



# Comparative techno-economic analysis of biofuel production through gasification, thermal liquefaction and pyrolysis of sugarcane bagasse

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## ABSTRACT

Techno-economic models for three plants in Queensland, Australia using thermochemical conversion of 10 tonnes per hour of sugarcane bagasse to liquid crude biofuels were developed. Thermochemical conversion was chosen to maximise yield from second-generation feedstock. The process models highlighted the differences in mass and energy flows of each process. Factory models were generated reflecting current methods in heat and material recovery. Liquefaction generated the highest amount of product per kg feed, followed by pyrolysis and gasification. Key parameters affecting plant economics were also highlighted. Based on net present values, the profitability was ranked as follows: pyrolysis > liquefaction > gasification. The plants were all sensitive to product price, thermochemical conversion ratio and refining conversion ratio, thereby affecting profitability the most. Conversion ratios sharply affect the minimum selling price of products, but attenuated by high product volumes. Varying tax rates and capital costs do not affect the minimum selling price as much as conversion ratios, therefore, incentives around improving conversion rates and increasing product volume are recommended.

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## 1. Introduction

The increase in demand, the imperative to reduce greenhouse emissions, and the drive for sustainable production present challenges to develop alternative energy sources and production processes that consider affordability, accessibility, sustainability and equity (Sovacool and Dworkin, 2015). A significant focus has been on biofuels since transportation is responsible for 23% of energy-related CO<sub>2</sub> emissions mostly due to increasing transportation activity (Sims et al., 2014). Biofuels have been shown to be a low carbon intensity alternative to fossil fuels based on life-cycle greenhouse gas savings (Hoeftnagels et al., 2010), therefore the development of biofuel processes and products using suitable feedstock is an active research field.

Among the various feedstock identified for biofuel production, waste agricultural and process biomass have better environmental impacts over the life cycle. Compared with the use of dedicated crops, the use of residues as feedstock instead of disposal result in lower net impacts and emissions. This may offset fossil fuel inputs in biofuel processing (Cherubini et al., 2009). The use of non-edible

feedstock also avoids complications related to food supply (Tilman et al., 2009), which are present in first-generation biofuels. Moreover, the conversion of agricultural residues to biofuels is value-adding as additional revenue could be generated from selling the fuel. For instance, using residues from sugar production (i.e. bagasse) reduces potential costs related to disposal while taking advantage of the reliability of feedstock supply. Established agricultural methods and the ubiquity of sugar production enables bagasse supply to be consistent (O'Hara et al., 2013). Sugarcane bagasse can also be easily collected from the sugar factory and stockpiled to offset the effect of variations in harvesting periods (Rainey et al., 2013). Globally, it was estimated that 540 million t/y of sugarcane bagasse is produced (Bezerra and Ragauskas, 2016), demonstrating potential for large-scale biofuel production in sugarcane areas. A biofuel plant can potentially be co-located with the sugar mill to minimise transportation cost and take advantage of facilities such as steam supply, and auxiliary infrastructure to manage costs.

In Australia, there is an opportunity to use bagasse to augment biofuel production. Bagasse has been demonstrated as a reliable source of 102 PJ of heat and power in 2015–16 from 10.7 million t/y of bagasse (2017a). This represents only 1.7% of the total Australian energy mix. In contrast, biofuels (i.e. ethanol and biodiesel) only account for 0.5% of road transport fuels in the same period.

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Although biofuels have been used for transport in Australia for a decade, the reliance on fossil fuels are reflected resoundingly in the transport energy mix (2017h). This presents an opportunity to shift or add the use of bagasse to produce liquid fuels. Its potential among other biomass for biofuels was also emphasised by (Kosinkova et al., 2015a).

The use of appropriate technologies to convert solid biomass to liquid biofuels is a crucial part of the supply chain (Azad et al., 2015). Compared to first-generation biofuel processing, second-generation biofuel processing has been met with challenges around the heterogeneity of biomass composition. The processes require several methods to break the complex structures of cellulose, hemicellulose, and lignin, and extract sugars or lipids for an energy dense product. Therefore, thermochemical processes have been promoted as suitable conversion technologies for lignocellulosic materials due to its indiscriminating mechanism to convert solids to fluids that can be further refined to make fuels. Table 1 presents a comparison of three key thermochemical conversions: gasification, thermal liquefaction and pyrolysis.

The common mechanism for gasification, liquefaction and pyrolysis is breaking down the large structural biomass molecules. The high temperature in which the processes occur provides the enthalpy to break bonds. In pyrolysis and gasification, components are vapourised in the early stages, then crack into light hydrocarbons, aromatics and oxygenates (Evans and Milne, 1987). In pyrolysis, the intermediate products in gaseous form are rapidly condensed after a brief residence time (Mohan et al., 2006). Higher gasification temperatures progress the conversion to light olefins, CO and more aromatics, then to production of light gases such as CO, CO<sub>2</sub>, H<sub>2</sub> and CH<sub>4</sub> and polynuclear aromatics that form soot (Evans and Milne, 1987). Liquefaction on the other hand, takes advantage of solvent properties at high pressures and temperatures (Peterson et al., 2008) such as reduced mass transfer resistances and higher penetration of the solvent into the biomass structure (Chumpoo and Prasassarakich, 2010). Following depolymerisation of the biomass, the components decompose through cracking, dehydration, and decarboxylation and the reactive fragments recombine to produce other compounds (Toor et al., 2011).

The differences in operating conditions affect the products of the three processes. The high gasification temperature produces mainly permanent gases and tars that make up syngas, while the immediate cooling following pyrolysis results in a condensed phase and a gas product. Liquefaction produces a higher yield liquid product alongside the solvent. All of the processes produce gaseous,

liquid and solid by-products that require separation from the main product.

Another common feature of the thermochemical processes is the need for further upgrading and refining of their products into a fuel that can be readily used. Syngas needs to be cleaned of impurities prior to fuel production (Woolcock and Brown, 2013). Following cleaning and before fuel synthesis, the syngas needs its composition to be conditioned to a proper H<sub>2</sub>/CO ratio to achieve desired results (Im-orb et al., 2016; Zhang, 2010). The syngas then will be processed in a catalytic reactor where it is converted to fuel (Swanson et al., 2010). Biocrude and bio-oil have similar physical and chemical properties. Both are liquid in ambient temperature; however pyrolysis bio-oil has a higher moisture and oxygen content (Huber et al., 2006). Both products also demonstrated changes in physical and chemical properties over time (i.e. aging) (Diebold, 2000; Kosinkova et al., 2016). These undesirable fuel properties may be corrected through catalytic hydrotreatment where oxygen is removed and hydrogen is added (Bridgwater, 2012; Grilc et al., 2014), although bio-oil requires a mild hydrogenation step to reduce formation of coke at severe conditions (Diebold, 2000), while biocrude can be directly processed in a severe hydrodeoxygenation process (Elliott and Baker, 1986).

There are several techno-economic studies on pyrolysis and gasification since they have been developed previously. Studies that compare pyrolysis and gasification to each other and to other processes have also been published (Patel et al., 2016). However, as liquefaction is relatively novel, a comparative study between gasification, pyrolysis and liquefaction has not been performed previously. To this end, this study laterally compares these thermochemical processes that can be employed to produce liquid fuels from sugarcane bagasse through a techno-economic analysis. This is a common method to adequately compare the economic implications of employing biomass to energy processes (Bridgwater et al., 2002; Klein-Marcuschamer et al., 2013; Tzanetis et al., 2017). Development of mass and energy balances are also widely used as the initial step in life cycle analysis (Ojeda et al., 2011).

The inputs, requirements and configurations of each process differ from the others; however, these all produce a biofuel product. Each process has advantages and drawbacks, such as the cost of drying, differences in conditions, etc.; however, these differences have not been adequately quantified and their effects to profitability have not been sufficiently analysed. In this study, the ASPEN Plus models and economic models of each process demonstrate

**Table 1**  
Comparison of thermochemical processes from (Bridgwater, 2012; Mohan et al., 2006; Rackemann et al., 2011; Ramirez et al., 2015).

Thermochemical Process	Gasification	Thermal liquefaction	Pyrolysis
<i>Operating parameters</i>			
Temperature, °C	800–1000	250–330	280–630
Pressure, bar	1–20	1–240	1–5
Solvent requirement	None	Water or organic solvent	None
Feedstock requirement	Dry, size-reduced	Dry, moist or in slurry	Dry, size-reduced
<i>Product profile</i>			
Phase (main products)	Gas	Liquid	Liquid, gas
Product name <sup>a</sup>	Syngas	Biocrude	Bio-oil
Chemical composition	Carbon monoxide (CO), carbon dioxide (CO <sub>2</sub> ), hydrogen (H <sub>2</sub> ), methane (CH <sub>4</sub> )	Phenolic compounds, aromatics, molecules with carbonyl groups	Molecules with carbonyl groups, sugars, dehydrosugars, phenolic compounds
Heating Value, MJ/kg	27–33	28–36	16–19

<sup>a</sup> Typical term used in literature; terminology to be used in this study.

their technical and economic viability, compare economic indicators, and lead to the comparative analysis of these processes.

## 2. Methodology

Bagasse to biofuel plants employing gasification, liquefaction or pyrolysis were modelled in ASPEN Plus version 8.4. This modelling tool was shown to be suitable in earlier work (Anex et al., 2010; Magdeldin et al., 2017). Each plant was modelled with the different conditions of each process and generating different products. The refining and upgrading processes for each thermochemical plant to generate a crude-oil like product were also modelled and are described in the following subsections.

For each plant, 10,000 kg/h of sugarcane bagasse was the main input. The feedstock was modelled to have 48% moisture as received and uniform composition. Sugarcane bagasse is usually produced in sugar factories 5–9 months in a year and might be collected in stockpiles where moisture levels can vary (Rainey et al., 2013), however, for this study, it is expected that the average values used in the models are representative of the likely value of these properties when sampled. The variations in feedstock composition can affect the yield and composition of the product; however this variation is manageable since the biomass composition will not vary widely due to having being sourced from a single plant species. Furthermore, the stockpile of bagasse can be used to constitute a feed concentration that matches the process as designed. The plant capacity for this study was also set such that bagasse can be collected from a single source and can be more or less consistent in composition. Other management and operational controls can be used to limit the variations, however these are not within the scope of this study. The properties of the feedstock used are presented in Table 2 (Kosinkova et al., 2015b).

It is expected that the size profile changes from mill to mill; however, biomass particle size does not have a significant effect on liquefaction yield (Akhtar and Amin, 2011), and bagasse is milled and screened for both gasification and pyrolysis, where particle size matters (Bridgwater, 2012; Buekens and Schoeters, 1985). The different plants was referred to as gasification, liquefaction and pyrolysis, which is labelled after the central thermochemical process and encompasses refining and upgrading processes, separation and auxiliary operations. To differentiate the products of the thermochemical processes and refined products, the products of the gasification plant was referred to as syngas and FT liquids; that of liquefaction was labelled biocrude and HDO biocrude; and that of pyrolysis was named bio-oil and HT bio-oil.

**Table 2**  
Properties of sugarcane bagasse used as feedstock in this study.

Bagasse Property	Value	Unit(s)
Proximate Analysis		
Ash Content	6.7	% db
Volatile Matter	76.5	% db
Fixed Carbon	16.8	% db
Moisture	48	% as received
Ultimate Analysis		
Carbon	43.16	% db
Hydrogen	5.47	% db
Nitrogen	0.51	% db
Sulphur	2.15	% db
Oxygen	42.57	% db
Particle size distribution (Rainey, 2009)		
>12.5 mm	0.25	mass fraction
4.0–12.5 mm	0.35	mass fraction
<4.0 mm	0.4	mass fraction

### 2.1. Description of process models

Fig. 1 presents the simplified process flow diagrams of the three processes modelled in this paper. A separate model has been prepared for each of the thermochemical process streams, although the feedstock for all models are the same. The crude product was determined by the result of the thermochemical process and subsequent refining processes.

#### 2.1.1. Gasification

Bagasse as received enters the dryer where its moisture content is reduced to 7%. The dried bagasse is then ground and screened to ensure that the particle size entering the gasifier is no larger than 2 mm. The dried and ground feedstock is fed with steam to the gasifier and converted to gaseous products and ash at 900 °C and 1 bar. This process has been modelled as a combination of an RYield block separating the elements comprising the feedstock and feeding it into an RGibbs block that generates products through an equilibrium path with the minimal Gibbs free energy at a specified temperature, pressure and number of phases. Following gasification, ash is removed in a cyclone, and the gas enters a tar reforming reactor to convert hydrocarbons and ammonia to hydrogen, carbon dioxide and nitrogen. The tar reformer was modelled with an RStoic block with the reforming reactions and the specified conversion rates (Table 3). Heat is recovered for heating other sections of the plant at the same time cooling the gas product stream in preparation for gas cleaning.

The gasified products proceed to the gas cleaning section where the syngas is scrubbed with methyldiethanolamine (MDEA) to remove hydrogen sulphide (Mitra, 2015), which can poison catalysts in the fuel synthesis section (Woolcock and Brown, 2013). Carbon dioxide is also removed to reduce inert components in syngas that can increase reactor volume unnecessarily. Scrubbers were modelled as an 8-stage RadFrac column where impurities are both physically and chemically absorbed into the MDEA-water stream, which ASPEN calculates with the ELECNRTL property method to handle the electrolyte-water system. The H<sub>2</sub>S- and CO<sub>2</sub>-rich amine proceeds to a stripper to separate the gases from the liquid stream and produce the lean amine for recycling into the gas scrubbers. The stripper was modelled using a RadFrac with a reboiler, heat exchangers and flash separators. The clean syngas is then compressed and preheated for the Fischer-Tropsch (FT) reactor.

Syngas enters the FT reactor where the synthesis of hydrocarbons from carbon monoxide and hydrogen in presence of a cobalt catalyst occurs at 200 °C and 20 bar (Van der Laan and Beenackers, 1999). A water-gas shift reaction was also modelled to ensure the proper H<sub>2</sub>/CO ratio of 2 that is required for the FT reaction. The FT reactions to produce paraffins and olefins from C1–C25 were modelled with chain growth probability factor,  $\alpha = 0.9$ , using the Anderson-Schulz-Flory distribution (Eq (1)), where  $M_n$  is the mass fraction of a hydrocarbon product with  $n$  carbons (Im-orb et al., 2016). A CO conversion of 70% was used (Ail and Dasappa, 2016). The FT reactor was modelled using an RStoic block.

$$M_n = \alpha^{n-1} (1 - \alpha) \quad (1)$$

The products of the FT reactor were cooled and separated. 95% of the gaseous products were recycled to the gas cleaning section to maximise the hydrocarbon production from the available CO. The rest of the gaseous products were used in a syngas power generation unit to supply electricity to work-requiring processes. The liquid products were decanted to separate the water and gases and light components were separated from heavier components in a distillation column with 15 stages, reflux ratio of 2 and a distillate to

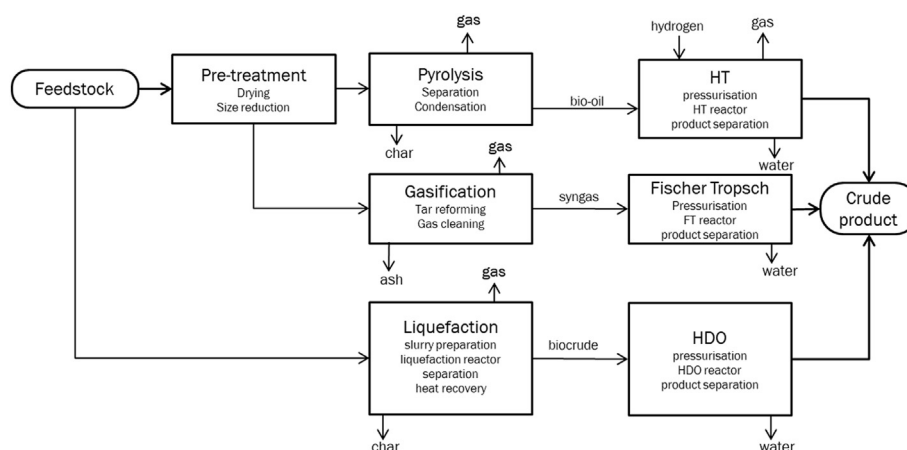


Fig. 1. Process Flow Diagram (simplified) for the three plants modelled.

Table 3

Modelling data used in the gasification model.

Quantity	Value	Unit(s)	Source
Drying target	7	% moisture content	
Grinding and screening target	2	mm	
Steam to biomass ratio	1.8	kg steam/kg biomass	Sandeep and Dasappa (2014)
Tar reforming conversion			
Methane	0.80		(2006)
Ethane	0.99		(2006)
Ethene	0.90		(2006)
Benzene	0.99		(2006)
Ammonia	0.90		(2006)
MDEA:Water ratio	0.25	kg MDEA/kg water	Polasek and Bullin (1994)
Product Separation			
Gaseous product (distillate) rate	0.2	mass proportion of product distillation feed	
Liquid product (bottoms) rate	0.8	mass proportion of product distillation feed	

feed ratio of 0.2. The distillation bottoms were cooled and collected as a product. Auxiliary processes such as heat exchange, steam and power generation were also modelled in ASPEN. Power was generated through the expansion of flue gases and steam in a Rankine cycle. The thermodynamic methods used for the gasification model was UNIQUAC, then UNIQU-RK following the Fischer-Tropsch reactors. The parameters used in this process model are given in Table 3.

### 2.1.2. Liquefaction

Bagasse as received is mixed with ethanol to prepare a slurry of 5% solid content for liquefaction. Ethanol was chosen to be a solvent due to its desirable thermal properties leading to more desirable yield and biocrude heating value results (Kosinkova et al., 2015b). The slurry is pumped and preheated to the continuous reactor where the bagasse is liquefied at 300 °C and 165 bar. The products are then separated in gaseous, liquid and solid streams. The liquid stream enters a solvent recovery section where ethanol is recovered as a recycle stream. The concentrated biocrude then enters the hydrodeoxygenation reactor where it reacts with hydrogen over a solid catalyst at 300 °C and 80 bar. A significant amount of oxygen is removed as water and carbon dioxide. The HDO products are separated and the liquid product is collected and cooled. A further process description is available in a previous study (Ramirez et al., 2018). The thermodynamic method used in the liquefaction model was RK-SOAVE. The parameters used in this process model are given in Table 4.

### 2.1.3. Pyrolysis

As in gasification, bagasse as received enters the plant through the dryer and grinder, where it is pre-processed to 7% moisture and particle size of no larger than 2 mm. The dried and ground bagasse is fed to the pyrolysis process where it is converted to condensable and non-condensable gaseous products, char and ash at 500 °C and 1 bar. The pyrolysis reactor was modelled with an RYield block using product yields and chemical profile by Varma and Mondal (2017) matched to the model compounds used by Jones et al. (Jones et al., 2013) to represent functional groups of chemicals in the bio-oil.

The pyrolysis products are then separated in an electrostatic precipitator where solid residues and liquid droplets that might condense are separated from the hot gases. The heavy components proceed to the filter where the liquid is separated from the char. Filtration was modelled such that 20% of the liquid remains with char (Paenpong et al., 2013). The filtrate then proceeds to the hydrotreatment section. From the electrostatic precipitator, the gaseous bio-oil proceeds to a contact tower where condensed bio-oil cools the bio-oil vapour. The contact tower was modelled with a RadFrac column with 10 stages, a partial vapour condenser and a reflux ratio of 3. The condensed bio-oil exits the tower and is cooled further and 70% of the cooled bio-oil is recycled back to the contact tower, while the rest moves on to the hydrotreatment section. The exiting cooled vapours from the tower are further flashed to separate non-condensable gases from bio-oil that may have condensed further.

All streams of cooled and separated bio-oil are combined and proceed to the hydrotreatment section. The bio-oil is pressurised



**Table 4**  
Modelling data used in the liquefaction model.

Quantity	Value	Unit(s)	Source
<i>Liquefaction Yields</i>			
Gas	19.34	% of dry feed	Kosinkova et al. (2015b)
Solid	22.9	% of dry feed	Kosinkova et al. (2015b)
Biocrude	57.76	% of dry feed	Kosinkova et al. (2015b)
Recovered ethanol	99	% of input ethanol	Manohar Rao (1997)
Biocrude and ethanol retention on filter cake	2	% of total liquid products	(Silverblatt and Dahlstrom, 1954){Silverblatt, 1954 #550}
HDO stoichiometric conversion	100	% reactive components	

and preheated before it enters the first hydrotreatment reactor. Pressurised hydrogen is also combined with the bio-oil upon entering the reactor. The first hydrotreatment reactor proceeds at 250 °C and 140 bar. In this mild hydrotreatment stage, the bio-oil is partially hydrogenated and deoxygenated. The first stage products move to the second stage, but not before preheating. The second stage proceeds at 400 °C and 140 bar, with pressurised hydrogen added in a stoichiometric amount. At this more severe stage, significant amounts of oxygen are removed and hydrogenated to water or removed as carbon dioxide. Some bio-oil species cannot be fully deoxygenated but are converted into hydrofurans, cyclic alcohols, diols and glycols. The hydrotreatment reactors were modelled using RStoic blocks, using reactions from literature, with reaction conversions between 98 and 100% (Adjaye and Bakhshi, 1995; Bernas et al., 2012; Bindwal and Vaidya, 2013; Duarte et al., 2016; Fisk et al., 2009; LaVopa and Satterfield, 1987; Li et al., 2013). The reactivity of the bio-oil components in mild and severe reactor stages were determined through the reactivity scale laid out by Elliott (2007).

Following hydrotreatment, the products are cooled in the hydrotreatment preheating heat exchangers to recycle heat. The products are also expanded in a turbine and further cooled. Solid residues and gaseous fractions are separated from the hydrotreated liquid product. Water is removed before the liquid product is cooled and collected for storage.

Solids from the pyrolysis reactor and non-condensable gases from the plant is used to generate heat and power for use in the plant, which was modelled in ASPEN as a Rankine cycle. A simple heat recovery network was also designed for major cooling and heating requirements. The pyrolysis model used UNIQUAC as the base thermodynamic model, and SRK for blocks preceding the hydrotreatment process. The parameters used in this process model are given in Table 5.

## 2.2. Modelling

The mass and energy models of the three plants were constructed using pre-defined ASPEN blocks. Separations were carried

out as programmed in ASPEN, although in some instances assumptions were made in order to efficiently build process models that maximises the product streams.

Following the process model, deterministic economic models were also built to analyse the profitability of the plants. Using mass and energy balances, the equipment sizing and capital costs, raw material, energy, catalyst, labour and maintenance costs were estimated from various data sources. The revenue from product sales was also calculated using the product streams and market price of the product. The plant was modelled to be located in the South East Queensland region in Australia, as a representative setting where feedstock such as sugarcane bagasse is readily available (Kosinkova et al., 2015a), and policy and market conditions may be ideal (2016). The relevant local rates and prices were used in the economic model. Capital costs were determined through calculations in cost databases (2014; Peters et al., 2002), and adjusted using the Chemical Engineering Plant Cost Index (CEPCI) to 2017 values. Adjustments based on scale used the commonly-used power rule with a higher than average exponent of 0.7, to account for high operating temperatures and moderate pressures of equipment (Peters et al., 2003). Direct and indirect costs were estimated using ratio factors based on delivered-equipment costs for a solid-liquid plant (Peters et al., 2003). A tax rate of 30% and a straight line depreciation schedule based on the method prescribed by the Australian Taxation Office (2017g) was used. Net present value (NPV) and internal rate of return (IRR) were calculated using a discount rate of 10% and used as the primary economic indicator for this study. The minimum selling price (MSP) was calculated by determining the product price where in the NPV is zero at the discount rate. The economic parameters used in the economic models are presented in Table 6.

Sensitivity analyses were also conducted to present the effect of fluctuating conditions on NPV and the MSP. The parameters were varied  $\pm 50\%$  of their base values. Thermochemical conversion and refining conversion were the process parameters changed, while the prices of products, feedstock, natural gas, chemicals (ethanol and MDEA) and hydrogen were the economic parameters varied. The tax rate was also varied to illustrate the effect of policy-driven

**Table 5**  
Modelling data used in the pyrolysis model.

Quantity	Value	Unit(s)	Source
<i>Drying target</i>	7	% moisture content	Wright et al. (2010)
<i>Grinding and screening target</i>	2	mm	Bridgwater (2012)
<i>Pyrolysis product breakdown</i>			
Bio-oil	36	% total product	Varma and Mondal (2017)
Water	9	% total product	Varma and Mondal (2017)
Gas	26	% total product	Varma and Mondal (2017)
Solids	29	% total product	Varma and Mondal (2017)
<i>Bio-oil retention on filter cake</i>	20	% bio-oil	
<i>Bio-oil quenching</i>			
Recycled to quench <sup>a</sup>	0.7	Mass proportion of cooled bio-oil	
For hydrotreatment	0.3	Mass proportion of cooled bio-oil	

<sup>a</sup> Selected ratio to adequately provide cooling in the bio-oil quench contact tower.

**Table 6**

Economic modelling data used in this study.

Economic quantities <sup>a</sup>	Value	Units	Source
Plant size	84000	tonnes per year feed as received	
Operating days per year	350	days	
Exchange rate	0.78	US\$/AU\$	(2017d)
Power law scaling factor	0.7		Peters et al. (2003)
CEPCI for study year	553		(2017b)
Location factor, Australia	1.4		Humphreys (2005)
Feedstock price	46.80	US\$/t dry feed	O'Hara, (2011)
Hydrogen price	1.56	US\$/kg	Hinkley et al., (2016)
Ethanol price	0.64	US\$/L	(2008)
FT Catalyst price <sup>b</sup>	38.49	US\$/kg	(2007)
Tar Reforming Catalyst price <sup>b</sup>	20.53	US\$/kg	Swanson et al. (2010)
HDO Catalyst price <sup>c</sup>	17.33	US\$/kg	
HT Catalyst price <sup>c</sup>	17.33	US\$/kg	
MDEA price <sup>b</sup>	2.60	US\$/kg	Protection (2009)
Natural gas price	4.63	US\$/GJ	(2017i)
Electricity price	38.10	US\$/MWh	(2017c)
Water price	2.20	US\$/kL	(2017f)
Trade waste handling price	0.76	US\$/kL	(2017f)
Crude oil price	0.327	US\$/L	(2017e)
Maintenance rate	2	% FCI	
Discount rate	10%		
Company tax rate	30%		(2017g)

<sup>a</sup> Values as of July 2017.<sup>b</sup> CPI adjusted to 2017 values.<sup>c</sup> Vendor information.

changes to profitability.

### 3. Results and discussion

#### 3.1. Process model

The main inputs and outputs of the three processes are comparable. Conceptually, fibrous feedstock enters the plant and a liquid hydrocarbon-like product is produced. The major inputs are heat for the thermochemical conversions, separations and refining processes, chemicals and steam. The emissions are flue gases and waste water. Table 7 presents the summary of the major inputs and outputs of the three processes (see Table 8).

Among the three thermochemical processes, gasification had the highest yield from the thermochemical conversion. This was expected due to the addition of steam to the gasifier and with most of the mass exiting the gasifier as syngas. There is only one by-product, solid char and ash produced at 0.09 kg/kg dry feed. Liquefaction yield was less than gasification due to the products segmenting into biocrude, gases and solids, which was the same case as pyrolysis; however, a significant part of the ethanol solvent

**Table 7**

Process model results for the three plants.

Stream	Consumption/Production (per kg feed dry basis)			Units
	Gasification	Liquefaction	Pyrolysis	
Thermochemical product	0.74	0.67	0.41	kg
Refined product	0.20	0.47	0.27	kg
Hydrogen	N/A	0.03	0.02	kg
Steam	N/A	1.92	N/A	kg
Natural gas heating	0.03	0.006	N/A	GJ
Electricity consumption	0.012	0.18	0.004	kWh
Combustion and drying air	39.8	6.54	37.8	kg
Chemical replacement	0.00009 <sup>a</sup>	0.12 <sup>b</sup>	N/A	kg
CO <sub>2</sub> e process emissions	2.89	4.40	1.63	kg
Wastewater	1.54	2.23	0.12	kg

<sup>a</sup> Amine.<sup>b</sup> Ethanol.**Table 8**

Economic results of the modelling of gasification, liquefaction and pyrolysis processes.

Quantity	Gasification	Liquefaction	Pyrolysis
Plant capacity, t/y feed as received	84000	84000	84000
Capital cost estimates, million US\$			
Total Installed Cost	16.77	17.87	23.20
Location-adjusted Direct Cost	41.57	44.46	38.42
Total Indirect Costs	15.09	16.20	4.95
Working Capital	11.33	12.13	8.67
Total Capital Cost	67.99	72.79	52.05
Operating Costs, million US\$/y			
Feedstock Cost	2.04	2.04	2.04
Electricity	0.02	0.30	<0.01
Heating	6.27	1.01	–
Ethanol or Amine Make-up	0.01	3.37	–
Catalyst replacements	0.07	0.12	0.31
Hydrogen	–	1.77	1.17
Steam Supply	–	2.07	–
Trade Waste Handling	0.05	0.06	<0.01
Water	0.94	1.46	0.60
Labour	1.81	1.17	1.17
Maintenance (2% FCI)	1.13	1.21	0.87
Total Operating Costs	12.35	14.57	6.16
Total Products, million L/y	11.50	25.78	11.58
Revenue, million US\$/y	3.76	8.44	3.79
Base Economic Indicators			
Annual Cash Flow, Million US\$	–8.59	–6.13	–2.37
NPV, Million US\$	–128.3	–113.7	–65.7

reacted with the biomass and augmented the product mass in liquefaction. Pyrolysis also has a tendency to produce larger amounts of light gaseous products due to the low pressure in which it operates. On the other hand, production of liquid products are enhanced in liquefaction due to high pressure (Evans and Milne, 1987).

Following upgrading or synthesis, liquefaction has the highest refined product yield among the three, because the upgrading process involves the addition of hydrogen into the biocrude. The refined HDO biocrude yield is lower than raw biocrude due to the removal of oxygen and entrainment of light gaseous products in HDO off-gas. Pyrolysis HT bio-oil yield follows liquefaction HDO

biocrude yield, similarly caused by hydrogen-addition and oxygen removal. However, there was less raw bio-oil coming into the hydrotreatment process compared with the raw biocrude fed into HDO, so it is expected that the HT bio-oil yield is less than HDO biocrude. For gasification and FT, the FT liquid yield drops from the syngas yield due to the 70% conversion of CO assumed in the model (Ail and Dasappa, 2016). The converted products were also split into FT liquids, water and gas, thereby diminishing the final product yield. Hydrogen consumption follows the upgraded product trend: more hydrogen was required for upgrading biocrude due to the higher flow rate. FT does not require addition of hydrogen so none was reported for the gasification case. The addition of steam was unique for liquefaction, which was used in solvent recovery. Gasification also had steam added in the thermochemical process, however, it was modelled to be generated in-house using recycled heat and natural gas.

Gasification had a much higher natural gas heating requirement brought about by the high gasification temperature, which cannot be completely maintained using heat from burning char. Another large heat draw was rich amine stripping, which handles 83.3 t/h of rich amine (75% water). Liquefaction had a relatively smaller natural gas requirement, which was offset by the heat supplied by combustion of off-gases and char. Most of the liquefaction heating requirement was to raise 111 t/h of the liquefaction slurry to the liquefaction temperature of 300 °C. Pyrolysis generated enough char to supply the heat required for both pyrolysis and drying and thus, did not need additional heat from burning gas. The electricity requirements are also lowest for pyrolysis, with only minimal make-up electricity required for size reduction. The requirement for pumping bio-oil and compressing hydrogen was supplied by in-house electricity produced from burning char and off-gases and expanding HT products. Gasification had higher electricity requirements from syngas compression prior to the FT reactor. Liquefaction required much more electricity to pump liquefaction slurry at a high flow rate to liquefaction pressure, with all the combustion energy directed to supplying heat rather than producing electricity. In terms of total energy input (heat and electricity) on a dry feed basis, liquefaction had the largest requirement.

The high feedstock moisture content required large amounts of drying air for both gasification and pyrolysis, while liquefaction did not have the same needs. Only combustion air was required for the liquefaction model and thus its air requirement is only a sixth of gasification and pyrolysis. Gasification needed slightly more than pyrolysis to supply a stoichiometric amount of air in combustion. Chemical replacement rates for the three models are also vastly different. Gasification was modelled to have 20% replacement of MDEA over one year of operation. For liquefaction, it was assumed that the solvent does not react in liquefaction and 1% of ethanol is lost in recycling (Manohar Rao, 1997) so make-up ethanol was required. The pyrolysis model does not include chemicals other than hydrogen, therefore there was no chemical replacement required.

For both CO<sub>2</sub> emissions and waste water, the order is liquefaction > gasification > pyrolysis. The CO<sub>2</sub> emissions of the processes, as calculated by ASPEN are mostly from the use of natural gas and combustion of by-products. The CO<sub>2</sub> emissions over the life-cycle can be affected by the application in which the products are used, and the nature of electricity generation supplying the plant. However, since there can be huge variations in application and power supply, these emissions were not included in the scope of the study. Most of the waste water from liquefaction was from the solvent recovery section where some steam added to the biocrude-ethanol mixture condensed, while for gasification, the waste water was water condensed from syngas after gasification and tar reforming. In pyrolysis, waste water was collected at product separation by

decanting water generated in HT.

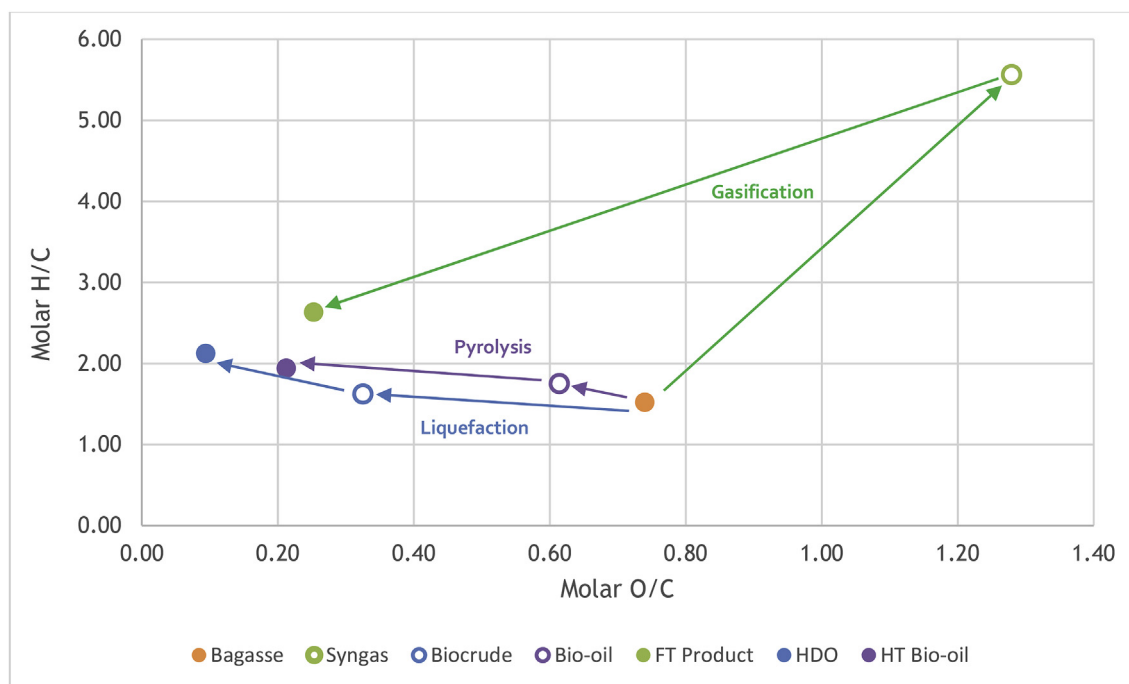
### 3.2. Product properties

Fig. 2 presents the elemental composition of the main streams for the three processes. The lines show the progression from feedstock to intermediate product to refined product.

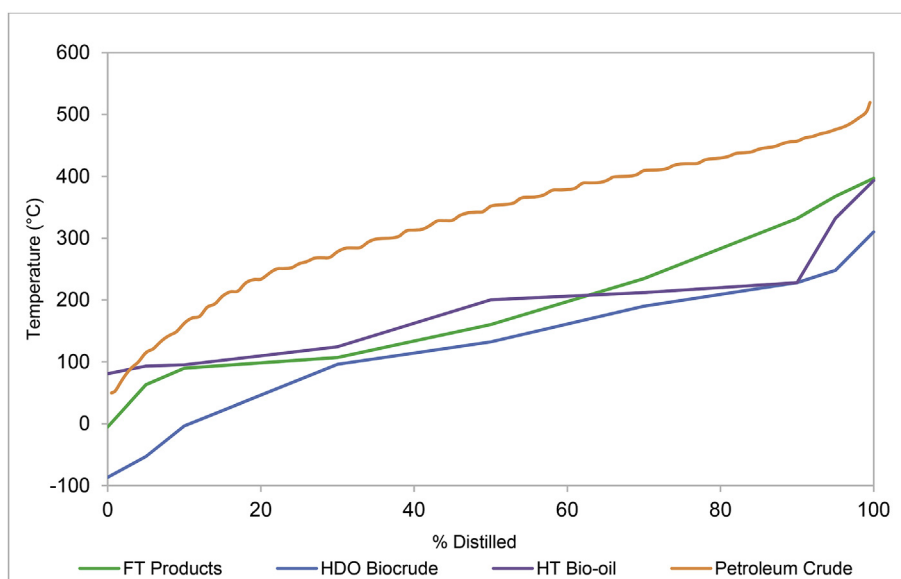
In Fig. 2 the progress of the elemental composition of the main biomass/biofuel streams through each process is shown. In gasification, the O/C and H/C ratio both increased from feedstock to syngas as some carbon is removed as char and the added steam reacts to form CO, CO<sub>2</sub> and H<sub>2</sub>. In the Fischer-Tropsch reactor, hydrocarbons are formed and oxygen and some hydrogen leave as water. A drastic reduction of both ratios can be observed. In liquefaction, the O/C ratio decreases from feedstock to biocrude by forming water and CO<sub>2</sub> removed in the gaseous phase. After HDO, the H/C ratio is significantly higher and the O/C ratio also decreases by the formation of water as a by-product. In pyrolysis, there is a slight reduction of O/C from feedstock to bio-oil, with some of the oxygen going into gaseous by-products. In hydrotreatment, hydrogen is added and some oxygen is liberated as water, causing lower O/C and higher H/C.

The refined products generated from these processes can be further upgraded or separated to fractions analogous to their petroleum counterparts as blendstock for diesel or gasoline products. For instance, decanting water from the hydrocarbons generated in gasification (FT products) can bring the O/C ratio down to nil and the H/C value to 2.15. The hydrocarbon composition also suggests a product similar to gasoline. It is possible for the products to be co-processed in refineries if the properties match. The refining processes in this study follow recommendations to reduce oxygen content and approximate the properties of petroleum crude oil to enable co-processing (de Miguel Mercader, 2010; Jensen et al., 2016). The modelling results show that the oxygen contents of HDO biocrude and HT bio-oil approach the very low oxygen content of crude oil (Speight, 2011a). This result is in agreement with numerous HDO and HT studies (Al-Sabawi and Chen, 2012; de Miguel Mercader, 2010; Jensen et al., 2016). Despite the good agreement in the oxygen content, it has been presented that the chemical profile of HT bio-oils and HDO biocrudes are still different compared with crude oils (Zacher et al., 2014). The different chemical moieties have implications in energy and hydrogen requirements (Grange et al., 1996), thus the efficiency of the refinery processes may be affected. Instead of blending the bio-based crudes with hydroprocessing feed, it has been suggested to blend them with distillation feed in order to generate fractions that may be more homogeneous in chemical composition (Hoffmann et al., 2016; Lavanya et al., 2016). The suitability of the products from these thermochemical processes to be processed in distillation units can be determined by analysing their distillation curves. Fig. 3 presents the ASPEN-generated true boiling point (TBP) distillation curves of the products compared with that of a typical crude oil (see Fig. 4).

The distillation curves of the FT products, HDO biocrude and HT bio-oil appear to be in the same range albeit around 100 °C lower than the crude oil curve. Among the bio-based products, HT bio-oil has the narrowest range of 312 °C, followed by HDO biocrude (397 °C) and FT products (402 °C). In contrast, the crude oil in this graph has a range of 470 °C. It is also important to note that the crude oil's initial boiling point (IBP) is 50 °C, while both FT products and HDO biocrude have sub-zero IBPs. Therefore, a degassing step to remove light components can be useful for better integration. HT bio-oil on the other hand, has a similar IBP with petroleum crude oil; however its final boiling point is 207 °C below that of petroleum crude oil. The differences in boiling range of the bio-based products



**Fig. 2.** Diagram showing the elemental composition of the feedstock (orange dot), intermediate products (ring) and refined products (solid dot) of the processes in this study. Arrows show progression through gasification (green), liquefaction (blue) and pyrolysis (purple). (For interpretation of the references to colour in this figure legend, the reader is referred to the Web version of this article.)



**Fig. 3.** Distillation curves of the products of the thermochemical processes in this study, compared with the distillation curve of crude oil obtained from (Ramirez et al., 2017).

with that of the petroleum crude oil mean that the distillation column processing this particular crude oil might not be appropriate for the bio-based products. The distillation curves show that 100% of the bio-based crudes distil at a temperature where only 50–70% of the petroleum crude oil has distilled. For instance, if any of the bio-based products were blended with crude oil, there might be a significant increase in the amount of lighter fractions obtained. Processes downstream of the fractionator can be affected. It is typical that refinery feedstock undergo light end removal (Speight, 2011b) so the matching of the bio-based products to the distillation properties of this particular petroleum crude oil can be improved.

### 3.3. Economic results

The economic results for the three plants were widely different from each other, as shown in Table 8. Liquefaction had the largest capital costs due to it having the highest total purchased cost, from which the other direct costs are factored. Gasification follows liquefaction closely and pyrolysis had the smallest total capital cost. Pyrolysis has higher total installed costs compared with the other plants, but had lower location-adjusted direct costs due to a higher installed cost of pyrolysis equipment obtained from a vendor. The additional costs for instrumentation, piping, electrical and civil for



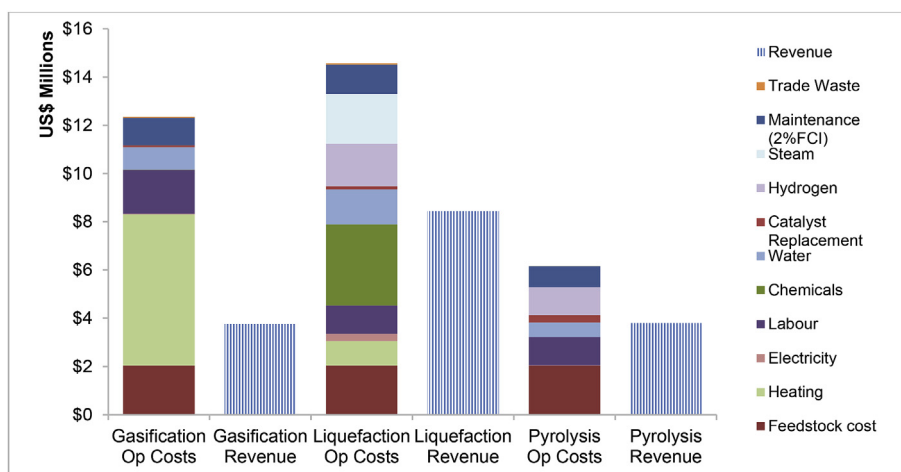


Fig. 4. Operating cost breakdown of each plant, compared to annual revenue generated. Revenue was calculated using a unit price calculated from the crude oil price.

the quoted pyrolysis equipment could be less than the estimate if the factors were used. The costs calculated in this study were comparable to the capital costs calculated from other studies, with adjustments based on scale and year of cost estimation.

The gasification plant in this study had a total capital cost of US\$ 68 million, which is similar to the scaled down, adjusted costs of US\$ 71 million from the plant described by Swanson et al. (2010). The key differences between the two plants were oxygen was added to gasification via air separation in addition to steam, FT products were hydroprocessed and sulphur was recovered using a catalytic recovery system. The aforementioned study also had costs based in the United States of America. Another study by Tijmensen et al. (2002) in a Dutch context had a slightly lower capital cost of US\$ 61 million (scaled, 2017 US\$), employing a gasification process with oxygen added from an oxygen plant, slightly different gas cleaning equipment, and a different gasification process taking in feedstock at 15% moisture, dried from 30%. The high capital costs are usually influenced by stringent syngas impurity levels in the order of 10 ppb for Fischer-Tropsch synthesis (Woolcock and Brown, 2013).

The liquefaction plant, on the other hand, was estimated to cost US\$ 73 million, comparable to the estimate by Zhu et al. (2014) of US\$ 77 million (scaled, 2017 US\$). The main difference of the two plants is that the liquefaction plant in this study involved extensive solvent recovery equipment, while that of Zhu et al. had a hydrogen production plant.

The cost to build the pyrolysis plant was the least among the three plants, since the separation of by-products and impurities is very little in pyrolysis compared to the two processes. The estimated capital cost was US\$ 52 million, which is higher than a plant modelled by Anex et al. (Anex et al., 2010) in 2007 costing US\$ 30 million (scaled, 2017 US\$). An update to Anex et al., in 2011 (Brown et al., 2013) adjusted the capital costs to US\$ 61 million for almost the same plant, with the significant differences around sizing of power generation and in how the costs were estimated. The updated power generation section was similar to the set-up in the pyrolysis plant in this study.

The breakdown of annual operating costs were also different from plant to plant. As anticipated from the mass and energy balances, liquefaction had the highest annual operating cost, while gasification was second. Pyrolysis had around half of gasification's total operating cost, due to the absence of heating costs. Fig. 4 shows the detailed breakdown of each process' annual operating costs and its comparison with annual revenue.

Each plant has different operating cost breakdowns, but have the same feedstock cost as a given for the study. For pyrolysis, feedstock cost has the highest contribution to operating costs (33%). This is common in similar studies (Jones et al., 2013; Wright et al., 2010). Moreover, feedstock cost has been identified as a major factor in predicting production cost of bio-oil (Bridgwater, 2012). For both gasification and liquefaction, feedstock costs were determined to contribute around 15% to the total operating cost. The largest cost in gasification is heating, making up more than half of the total cost. In liquefaction, the largest contributor is ethanol replacement, bearing 23% of the total cost. Labour costs are comparable, but the gasification plant required US\$ 642,000 more in annual labour costs due to the numerous columns in gas cleaning for which attention is needed. The other costs correspond to the mass and energy balances discussed in Section 3.2.

The revenue of the three plants were all based on the same crude oil unit price. Each plant differs in annual product volume, and consequently, annual revenue. From the revenue calculations, it was determined that liquefaction had the highest revenue, followed by pyrolysis and closely behind, gasification. However, since each plant produces different annual volumes of products, it may be difficult to compare them based on annual values of revenue and operating cost. For this reason, the unit margin/loss value as calculated using Eq. (1).

$$\text{Unit Margin} \left( \frac{\text{US\$}}{\text{L}} \right) = \frac{\text{Annual Revenue} - \text{Annual Operational Costs}}{\text{Annual Production Volume}} \quad (2)$$

The unit margin summarises the earnings or losses before depreciation and tax, or annual cash flow of each plant per unit volume of fuel produced. Due to all plants having a revenue that was less than the operating costs, the annual cash flows, and consequently, margins were calculated to be negative. A smaller negative value can be viewed as more favourable since there is a smaller hurdle that needs to be overcome to achieve profitability. The unit margin for pyrolysis was deemed the best result at US\$ −0.21/L, while liquefaction was not far behind at US\$ −0.24/L. Due to a high operating cost and low production volume, gasification achieved a margin of US\$ −0.75/L, more than thrice of the other cases. The NPVs also followed this trend, with gasification having the most negative NPV of US\$ −128.3 million, liquefaction following with US\$ −113.7 million and pyrolysis with US\$ −65.7 million. The difference in NPV widened between the three plants

due to the effect of the large capital outlay required for the gasification and liquefaction plants compared with the pyrolysis plant, and the significant difference between the three plants' annual cash flows.

### 3.4. Sensitivity analysis

The capital costs determined in this study can be considered a study estimate where only major items were considered; therefore, it can be expected that the estimates can be  $\pm 30\%$  accurate (Peters et al., 2003). Furthermore, installation and indirect costs were estimated using ratio factors, which can be  $\pm 20\%$  accurate. It follows that the capital cost estimates can vary by 50% due to a number of factors such as market fluctuations or unforeseen costs. From the sensitivity analysis, it was determined that the NPV for each plant was most sensitive to the capital costs. This presents a financial risk from this determination up to the plant's construction phase. Fig. 5 shows the sensitivity of NPV of the three processes to the capital costs.

Among the three plants, pyrolysis was most sensitive to changing capital costs. This is due to the large changes in cash flow for each case. When the capital cost changes, the operating costs also change since the maintenance cost is estimated as 2% of the capital cost. While the capital costs change at the same rate, the cash flows change largest for pyrolysis, so the NPV will bear a larger change.

Upon operation, the profitability of the plant will change if parameters fluctuate due to key processes operating below or above expected levels, or price changes in the market. The sensitivity of the NPV of each process to these key technical and economic parameters have been determined and is presented in a tornado diagram shown in Fig. 6.

For all processes, the product price is a significant parameter, although it affected the NPV of liquefaction the most. This is due to the larger volume of product from the liquefaction process compared with gasification and pyrolysis. It is expected that the product price will change according to market factors, although it could be possible that incentives and tariffs to support biofuel production can augment revenues and make the venture profitable.

Thermochemical conversion was also a major parameter for liquefaction and pyrolysis. This is expected since the amount of biocrude and bio-oil dictate the amount of refined product produced. An opposite but much smaller effect happens to gasification due to the increase in operational costs involved in cleaning higher flows of syngas, and the increase in revenue is dampened by the FT conversion less than unity. With any production process, improvements are focused largely on obtaining greater yields from

the same amount of raw material, however, as seen in gasification, there should also be a focus on increasing the cost efficiency of gas cleaning processes to maximise the positive effect of higher yields to NPV. Another cost reduction prospect is to develop synthesis processes and catalysts that require less stringent gas cleaning.

As expected, the natural gas price affects gasification immensely, due to the large requirement for heating, while pyrolysis was not at all sensitive due to its self-sustaining design. The effect on liquefaction NPV by natural gas price was moderate. The development of more energy efficient processes can ameliorate both the costs of heating and the fossil fuel-related CO<sub>2</sub> emissions. Furthermore, the use of by-product streams for heat and power production can be considered.

Refining conversion affected gasification significantly as a change in FT conversion proportionally affected the volume of product. The effect to both liquefaction and pyrolysis were different, primarily due to the base case modelled at 100% conversion, making the base value the concurrent with the "high case" in the sensitivity analysis. The incomplete conversion had different effects too. Upon comparison of boiling point distributions, it was determined that for liquefaction, 5% of the unreacted biocrude is beyond the boiling point range of the HDO biocrude product, while for pyrolysis it was 26%. These values allowed for an estimation of the "off-spec" amount in the product, and using this number to determine the value-reduction factor for the sensitivity cases, with which the base case revenue was multiplied. For liquefaction, the revenue was reduced by 96% for 50% HDO conversion, while for pyrolysis it was 87% reduction for 50% HT conversion. These values can be seen affecting NPV proportionately. The focus on improving conversion values are both on product yields and quality of products. This encompasses many different considerations such as catalyst activity, reactor design, fouling and deactivation of catalysts and process effectiveness relating to deoxygenation, cracking and formation of by-products. In all three processes, it was determined that catalyst costs were not critical, however, for a larger plant with larger catalyst beds, the cost of new catalysts with superior performance can affect profitability. The cost of more effective catalysts should be balanced with the effect of resulting higher conversion ratios. An improvement in quality of the refined product can also affect product price, which has been demonstrated to be a very significant parameter in profitability.

The effect of the prices of inputs can be seen at the bottom of the diagram. The price of ethanol affects liquefaction NPV due to the significant amount of annual replacements, even with a process loss of only 1%. This presents an opportunity for cost savings by focusing process improvements on ethanol recovery, or maintaining a supply of stably-priced ethanol by hedging contracts, or in-house supply from an integrated liquefaction-cellulosic bio-ethanol plant. In contrast, the amine price affects gasification NPV very slightly. The variation of the feedstock price within 50% of the base case price generates a consistent change of US\$ 7.90 million across the three processes. This was interesting considering the different cost contributions of feedstock between these processes. The variation in feedstock cost could be managed by stockpiling to reduce the effect of the seasonality of sugarcane supply. Hydrogen is another key input for liquefaction and pyrolysis, although the variation in price does not affect NPV as much as other raw materials. Nonetheless, its effect on the NPV was higher than HT or HDO conversion. From the process side, the effect of the varying hydrogen price can be managed by process improvements related to hydrogen uptake. In the models used in this study, the hydrogen requirements in HDO and HT were determined by stoichiometric quantities required to carry out the reactions. The hydrogen requirement may increase if catalysts were developed to increase the deoxygenation of the biocrude or bio-oil components, and may

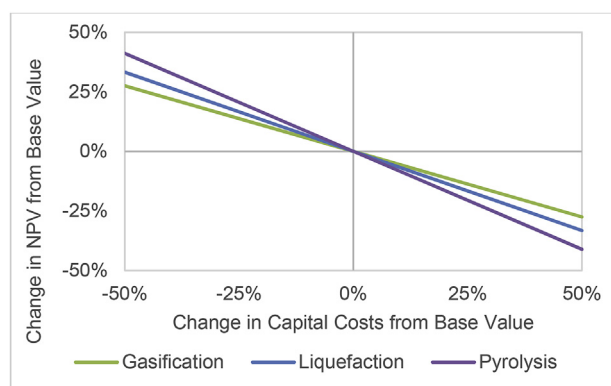
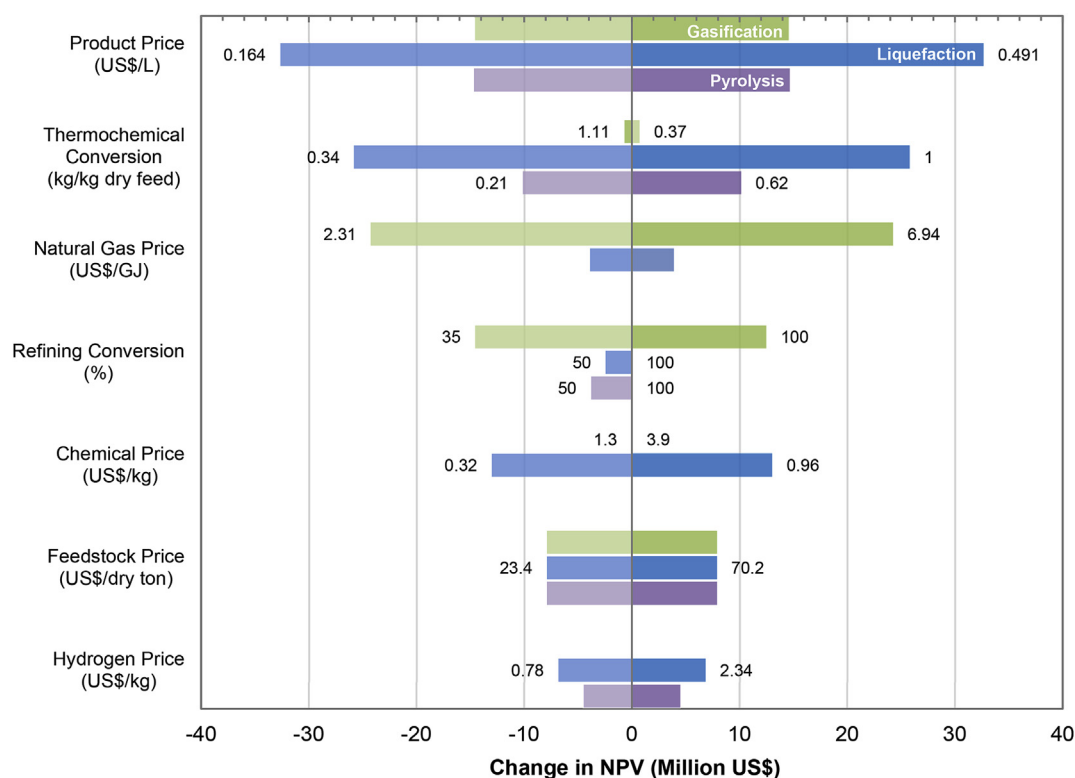
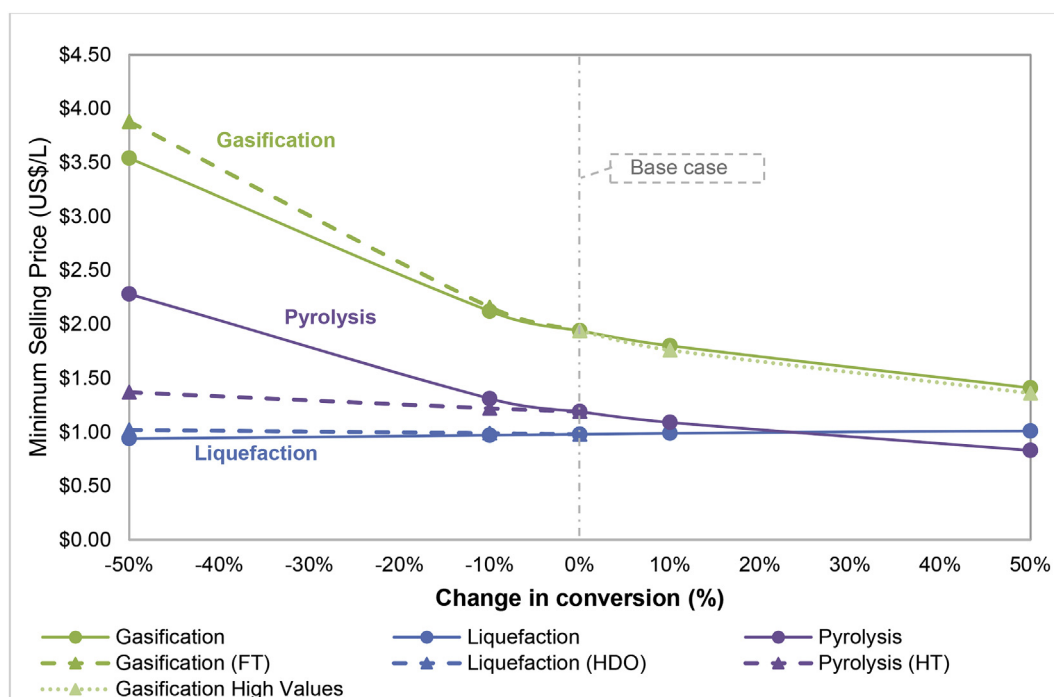


Fig. 5. The sensitivity of NPV to prospective changes in capital costs.



**Fig. 6.** Sensitivity of the NPV to selected process and economic variables ( $\pm 50\%$ ). Green bars represent gasification, blue bars represent liquefaction and purple bars represent pyrolysis values. Change in NPV values were derived from the base case value for each process. (For interpretation of the references to colour in this figure legend, the reader is referred to the Web version of this article.)



**Fig. 7.** Effect of varying thermochemical conversion (circles, solid lines) and refining conversion (triangles, dashed lines) to the minimum selling price of the product.

cause the NPV to be more sensitive to the hydrogen price. There can also be improvements in the thermochemical process where the biocrude or bio-oil generated will have lower oxygen content and

will require less hydrogen in refining. This can reduce the effect of the hydrogen price. Furthermore, a hydrogen plant can be integrated with the liquefaction or pyrolysis plant; however, the effect

of an increased capital cost vis-à-vis decreased operational cost should be further analysed.

### 3.5. Minimum selling price analysis

The minimum selling price was calculated as a measure of product value based on costs and revenues. This allows a parametric prediction of the product price and enables development of profitable and non-profitable operating regimes based on key process and economic parameters and product price. The base case MSP of the three processes are the following: gasification, US\$ 1.94/L, liquefaction, US\$ 0.98/L, and pyrolysis, US\$ 1.19/L. The following figures show the variation of the MSP with varying parameters. The area above the MSP lines is the profitable operating region, while the area below the line is the non-profitable operating region.

The thermochemical and refining conversions are palpably critical. For gasification, there is a sharper change between the low values (−50% and −10%) and base case values than the high values (+10% and +50%), although the effect of thermochemical conversion was slightly gentler at 50% reduction compared with FT conversion. Pyrolysis conversion has a similar effect, although the MSP values are lower. Lower HT conversion values increase the price, but not as much as the pyrolysis conversion. The small product volumes cause a sharp increase for lower conversions, with higher prices required to achieve profitability. The effect of a lower HT conversion, however, is dampened because a 50% reduction in conversion corresponds to only 13% reduction in volume. For liquefaction, the MSP does not move greatly due to a larger product volume (see Fig. 7).

The effect of the input prices are shown in Fig. 8. Gasification product MSP had the largest variations among the three processes. This is ostensibly due to the smaller volume of product. Natural gas

price greatly affects the MSP since the process is dependent on a large amount of heating. The feedstock price, as seen in the NPV sensitivity analysis, moderately affects the MSP; however, a slightly larger effect is shown for gasification and pyrolysis, compared with liquefaction, again influenced by the product volumes. The pyrolysis MSP values change between US\$ 1.10/L to US\$ 1.28/L, which is moderate, compared with gasification, which swings between US\$ 1.67/L and US\$ 2.21/L. At lower values, hydrogen price drives a higher MSP compared with feedstock price, which is the opposite at higher values. Feedstock cost, which comprises a third of the operating cost of pyrolysis is expected to vary the MSP more, compared with hydrogen costs. On the other hand, liquefaction MSP lines are comparably docile with the other processes, due to the diminished contribution of feedstock, natural gas and hydrogen prices and the relatively larger product volume.

Fig. 9 shows the effect of different corporate tax rates and variations in capital costs to the minimum selling price. This provides some insight on potential policies on taxation and incentives for thermochemical biofuel plants. This approach, compared with NPV analysis, can be useful since the MSP is more relevant to the public, while the NPV is more relevant to companies and investors. The corporate tax rates were applied only to income, therefore the effect on profitability was in the final annual cash flows. Capital costs, on the other hand, are applied as a single negative cash flow at the start of the plant life. Moreover, it affects operational expenses through the maintenance costs, which consequently affect future cash flows.

The three processes exhibit the same trend for both tax rate and capital cost. Higher values increase MSPs. Compared with the base case tax rate, the change in tax rate influenced the MSP very slightly. A reduction from 30% to 0% tax rate changed the MSP by only 8–11%, while an increase to 50% increased the MSP by 11–15%.

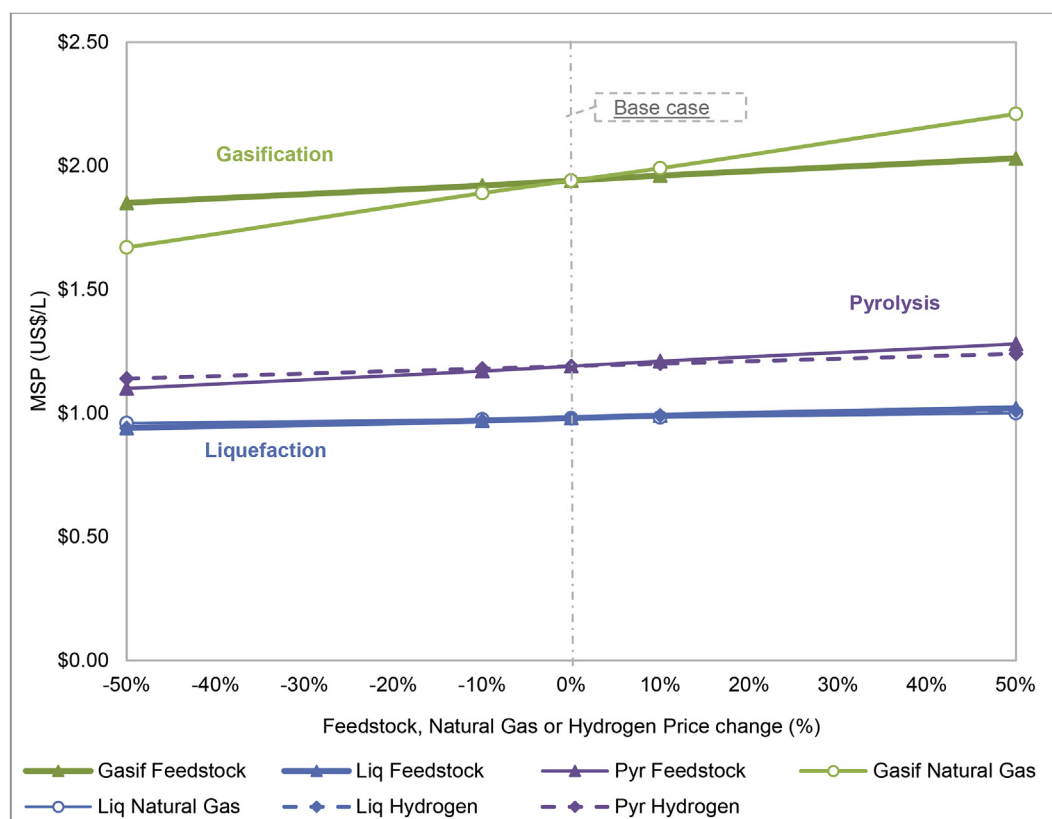
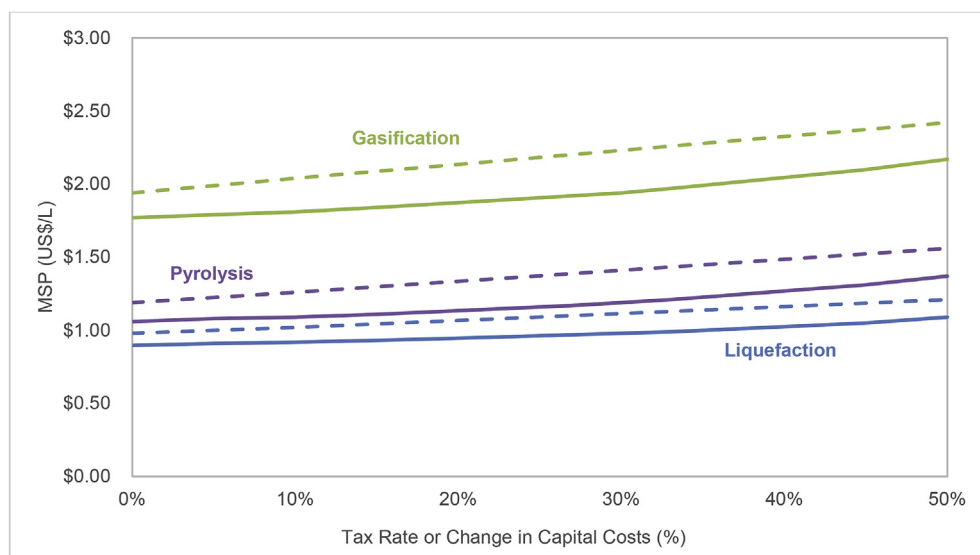


Fig. 8. Effect of varying feedstock price (triangles, solid lines), natural gas price (circles) and hydrogen price (diamonds, dashed lines) to the minimum selling price of the product.



**Fig. 9.** The effect of varying tax rate (solid line) and capital costs (dashed line) to the minimum selling price for gasification (green), liquefaction (blue) and pyrolysis (purple). Zero percent in capital costs refers to the base case cost. Base case tax rate is 30%. (For interpretation of the references to colour in this figure legend, the reader is referred to the Web version of this article.)

Considering that tax rates primarily affect company profits, changing the amount of profit does not translate to equitable cost savings for consumers, although it can be argued that lower tax rates ease the burden of unfavourable cash flows in the early years of the plant's life, and higher tax rates compensate for externalities that production might incur. The varying capital costs show similar trends with changing tax rates. Therefore, for the three plants in this study, capital-related incentives or tax holidays could have the same effect to the MSP. Of course, if both measures were applied, the MSP can further decrease, making the biofuel products more competitive and accessible to consumers.

#### 4. Conclusions

This study presented a comparison of three thermochemical process plants that can potentially produce biofuels from sugarcane bagasse, producing cleaner fuels and reducing solid wastes from sugar production. This serves as a case study for biofuel production plants that use similar lignocellulosic feedstock in thermochemical pathways. The technical and economic models presented key production and profitability aspects of the plant and their relationship to economic indicators were analysed. The process models returned unfavourable values of net present value, reflecting on the higher minimum selling price compared with the current market price. Moreover, the models highlight opportunity areas for future work to improve cost efficiency for each plant. Heating costs affect gasification greatly, while ethanol costs make up a large portion of liquefaction operating expenses.

The profitability of the three plants were sensitive to product price, thermochemical conversion and refining conversion, while input prices affect the NPV moderately. The process parameters influence the minimum selling price the most, with low production volumes intensifying MSP increases. Tax rates and capital costs influence product price but not to the extent conversion ratios do. Therefore, to promote these cleaner production technologies, it could be advantageous to focus policies on incentives towards improving process conversion ratios, increasing thermal efficiency and reducing process losses. This might be in form of grants or incentives for producers that enable them to reinvest in research to

improve production or develop ways to increase efficiency. Future work stemming from this study will be co-processing of feedstock to manage seasonal variations in supply, and the refining of crude products to match current fuel standards. Potential integration of these processes to produce biofuels for use in industry can also be investigated. Finally, stochastic methods can be used to model real variations in critical parameters and demonstrate their effect to viability.

#### Acknowledgements

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#### Appendix A. Supplementary data

Supplementary data to this article can be found online at <https://doi.org/10.1016/j.jclepro.2019.05.017>.

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