grade AA methanol is the final product.

Process Modelling Overview

The final configuration that was modelled is laid out in full in Figure 11**.** The modelling was subdivided into five blocks, namely, drying, gasification and gas clean-up (including the production of

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[11] A planned 200 MW SNG gasification project in southern Sweden, which is now halted [84]

oxygen in an air separation unit), water-gas shift, acid gas removal and methanol synthesis. Technology selection, process settings and points of interests are discussed below in greater detail for each block. The model was run on a woody biomass input of 647 MW on a HHV basis and all process units were sized accordingly. Note, however, that evaluation of cases (chapter 4) is done using generic balances without considering a specific scale. Hence, balances calculated are assumed to be representative also when scaled up or down.

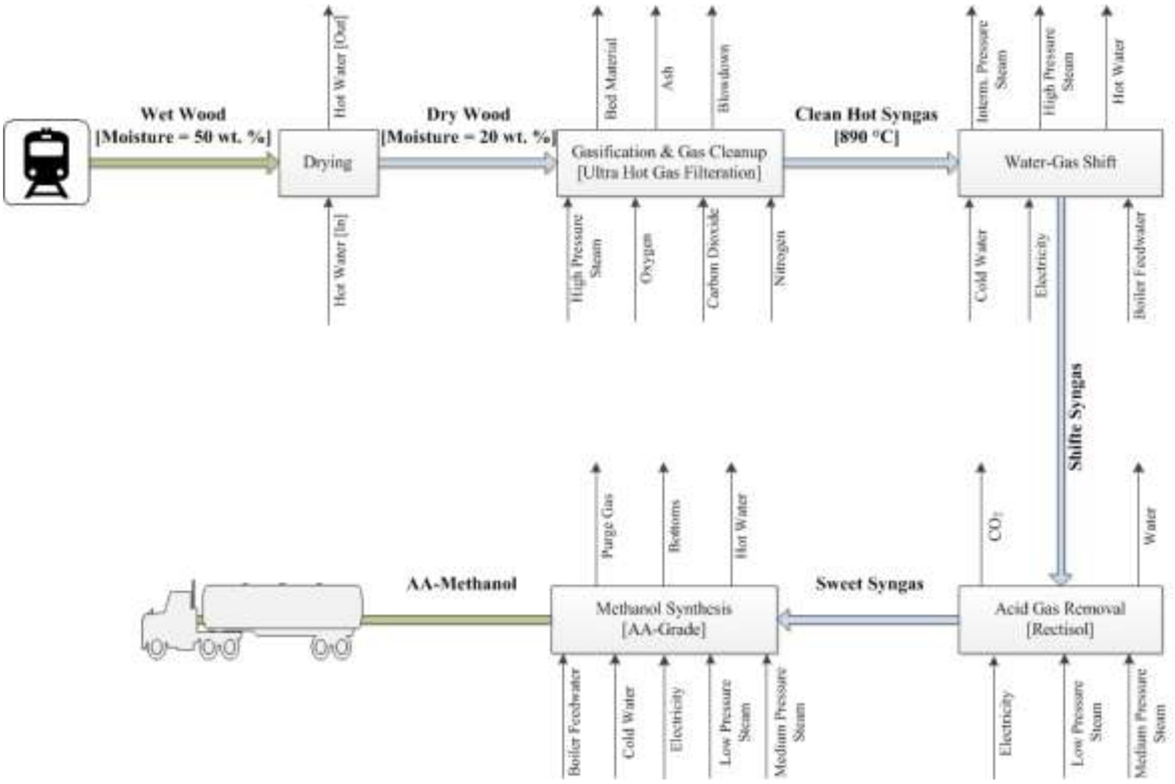

Figure 11. Process flow of forest residue gasification for methanol production.

Feedstock Drying and Gasification

In addition to the gasifier (pressurized oxygen and steam-blown bubbling fluidized bed), the block also contains an air-separation unit, an ultra-hot gas filter and a high temperature tar reformer.

The drying of woody biomass is modelled after the GTI/Andritz/HTAS/UPM concept, in which it is handled by six parallel closed loop type belt dryers that are operated at atmospheric pressure. The moisture content of biomass is reduced from 50 wt% to 20 wt%. Hot air is used as the drying agent. It is heated from 10 °C to 45.6 °C in coil heat exchangers with water, which enters at 90 °C and exits at 60 °C. The water demand for heating the air is substantial but it is wholly met with the water that is heated to the required temperature in syngas coolers at various stages of the upgrading process. The dried biomass is gasified at 850 °C and 9 bar.

Although both the VTT FBG concept and the GTI/Andritz/HTAS/UPM commercial case are based around a pressurized oxygen and steam-blown fluidized bed gasifier, their respective designs differ somewhat in operating principles. The former proposes a circulating fluidized bed (CFB) variant that is pressurized to 5 bar in the most optimal case for methanol synthesis. The operating temperature is 850 °C and the molar steam-to-oxygen ratio is 1. In contrast, the latter uses a bubbling fluidized bed (BFB) reactor that has the same nominal operating temperature of 850 °C but is pressurized to 9 bar. The steam-to-oxygen ratio in the gasifier is thus greater than 1 in order to avoid ash sintering issues. Most significantly, the final synthesis product from the GTI/Andritz/HTAS/UPM process is petrol instead of methanol. The process is designed so that over 10 % of syngas is recirculated back to the gasifier from the gasoline synthesis unit. An important consequence is that the circulation-free mass flow rate out of the gasifier is not available and a key parameter required to techno-economically evaluate the process, namely, gasifier cold gas efficiency (CGE)[12], cannot be directly determined from available data. On the other hand, the full composition of syngas after the tar reformer is available for the GTI/Andritz/HTAS/UPM process but not for the VTT FBG cases.

The two concepts also share some important similarities. Both propose the use of hot gas ceramic candles filters (UHGF) operating at 850 °C to remove fines from the syngas. The ability to operate at such a temperature eradicates the need for syngas cooling prior to filtration, which has been shown to improve cold gas efficiency by more than 5 percentage points [86]. Filtering syngas at gasifier exit temperature does come with the risk of catalyst poisoning and metallic deposit build-up. GTI/Andritz/HTAS/UPM tested filter operation at 850 °C in their demonstration facility for several dozens of hours at part load and found the performance satisfactory [85].

In the GTI/Andritz/HTAS/UPM concept, the tar-reformer is a fixed-bed reactor containing pelletized catalyst. The outlet temperature is 890 °C, while in the VTT case the tar reformer is a staged reactor with an outlet temperature of 950 °C. Importantly, in both cases, the methane fraction of the exiting syngas was 0.41-0.43 and higher hydrocarbon conversion was nearly 100 %. In the absence of syngas composition in VTT data and CGE in GTI/Andritz/HTAS/UPM data, a composition approach was used for modelling the process. The CGE for the most efficient VTT case on an LHV basis, 83.3 %, which was attained with UHGF at 5 bar, was used in conjunction with syngas composition from the GTI/Andritz/HTAS/UPM commercial case to scale the gas flow after the tar reformer for the chosen feedstock input, which was based on the GTI/Andritz/HTAS/UPM comercial case as well. Furthermore, as syngas composition data for the GTI/Andritz/HTAS/UPM commercial case was only provided for dry syngas at the outlet of the compression and cooling stage, it was assumed that commercial case had the same water fraction as the pilot case, the figures for which were available.

Calculating the wet syngas flow in the manner outlined above likely results in an overestimation, as the substantially greater heat loss in the pilot plant means that syngas contains proportionally more water than a commercial equivalent. However, as the water is condensed out in the next step, the effect of this overestimation on the remainder of the process is relatively limited. Note that according to the block flow diagram of the pilot-scale GTI/Andritz/HTAS/UPM plant, the methane and tar content in syngas were 8.11 mol% and 0.99 mol%, respectively at gasifier exit prior to the reformer. For this study, syngas at reformer exist was assumed to be free of tars on the basis of GTI/Andritz/HTAS/UPM commercial case data, while the molar fraction of CH_4 was calculated to be 0.28 mol%.

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[12] Cold gas efficiency is the most common measure of gasification efficiency. It is defined as the ratio of the heat flow of the synthesis gas leaving the gasifier ad the heat flow of the gasification feedstock.

Syngas Upgrading and Methanol Synthesis

Upon entrance to the water-gas shift block, syngas leaving the reformer at 890 °C is cooled to a temperature of 280 °C. Then, it is split into two streams: a certain amount is sent down the water-gas shift reactor, while the remainder is put through the bypass stream. The streams are recombined immediately after the WGS reactor and pressurized to 25 bar. They are then led through two sets of heat exchangers that sequentially reduce the temperature down to 36 °C and yield Intermediate Pressure Steam at 30 bar and hot water at 90 °C in the process.

In the water-gas shift reactor, 40 000 kg/hr of Intermediate Pressure Steam [30 bar] is used to shift the equilibrium of the water-gas shift reaction so that the molar $[H_2\text{-}CO_2]/[CO+CO_2]$ ratio in the syngas stream that enters the AGR block is 2.00.

The purpose of the acid gas removal (AGR) unit, which is modelled after the Rectisol process, is to remove CO_2 and H_2S from the syngas. H_2S is reduced to 0.1 ppmv to avoid poisoning of catalysts that are used to synthesize methanol. The CO_2 content of the so-called sweet gas is reduced to around 2.1 %; retention of small quantities is considered beneficial for methanol synthesis. The synthesis of grade AA methanol from sweet gas is modelled after assumptions described in section 3.2.1 to a purity of greater than 99.85 %.

Steam and Electricity Distribution

Steam demand in the methanol plant is fully met by the steam that is produced from cooling syngas in the various different process steps. Hot water at 90 °C is generated in the WGS and MeOH synthesis units. Steam was produced and used as follows: LPS at 150°C/4.5 bar, MPS at 200°C/10 bar, IPS at 235°C/30 bar and HPS at 268°C/53 bar. The distribution of steam to the different units is presented schematically on Figure 12. While some of the electricity requirement is met by expanding both LPS and HPS as well combusting purge gas from methanol synthesis for steam production, the plant is a net consumer of electricity. The deficit is met by importing electricity from the grid. A schematic overview of the distribution and generation of electricity at the plant is provided in Figure 13. An isentropic efficiency of 85 % was used to model the steam turbines, while the mechanical and pump efficiencies were assumed to be 95 % and 80 %, respectively. The procedure for estimating the electricity usage in the process units is listed in Table 18. As noted above, it would have been possible to use an existing pulp mill steam turbine (provided it has sufficient capacity) to obtain very similar electricity production from biofuels plant waste heat.

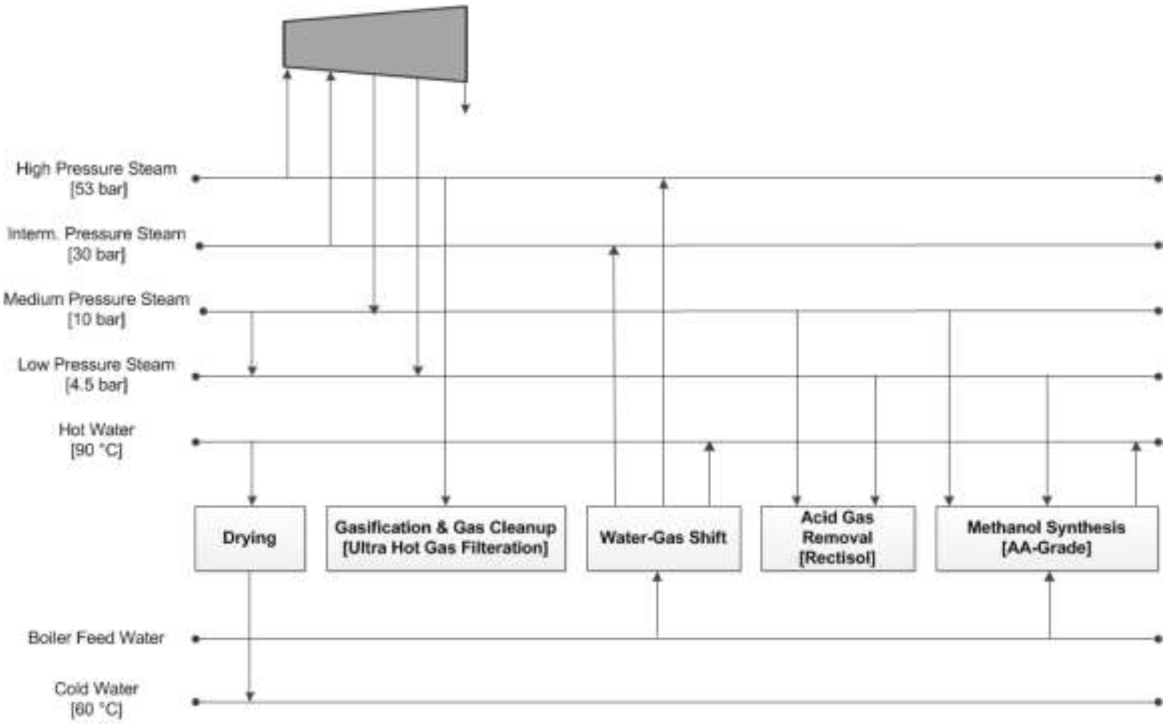

Figure 12. Production and distribution of steam in the methanol plant for the BMG case.

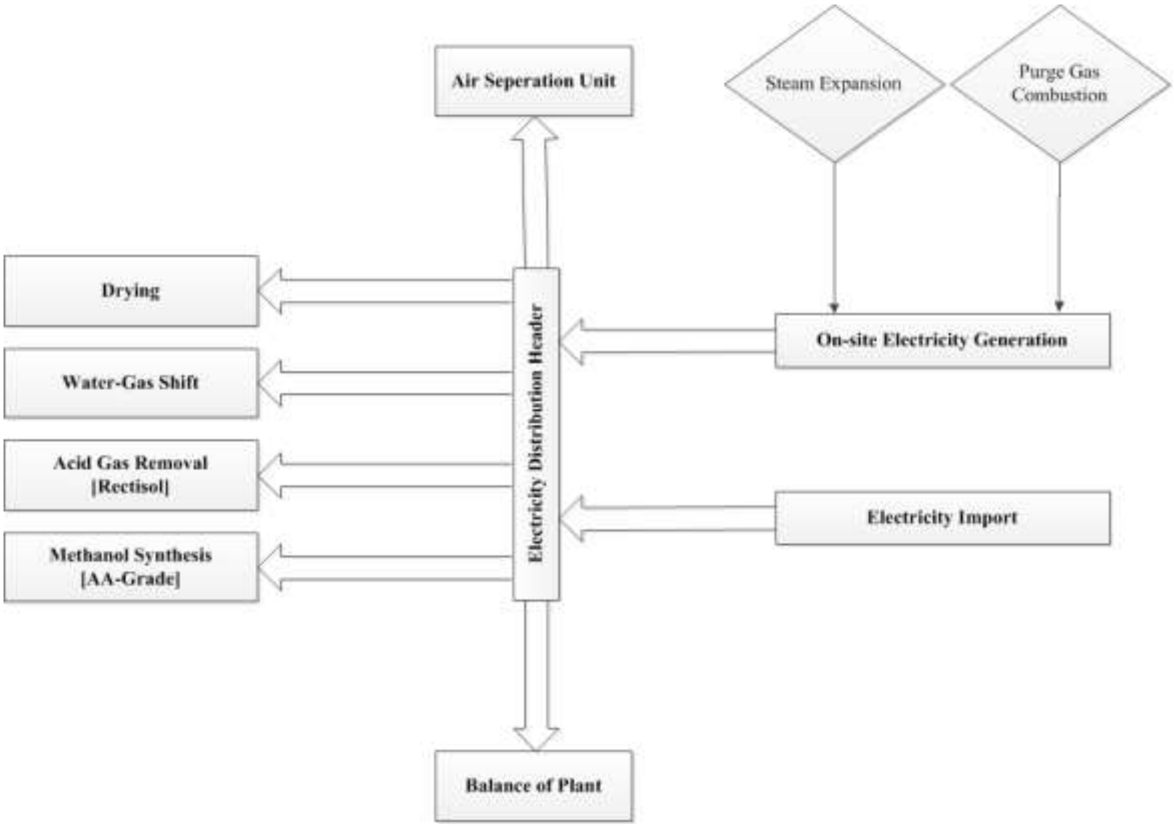

Figure 13. Overview of plant electricity distribution and generation

Table 18 Estimation of electricity use in the methanol plant for the BMG case.

Unit	Estimation Source
Compressor – WGS	From ASPEN model [modelled after [48]]
Compressor – Methanol Synthesis	From ASPEN model [modelled after [48]]
ASU	Based on [48]
AGR	Estimated with the mass flow rate of CO_2 scaled to [48]
Balance of plant (BOP)	Estimated with the mass flow rate of crude MeOH scaled according to [48]

4 RESULTS AND DISCUSSION

4.1 ENERGY BALANCES

Figure 14 and Figure 15 show energy balances (HHV basis) for each of the evaluated cases normalized to 1 MW biofuel products. Internal flows that illustrate the primary conversion steps in the process chain are also shown.

At existing kraft pulp mills, black liquor is combusted in the recovery boiler to recover pulping chemicals and to generate electricity. Figure 14 shows that taking out lignin from black liquor for the production of bio-oil, as practiced in case 1 MSL-HDO, has the effect of decreasing the amount of electricity generated at the pulp mill. As noted above, most modern pulp mills run an electricity surplus, which is exported to the grid. This essentially means that, from a system perspective, the (lost) electricity effectively constitutes the feedstock for case 1. Owing to the low electricity generation efficiency of a typical condensing turbine, the efficiency of conversion from electricity to biofuel comes out as very high, when evaluating the process from this perspective. As shown in Figure 14 (a), 0.33 MW of power (net lost, including consumption in new unit) is used in combination with 0.36 MW of hydrogen to produce 1 MW of biofuel products. At the same time, the hydro-deoxygenation and cracking of bio-oil releases heat and NCGs that can be combusted to reduce the use of fuel oil by 0.50 MW for every MW of biofuels produced. These numbers translate to a high efficiency, as further discussed below.

For the BLG case (case 2), the recovery boiler is replaced by a gasifier that performs the functions of the recovery of pulping chemicals and the production of syngas, which is upgraded to methanol onsite. The gasification and methanol synthesis steps release more heat than is used in the syngas conditioning and cleaning processes and this excess heat, in the form of steam, can be exported to the pulp mill to partly cover the lost steam production in the recovery boiler. In addition, purge gas from the methanol plant can substitute 33 % of the bark that is used as a fuel in the lime-kiln. The bark thus released can in turn be used in an expanded power boiler for steam generation. However, the quantities of steam available are not enough and for every MW of methanol produced, approximately 0.62 MW of forest residues have to be added to cover the steam demand of the pulp mill. While some electricity is generated from the expansion of the high pressure steam in the back pressure turbine, in stark contrast to the stand-alone pulp mill that is a net electricity exporter, the integrated pulp mill needs to import electricity from the grid to satisfy its electricity demand. Hence, for the BLG case, there are two feedstocks to be considered from a system perspective: fuel biomass to the pulp mill boiler and the net change in the electricity balance of the pulp mill compared to the non-integrated case, which amounts to 0.27 MW per MW biofuel product. Note that the 0.27 MW figure is the sum of lost export, the electricity sent to the grid by the mill without biofuel production, and import, the electricity imported from the grid in the integrated case evaluated here.

The first step in the production of renewable petrol and diesel, the pyrolysis of forest residues is common to both Pyr-HDO (case 3a) and Pyr-FCC (case 3b). However, refinery integration takes significantly different forms. The pyrolysis yield is 54 % on an energy basis. Owing to the composition characteristics, i.e. the ash content, of the forest residues used as feedstock, this yield is lower than what is normally obtained with stem wood feedstock. At the mill, the net effect on the electricity balance is an increase in the amount sold to the grid by 0.11 MW for every MW of biofuels produced. For case 3c Pyr-HDO, hydrodeoxygenation and cracking of the bio oil consume

0.51 MW/MW biofuel of hydrogen, which is higher than that in the MSL-HDO case (0.36 MW/ MW biofuel) owing to the higher oxygen content of FPO compared to lignin oil (LO). Heat released in the HDO reactor is approximately proportional to the amount of oxygen removed. Consequently, the use of this heat and that obtained from the combustion of released NCGs can reduce the need for fuel oil in the crude oil refinery considerably. As noted in section 2.1.1, the electricity consumption in the pyrolysis and upgrading steps has been neglected.

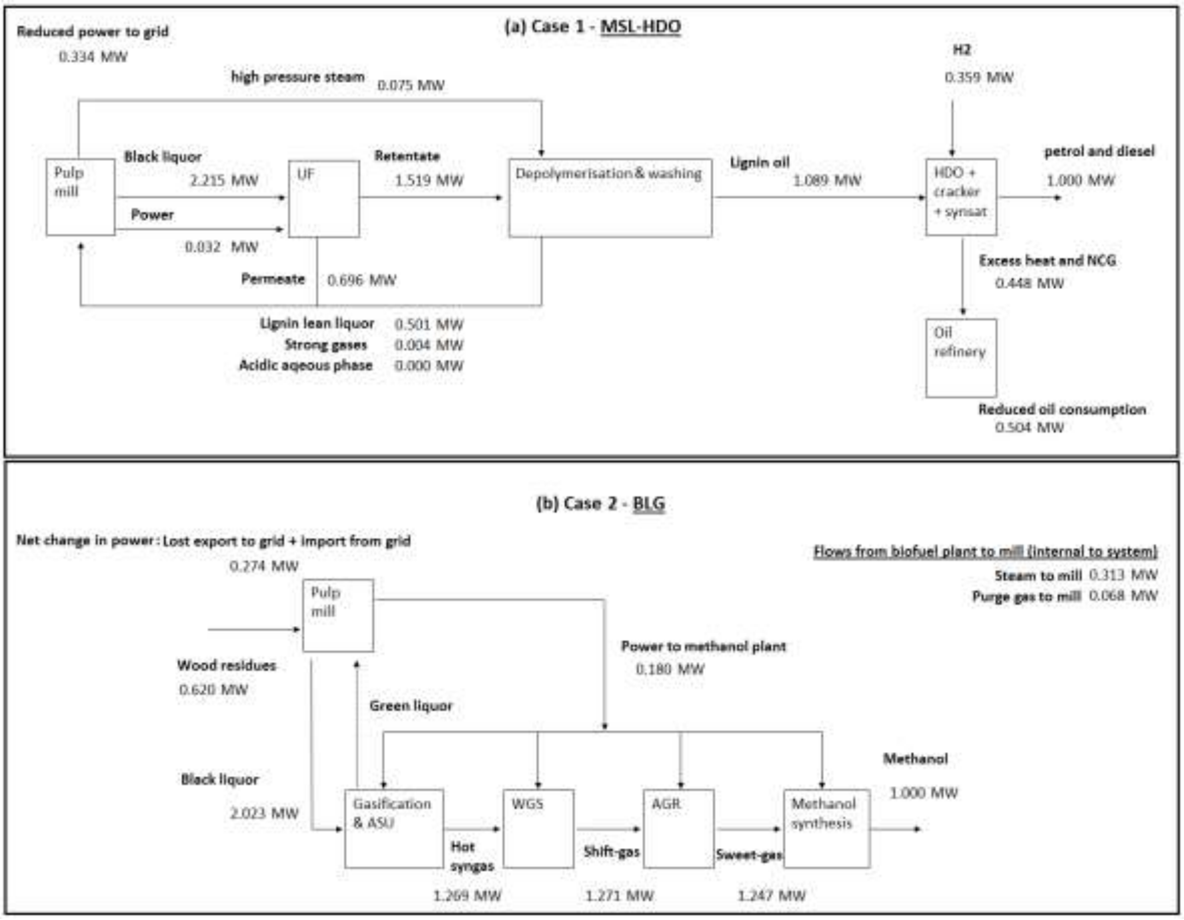

Figure 14. Energy balances for the production of 1 MW of biofuels: (a) case 1 MSL-HDO, (b) case 2 BLG.

In the Pyr-FCC case, unrefined bio-oil is co-processed with crude fossil oil in the fluid catalytic cracker (FCC). The cracking reactions are endothermic and the required heat is normally supplied by burning the coke produced in the process. NCGs from the FCC are also used as a source of energy. However, in this case, a shift in the product distribution towards higher yields of petrol and diesel means that the combined volumes of fossil and renewable coke and NCGs available are not enough to satisfy the refinery's heat demand. This leads to an increase in the use of heating oil by 1.2 MW for every MW of renewable petrol and diesel produced. However, it should be noted that the case also yields an additional of 1.336 MW of fossil petrol and diesel per MW biofuel and that the main part of the increased fuel oil demand is the consequence of the fossil yield shifts from NCG to liquid products, which results in this increase.

Bio-oil production has a very marginal impact on the mill's electricity balance for case 3c Hydropyr. Some steam is required for biomass drying, but that is partly compensated by steam generated in the exothermal process and by export of char to the pulp mill boiler. Hence, electricity production is only reduced by 0.008 MW for every MW of biofuel produced. Similarly, only small quantities of hydrogen are required for fuel refining in the refinery, in the SynSat reactor. As noted in section 2.1.1, the electricity consumption in the hydropyrolysis has been neglected.

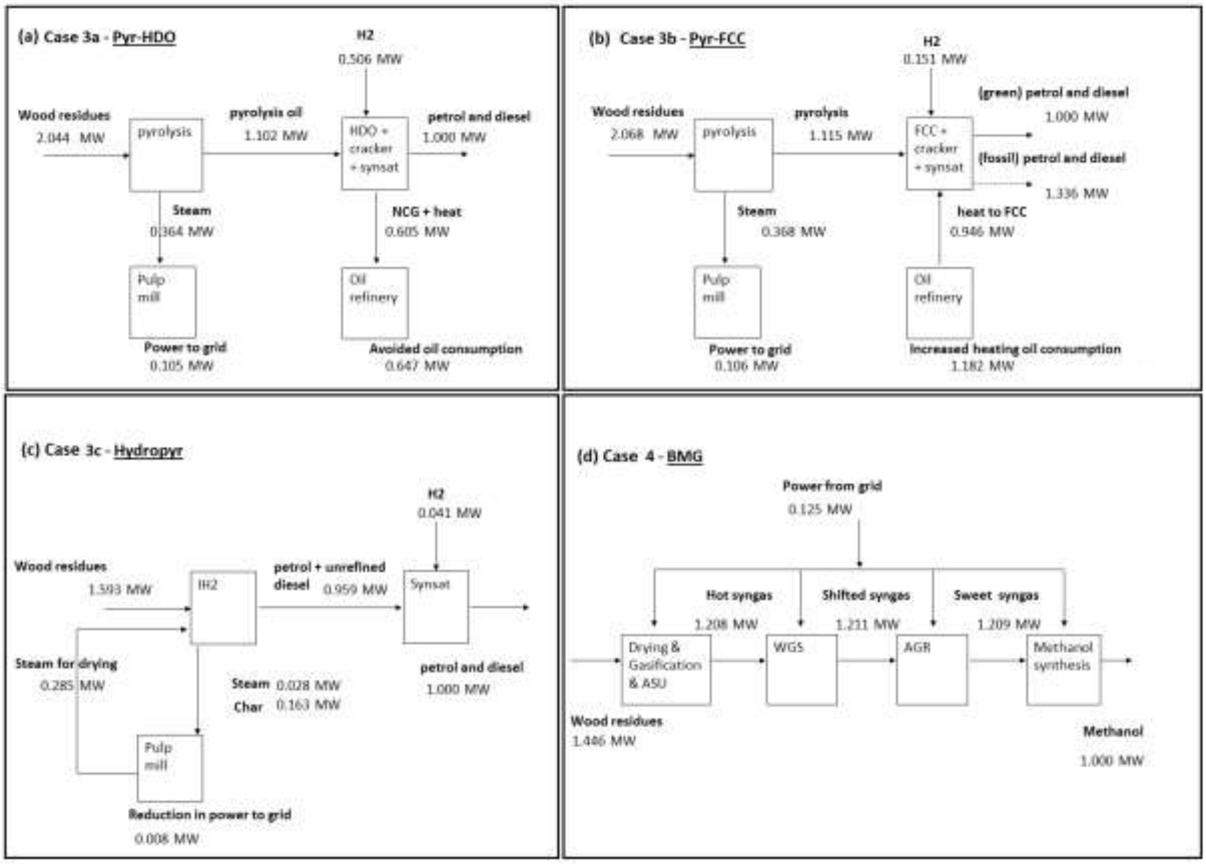

Figure 15. Energy balances for the production of 1 MW of biofuels: (a) case 3a Pyr-HDO, (b) case 3b Pyr-FCC, (c) case 3c Hydropyr, (d) case 4 BMG.

Finally, the BMG case (case 4) exemplifies a standalone facility for methanol production from forest residues. As noted above, the reason that this case is not integrated with a pulp mill is the high risk of insufficient pulp mill steam turbine capacity given the typically large scales of gasification based biofuels plants. The net balance presented in Figure 15d would, however, be very similar if this case was integrated with a pulp mill that had sufficient turbine capacity.

The process uses 0.18 MW of electricity to produce 1 MW of methanol. Nearly one-third of this electricity is produced on-site from the expansion of high pressure steam from excess process heat and purge gas combustion and some low pressure steam from excess process heat. The remainder, around 0.125 MW for every MW of methanol is taken from the grid. Approximately 1.446 MW of biomass feedstock is used per MW fuel product, giving a high biomass-to-product efficiency of 69 %.

4.1.1 Energy efficiencies

Table 19 shows the performance of the evaluated cases under the three different measures of energy efficiency used in the analysis (see section 2.2.1 for a detailed description). $\eta_{conversion}$ is the biofuel energy yield from biomass feedstock (black liquor, lignin retentate or forest residues), to methanol or refined petrol and diesel components. η_{system} is a measure for the system efficiency of biofuel production taking into account all direct energy inputs. $\eta_{system\text{-}exp}$ is calculated with energy flows in and out of the integrated system. Thus, it takes into account the effects of biofuel production on electricity generation at the mill and heating oil use at the refinery. $\eta_{system\text{-}el}$ and $\eta_{system\text{-}exp\text{-}el}$ represent electricity equivalent efficiencies for η_{system} and $\eta_{system\text{-}exp}$, respectively. Note that the thermal efficiency of conversion from the intermediates, syngas or bio-oil, to the biofuel products can be calculated from the energy balances presented in Figure 14 and Figure 15.

BMG returned the highest efficiency of thermal conversion, which is partly due to the incorporation of state-of-the-art gas filtering and tar reforming technologies that, however, have not yet been demonstrated commercially. As BMG is a standalone case with electricity as the only other input besides the biomass feedstock, both the system and the expanded system efficiencies were the same. A similar result was seen for Hydropyr. Owing to the nature of the bio oil production step, this case requires the use of relatively small amounts of hydrogen during refinery based final fuel upgrading. Although there is integration with both the mill and the refinery the only integration effect of consequence was a very minor decrease in the export of electricity from the pulp mill to the grid.

MSL-HDO and BLG have the advantage that the feedstock used is a pulping by-product that is currently utilized to generate electricity in condensing turbines. The energetic efficiency of conversion is relatively low for this process, although with electricity as the end product, the exergetic efficiency is higher. It can be noted that in the alternative case of integrating these processes with a mill that is energy deficient and does not export electricty, for example an integrated pulp and paper mill, would give different performance since this advantage would not be present. From an economic and energy efficiency perspective, the refining of BL lignin by-product to a higher value product without the use of large quantities of additional feedstock significantly bolsters the system energy efficiency in the integrated pulp mill-biofuels plant. Accordingly, for the MSL-HDO case, the expanded system efficiency returned an eye-popping figure of >500 %. This can be explained by the fact that the feedstock lignin retentate, being a by-product of the pulping process, is not an energy input under the definition of expanded system efficiency used in this report. For this case, the reduction in electricity export from the pulp mill and the import of hydrogen at the refinery represent the only two inputs by the expanded system efficiency matrix. Black liquor, which is the feedstock for BLG, is also a by-product and thus an “internal” flow in the expanded system. However, the import of substantial quantities of biomass besides electricity meant that $\eta_{system\text{-}exp}$ was significantly lower, although still high, at 112 %. Interestingly, a comparison of the electricity equivalent efficiencies shows that while BLG comes out ahead of MSL-HDO on the $\eta_{system\text{-}el}$ measure, it is outperformed by the latter on the $\eta_{system\text{-}exp\text{-}el}$ measure. These results underscore the importance of integrating the upgrading stage in the MSL-HDO case with refinery operations. It is the resulting reduction in heating oil use, in combination with the use of a by-product, which is responsible for the excellent expanded system performance of this case.

The two fast pyrolysis based cases, 3a Pyr-HDO and 3b Pyr-FCC, had very similar thermal efficiencies of 49 % and 48 %, respectively. However, they performed contrastingly on the system and expanded-system efficiency measures. Compared with case 3a Pyr-HDO, which had a η_{system} of 39 %, case 3b Pyr-FCC yielded a higher efficiency of 45 %, which can be attributed to the use of large quantities of hydrogen in the cracking process for the former case. However, case 3a Pyr-

HDO performed better on the $\eta_{system\text{-}exp}$ measure as the hydrogen consumption was compensated by a reduction in the amount of heating oil at the refinery. This is in contrast to case 3b Pyr-FCC, where changes in end product distribution have the effect of increasing the heating oil consumption substantially. In general, the example of these two cases serves to highlight the importance of taking into account integration effects, when assessing process efficiencies from an overall system perspective. It is also worth noting that although case 3b Pyr-FCC has the lowest overall efficiency of the cases by the two expanded system efficiency measures, the definition does not take into account the additional fossil petrol and diesel yield and the energy efficiency measures used here do not fully capture the effect of uncertainties around product distribution for the complex 3b Pyr-FCC case.

It can be noted that the results presented here would be partially different if the mill with which the fast pyrolysis process of cases 3a Pyr-HDO and 3b Pyr-FCC is integrated was energy deficient instead of exporting electricity (compare 2.1.1). The steam exported to the mill (see Figure 15a-b) would then replace around the same quantity of biomass instead of producing ~30 % electricity, giving higher efficiencies when system expansion is considered.

Table 19. Energy efficiency measures for evaluated cases (HHV basis).

	1 **MSL-HDO**	**2** **BLG**	**3a** **Pyr-HDO**	**3b** **Pyr-FCC**	**3c** **Hydropyr**	**4** **BMG**
$\eta_{conversion}$	66%	49%	49%	48%	63%	69%
η_{system}	52%	45%	39%	45%	61%	64%
$\eta_{system\text{-}exp}$	529%	112%	56%	30%	61%	64%
$\eta_{system\text{-}el}$	77%	88%	35%	42%	58%	69%
$\eta_{system\text{-}exp\text{-}el}$	134%	98%	51%	30%	57%	69%

4.2 PRODUCTION COSTS & SPECIFIC INVESTMENT MARGIN

Figure 16 provides a breakdown of commodity costs and revenue from biofuel sales in the 2018 energy market scenario. The dark mono-coloured columns denote aggregate expenditure, while the dashed red lines show the average spot price of European biodiesel and ethanol in 2016. All data are plotted in SEK/MWh for ease of comparison. For each case, the difference between income from biofuel sales and expenditure on commodities represents the available *specific investment margin*, which needs to cover both capital costs and operation and maintenance, as noted in section 2.1.2.

It is evident from the figure that the projected annual income from biofuel sales can cover the cost of commodities for all cases. Specific investment margins range from 155 SEK/MWh for the Pyr-FCC case to 607 SEK/MWh for the BLG case. The corresponding figures for the MSL-HDO, Pyr-HDO, Hydropyr and BMG cases are 377, 185, 329 and 495 SEK/MWh, respectively.

While most expenditure categories are self-explanatory and applicable to all cases, two require some elaboration. “Chemicals” is used as a catch-all term for consumables such as CO_2 and NaOH that are of relevance primarily in the production of bio-oil from kraft lignin. “Fossil production” captures the net cost of the substitution of a fraction of fossil production with biofuel production in an at-capacity refinery, as discussed in section 2.1.1. For all cases except case 3b Pyr-FCC (discussed below), “fossil production” is thus the price for using capacity in the existing refinery equipment. It can be seen from Figure 16 that it makes up a small but noticeable fraction of the total production cost of the bio-oil cases (cases 1, 3a and 3c).

For the Pyr-FCC case (case 3b), fuel oil was the largest single cost component, followed closely by biomass. This can be understood by recalling the effect of co-processing bio-oil and fossil crude oil in the FCC on the end product distribution of the fossil component. There is a shift in the balance of products towards increased production of petrol and diesel, which comes at the expense of coke and NCGs. Consequently, the consumption of fuel oil has to be increased to satisfy the refinery heat demand. The expense of doing so is, however, balanced out by the income from the additional fossil diesel and petrol, which is allocated to the “fossil production” category and without which the case would have struggled to return a meaningful investment margin.

The hydrodeoxygenation and cracking of bio-oil is carried out at the refinery for the Pyr-HDO and MSL-HDO cases. As discussed above (section 4.1), the excess heat from these processes is at high enough temperature to be of use internally, while the NCGs generated can be combusted onsite to reduce heating oil usage. The savings thus incurred can lead to a notable improvement in the specific investment margin (light blue negative bars in Figure 16). Catalysts represent another significant cost category that is common to MSL-HDO and Pyr-HDO. In this study, the costs of catalysts for these cases are based on 2016 price estimates by PNNL (see section 2.1.3). Moreover, for both these cases, but in particular for Pyr-HDO, the expenditure on hydrogen, which is used to remove oxygen from bio-oil, also represents a significant outlay.

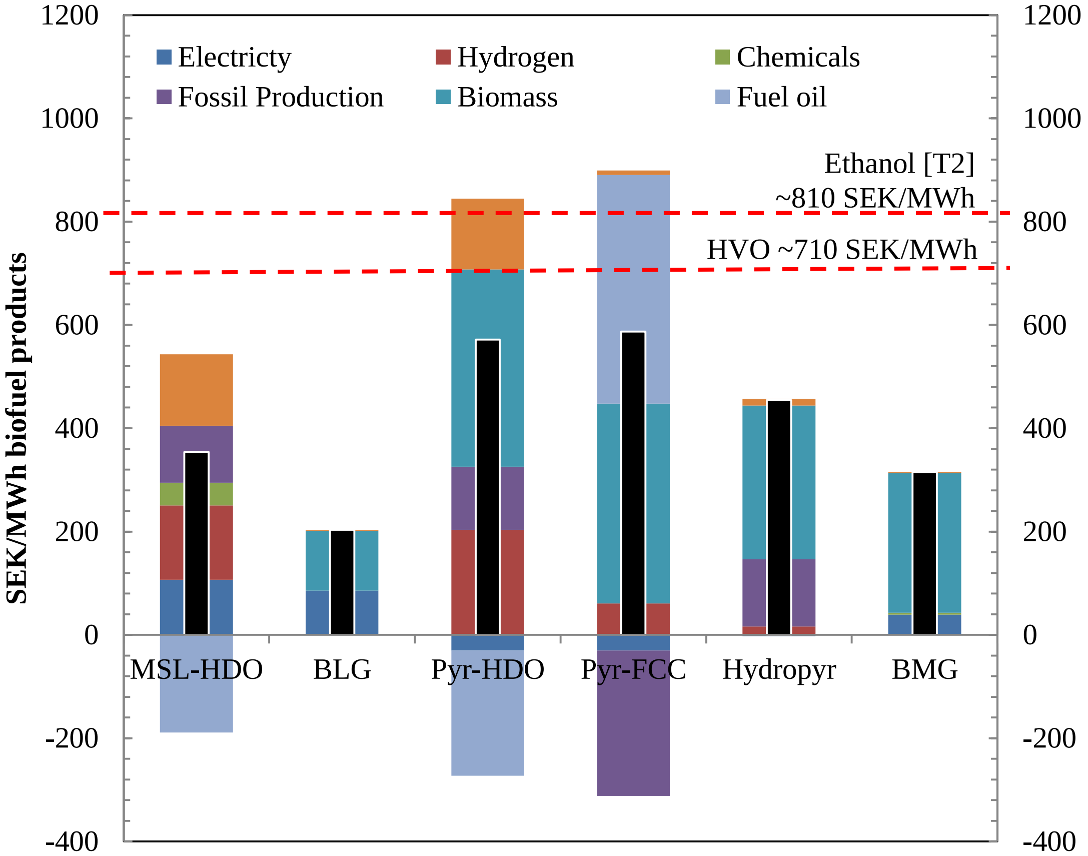

Figure 16. Cost and revenues from consumables and by-products for all cases in the 2018 energy market scenario, indicating the remaining margin (“Specific Investment Margin”) for capital expenditures and operation and maintenance costs.

One difference between the two HDO cases is that, unlike for MSL-HDO, the investment margin for Pyr-HDO is also very sensitive to the cost of biomass, which represents the single largest cost component ahead of both hydrogen and catalysts. The cost of biomass is also the decisive factor for the investment margin in the Hydropyr and BMG cases, and to a somewhat lesser extent, the BLG case. In the latter instance, the integration of methanol production with pulping operations means that the pulp mill goes from being a net exporter of electricity to a net importer. However, this is more than offset by the income from the sale of methanol, which is the more valuable commodity.

So far the discussion has been confined to the 2018 scenario, which only captures the current price of commodities. The revenues and expenditures in the 2030-CP and 2030-450 scenarios are provided in Figure 17 and Figure 18. In both scenarios, the price of biofuels is projected by the ENPAC model to be considerably higher than is the case today. The same is also true for electricity, biomass, natural gas and heating oil (see section 2.1.3). The costs of the first two commodities are the same in both scenarios. Meanwhile, natural gas, and by extension, hydrogen, is projected to cost more in the 2030-450 scenario than in the 2030-CP one, while it was the other way around for heating oil.

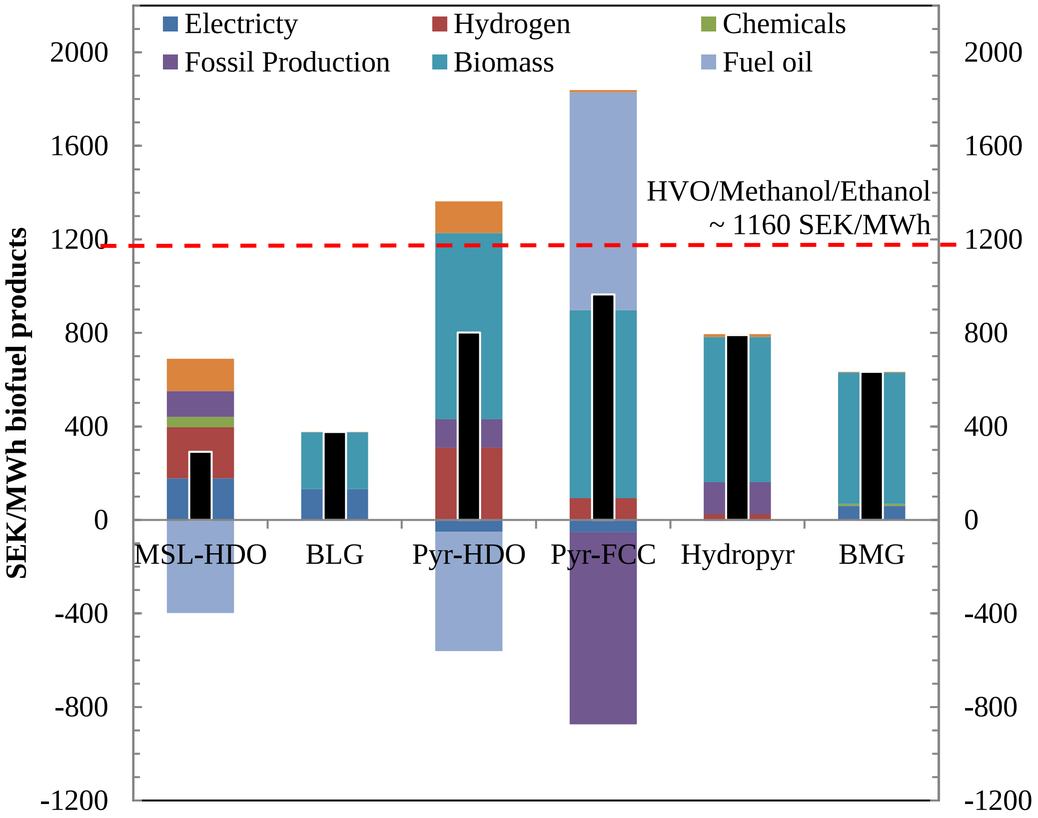

Figure 17. Cost and revenues from consumables and by-products for all cases in the 2030-CP energy market scenario, indicating the remaining margin ("Specific Investment Margin") for capital expenditures and operation and maintenance costs.

In the 2030-CP scenario, increased revenues from the sale of biofuels at a higher price of 1160 SEK/MWh were found to outstrip the rise in expenditure. Therefore, all cases returned a specific investment margin that was greater than the corresponding figures in the baseline 2018 scenario. Case 1 MSL-HDO had the highest margin, followed closely by case 2 BLG, and then by case 3c Hydropyr, case 3a Pyr-HDO and case 3b Pyr-FCC (in that order). The picture was somewhat different for the 2030-450 scenario. While the specific investment margin still remained positive for all the cases with a projected biofuel price of 1100 SEK/MWh, a combination of factors meant that some cases, such as case 3a Pyr-HDO, experienced a significant reduction, while others, such as case 3b Pyr-FCC, saw a marked increase when compared to the baseline and the 2030-CP scenario. For this scenario, the highest investment margin was obtained for case 2 BLG, followed closely by case 1 MSL-HDO. All investment margins are presented graphically in section 4.6.

It may be illustrative to contrast the results for case 3a Pyr-HDO and case 3b Pyr-FCC in the future energy market scenarios to understand some of the most influential factors behind the results. Case 3a Pyr-HDO permits a reduction in the use of fuel oil at the refinery. The margin for this case therefore improves with increasing fuel oil price. This can offset any rises in the cost of hydrogen, although the need to pay for feedstock means that biomass prices play an important role in setting the margin. On the other hand, case 3b Pyr-FCC must bear the cost of an increase in fuel oil usage at the refinery. Its margin is therefore greatest when, such as in the 2030-450 scenario, increased earnings from both fossil and renewable sales outstrip the rise in the cost of consumables.

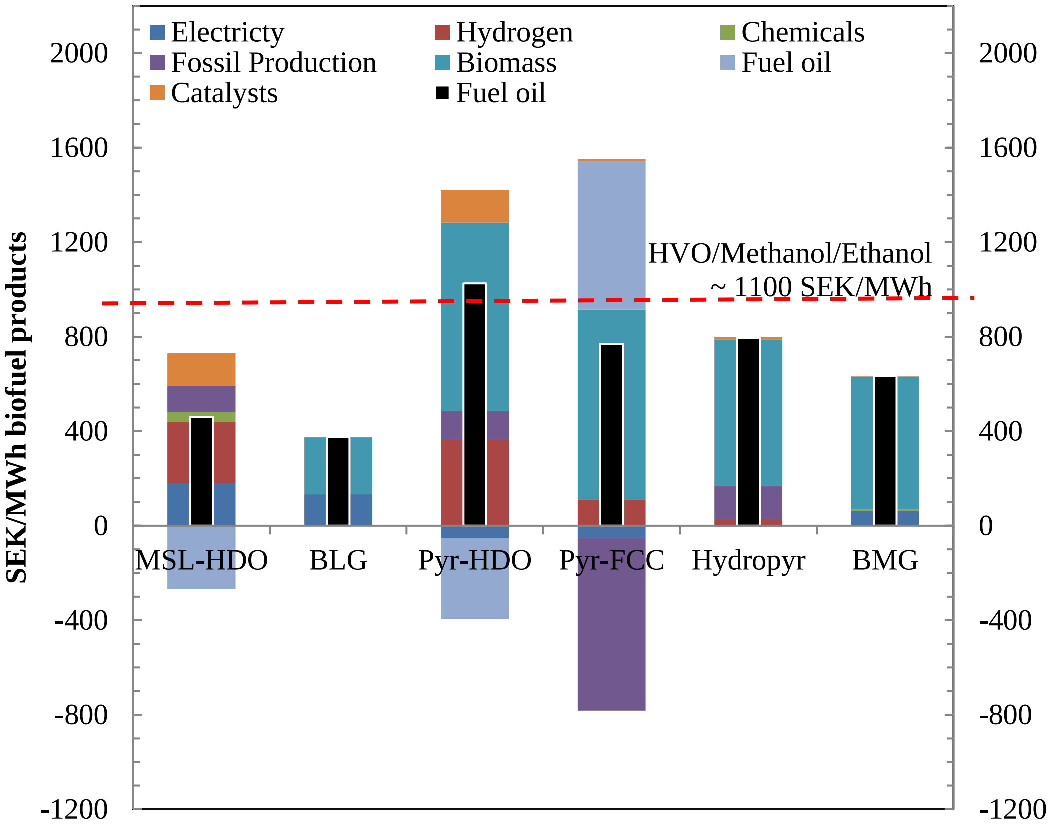

Figure 18. Cost and revenues from consumables and by-products for all cases in the 2030-450 energy market scenario, indicating the remaining margin ("Specific Investment Margin") for capital expenditures and operation and maintenance costs.

The investment margin for case 1 MSL-HDO is high under all three of the energy market scenarios. It benefits greatly from integration with the pulp mill, which enables the production of biofuels from excess pulp mill energy that is currently used to produce electricity with low efficiency. This case is thus not associated with any feedstock procurement. It should be noted that this positive effect is only valid under the assumption of an energy surplus at the mill that is assumed to be used to produce electricity with (relatively) low efficiency. To a somewhat lesser extent, the same applies to case 2 BLG, even though a certain amount of biomass needs to be purchased to satisfy the steam demand at the pulp mill. For case 4 BMG, the margin is also completely set by the price of biomass; all other costs pale into insignificance for both 2030 scenarios.

The cost of biomass also dominates the total cost of production for case 3c Hydropyr. Note that although the cost of catalysts represents only a small fraction of the total cost for this case, the price of Hydropyr catalyst is subject to some uncertainly. The publically available price estimates are several years old and catalyst development in this field has undergone significant progress in the intervening years but longer experimental runs may also have revealed catalyst deactivation problems, meaning that catalyst cost can be either higher or lower than the estimate we have used.

4.2.1 Sensitivity Analysis

A sensitivity analysis was carried out to determine the influence of some factors for which input data was considered uncertain or was suspected to have a large effect on the available investment margin. The factors studied are listed in Table 20. Figure 19 shows the effect of variations in the values of selected factors on the specific investment margin for certain cases in the 2018 energy market scenario. Note that although sensitivity analysis of investment cost is common in comparative techno-economic evaluations, it is not included here as investment costs are not used in estimating the specific investment margin.

Table 20. Factors studied in the sensitivity analysis of specific investment margin under the 2018 energy market scenario.

Parameter	Range	Case(s)	Description
Hydrogen cost	±26 %	1, 3a, 3b, 3c	Range of hydrogen-to-natural gas energy cost ratio 3.56±26 % [87–89]; the span can alternatively be interpreted as the variation in natural gas price.
Catalyst cost	-50 %, +100 %	1, 3a, 3b, 3c	The upper limit is the recorded 2015 price, while the lower limit is the 2017 projection in Jones et al [30].
Chemicals	+182 SEK/MWh	1	20 % of the chemical cost in the LBL-HTL case in Anheden et al [32], see discussion below.
Cold gas efficiency	-7.1 p.p.	2, 4	Represents operation of BMG hot gas filter at 550 °C, as opposed to a projected 850 °C demonstrated in pilot-scale[86]. The same range is used for both BMG and BLG.
HDO petrol yield	±25 %	1, 3a	Evaluation of uncertainty in product distribution. The upper limit is the high petrol/low NCG scenario, while the lower limit represents the low petrol/high NCG case.
Biofuels yield	±10 %	3c	Evaluation of the impact of lower-than-expected biofuel product yields. Of relevance owing to the low levels of technological demonstration reported at the time of writing.
FCC bottoms distribution	-50 % renewable bottoms	3b	The of fossil yield shifts on investment margin is evaluated by allocating 50% of the additional volumes of bottoms obtained from co-processing to the fossil component. Conclusive experimental evidence for FCC product distribution is unavailable in the literature.

In the present study, hydrogen is assumed to be produced by steam reforming of natural gas onsite at the refinery. The base price and range in sensitivity analysis was set on the basis of published data [87–89]. It is apparent that specific investment margin for case 3a Pyr-HDO and case 1 MSL-HDO would certainly benefit from a reduction in the cost of hydrogen production. As discussed below (section 4.3), alternative renewable hydrogen production technologies are of interest from a sustainability perspective. The sensitivity analysis of hydrogen cost presented here can be used in such an analysis, although that is outside the scope of this project.

Given the commercial considerations involved, catalyst costs represent some of the more uncertain data in the analysis, since reliable experimental data on the performance of HDO processes in relevant scales and with industrially applicable on-stream time does not currently exist. In recent years, the costs of hydrotreating catalysts have been projected to decrease by a factor of 2 year-on-year [30], which is reflected in the sensitivity analysis. Such a change can increase the available investment margin for the MSL-HDO and Pyr-HDO cases by a factor of one-third and lend a significant boost to their economic viability. This result highlights the importance of continued catalyst and process development for HDO.

The cost of chemicals was found to constitute a significantly greater proportion of the total production cost for another less technologically mature pathway in the lignin conversion track that did not make it into the final case-set (see discussion in section 3.1 and Anheden et al [32]). Lignin conversion is represented by the MSL-HDO case and chemicals contributed a relatively modest ~ 12 % to the total cost in the 2018 energy market scenario. The sensitivity analysis shows the effect of what would happen to the specific investment margin if the cost of chemicals in the MSL-HDO case rose to equal 20 % of the costs in the discarded track. The effect is dramatic as the available margin nearly drops by a factor of 2 from 375 SEK/MWh to <200 SEK/MWh. This highlights the importance of keeping the chemical costs low to ensure the economic viability of technologies in this emerging technology track.

Three parameters in the sensitivity analysis evaluate uncertainty in product yields. For the Hydropyr case (3c), lower-than-expected biofuel yields could potentially reduce the available specific investment margin by ~40 SEK/MWh. For the HDO cases (1 and 3a), the petrol yield parameter captures the effect of a higher NCG yield at the cost of petrol, which is insignificant for case 1 MSL-HDO as diesel is the dominating biofuel fraction and ~30 SEK/MWh for case 3a Pyr-HDO.

In the sensitivity analysis of case 4 BMG, the lower CGE value represents the operation of the hot gas filter at 550 °C instead of the 850 °C used in the default scenario. The net effect, taking into account changes in the steam and electricity balances, is a relatively marginal reduction in investment margin of ~10-15 SEK/MWh. An analysis of a similar drop (in p.p. terms) in CGE for case 2 BLG shows a nearly negligible drop in the investment margin of ~10 SEK/MWh. The decrease in case 2 CGE leads to a drop in the revenue from methanol sales, which causes to margin to shrink. However, this is countered to a certain extent by the increased production of LPS in the gasifier, which can be used in the biomass boiler to reduce the expenditure on biomass imports. Another contributing fact to the somewhat unexpectedly small effect of CGE decrease on case 2 and 4 specific investment margin, is the fact that the margin is calculated on biofuel production basis, which leads to a certain amount of “self-compensation” when biofuel production decreases. This is not necessarily reflected in the actual investment cost, however, since the gasifier investment scales better to fuel input and the gas cleaning and conditioning units to total gas flow and removed acid gases, respectively.

Finally, the effect of the uncertainty around fossil yield shifts is also evaluated. As discussed in section 3.3.2, the approach used in this report assumes somewhat optimistically that the additional amounts of bottoms, and the diesel obtained from their cracking can be assigned wholly to the green part. The FCC bottoms distribution parameters measures the effect on the investment margin if 50 % of these additional bottoms volumes were assumed to originate from fossil VGO instead of renewable bio-oil. Under such a case, the total yield of diesel from the FCC would remain the same but the diesel obtained from the cracking of the VGO-based bottoms would be assigned to the fossil diesel fraction. Figure 19 shows that loss of revenue from the decrease in renewable diesel yields under the above would reduce the available margin by more than half to 71 SEK/MWh in the 2018 energy market scenario. These results show that not only are the extra revenues from the additional diesel and petrol yields in Pyr-FCC vital to economic viability of this case, but the actual distribution of these yields between the fossil and the renewable fractions is also of tangible importance. Dispelling the uncertainty surrounding both the presence and the origin of these additional petrol and diesel volumes likely requires results from demonstration of the concept in a representative scale, which is yet to be established.

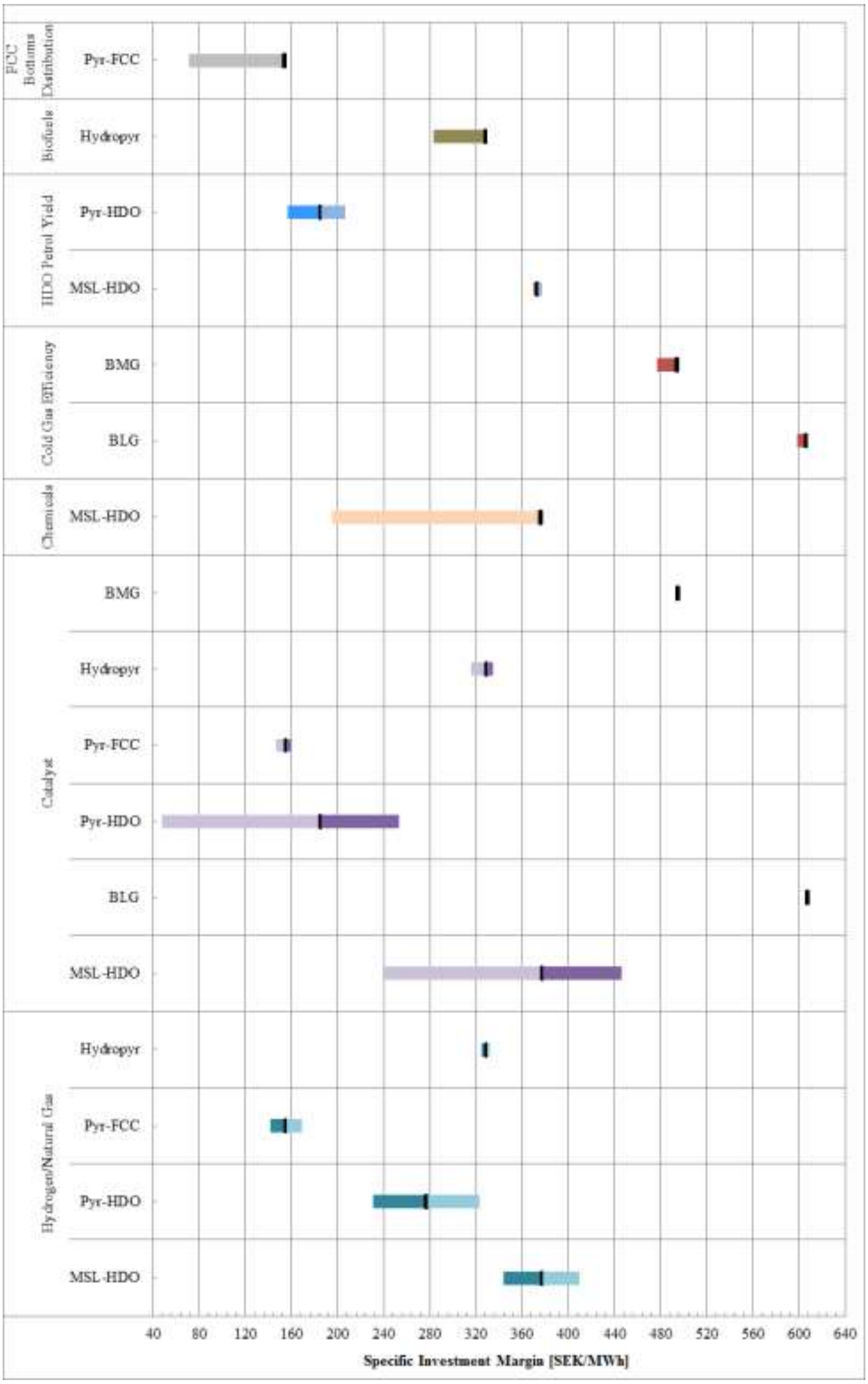

Figure 19. Sensitivity analysis of Specific Investment Margin for selected cases in the 2018 energy market scenario. Black bars represent the base case Specific Investment Margin.

4.3 GREENHOUSE GAS EMISSION PERFORMANCE

The GHG footprint of the production pathways was estimated based on the factors known to have the largest influence as described in section 0 above. There are a number of methodological choices which have a large influence on the results, especially for the refinery integrated tracks (1, 3a, 3b) as discussed below. Note that LHV energy basis is used in this section, which is an exception as HHV basis in used in the remainder of the report. The reason for this is to make GHG emissions on an energy basis, i.e. as g CO_2eq/MJ, directly comparable with estimates in the literature such as the JRC-EUCAR-CONCAWE collaboration report [90] and the EU Renewable Energy Directive [91].

Figure 20 shows the estimated GHG footprints for all cases when using an allocation methodology that is consistent with the Renewable Energy Directive. For the HDO cases (1 and 3a), it has been assumed that emissions can be allocated to the NCG formed in the process but not to the excess heat that is recovered[13]. Note that the GHG footprints for cases 1, 3a and 3b are estimated using alternative accounting methods as discussed further below. Emissions are averaged over all biofuel products, i.e. we have not differentiated between biofuel products that substitute petrol and diesel respectively. The relevant fossil references are 93.5 and 95.5 g CO_2eq/MJ for petrol and diesel, respectively.

Cases 1 and 3a, which both involve the HDO step, give GHG emissions savings of 70 % and 66 %, respectively compared to the fossil reference. This means that these technology tracks, using the evaluated process designs, risk not meeting the EU Renewable Energy Directive sustainability criteria, which is proposed to require at least 70 % GHG savings for new plants from 2021 [42,91]. The reason for the large GHG footprint is mainly the emissions associated with natural gas-based hydrogen production. However, the loss of electricity generation in the pulp mill because of lignin extraction (case 1) and the feedstock supply (case 3a) also play significant roles. Note that without the allocation of emissions to the NCGs that replace fuel oil at the refinery (negative violet bars in Figure 20), the greenhouse gas reduction would be only 60 % and 50 %, respectively for cases 1 and 3a.

The use of a hydrogen production technology with a lower GHG footprint would decrease the GHG footprint substantially. Potential options are water electrolysis or natural gas reforming with carbon capture. The electrolysis option requires that the imported electricity mix has a low GHG footprint. The GHG footprint of hydrogen generated by water electrolysis using the Swedish electricity mix is approximately 20 g CO_2eq/MJ[14], which is almost 80 % lower than the natural gas reforming-based footprint of 91.4 g CO_2eq/MJ. For the cases involving HDO (1 and 3a) it is also technically possible to use a part of the NCGs from the HDO process for the production of (green) hydrogen. However, in the present work, these are already given credit for replacing fuel oil, so the effect on the GHG footprint would not be large.

It must be noted, however, that the use of other technological options to reduce the hydrogen GHG footprint can also lead to increased production costs. These options have not been investigated further in the present work but the analysis by Anheden et al [32] can give some guidance.

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[13] For case 1 and 3a, NCGs corresponds to 19% and 30%, respectively of the total product yield from HDO, on an energy basis.

[14] Based on the assumption of 13.1 g CO2eq/MJ electricity (see 0) and electrolysis efficiency of 65%.

For the two gasification-based cases 2 and 4, the estimated GHG footprint is 5 g CO_2eq/MJ, which translates to GHG savings of almost 95 % GHG savings compared to the fossil reference. This is in agreement with the low footprint for black liquor and waste wood-based methanol presented by the JRC/EUCAR/CONCAWE collaboration [90]. Cases 3b and 3c also give low estimated GHG footprints with savings of 88 % and 93 %, respectively. There is, however, significant complexity present in the choice of methodology for case 3b Pyr-FCC, as discussed below. The numbers presented in Figure 20 do not account for fossil yield shifts. It can also be noted that the estimated GHG savings for the Hydropyr products (93 %) are somewhat higher than what has previously been calculated for an integrated hydropyrolysis-based track. The values in the literature are in the range 67-86 %, and the differences depend mainly on feedstock transport and the source for H_2 (internal or external NG based) [92].

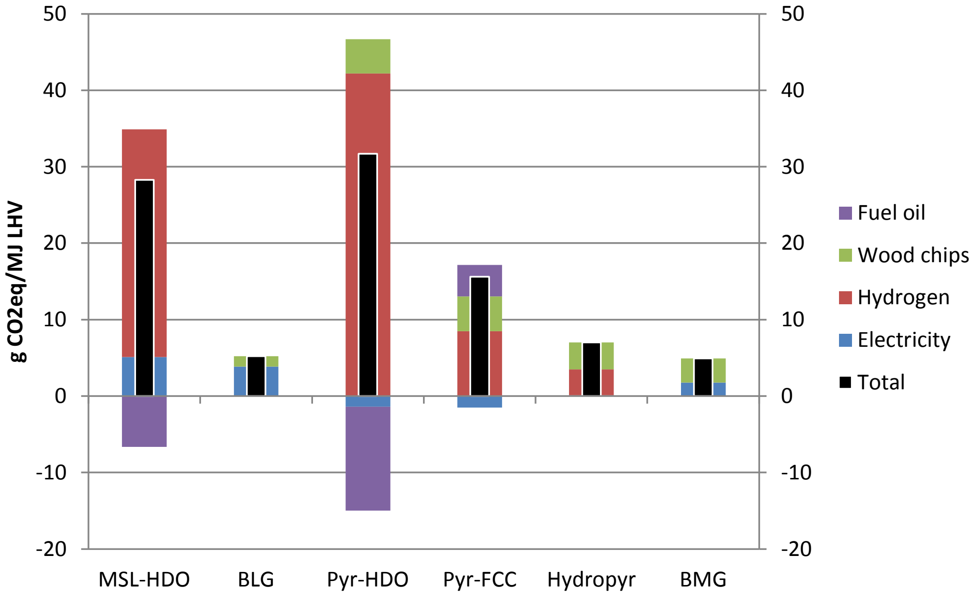

Figure 20. GHG footprint of the studied production pathways following RED methodology.

As noted above, the large GHG footprint for cases involving HDO is a consequence of using natural gas-based hydrogen. Large amounts of heat are released in the process, which can replace combustion of fuel oil as a source of process heat in the (fossil part of the) integrated refinery. Thus, if the allocation methodology is expanded to include the integration effect discussed above, the savings associated with the withdrawal of fuel oil represent a large negative GHG contribution.

Figure 21 shows GHG footprints for cases 1, 3a and 3b when calculated according to alternative (not according to RED) GHG accounting methodologies. As evident from the figure, applying system expansion gives overall negative GHG footprints for both cases 1a and 3a. These results are in strong contrast to the GHG performance figures obtained with the RED accounting methodology. The current RED GHG allocation methodology is thus unable to capture fully the effect of system integration at the refinery, as it does not permit the allocation of emissions to the released process heat and the consequent benefits from its use in other refinery processes.

For case 3b, which involves co-feeding of FPO in the refinery FCC unit, the complexity is even greater since another complicating effect is present in the form of the influence of FPO co-feeding on fossil yields. If the system is expanded to include the shift in fossil yields from light fuel gases to liquid fuel products, the overall GHG footprint for this case also turns negative, as shown on

Figure 21. This can also be understood by recalling that shifting (fossil) fuel gas yield to (fossil) petrol and diesel yield gives a lower increased fuel oil demand than the crude oil that would have been required to produce the increased petrol and diesel production. Hence, the credit for increased petrol and diesel production (cyan and orange in Figure 21) is larger than the burden from the increased fuel oil consumption (purple in Figure 21). It can also be noted that the two large numbers being compared here can be considered significantly uncertain due to the uncertainty surrounding the whole phenomenon of yield shifts. Accordingly, this means that the resulting negative GHG footprint calculated using system expansion for case 3b should be considered highly uncertain.

Another way to account for the fossil yield shifts for case 3b is to do energy allocation of the GHG burden to the marginal changes in both “green” and fossil FCC products (case “RED w alloc” in Figure 21). This entails the following of RED methodology for the main part of the analysis, but also allows the inclusion of changes[15] in the fossil part of the FCC process and not only the fate of the green atoms as in the RED method (Figure 20, “RED only green”). Using this methodology, the benefits of feeding biomass-based feedstock to the FCC are divided between the green products and the increased fossil yield, resulting in a GHG reduction that is smaller than if only the green part of the process is considered (as in Figure 20).

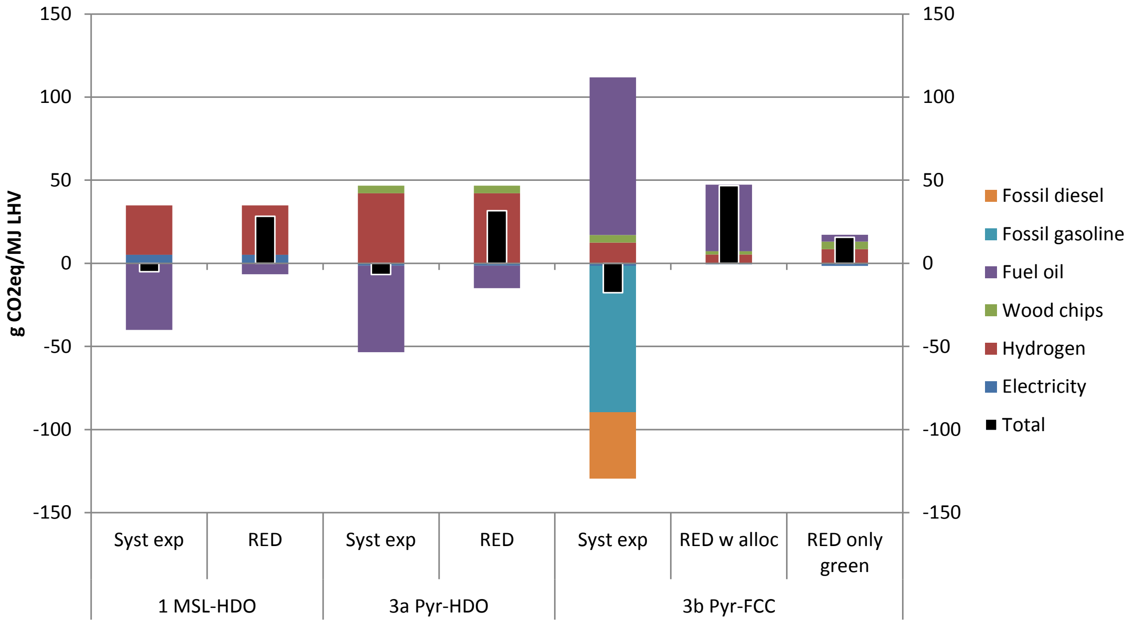

Figure 21. Different GHG accounting methodologies for cases 1, 3a and 3b. The results for cases denoted “RED” and “RED only green” is identical to results in Figure 20.

The discussion above illustrates the large effect of applying different accounting methodologies. The fact that the type of effects that are observed when co-feeding FPO to an FCC unit (fossil yield shifts) were not considered when the GHG accounting methodology in RED was decided leads to uncertainties regarding which approach to use. However, it is opined that a GHG accounting meth-

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[15] Note that only the biomass based part of the FCC process and changes in the fossil FCC process are considered in this analysis. The full FCC process, with 95 % fossil VGO feed, is not considered, since it is treated as the reference/baseline.

odology that concerns itself only with the fate of green atoms in the FCC provides the greatest consistency with RED by analogy to other processes and, hence, that was the approach used in Figure 20.

In summary, the GHG performance of the two gasification cases (2 and 4) and the hydropyrolysis case (3c) was found to be very good. For cases 1 and 3a involving HDO, the current process design that uses natural gas-based hydrogen offers some challenges in terms of GHG performance but alternative hydrogen production methods can change this picture. For case 3b, with FCC co-feeding, there is significant methodological complexity but an approach following RED considering only the fate of green atoms gives good GHG performance.

4.4 TECHNOLOGY MATURITY

A technology maturity assessment was carried out for all cases evaluated in the project. As noted in the description of the methodology (section 2.3), two approaches are used:

- A weighted average approach, giving an estimate of the overall maturity of the chain of individual process technologies.
- A ”weakest link” approach, in which the main process step with the lowest maturity is used to represent the chain, since it can be considered the limiting factor with respect to development and application.

Figure 22 shows TRL estimates of both types. The underlying estimates for each process step and the weights used in the estimation can be found in Appendix B.

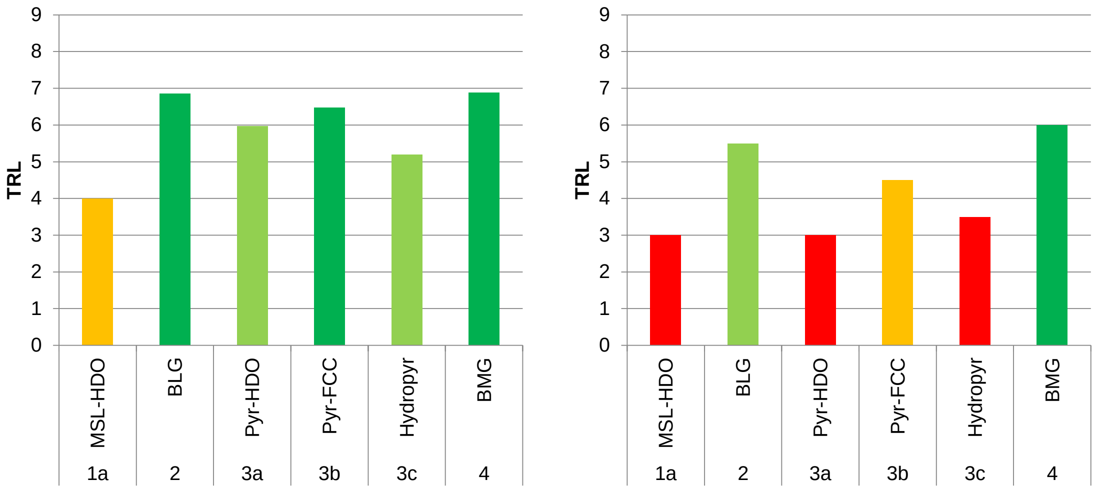

Figure 22. TRL assessment for the cases using weighted average approach (left) and weakest link approach (right). Colouring of bars: red < 4 ≤ orange < 5 ≤ light green < 6 ≤ dark green.

Three cases, 2 BLG, 3b Pyr-FCC and 4 BMG, have a weighted TRL larger than 6, which corresponds to “Technology demonstrated in relevant environment (industrially relevant environment in the case of key enabling technologies)” on the TRL scale. However, if the “weakest link” is considered, this only holds for case 4 BMG, with case 2 BLG coming close. Cases based on refinery integrated processes, cases 1 and 3 generally have lower TRL with case 3b Pyr-FCC being the most developed.

Overall, case 1a MSL-HDO is the least mature with TRL estimates of 4 and 3 using the two approaches, respectively. This clearly indicates that there is substantial R&D work required to bring this technology track to commercial application. For this case, there are several process steps that have low TRL, namely, depolymerisation of lignin, purification of crude "lignin-oil" and hydrodeoxygenation and cracking. Research is actively being pursued or has been commissioned for all these processes by a number of commercial actors, which means that results are typically not available in the literature publically. Hence, there is a possibility that research has progressed further than is indicated in the present analysis, which is based on the open literature and the information that the commercial developers have chosen to reveal.

For the FPO upgrading cases, 3a-3b, the least mature step is the step at which the FPO is introduced in the refinery. In case 3a Pyr-HDO, this is done in a 2-step catalytic HDO unit, for which the state-of-the-art [30] is only based on data from small-scale catalytic experiments in 30-400 mL reactors. For scale-up, temperature control in the exothermal deoxygenation process is crucial and typically not considered in small, lab-scale studies. A commercial actor, UOP, has several patents [71] in this area, covering recycling of HDO product and an initial mild HDO stabilization. In combination with the fact that most work has been done on FPO that is derived from stem wood, which generally has lower levels of alkali than FPO from forest residues, this gives a TRL estimate of 3 for this step. All other process steps in this track have a fairly high TRL, which explains the large difference between the two types of TRL estimates (3 and almost 6 respectively, see Figure 22).

In case 3b, FPO is introduced in an existing FCC as a co-feed with fossil vacuum gas oil. This is an "easier" task, e.g. due to dilution, which facilitates temperature control. The co-feeding has been experimentally demonstrated on a reasonable scale (200kg/h) but with relatively low accumulated operating on-stream time, which leads to a TRL of 4-5 for this step. Similarly to case 1, research is actively being pursued or has been commissioned by commercial actors, which means that the knowledge front may have advanced beyond what is publically available. Claims of commercial trials have been made by UOP; however, results from such trials have not been made available.

Case 3c is quite different from cases 3a and 3b, even if it is also based on pyrolysis and refinery integration. The main difference is that the refinery integration is much weaker, with the bio oil being introduced in the refinery only for final upgrading after it has been deoxygenated in the integrated hydropyrolysis/HDO process. Hence, the key to this process is the successful development of this integrated process, which has, to the authors' knowledge, so far only been demonstrated in a small pilot scale unit of 50 kg/d at the GTI facility in Chicago. There is a lot of work currently being put into developing this process further including a demonstration plant with a feedstock capacity of 5 t/d, which is under construction in India. The commissioning and continuous operation of that plant, including the use of new waste-based feedstocks, represents a very important step for this technology. If successful, it will denote a significant climb on the TRL scale for this case.

The gasification cases, 2 and 4, generally have higher TRL compared to the refinery-integrated cases. For both gasification cases, the gasification process step itself is the step with the lowest TRL, since synthesis gas conditioning and upgrading processes are based on technologies that have been commercially available for fossil feedstocks for many decades. Nevertheless, both gasification technologies studied have been demonstrated in significantly large scales (20 t/d) for extended periods of time (>26 000 h and >3 000 h respectively). This leads to the overall higher TRL levels for these cases, both when using the weighted approach and the weakest link approach.

In summary, technology maturity is often often very different for the steps in a specific production chain. Aggregated results indicate that cases 2 BLG, 3b Pyr-FCC and 4 BMG are ready for a commercial-scale demonstration. Cases 1 MSL-HDO, 3a Pyr-HDO and 3c Hydropyr have lower TRLs (below 4), indicating that substantial R&D effort is still needed to bring these technology tracks to commercial use. This is a paradox considering that drop-in biofuel production from technologies represented by cases 1 and 3a-c is a short-term priority according to the strategy of the Swedish Energy Agency.

4.5 PRODUCTION POTENTIAL

As described in the methodology section (2.4), the technical production potentials were estimated based on domestic feedstock resources and the biomass-to-biofuel conversion efficiencies from the resulting mass and energy balances for each production pathway, with some additional technology constraints considered as well. Table 21 summarises the resulting ranges of different feedstocks and technology boundaries as used in the overall potential estimations for the technology tracks, and also gives a brief description of the key assumptions behind those ranges. For details, see Appendix C.

Table 21. Summary of estimated ranges for feedstock potentials and technology constraints (FCC), with brief descriptions of key assumptions.

Parameter	Range (TWh/y)	Max. estimate	Min. estimate	Case(s)	Sources
Black liquor (BL)	24-54	All BL in all kraft pulp mills, with annual prod. increase of 1.3 % (to 2030)	All BL in kraft pulp mills (current prod.) with recovery boilers (RB) built before 1995	2	[93–95]
Black liquor retentate	6.7-16 [a]	30 % of max BL potential assumed extractable (all mills, increased prod.)	Individual max extraction potential, to ensure no negative effects on RB operation	1	[94]
Forest residues (FR) (branches and tops)	13-37	"Techno-ecological" harvesting potential from SKA 15, reduced by estimated use in other sectors (2030)	Same as for max, but reduced by further ecological restrictions to avoid conflicts with environmental quality objectives	3, 4	[94,96,97]
FCC co-processing capacity	0.30-0.50 [b]	5 wt% co-feed in existing FCC	3 wt% co-feed in existing FCC	3b	[c]

[a] Both estimates would require additional biomass fuel to the pulp mill to account for the energy in the extracted lignin. This has not been considered in the potential estimations.

[b] PO feed to the FCC.

[c] Personal communication with Åsa Håkansson, Preem (6 October, 2017).

Figure 23 shows the resulting production potential ranges for the considered technology tracks. As can be seen, the ranges are for most cases very broad, which can be explained by the uncertainties with which the feedstock potential estimations are associated.

For case 2 BLG, the technical maximum available amount of feedstock would be all black liquor from all chemical pulp mills in Sweden. However, converting all kraft pulp mills to black liquor gasification is for obvious reasons extremely unlikely. The lower end of the potential range (12 TWh biofuel per year), where only pulp mills with older recovery boilers were considered, should thus be seen as a more likely indication of the practical potential for the BLG case.

Also for case 1a MSL-HDO, the upper estimate can be considered as quite optimistic. This estimate builds on the assumption that 30 % of all black liquor in all kraft pulp mills (assuming also an overall kraft pulp production increase of 1.3 % per year) can be extracted, which would give a total black liquor retentate potential of 16 TWh/y. As a contrast, the lower estimate builds on a mill-by-mill estimate limited by the chemical recovery operation in the recovery boilers(in particular the operating temperature), in order to prevent negative effects [94]. This estimate was done based on experience figures regarding reliable energy and chemical balances in the recovery boiler for different pulp types, and resulted in relatively modest 6.7 TWh black liquor retentate per year [94], which corresponds to only 15 % of the current total black liquor volumes in Sweden, on an energy basis. After application of the biomass-to-biofuel conversion efficiency, a likely practical potential for the MSL-HDO case is thus in the mid-to-lower end of the total range (4-8 TWh biofuels per year).

For forest residues (cases 3 and 4), the full technical harvesting potential from final felling as well as commercial thinning was considered as the upper boundary, reduced only by the set of restrictions and recommendations regarding environmental considerations issued by the Swedish Forest Agency [98]. Conversely, for the lower boundary, further restrictions on the harvesting were considered in order to avoid conflict with different environmental quality objectives that may be impacted negatively by increased use of forest biomass for energy purposes, as discussed by de Jong et al. [97]. As a consequence, the forest residue-based biofuel production potential may be as low as 6-9 TWh/y, or as high as 18-25 TWh/y, assuming no significant increases in biomass demand in other sectors, as compared to today’s levels. Since the potential estimations do not include any economic considerations (e.g. for harvesting or transportation of forest residues), the maximum values in the resulting potential ranges are probably not feasible, at least not within the timeframe considered in this study.

As the figure shows, case 3b Pyr-FCC has a notably lower potential than all other studied technology tracks (0.27-0.45 TWh biofuel per year). The reason is that for this case, the total FCC capacity in Sweden in combination with the maximum blend-in potential (set at 3-5 wt% PO) was assumed to be limiting. Without the FCC capacity restriction, the Pyr-FCC case would have comparable potential as case 3a Pyr-HDO.

When making a pairwise comparison over feedstocks, the biofuel potentials for the forest residue-based tracks (3 and 4) are in the same order of magnitude, with the exception of the aforementioned case 3b Pyr-FCC. These cases are also subject to the same key limitations (forest residue potential). Conversely, for the kraft lignin-based tracks (1 and 2), case 1 MSL-HDO shows significantly lower potential than case 2 BLG, due to limitations on the maximum lignin extraction rate. The potentials of the lignin-based cases (1 and 2) can also be noted to be limited by fundamentally different aspects. Where a key assumption for the BLG track is the transition from recovery boiler to gasification based chemical recovery, which means that the number of mills possible to convert to biofuel production constitutes the principal restriction, the corresponding key assumption for the MSL-HDO track is the preservation of core pulp mill functionality (recovery boiler), which makes the technical extraction potential the principal restriction.

In summary, the overall range of biofuel production potential from the studied technology tracks (with the exception of 3b Pyr-FCC) thus corresponds to 4.4-27 TWh/y, although a more feasible upper limit given the timeframe evaluated here would be around 12-15 TWh/y. It can also be noted that the technology tracks with the highest TRL, again with the exception of 3b Pyr-FCC, are also the tracks with the highest production potentials (case 2 and 4).

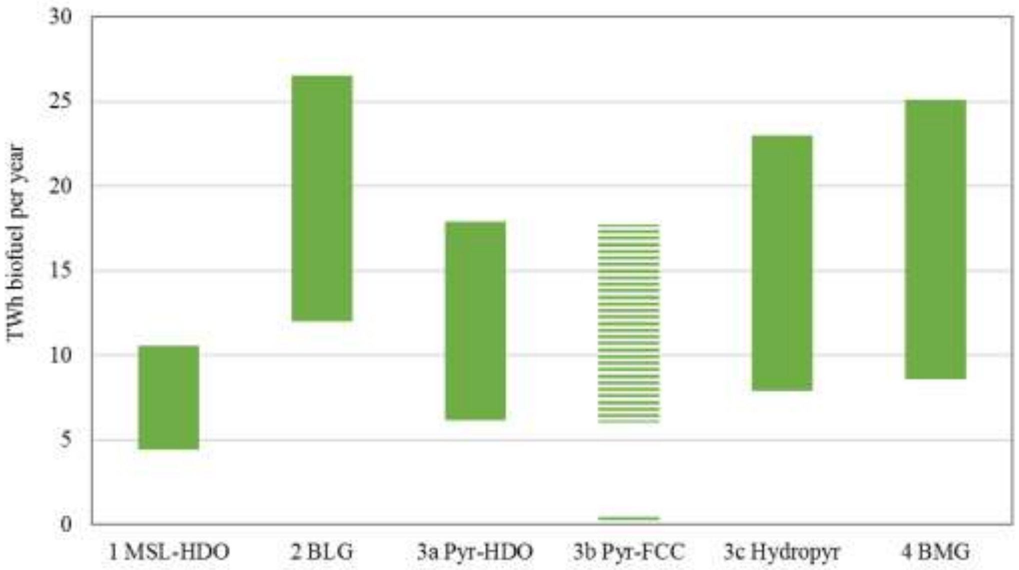

Figure 23. Resulting ranges of total production potential in Sweden from the considered technology concepts. For case 3b (co-processing of PO in FCC), the FCC and co-blending capacities are assumed to set the limitation. Nevertheless, the total capacity assuming no FCC capacity restriction is also shown (hatched).

4.6 INVESTMENT OPPORTUNITY

This section builds on the specific investment margins, calculated for each case and scenario in section 4.2. First, literature estimates of investment costs are presented (4.6.1) and compared with the available specific investment margins (4.6.2). Second (4.6.3), the effect of plant size on the feasibility of investments is assessed by scaling investment cost estimates from the literature.

4.6.1 Investment cost estimates

With the exception of case 1 MSL-HDO and case 3b Pyr-FCC, multiple cost estimates were found in the literature for all the cases. An attempt was made to pick at least two sources that were independent of each other for each, though that was not always possible.

Table 22 and Table 23 give investment cost estimates for cases 1 and 3.Only a FOAK estimate was available for the lignin oil production unit in case 1 MSL-HDO, originating from the technology developer through Anheden et al [32]. This was used together with a NOAK estimate for the total installed costs of an HDO based HTL oil upgrading unit [99]. Given that the cost of the lignin oil production facilities stand for more than 85 % of the total investment cost, the estimate was treated as an FOAK plant. Owing to doubts over the accuracy of the installed costs for the HDO unit presented in the source study, the installed costs used were instead calculated from the provided uninstalled costs by multiply them with the ratio of the installed and uninstalled costs for another hydrotreating unit in the same source. The cost of the hydrogen production unit was excluded from the total investment cost in the estimates for case 3a [74,100,101].

No estimates could be found for case 3b Pyr-FCC, which can be broken down into two steps: pyrolysis production and bio-oil upgrading. Several investment cost estimates for the first step, which is also common to the Pyr-HDO case, are available and two different sources were selected for evaluation. Cost estimates of modifications that would enable co-feeding of HDO and VGO in existing FCC units were not found in the open literature. Based on the knowledge that the FCC itself requires only minor modifications (see section 3.3), the most important of which are changes to the in-take, the tanks and the feed system, the cost of required modifications was estimated to be roughly 70 MSEK for a system with 500 GWh/y FPO feed (around 60 MW). This estimate is very uncertain, but as can be noted from Table 23, the required investment for the FPO production part of the process is much larger, thereby making the total investment for this case relatively insensitive to the estimated cost of FCC modifications.

Table 24 gives investment cost estimates for cases 2 and 4. For case 2 BLG, nth-of-a-kind specific investment costs were provided for several plant sizes in the original source [10]. The specific cost for the smallest plant size was used as it was the point from which data was scaled in that study.

It is worth remarking that as investment cost estimates were collected from several independent sources, underlying assumptions such as contingencies and indirect costs sometimes differed greatly and were not always made explicit. Adjusting the estimates so that they would follow a standardised method was considered both challenging and outside the scope of the present work. The specific investment costs used in the economic comparison thus incorporate the effects of the above-mentioned methodological discrepancies and the interpretation of the result should be made with this in mind.

Table 22 Details of source studies for estimation of specific investment cost - Cases 1 and 3a.

Case	**1 MSL-HDO**	**3a Pyr-HDO**			
Reference	Anheden et al [32], Tews et al [99]	Jones et al. [74]	Shemfe et al [102]	Dutta et al [100]	Carrasco et al [101]
Total Capital Investment (from sorurce)	400 MSEK$_{2017}$+ 6.4 MUSD$_{2014}$^{A}	578 MUSD$_{2011}$	17 MGBP$_{2014}$	480 MUSD$_{2014}$	354 MUSD$_{2016}$
Total Capital Investment (harmonized)	456 MSEK$_{2017}$	5347 MSEK$_{2017}$	184 MSEK$_{2017}$	4218 MSEK$_{2017}$	3070 MSEK$_{2017}$
Biofuel Production Capacity [MW HHV]	34	275	7	270	168
Specific capital investment [MSEK$_{2017}$/MW]	15.0	23.6	26.3	15.6	18.3
Technology Learning	FOAK/NOAK	NOAK	NOAK	NOAK	NOAK

^{A} 400 MSEK (2017) is the cost of the lignin depolymerisation step, 6.4 MUSD (2014) is the cost of the HDO step.

Table 23 Details of source studies for estimation of specific investment cost - Cases 3b and 3c.

Case	**3b Pyr-FCC^{A}**		**3c Hydropyr**		
Study	Benjaminsson et al [69] -modified	Jones et al [74] - modified	Tan et al [31]	Meerman and Larson [103]	Meerman and Larson [103]
Total Capital Investment (from source)	301 MSEK$_{2013}$+ 67 MSEK$_{2017}$	607 MUSD$_{2011}$+ 67 MSEK$_{2017}$	264 MUSD$_{2007}$	306 MUSD$_{2014}$	612 MUSD$_{2014}$
Total Capital Investment (harmonized)	685 MSEK$_{2017}$	2747 MSEK$_{2017}$	2595 MSEK$_{2017}$	2163 MSEK$_{2017}$	4852 MSEK$_{2017}$
Biofuel Production Capacity [MW HHV]	56	275	260	493	493
Specific capital investment [MSEK$_{2017}$/MW]	12.2	10.0	10.0	4.4	9.8
Technology Learning	NOAK/NOAK	NOAK/NOAK	NOAK	NOAK	FOAK

^{A} Data is divided into FPO production and refinery modification, see text.

Table 24 Details of source studies for estimation of specific investment cost - Cases 2 and 4.

Case	**2 BLG**		**4 BMG**	
Study	Andersson et al [9]	Andersson et al [10]	Hannula et al [86]	Udengaard et al [85]
Total Capital Investment (from source)	345 MEUR_{2014}	141 MEUR_{2014}	328 MEUR_{2013}	410 MEUR_{2015}
Total Capital Investment [MSEK_{2017}]	3349 MSEK_{2017}	1371 MSEK_{2017}	2509 MSEK_{2017}	4032 MSEK_{2017}
Biofuel Production Capacity [MW HHV]	110	82	229	447
Specific capital investment [MSEK_{2017}/MW]	30.4	16.7	11.0	9.0
Technology Learning	FOAK	NOAK	NOAK	NOAK

4.6.2 Specific investment costs and available investment margins

Figure 24 shows the specific investment margins under the three energy market scenarios investigated in this report (orange bars) and compares them with three annualized specific capital cost estimates for each case (green bars). Two of these estimates were based on IRRs of 15 % and 10 %, corresponding to annuities of 0.16 and 0.12, respectively (see 2.1.2). The third annualized specific capital cost estimate evaluates the effect of technology maturity on the economic viability of the cases. This is done by correcting each specific capital cost estimate using a plant availability factor indexed to the TRL assessment. Thus, the cases with the lowest TRL scores, namely 1 MSL-HDO, 3a Pyr-HDO and 3c Hydropyr, were assigned the lowest plant availability factor of 0.80, while the corresponding figures for 2 BLG, 3b Pyr-FCC and 4 BMG were 0.90, 0.85 and 0.95, respectively (see section 4.4 for a discussion of case-specific technology maturity).

Some of the cost estimates used in the comparison were developed for first-of-a-kind (FOAK) plants and costs for these are generally significantly higher than for nth-of-a-kind (NOAK) plants. Hence, a distinction is made between FOAK and NOAK estimates in the discussion. Note that the specific investment costs are not corrected for plant size, which means that certain cases benefit disproportionately from the economy-of-scale effect. The effect of scale is discussed further in section 4.6.3 below. It is immediately obvious from Figure 24 that although the available specific investment margins for BMG are not the highest among the cases evaluated, they easily cover the annualized specific investment costs from both source studies in all three energy market scenarios. However, the case benefits from the use of NOAK plant costs and from the relatively large plant sizes used in the source estimates.

Similarly, the specific investment cost for a NOAK BLG plant also compares favourably with the available investment margins in all the scenarios. Even in this case, the margin could conceivably have benefitted from economies-of-scale, as the plant capacity that is used as the basis for the investment cost calculation is on the lower side for the technology concerned. Somewhat surprisingly, even the FOAK plant appears to be competitive, albeit marginally and only at a lower annuity, in the 2030-CP energy market scenario.

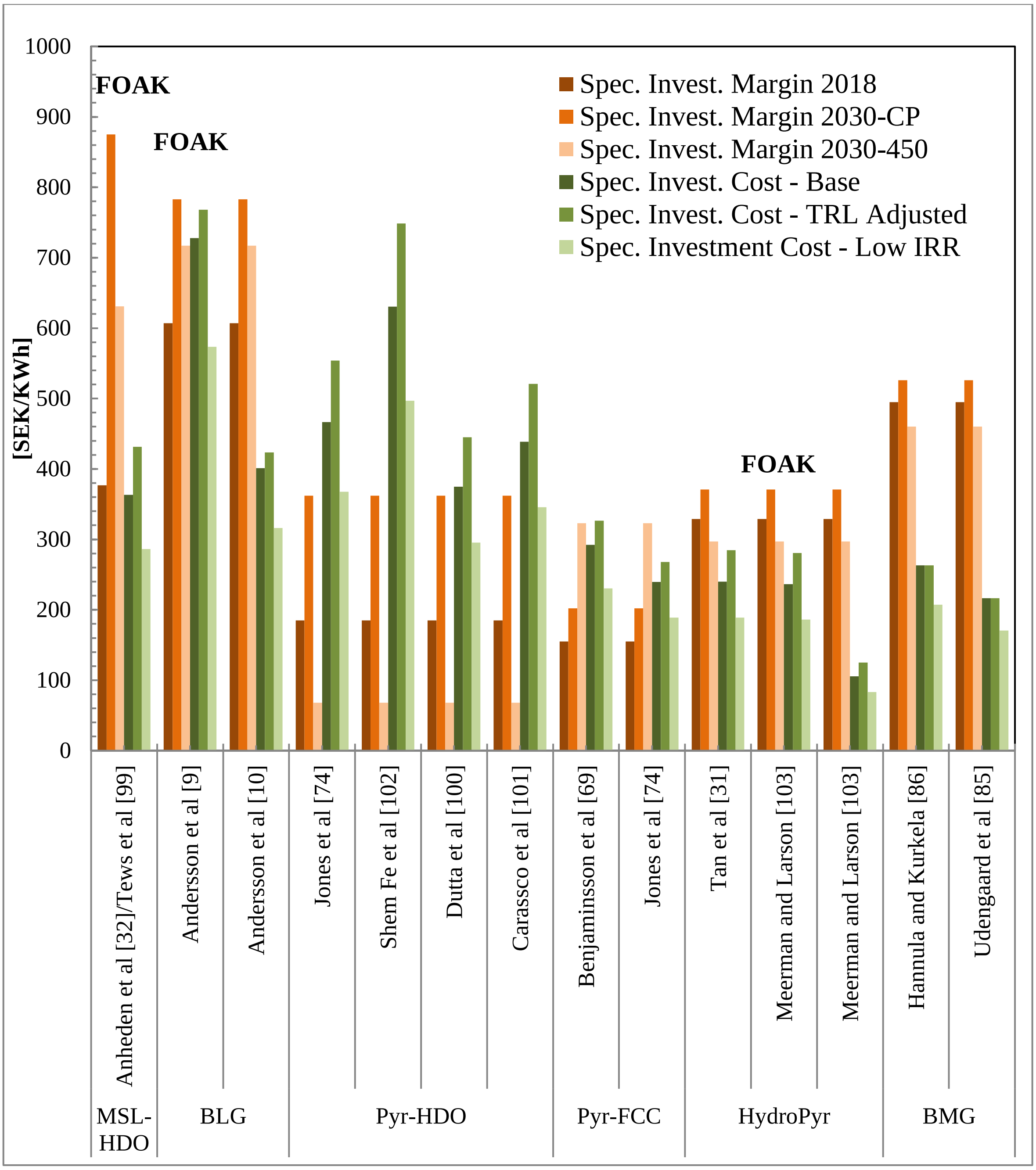

Figure 24. A comparison of specific investment costs for all cases with available specific investment margins in a variety of energy market and financial scenarios. FOAK denotes first-of-a-kind plant cost estimates.

For the Hydropyr case (3c), the prospects of covering investment costs seem good in general. The specific investment cost for the very large Meerman and Larson [103] NOAK plant was approximately half that of the available margin in all three scenarios. The specific NOAK investment cost for Tan et al. [31] was significantly higher than Meerman and Larson [103] NOAK, although it could still be covered by the available margin. The production capacity of the Meerman and Larson [103] estimate was nearly twice that of Tan et al. [31], which may explain some of the difference. The FOAK Meerman and Larson [103] estimate, when adjusted to account for the relatively low technology readiness level, also appeared to come out favourably for this case.

The annualized specific investment costs for the Jones et al., estimate in Pyr-FCC case were either lower than or in the vicinity of the available specific investment margin in all the scenarios considered. The large effects of changes in fossil yields and the uncertainty associated with them, as discussed in section 4.2 above, should be kept in mind when interpreting the economic viability of this case. It can be noted that the specific investment costs for the Benjaminsson et al. [69] estimate were somewhat higher than for Jones et al. but even in this instance the 2030-450 energy market scenario came out as economically viable in the base and the low-IRR scenarios. It is worth noting that despite very similar results, the biofuel production capacity in the Benjaminsson et al. [69] estimate is approximately 1/5th that of in the Jones et al. [74] estimate. Consequently, this is another instance of a case that would stand to benefit greatly from economies-of-scale.

None of the four different Pyr-HDO (3a) estimates appeared to be economically viable in the current energy market. On the other hand, the margin in the future 2030-CP scenario was more favourable to the specific investment costs of the three medium-sized estimates. It must be noted that the differences in the literature investment cost estimates span a wider range than can be explained by scale alone and, hence, that the investment costs for this technology must be considered very uncertain, which is unsurprising given its relatively low technical maturity. Still, the general conclusion from the analysis of the data for case 3a Pyr-HDO in Figure 24 is that it is questionable if this track can be considered economically viable in the form analysed here.

Even though the annualized specific investment cost in the MSL-HDO case was derived from a FOAK estimate, it was covered by the specific investment margins in all three energy market scenarios. The case was also on the threshold of economic viability even when the investment estimate was adjusted to reflect the case’s low TRL score in the weakest link assessment .It is likely that the invest cost for an MSL-HDO plant would be lower if the effect of technology learning was taking into account. However, the cost estimate for the lignin oil production unit used here represents the only such figure available in the open literature. By analogy with the discussion about case 3a Pyr-HDO, a case which is technically similar to case 1 MSL-HDO and has a similarly low technology maturity, it is suspected that the only available estimate may have an enhanced risk of cost escalation. Thus, the (only) capital cost estimate available for case 1 MSL-HDO is fairly uncertain.

4.6.3 Plant size and economies of scale

The effect of economies-of-scale on the specific investment cost of both NOAK and FOAK plants is illustrated in Figure 25 and Figure 26. Specific investment margins in the 2018 energy market scenario are also included for comparison. Note that two of the four Pyr-HDO estimates were excluded from the analysis: one was clearly financially unviable, while the other had yielded a specific investment cost that was broadly similar to the Dutta et al., estimate that was selected for further analysis.

For each estimate, the effect of scale is indicated by the span of the error bars on the specific investment cost column in Figure 25 and Figure 26, which corresponds to the range 25-600 MW biofuel production. It has to be emphasized that the scaling has been carried out in a simplistic manner using an exponent of 0.65 for the aggregate investment. Given the heterogeneity and uncertainty of the original investment cost estimates, attempting the use of a more thorough approach is not believed to add more meaningful insights. It is interesting to note that the specific NOAK investment costs, after correction for scale given as ranges in Figure 25, are very similar for a plurality of technologies. The range of 200-600 SEK/MWh in the size range of 25-600 MW scale is common for most cases and investment cost sources. The exceptions are the two Pyr-HDO estimates (3a), the Benjaminsson et al [69] Pyr-FCC (3b) estimate and the Meerman and Larsson Hydropyr (3c) estimate. Of these, the Pyr-HDO (3a) estimates had specific investment costs that were significantly greater than those for the other two estimates.

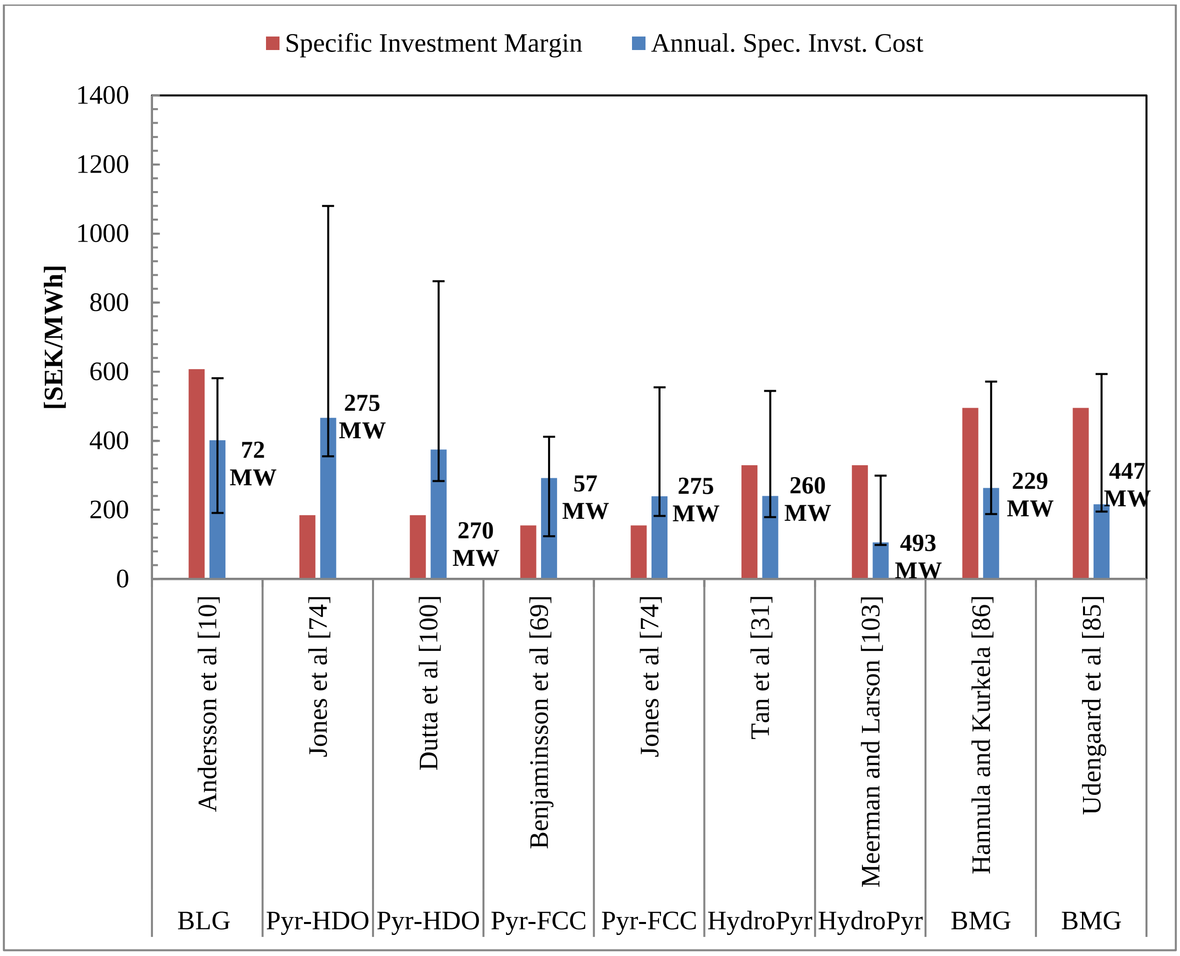

Figure 25. Effect of plant size on economic viability in the 2018 energy market scenario for cases with NOAK Specific Investment Cost estimates. Red bars: specific investment margin (from 4.2 above). Blue bars: capital cost estimate with specified plant size for, which the estimate is valid and "error bars" showing scale specific capital costs for 25 MW and 600 MW biofuel production capacity.

The approximate plant size at which each NOAK and FOAK estimates would reach break-even in the 2018 energy market scenario is given in Figure 27 and Figure 28, respectively. These sizes can be seen as the smallest economically viable scale for the selected cases, while the accompanying total investment costs provide an idea of the magnitude of the investment required in absolute terms. Again, uncertainties in investment cost estimates means that the absolute values needs to be treated with a high degree of caution. Also it should be noted that the generic approach used for scaling investment costs in this work does not take into account technology specific factors, such as equipment operating in parallel due to scale limitations. Likewise, it does not account for effects that can occur at small scale, which means that investment cost at small scale may be higher than indicated by the calculations, especially for cases when the original estimate is at very large scale.

An immediate conclusion from the analysis of the NOAK estimates is that the economic case for the cases 2 BLG, 4 BMG and 3c Hydropyr is convincing over a wide range of plausible plant sizes. Feasible plant sizes <50 MW indicated in Figure 27 for these cases are not to be interpreted as a size for which a plant would be designed but merely indicates that all realistic plant sizes would be profitable under the simplified assumptions used in the analysis.

Another interesting result is that, relatively small BLG units still appear to be viable even if the absolute numbers in Figure 27 should be treated with caution as noted above. This can represent an opportunity for pulp mills to de-bottleneck their existing recovery boilers by producing methanol from surplus black liquor for the transportation market. Similarly, the break-even threshold for Hydropyr appears to be no higher than 100-120 MW, which improves prospects for the establishment of medium-sized facilities in areas where large quantities of biomass may not be easily or cheaply available. The numbers for case 3a Pyr-HDO look less encouraging. As evident from Figure 25 and Figure 27, neither of the two estimates appears to be economically viable at realistically achievable plant sizes. The Carassco et al [101] estimate yielded a similar result (not shown), with a breakeven plant size of above 2000 MW.

The most promising case among the FOAK plants evaluated appears to be 1 MSL-HDO. The break-even plant size is ~30 MW. From an economic perspective, much like in case 2 BLG, the viability of small scale MSL-HDO facilities (case 1) denotes a good opportunity for pulp mills to expand their revenue base. The break-even sizes for FOAK BLG and Hydropyr plants, ~400 MW and ~200 MW, respectively, are larger than what would typically be considered for a pioneer plant, indicating that financial support for a FOAK plant using any of these technologies would likely be required.

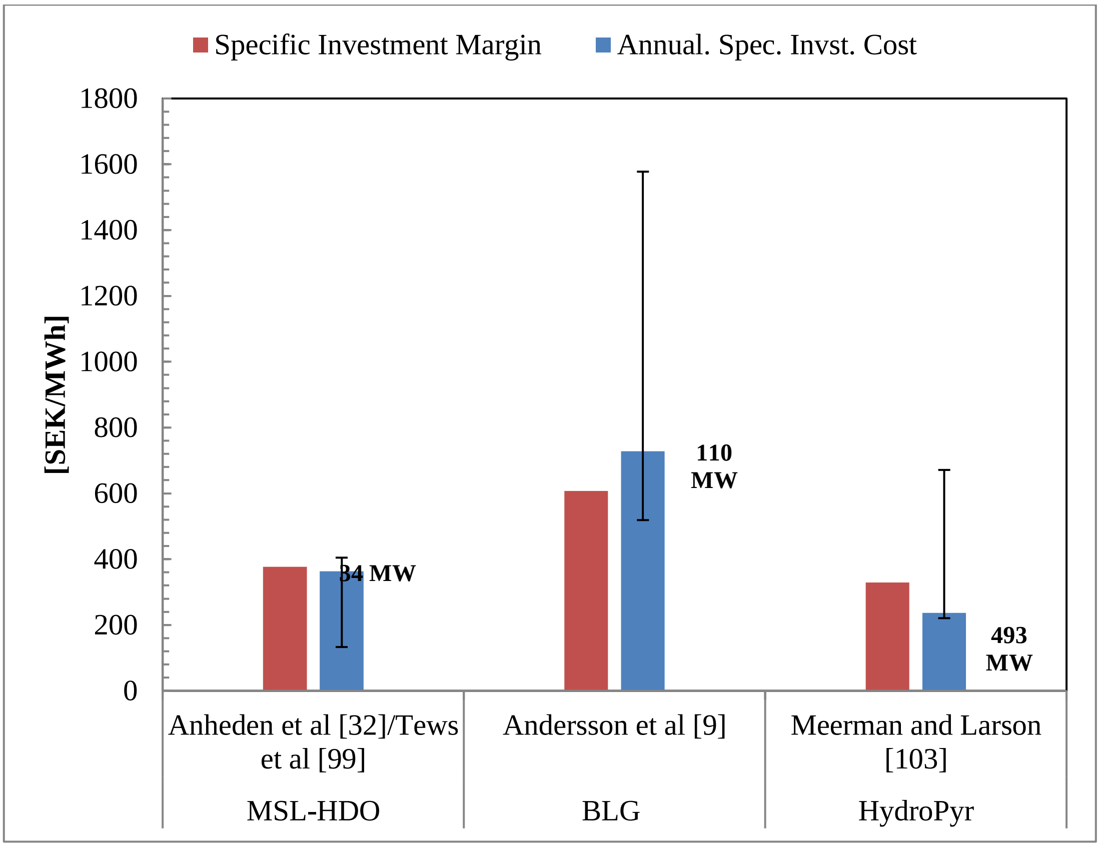

Figure 26. Effect of plant size on economic viability in the 2018 energy market scenario for cases with FOAK Specific Investment Cost estimates. Red bars: specific investment margin (from section 4.2 above). Blue bars: capital cost estimate with specified plant size for, which the estimate is valid and "error bars" showing scale specific capital costs for 25 MW and 600 MW biofuel production capacity.

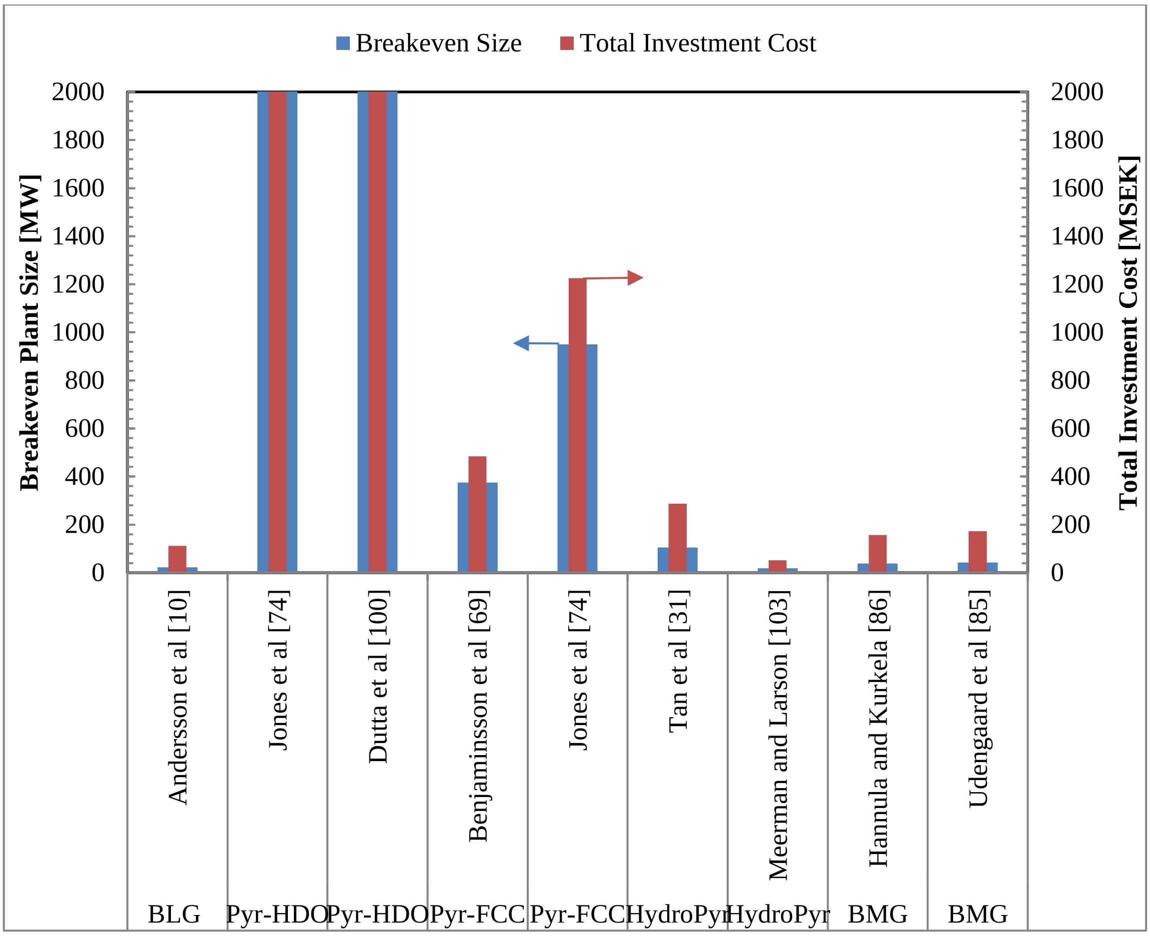

Figure 27. Total investment cost for nth-of-a-kind estimates at the breakeven plant size. Red bars specify total investment cost in 2017 MSEK, while the blue bars represent breakeven plant size in MW biofuels. Breakeven plant size and total investment for Jones et al. [74] (not visible on the figure) are 6700 MW and 10315 MSEK, respectively.

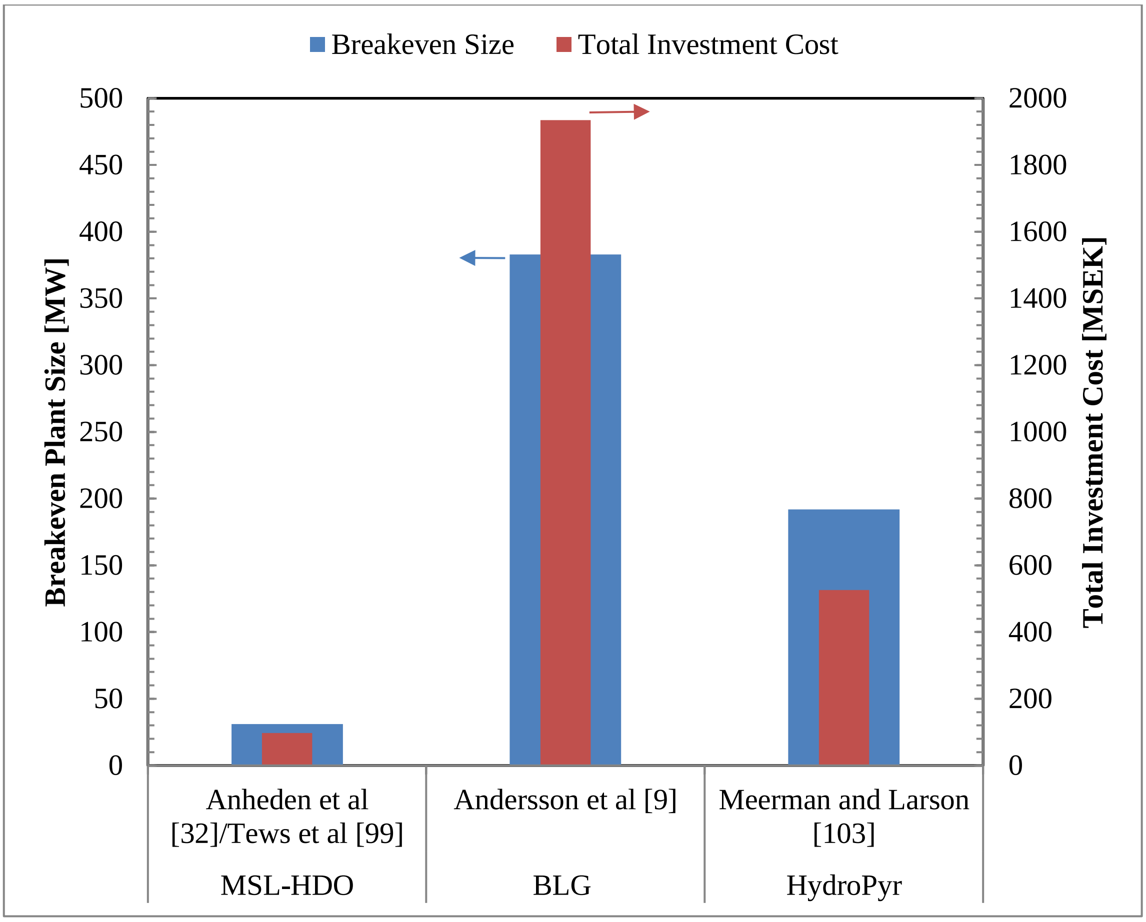

Figure 28. Total investment cost for first-of-a-kind estimates at the breakeven plant size. Red bars specify total investment cost in 2017 MSEK, while the blue bars represent breakeven plant size in MW biofuels.

5 CONCLUSIONS

The primary aim of this project has been to deliver a comparative techno-economic evaluation of four different technology tracks for the production of biofuels from forest-based feedstocks. The technology tracks were selected to be able to compare refinery integration-based drop-in fuel production with gasification-based production of methanol. Drop-in fuel production has been pointed out as a short-term strategic priority by the Swedish Energy Agency, while gasification-based production of methanol (and of similar fuels such as SNG and DME) is considered a long-term strategy due to higher biomass feedstock utilisation efficiency. Two biomass feedstocks were considered: black liquor (or kraft) lignin and forest residues.

While the main focus of the project was on techno-economic evaluation, an overall assessment of the technology tracks was also performed and the most important results are summarized in Table 25. Generally, the scale used in the assessment (+, 0, -) is relative and shows the merits of each case compared to those of the others. The only exception to this is GHG performance, which is measured on an absolute scale against RED sustainability criteria.

Table 25. Overall assessment of the technology tracks investigated in this study.

	Energy efficiency	Profitability [a]	Investment requirement	Potential	GHG perform.[b]	Technology maturity
1 MSL-HDO	+	+	+	0	0	-
2 BLG	+	+	0/- [c]	+	+	+
3a Pyr-HDO	0	-	-[d]	+	-	0
3b Pyr-FCC	-	-	+ [c]	-	+	+
3c Hydropyr	0	+	+	+	+	0
4 BMG	0	+	+	+	+	+

[a] Based on aggregation of results in Table 26.

[b] GHG performance assessment was made against RED sustainability criteria rather than on a relative basis.

[c] Assessment of investment requirement for case 2 BLG and case 3b Pyr-FCC were not made based on economies of scale. A relatively large investment requirement exists for case 2 BLG because of the fact that that the investment cost estimate is based on the assumption that the recovery boiler is fully replaced with a gasifier. For case 3b Pyr-FCC, the investment size is limited by the capacity of the FCC in the refinery.

[d] Very varying specific investment cost estimates were found for this case, the assessment reflects the overall picture.

Various measures of energy efficiency were calculated for each case to evaluate performance from different viewpoints. From a systems perspective, the most relevant mode of assessment is based on system expansion that includes the effect of refinery and mill integration, which is thus what the overall efficiency assessment shown in Table 25 is based on. The good performance for the cases that utilise black liquor lignin as direct feedstock (cases 1 and 2) is mainly due to the assumed integration with a pulp mill with energy surplus, which can be converted to biofuels with high efficiency instead of to electricity with low efficiency. It can be noted that in the alternative case of integrating these processes with a mill that is energy deficient and does not export electricity would give different performance since this advantage would not be present. For case 2 BLG, efficient use of excess heat from the biofuels plant in the pulp mill is also an important factor. The poor efficiency for case 3b Pyr-FCC is due to the complex and uncertain influence of fossil yields, which can increase the fuel oil demand in the refinery. Hence, a more detailed assessment is recommended for stakeholders with a particular interest in this track.

The profitability assessment in Table 25 is an aggregation of the assessments summarized in Table 26. It can be concluded that practically all technology tracks benefit from the projected changes in energy prices in the 2030 scenarios compared to 2018. A major factor behind this is the projected increase in biofuel product price from the current 700-800 SEK/MWh to approximately 1100 SEK/MWh, which more than compensates for price increases in the process inputs, including a 100 % increase in the price of forest residues and a ~ 50 % increase in the electricity price. Investment margins are highest for cases 2 and 4, while the investment size is smallest for cases 1 and 3c; all of the cases 1, 2, 3c and 4 are projected to give good profitability for a NOAK plant investment both in 2018 and 2030.

Technologies based on integration of FPO in an oil refinery, represented by cases 3a Pyr-HDO and 3b Pyr-FCC, do not appear to be feasible economically as evaluated in this work. Using higher grade feedstock than forest residue, such as saw dust, would increase yields and thereby potentially improving economics. However, with such a feedstock change, production potentials used in this study would no longer be valid.

In addition to profitability, the absolute size of the investment is an important aspect in realizing biofuels plants in Sweden, especially for FOAK plants, for which the technical risk is higher. For FOAK plants, investment estimates were available for case 1 MSL-HDO, case 2 BLG and case 3c Hydropyr. This data is very uncertain but indicates that FOAK plants for MSL-HDO can be built without substantial support once the technology is sufficiently mature, while this is not the case for case 2 BLG and probably not for case 3c Hydropyr either.

Economies-of-scale mean that in the case of NOAK plants, all technologies with good indicated profitability can also be profitable at relatively small scale, thereby indicating that large investments are not necessarily required for viable returns. However, for BLG, the technology investigated assumes that the pulp mill recovery boiler is replaced completely with the gasification-based chemicals recovery and, hence, the size of the pulp mill will in practice determine the size of the biofuels plant.

The fact that the results indicate that gasification-based biofuels production can also be viable in small scales is interesting and goes against the common perception that gasification is only economically feasible in large scales. However, it must be noted that the scaling approach used was quite rough and that NOAK investment estimate for case 2 BLG benefits from the inclusion of an investment credit as the gasifier in the source study was assumed to replace an end-of-life recovery boiler. Instead of replacing the recovery boiler, smaller gasifiers could help debottleneck pulp mill chemical recovery. It is thus conceivable that smaller gasifiers could be given a so-called debottlenecking credit. This seems to motivate further work on the feasibility of biofuels plants based on a fraction of the black liquor flow, thus supporting debottlenecking the recovery area of the pulp mill in a manner similar to that for lignin extraction, e.g. as exemplified by case MSL-HDO, which would also significantly reduce the technology risk.

Table 26. Economic assessment of the technology tracks investigated in this study.

	Investment margin 2018	Investment margin 2030	Profitability 2018[a]	Profitability 2030[a]
1 MSL-HDO	0	+	0/+ [b]	+
2 BLG	+	+	+	+
3a Pyr-HDO	-	0/-	-	-
3b Pyr-FCC	-	0/-[c]	-	0/- [c]
3c Hydropyr	0	0	+	+
4 BMG	+	+	+	+

[a] The assessment concerns mature (NOAK) technology in the respective scenario.
[b] Investment assessment is only available for a FOAK plant. Profitability can be expected to be better for NOAK plant.
[c] Depending on scenario; 2030-CP or 2030-450, see **Fel! Hittar inte referenskälla.**.

The technical biofuel production potentials for all cases using forest residues as feedstock or as feedstock replacement (case 2 BLG) were found to be around the range 10-25 TWh/y. Compared with the rest, case 3a Pyr-HDO had a slightly lower potential because of a lower conversion efficiency. For case 1 MSL-HDO, the potential is limited by the lignin supply, while for case 3b Pyr-FCC, the current potential is constrained by a combination of restrictions on allowed bio oil co-processing quantities and installed cracker capacity. The production of 10-25 TWh/y of biofuel from domestic feedstocks is a chance to strongly contribute to the transformation of the transportation system. The ranges obtained in this study also appear to be in agreement with a number of 2030 biofuel demand projections, see for example Johansson [2]. It must be noted that the potentials presented here do account for current use of forest residue but not future increased use in other sectors. Future feedstock competition will be highly dependent on technical developments and strategic decisions in carbon dioxide intensive industries, such as steel, cement and chemicals, and the scope and effect of these decisions is very difficult to quantify today.

GHG performance was generally found to be satisfactory for most production technologies with GHG savings of 83-95 % compared to a fossil reference for cases 2, 3b, 3c and 4, when calculating by the RED method. For the cases that use natural gas-based hydrogen, case 1 and 3a, the GHG savings were only 70 % and 66 %, respectively. This means that these technology tracks, for the configurations evaluated in this report, clearly risk not meeting future RED sustainability criteria. This result is in contrast to HVO production that, because of lower hydrogen consumption, can probably be based on natural gas-based hydrogen and still meet the GHG threshold in the future RED criteria. For cases 1 and 3a investigated here, either carbon capture or another hydrogen production technology with lower carbon footprint would represent a means of meeting the specified threshold. It should be noted that this will, however, risk increasing the production costs and, hence, the economic assessment for these tracks may not be valid in such a case. An assessment of alternative hydrogen production technologies is presented in Anheden et al [32] and appears to be a very relevant topic for further research.

System expansion represents an alternative and more holistic way to calculate GHG performance, but that is not currently implemented in RED and national legislation concerning sustainability criteria. Calculating GHG performance with the system expansion approach gives a dramatically different picture for cases 1 and 3 with carbon negative footprints. This is caused by the use of excess heat from the HDO process in the fossil refinery processes to replace fuel oil combustion.

The technology maturity assessment based on the full production chains shows TRLs of 6-7 for cases 2 BLG, 3b Pyr-FCC and 4 BMG, indicating that the next step for these tracks is commercial scale demonstration. When the weakest link is considered, TRLs are a bit lower but still above 5 for cases 2 and 4 and above 4 for case 3b. For case 3b Pyr-FCC, UOP has communicated that there are commercial demonstrations in planned or in progress, meaning that the TRL may be higher than can be adjudged from the open information on which the analysis in this report is based. For cases 1 MSL-HDO, 3a Pyr-HDO and 3c Hydropyr, TRLs from a weakest link perspective are below 4, indicating that substantial R&D efforts are needed to bring those to a point at which commercial deployment becomes feasible.

In summary, an apparent paradox can be identified in that drop-in biofuel producing technology tracks that generally use technologies with the lowest technology maturity constitute the prioritized short-term tracks. There is currently substantial R&D effort being spent on these technologies. However, results would need to be obtained very quickly in order to enable these to be employed commercially to any substantial degree in a time span that can be considered consistent with a short term perspective. It can be noted that according to the efficiency measures used in this work, the technology track represented by case 1 MSL-HDO can potentially yield efficiencies in the same range as gasification technologies provided issues around GHG performance are resolved as discussed above, and provided that that ongoing technology development is successful and leads to technological demonstration at a level that meets the assumptions made in this study.

In addition, the results presented in this report have been used to draw conclusions about prioritized R&D&D efforts for the technologies and production pathways investigated as shown in Table 27.

Table 27. Prioritized R&D&D for the technologies investigated.

Case	Prioritized R&D&D
1 MSL-HDO	General technology and catalyst development to improve TRL, confirm yields, etc., followed by pilot scale demonstration; integration of alternative hydrogen production technologies for HDO
2 BLG	Commercial demonstration; investigation of technical and techno-economic feasibility of smaller-scale plants based on a fraction of black liquor at a mill, e.g. for recovery boiler debottlenecking
3a Pyr-HDO	General technology development to improve TRL of in particular the HDO-process step and improve process economy through for example higher yields or lower equipment cost; development and evaluation of alternative hydrogen production technologies for HDO. Expansion of feedstock base to include other biomass options such as saw dust with higher yields.
3b Pyr-FCC	Commercial demonstration; R&D in order to better understand yields in commercial scale equipment, R&D to increase maximum FPO fraction in FCC feed in order to increase potential. Expansion of feedstock base to include other biomass options such as saw dust with higher yields.
3c Hydropyr	Pilot scale demonstration to increase TRL (underway in India at the time of writing this report); R&D to better understand behaviour of residue feedstock
4 BMG	Commercial demonstration; investigation of technical and techno-economic feasibility of smaller scale plants

NOMENCLATURE/ABBREVIATIONS

Abbreviation	Definition
AGR	Acid gas removal
ASU	Air separation unit
BFB	Bubbling fluidized bed
BLG	Black liquor gasification
BMG	Biomass gasification
BOP	Balance of plant
CFB	Circulating fluidized bed
CGE	Cold gas efficiency
CPO	Catalytic pyrolysis oil
FBG	Fluidised bed gasification
FCC	Fluid catalytic cracker *or* Fluid catalytic cracking
FOAK	First of a kind
FPO	Fast pyrolysis oil
HHV	Higher heating value
HDO	Hydrodeoxygenation
HT	Hydrotreatment
HPS	High pressure steam
LCO	Light cycle oil
LGO	Light gasoil
LHV	Lower heating value
LO	Lignin oil
LPS	Low pressure steam
MPS	Medium pressure steam
NCG	Non-condensable gases
NOAK	Nth of a kind
TOFA	Tall oil fatty acid
VGO	Vacuum gasoil
WGS	Water gas shift

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APPENDIX A. REFINERY PROCESS NOMENCLATURE

Table 28. Refinery process nomenclature

Refinery process	Description
Cracking	Elimination reactions to break large molecules into smaller that are easier to process further. Can be thermal-, catalytic-, or hydro-, see below.
Thermal cracking	Thermal decomposition at ~500 °C. Forms coke due to hydrogen deficit.
Catalytic cracking	Catalytic decomposition that gives higher yields than thermal cracking. Often implemented as fluid catalytic cracking (FCC) using zeolite based catalyst with coke burnt off during catalyst regeneration. The zeolite catalyst can also deoxygenate biomass intermediates through elimination of CO, CO2.
Hydrocracking	Catalytic cracking in presence of hydrogen, which reduces coke formation.
Hydroprocessing	Reactions in the presence of hydrogen, general [73] – sometimes integrated process of hydrotreatment and hydrocracking [104].
Hydrotreating	Catalytic process to remove S, N, O, Cl as H2S, NH3, H2O, HCl. Also involves saturation of alkenes. Often relatively carbon efficient since heteroatoms are removed with hydrogen but at the expense of hydrogen consumption [104].
Hydrodeoxygenation (HDO)	Hydrotreating specifically aimed at removing oxygen.

APPENDIX B. DETAILED TRL ASSESSMENT

Table 29 and table 30 give detailed TRL estimates, including TRL for individual process steps and weights used. See sections 2.3 and 4.3 for discussion.

Table 29. Detailed TRL assessment for cases 1, 2 and 4.

Case	Process step	TRL		Weight	Comment	References
		Low	High	%		
1 MSL-HDO	Ultrafiltration	5	6	20	1.5 kg/h pilot operating in pulp mill environment >1000 h	SunCarbon[A]
	Depolymerisation via HTL	3	4	25	Batch and flow reactors tested in lab environment. Optimisation ongoing.	
	Purification of crude “lignin-oil”	3		20	Ongoing initial work in lab-scale	BioLi project[B]
	Hydrodeoxygenation and cracking	3	4	25	Work is being performed in lab scale. Concept demonstrated in lab scale for other bio-oils (see case 3)	BioLi project[B]
	Integration with pulp mill, permeate and lignin lean liquor	3	4	5	Difficult to mature when working on small slip-streams, but partly demonstrated on similar effluents streams	
	Integration with refinery	7	8	5	Relatively easy (steam and hot water, NCG combustion in refinery boiler). Proven in other applications; -(1-2) for new process source	
	Overall "weighted average"	**4.0**				
	Overall "weakest link"	**3.0**				
2 BLG	Black liquor gasification	5	6	40	Pilot operation 3 MW >26 000 h (est. 4-7 % of commercial scale)	[4,6]
	Gas cleaning and conditioning	8	9	25	Commercial technology (Rectisol and WGS), -(1-2) for biosyngas	
	Methanol synthesis and purification	8	9	25	Commercial technology (BWR), -(0-1) for biosyngas	
	Integration with pulp mill, green liquor and steam	4		10	Not demonstrated since GL flow from gasifier is negligible in pilot. GL upgrading to WL demonstrated in bench scale.	[105,106]
	Overall "weighted average"	**6.9**				
	Overall "weakest link"	**5.5**				
4 BMG	Biomass gasification and primary gas cleaning	6		50	GTI/Andritz pilot 20 t/d >3000 h	[85,107]
	Secondary gas cleaning and conditioning	7	8	25	Commercial technology (Rectisol and WGS), -(1-2) for biosyngas	
	Methanol synthesis and purification	8	8	25	Commercial technology (BWR), -1 for biosyngas	
	Overall "weighted average"	**6.9**				
	Overall "weakest link"	**6.0**				

[A] See http://www.suncarbon.se/2017/02/01/membrane-pilot-at-kraft-pulp-mill/

[B] Personal communication, BioInnovation BioLi project

Table 30. Detailed TRL assessment for cases 3a, 3b and 3c.

Case	Process step	TRL		Weight	Comment	References
		Low	High	%		
3a FPO-HDO	Fast pyrolysis of forest residues	7	8	30	Commercial operation 30 MW on pellets/sawdust, -(1-2) for forest residue feedstock	Empyro[A], Fortum[B]
	Hydrodeoxygenation	3	3	40	SOTA: 2-step HDO in lab scale <400 mL, -1 for forest residue FPO (alkali)	
	Final upgrading to transportation fuel	8	9	20	Commercial process, -(0-1) for product from new process	
	Integration in pulp mill	9		5	Only integrated through steam supply	
	Integration in refinery	7	8	5	Only integrated through use of NCG in refinery boiler and heat recovery in form of hot water and steam	
	Overall "weighted average"	6.0				
	Overall "weakest link"	3.0				
3b FPO-FCC	Fast pyrolysis of forest residues	7	8	35	See case 3a	See case 3a
	FCC co-feeding	4	5	40	Petrobras pilot 200 kg/h (<5 % of commercial scale) with approx. 250 h acc. op. time. UOP claims that commercial trials are underway but no results available. Not demonstrated with forest residue FPO (alkali).	[76,77]
	Final upgrading to transportation fuel	8		20	Commercial process, -1 for product from new process	
	Integration in pulp mill	9		5	See case 3a	
	Overall "weighted average"	6.5				
	Overall "weakest link"	4.5				
3c Hydropyr	Catalytic hydropyrolysis, incl. feeding	3	4	40	GTI pilot 50 kg/d, -(1-2) for forest residue	[61,65]
	HDO	4	5	25	GTI pilot 50 kg/d, -(0-1) for forest residue based vapours	(T. L. Marker et al., 2014; T. Marker, Linck, & Felix, 2010
	NCG reforming to H2	7	8	15	Integrated NCG reforming not demonstrated but very similar to commercial process, -(1-2) for gas from hydropyrolysis	
	Final upgrading to transportation fuel	8		10	Commercial process, -1 for product from new process	[62]
	Integration with refinery	7	8	5	Relatively easy integration with steam and hot water generation, proven in other applications	
	Overall "weighted average"	5.2				
	Overall "weakest link"	3.5				

[A] http://www.empyroproject.eu/

[B] https://www.fortum.com/en/mediaroom/pages/fortums-bio-oil-plant-commissioned-in-joensuu-first-of-its-kind-in-the-world.aspx

APPENDIX C. POTENTIAL ASSESSMENT ASSUMPTIONS

For black liquor, the technical maximum available amount of feedstock would be all black liquor from all chemical pulp mills in Sweden. The current (2015) production and use of black liquor is around 44 TWh/year [93]. As an upper limit for the total black liquor potential in 2030, an annual increase of the chemical pulp production of 1.3% from 2015 to 2030 was assumed, which is in line with the "Utvecklad bioekonomi" scenario from [108] and which would give a total black liquor volume of 54 TWh/year. This was used as the maximum technical potential for black liquor as feedstock for gasification based biofuel production (case 2), as well as for feedstock for black liquor retentate production, which in turn constitutes the feedstock in case 1. As lower value for the potential for black liquor as gasification feedstock, the current black liquor volumes in pulp mills with "old" recovery boilers (built before 1995) was used (24 TWh/year).

From the maximum technical black liquor potential, the maximum lignin extraction potential was estimated in order to give the maximum lignin retentate potential (case 1). The maximum lignin extraction potential was estimated at 30 % of all black liquor for all mills, based on assessments from Valmet LignoBoost, giving a total of 16 TWh/year. Thuresson and Andersson [108] instead assessed the lignin extraction potential for each chemical pulp mill individually, under the condition that no recovery boiler operation would be negatively affected. They arrived at a total potential of 6.7 TWh/year, which was used as the lower lignin retentate potential here.

Cases 3 and 4 both use forest residues (branches and tops, "grot") as feedstock. A number of assessments of the future domestic forest residue potential in Sweden have been published in recent years, e.g. [96,109]. Here, the "techno-ecological" potential for 2030 as estimated in SKA 15 [96] for the scenario "Today's forestry", was applied in order to estimate the upper limit for forest residues, giving a total of 49 TWh/year. In brief, the techno-ecological potential represents the total technical harvesting potential (final felling and commercial thinning), reduced by a set of restrictions and recommendations regarding environmental considerations, as devised by the Swedish Forest Agency. In order to estimate the lower limit, further restrictions on the harvesting were applied in order to avoid conflict with different environmental quality objectives that may be impacted negatively by increased use of forest biomass for energy purposes [97], which gave a lower forest residue potential of 25 TWh. The total forest residue potentials (upper and lower) were reduced by the estimated use in 2030 in other sectors (primarily for heat and electricity production), of 12 TWh/year [108], resulting in a final forest residue potential span of 13-37 TWh/year.

An additional technology constraint was added for the technology track involving co-processing of PO in the refinery's FCC (case 3b), where the total FCC capacity in combination with the maximum blend-in potential was assumed to be limiting. The total Swedish FCC capacity amounts to 1,600 kton (19 TWh) per year[16]. As boundaries for the co-blending ratio, a range of 3-5 wt% of PO in the FCC were considered. This would give a co-processing range of 0.30-0.50 TWh PO per year, which would in turn give 0.27-0.45 TWh green products (diesel and petrol) per year.

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[16] Personal communication with Åsa Håkansson, Preem (6 October, 2017).

APPENDIX D. TECHNOLOGY READINESS LEVEL DEFINITIONS

Table 31. Definitions used for TRL assessment.

TRL	European Commission Horizon 2020 program with additional explanations by Serbian Innovation Fund [a]	US DOE Fossil Energy Clean Coal Program [b]
1	**Basic principles observed.**	**Basic principles observed and reported.** Lowest level of technology readiness. Scientific research begins to be translated into applied R&D. Examples include paper studies of a technology’s basic properties.
2	**Technology concept formulated.**	**Technology concept and/or application formulated.** Invention begins. Once basic principles are observed, practical applications can be invented. Applications are speculative and there may be no proof or detailed analysis to support the assumptions. Examples are still limited to analytic studies.
3	**Experimental proof of concept.** Active research and development (R&D) is initiated. This must include both analytical studies to set the technology into an appropriate context and laboratory-based studies to physically validate that the analytical predictions are correct. These studies and experiments should constitute proof-of-concept validation of the applications/concepts formulated at TRL 2.	**Analytical and experimental critical function and/or characteristic proof of concept.** Active R&D is initiated. This includes analytical and laboratory-scale studies to physically validate the analytical predictions of separate elements of the technology (e.g., individual technology components have undergone laboratory-scale testing using bottled gases to simulate major flue gas species).
4	**Technology validated in lab.** The design, development and lab testing of technological components are performed. Following successful proof-of-concept work, basic technological elements must be integrated to establish that the pieces will work together. The validation must be devised to support the concept that was formulated earlier, and should also be consistent with the requirements of potential system applications. It is "low-fidelity" validation compared to the eventual system.	**Component and/or system validation in a laboratory environment.** A bench-scale prototype has been developed and validated in the laboratory environment. Prototype is defined as less than 5 percent final scale (e.g., complete technology process has undergone bench-scale testing using synthetic flue gas composition).
5	**Technology validated in relevant environment (industrially relevant environment in the case of key enabling technologies).** The basic technological components are integrated with reasonably realistic supporting elements so they can be tested in a simulated environment. Examples include “high-fidelity” laboratory integration of components.	**Laboratory-scale similar-system validation in a relevant environment.** The basic technological components are integrated so that the system configuration is similar to (matches) the final application in almost all respects. Prototype is defined as less than 5 percent final scale (e.g., complete technology has undergone bench-scale testing using actual flue gas composition).
6	**Technology demonstrated in relevant environment (industrially relevant environment in the case of key enabling technologies).** Representative model or prototype system, which is well beyond that of TRL 5, is tested in a relevant environment. Represents a major step up in a technology’s demonstrated readiness. Examples include testing a prototype in a high-fidelity laboratory environment or in a simulated operational environment.	**Engineering/pilot-scale prototypical system demonstrated in a relevant environment.** Engineering-scale models or prototypes are tested in a relevant environment. Pilot or process-development-unit scale is defined as being between 0 and 5 percent final scale (e.g., complete technology has undergone small pilot-scale testing using actual flue gas composition).

TRL	European Commission Horizon 2020 program with additional explanations by Serbian Innovation Fund [a]	US DOE Fossil Energy Clean Coal Program [b]
7	**System prototype demonstration in operational environment.** TRL 7 is a significant step beyond TRL 6, requiring an actual system prototype demonstration in a real relevant environment. The prototype should be near or at the scale of the planned operational system and the demonstration must take place in real environment. The goal of this stage is to remove engineering and manufacturing risk.	**System prototype demonstrated in a plant environment.** This represents a major step up from TRL 6, requiring demonstration of an actual system prototype in a relevant environment. Final design is virtually complete. Pilot or process-development-unit demonstration of a 5–25 percent final scale (e.g., complete technology has undergone large pilot-scale testing using actual flue gas composition).
8	**System complete and qualified.** Technology has been proven to work in its final form under the expected conditions. Examples include developmental test and evaluation of the system in its intended system to determine if it meets design specifications. In almost all cases, this level is the end of true system development.	**Actual system completed and qualified through test and demonstration in a plant environment.** The technology has been proven to work in its final form and under expected conditions. In almost all cases, this TRL represents the end of true system development (e.g., complete and fully integrated technology has been initiated at full-scale demonstration including start-up, testing, and evaluation of the system using actual flue gas composition).
9	**Actual system proven in operational environment (competitive manufacturing in the case of key enabling technologies).** Actual application of the technology in its final form and under target conditions, such as those encountered in operational test and evaluation. At this point, the technology is ready for commercial deployment.	**Actual system operated over the full range of expected conditions.** The technology is in its final form and operated under the full range of operating conditions (e.g., complete and fully integrated technology has undergone full scale demonstration testing using actual flue gas composition).

[a] See https://ec.europa.eu/research/participants/data/ref/h2020/wp/2014_2015/annexes/h2020-wp1415-annex-g-trl_en.pdf and https://www.inovacionifond.rs/TTF/Content/Public/08%20Explanation_for_Technology_Readiness_Level_vf.pdf

[b] References to absolute scale of CCS plants were removed as they are not relevant for the present study.

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