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Fischer-Tropsch route for the conversion of biomass to liquid fuels - Technical and economic analysis

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ABSTRACT

The techno-economics of biomass gasification systems for the production of Fischer-Tropsch (FT) based liquid fuels are analysed by estimating the overall mass and energy conversion of biomass to liquid (BTL) fuel. The investigation of BTL systems for 1000 kg/h biomass gasification system and an expected liquid hydrocarbon output of 1500 tonnes are estimated. The cost analysis, based on the annualized life cycle of the systems, includes a steam-oxygen based biomass gasification plant paired with the FT unit. The gasifier considered in this analysis is the downdraft reactor design, operating on oxygen-steam gasifying medium at an equivalence ratio of 0.1 and a steam-to-biomass ratio in the range of 0.8–1.2 to generate syngas with H_2/CO ratio of 2.1:1, ideally suitable for the cobalt based fixed bed FT reactor. The mass and energy balance reveal that for a once-through FT reactor configuration, substantial energy exists in the gas phase, which includes C1-C5 hydrocarbons and unconverted syngas. The study suggests that the product gas be utilized in an IC engine and converted to electricity, for in-house power demands and for the sale of excess electricity to the grid. The analysis indicates a market competitive liquid fuel production with CO conversion greater than 60%, at a cost ranging from INR 35–40/litre (0.5–0.6 USD/litre) alongside electricity as a major co-product in the BTL system. This study examines the economics of building economically affordable and environmentally favourable BTL systems of smaller throughputs with particular reference to India.

1. Introduction

The quandary of fluctuating oil prices and, the deteriorating environmental factors have sufficiently motivated the exploitation of renewable energy sources. For a country such as India, which imports nearly 75% of its crude oil, a primary requisite emerges to investigate and establish indigenous facilities for producing crude oil [1]. Biomass, is a promising renewable energy source for producing liquid fuels [2–4]. Amongst the indirect methods of producing synthetic oil from biomass, the Fischer Tropsch (FT) process is a propitious route to synthesize higher hydrocarbon compounds which can be further hydrocracked and upgraded to produce transportation fuels and chemicals [5,6]. Since the transportation fuel is predominantly bonded to the global CO_2 emissions, the use of biomass derived liquid fuels would greatly reduce the associated green house gas (GHG) emissions [7–9]. The most promising advantage of using biomass for the production of liquid fuels is that the process is carbon neutral, provided the biomass is used sustainably [10]. The production of biomass derived liquid fuels from waste biomass such as agro-residues, domestic and industrial wastes [11], perennials grown on waste lands and the use of short rotation forestry or straw are likely to greatly reduce the GHG emissions [12,13].

The recent developments in generation of syngas (combination of H_2 & CO) via biomass gasification, coupled with the FT process, has resulted in the large scale operations of biomass to liquid (BTL) fuel plants [5]. The FT reactions require syngas with H_2/CO ratio >2.1. The oxygen-steam gasification of biomass generates syngas which is ideal for FT reactions. For lower H_2/CO ratio an additional reforming reactor is required to increase the H_2 content and satisfy the FT stoichiometry. With air as a gasification medium the product gas comprises (vol%) H_2 - 18–22%; CO - 18–20%; CH_4 - 1–3%; CO_2 -

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Nomenclature

$\dot{m}_i$	mass flow rate of specie i (kg/h)
η_{FT-BTL}	Biomass to liquid fuel via FT efficiency (%)
$\eta_{gasification}$	Gasification efficiency (%)
ρ_{syngas}	syngas density (kg/m^3)
ξ_{CO}	CO conversion (%)
ALC	Annualized life cycle
C_m	Annual maintenance cost (INR)
C_{cap}	Annual capital cost (INR)
C_{ele}	Annual electricity cost (INR)
C_{op}	Annual operation cost (INR)
CRF	Capital recovery factor
CV_i	Calorific value of component i (MJ/kg)
d	Discount factor (%)
$E_{fuel-gas-output}$	Fuel gas power (MW)
$E_{i-biomass}$	Biomass input power (MW)
$E_{liquid-fuel-output}$	FT Liquid fuel power (MW)
$E_{o-syngas}$	Syngas output power (MW)
ER	Equivalence ratio
m_{cat}	Mass of catalyst (kg)
MW_i	Molecular weight of specie i (kg/kmol)
n	Component life (Year)
S_{C2-C5}	C_2-C_5 hydrocarbons selectivity (%)
S_{C6+}	C_{6+} hydrocarbon selectivity (%)
S_{CH4}	CH_4 selectivity (%)
SBR	Steam to biomass ratio
$WHSV$	Weight hourly space velocity (ml/h·g_{cat})
X_i	Volume fraction of specie i (%)

10–12% and balance N_2 [14]. However, using steam and oxygen as the gasification reagents, the water-gas reaction can be enhanced resulting in higher H_2 concentration in the exit syngas. With the use of steam-to-biomass ratio (SBR) in the range of 0.8–1.2, and an equivalence ratio of 1.0, one can get syngas with FT desired H_2/CO ratios. Furthermore, the synthesis of active FT catalysts has resulted in the increase in the yields of higher hydrocarbons that could feasibly enhance the overall BTL efficiencies [15].

The BTL fuel production economy determines the optimum scale of the BTL plant and, since the hydrocarbon generation process is associated with large investment costs, large scale production is essential for achieving effective returns. Larson et al. [16] performed techno-economic analysis on a single pass biomass-to-liquid fuel system via FT reaction. The syngas for the system is generated in a pressurized O_2 blown fluidized bed gasification system. The FT reaction is achieved in a low temperature, slurry based reactor. The analysis assumed approximately 200 tonnes/h of biomass input (893 MW_{th}) and net power generation of 228 MW_e using both steam and gas turbine. The liquid hydrocarbons were produced at a rate of 736 kl/day and the total efficiency (based on the energy output to energy input) of the plant was evaluated to be 57.3%. The total capital cost was estimated at 541 million US$ (2009) and, based on the annual levelized fuel production cost, the break-even FT oil production cost was estimated at 0.48 $/l (assuming a biomass cost of $ 50.9/t). Similar work reported by Tock et al. [17] estimated the BTL efficiency of 59.8% and a FT crude oil production cost of 0.9 €/L. The economic development of the established plant, by improved gasification systems, gas cleaning, cooling and purification sections and, the use of advanced high performance catalysts would increase the economic competitiveness in the market for BTL fuels. The techno-economic analysis performed by Anex et al. [18] compared the gasoline production cost for three conversion processes - pyrolysis, Fischer-Tropsch and the bio-chemical route (for 2000 tonne/day corn stover feedstock). The gasoline production cost via FT process was estimated at 1.21 $/l compared to 0.80 $/l via the pyrolysis process. The higher cost for the FT derived gasoline compared to that produced by the pyrolysis process was attributed to higher associated capital cost [18]. The well-to-wheel analysis of different types of FT based plants using coal, natural gas, biomass and mixed-feedstock fuels was conducted by Van Vliet et al. [19]. The analysis indicated that the biomass based liquid fuel production costs break even when the conventional oil prices are greater than 0.47 $/l. Further, the BTL based transportation fuels result in 32–63 g CO_2/km emission compared to 123 g CO_2/km for coal based FT plant [19]. The BTL techno-economic analysis based on low temperature (1143 K) fluidized bed gasification layout and high temperature (1573 K) entrained flow gasification system was performed by Swanson et al. [20]. The analysis showed a comparatively lower liquid fuel cost (1.06–1.32 $/l) for the high temperature gasification scenario owing to the higher syngas yields, resulting in higher biomass conversions. The sensitivity analysis results also showed distinctly significant effect of the capital and the feedstock cost on the product value [20]. Clearly, the literature reports a large variation in the FT based liquid fuel production cost, but most of the literature treats extremely large throughputs needing billion dollar investments. While the capital costs, the operation & maintenance cost and the biomass cost itself vary from one region to the other, the process technology coupled with its conversion efficiencies largely affect the overall economies [21].

In general, a vast majority of the literature have evaluated the BTL techno-economics for scales ranging from 80 to 300 tonnes per hour. The operation of BTL facilities at such voluminous scales, though advantageous, highlights the concerning unpredictability of the availability of biomass. The current work analyzes the economic feasibility of a biomass-to-liquid fuel plant for a one tonne per hour fixed-bed, oxy-steam biomass gasification system coupled to a fixed bed low temperature FT reactor. The overall economics of building such systems at smaller throughputs are examined with a particular reference to India. The cost of the liquid fuel is assessed using the annualized life cycle of the systems involved. The global conversions that include conversion of biomass to syngas are evaluated from the experimental work by Sandeep and Dasappa [22] and the FT activity and the hydrocarbon yield of cobalt based catalysts that comprise of the BTL building block are assessed using the experimental work by Ail and Dasappa [15] and Ail [23]. Detailed capital cost estimates for each section involved in the BTL system are set out, on the basis of which future commercial large-scale facilities producing liquid fuel or hydrogen are evaluated. Several elements of the total system are available as off-the-shelf items and the gasifier itself has originated from know-how and long experience of air biomass gasification at the Indian Institute of Science [24]. The total capital cost and the electrical input required for the various sub-systems were obtained from the costs of off-the-shelf items from e-market sources and subject to critical evaluation internally [25].

2. Process flows and, mass and energy balance

2.1. Biomass gasification system

The process layout considered in this study assumes the principles of open-top downdraft reburn gasifier with a lock-hopper arrangement for oxygen and steam injection [22]. Casuarina wood-chips are used as biomass feed to generate clean syngas that is further processed for generating liquid fuels via FT reaction. These wood chips are dried at 378 K to reduce the moisture content below 1%. Table 1 lists the proximate and ultimate analysis details of the selected biomass.

The gasifier is designed for 1000 kg/h biomass consumption rate and the details regarding the principle of operation are described in Ref. [22]. The gas composition evolving from a downdraft gasifier, as used in the current study, chiefly depends on the reactions that occur in its reduction zone and are described by the water gas reaction, Boudouard reaction, water gas shift reaction, and the methanation reaction.

The single reactor gasifier utilizes a lock hopper mechanism for fuel loading enabling the injection of oxygen and steam into the biomass bed by separating the downdraft reactor from ambient atmosphere. The ash extraction system at the base of the reactor assists in the removal of char/ash at regular intervals and facilitates in its steady operation. Steam is generated using a gas fired boiler with a peak generation capacity of 750 kg/h and oxygen is generated using a 250 kg/h air separation unit. The synthesis gas contains very low tar and particulate matter apart from inorganic and sulphur containing compounds. The tar and particulate matter removal system of the open top downdraft gasifier developed at Indian Institute of Science consists of cyclone separators, specially designed scrubbers, an additional chilled scrubber and a final fabric filter section to remove particulate matter less than 5μ. The cleaned gas has tar concentrations of 3 mg/Nm^3, which is within the acceptable limits for FT process [26]. Overall, the syngas purification system effectively reduces impurities to levels which minimize the catalyst poisoning and has no deteriorating impact on the overall conversion and product yields. An effluent treatment plant is used to treat the tar contaminated water.

Amongst the process conditions for oxy-steam biomass gasification, the SBR and ER have a significant effect. Increasing SBR, elevates the H_2 concentration in the syngas and at the same time, the adiabatic flame temperature reduces due to the endothermic nature of steam gasification (the water gas reaction). As described in Ref. [22], SBR must be in the range of 0.8–1.2, to achieve syngas with H_2/CO in the ratio 2.1:1, which is an ideal case for cobalt based FT system. Simultaneously, the ER affects the bed temperature, and has to be selected appropriately in order to maintain the H_2/CO ratio. An adiabatic temperature of 1200 K is preferable for enhancing the H_2 concentration in the syngas and therefore, an ER of 0.1 is maintained, and SBR ranging from 0.8 to 1.2. Fig. 1 shows the 10 kg/h oxygen-steam downdraft biomass gasification unit. Table 2 gives the mass flow rate of the components involved in the biomass gasification system. The gasifier is designed for 1000 kg/h biomass consumption rate, utilizing steam and oxygen at a rate of 660 kg/h and 230 kg/h respectively. The exit syngas flow rate is obtained at 1350 kg/h. The residual char and condensed water are produced at 20 and 515 kg/h respectively. The exit gas syngas comprises volume fraction of H_2 - 47 ± 2%, CO - 22 ± 3%, CO_2 - 27± 1.6% and CH_4 - 4 ± 0.5%. The equations for the mass and energy balance of the gasification system, and the gasification efficiency are described in Eqns. (1)–(5).

Table 1
Proximate and ultimate analysis of casuarina wood chips [23,36].

	Mass fraction (%, dry basis)
Proximate analysis	
Fixed carbon	18.38
Volatile matter	81.28
Ash content	0.34
Calorific value	18.2 MJ/kg
Ultimate analysis	
Carbon	42.83
Hydrogen	6.24
Oxygen	50.39
Nitrogen	0.12
Sulphur	0.42
Chemical composition	$CH_{1.7}\,O_{0.9}$
Molecular weight	23 g/mol

$$MW_{syngas}(kg/kmol) = \frac{\sum(X_i \times MW_i)}{100} \tag{1}$$

$$\rho_{syngas}\left(kg/m^3\right) = \frac{MW_{syngas} \times 10^5}{8.314 \times 298 \times 1000} \tag{2}$$

$$E_{i-biomass}(MW) = \frac{18.2 \times \dot{m}_{biomass}}{3600} \tag{3}$$

$$E_{i-syngas}(MW) = \frac{\sum(X_i \times CV_i) \times 1350}{MW_{syngas} \times 3600} \tag{4}$$

$$\eta_{gasification}(\%) = \frac{E_{i-syngas}}{E_{i-biomass}} \times 100 \tag{5}$$

Considering the biomass calorific value of 18.2 MJ/kg, and the evolving syngas calorific value of 10.6 MJ/kg, a 79% gasification efficiency is achieved.

The time variation of gas concentration under these process conditions, is shown in Fig. 2. The exit syngas composition as shown in the figure were recorded for a duration of six hours. The mass and energy balance for the techno-economic analysis consider these volumetric fractions of the syngas for FT reaction. The exit syngas is scrubbed of its CO_2 with the help of a commercially established Rectisol technology. Rectisol is a physical acid gas removal process which employs an organic solvent, such as methanol, at subzero temperatures and can purify synthesis gas upto 0.1 vppm total sulphur (including COS) and CO_2 to ppm range [27]. Though sulphur concentrations are low in biomass derived syngas, it is crucial to reduce the H_2S and COS concentration to sub-ppm levels to prevent any catalyst deactivation. CO_2 is known to dissociatively adsorb on the catalyst surface and is hydrogenated to form higher hydrocarbons. Yao et al. [28] demonstrated the effective hydrogenation of CO_2 on the cobalt catalysts leading to the formation of higher hydrocarbons. However, unlike CO hydrogenation, the CO_2 reaction produces methane-rich short chain hydrocarbons. Hence CO_2, if introduced into the FT catalyst bed, acts as an effective reactant, *significantly affecting the reaction stoichiometry*, varying the H_2/CO consumption ratio and consequently reducing the syngas conversion rates. The gas stream from Rectisol system consists of syngas with 68 vol% H_2 , and 32 vol% CO. The syngas is then compressed to 30 bar before making a single pass through the FT reactor.

2.2. Fischer-Tropsch unit

The FT unit considered for this study employs a single pass multi-tubular fixed bed reactor maintained at 503 K and 3 MPa. The advantages of a multi tubular fixed bed reactors include ease of handling and maintenance and, convenient product separation, since the heavy wax products trickle down the bed and are collected in the receiver pot. Most importantly, the performance and the behaviour of large scale plants can be predicted accurately based on the performance of pilot scale plants [29]. The reactor tubes are packed with combustion synthesized 20 wt% Co supported on silica doped alumina (SDA) catalysts. The enhanced

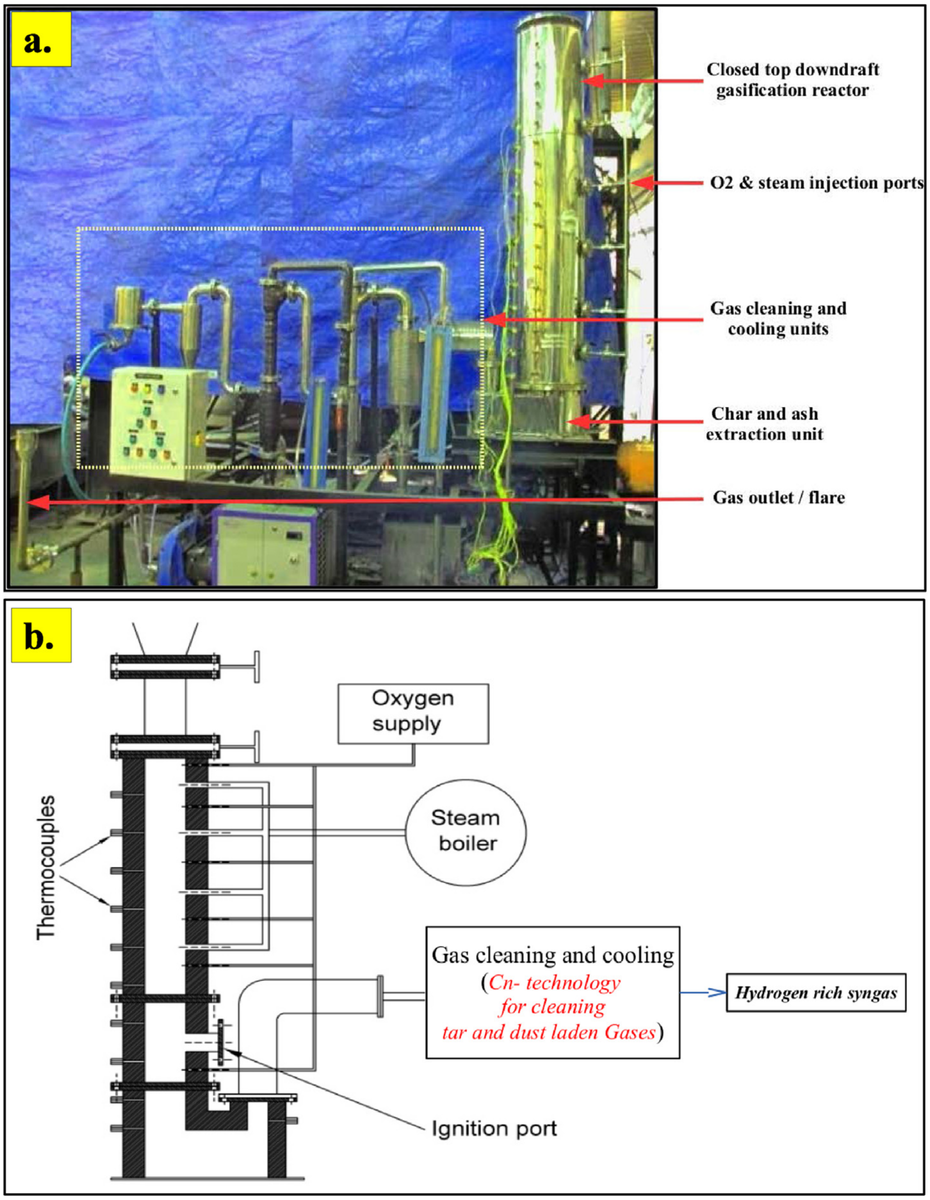

Fig. 1. (a). 10 kg/h oxygen-steam downdraft gasification system developed at Indian Institute of Science, Bangalore; (b.) Gasifier layout [22,23].

performance and higher C_{6+} yields with these catalysts are presented in detail in Ail and Dasappa [15]. The hydrocarbon product spectrum under the prevailing process conditions comprised of mixed fractions of waxes and middle distillates. The syngas conversion and product formation rates in the FT reaction section are considered for varying weight hourly space velocities (WHSV) - 2610 ml/(h·g_{cat}), 1730 ml/(h·g_{cat}), 1310 ml/(h·g_{cat}) and 873.3 ml/(h·g_{cat}). In each of the cases, the CO conversions, hydrocarbon selectivity and overall yield are considered from the lab-scale experiments investigated by Ail and Dasappa [15]. The yield of higher hydrocarbons (C_{6+}) per unit mass of syngas for varying WHSV are shown in Fig. 3. Obviously, the decreasing space velocity increases the residence time and improves the output of higher hydrocarbons. A reduced WHSV requires larger FT reactor and liquid fuel processing volumes. Simultaneously, increasing reactor volumes result in elevated capital costs and higher energy requirements.

Table 3 lists the FT reaction details. The packing density of catalyst is estimated to be 850 kg/m^3 and the mass of catalyst used in the FT reaction is calculated by dividing the volumetric syngas flow rate with the weight hourly space velocity (WHSV), as shown in Eqn. (6). As the WHSV is reduced from 2610 ml/h*g_{cat} to 873.3 ml/h*g_{cat} (i.e., varying the syngas residence time in the catalyst bed from 6 s to 18 s respectively), the catalyst mass correspondingly increases from 451 kg to 1348 kg. Proportionately, the CO conversion increases from 43% to 73% and the C_{6+} yield increases by 85%. Eventually, the ratio of mass of liquid fuel to the mass of gas phase hydrocarbons (referred to as fuel-gas, which also includes the unconverted syngas), increases. This ratio is termed as the ***fuel ratio***. Further, based on the mass of the catalyst required for syngas conversion, the reactor volume is established. The reactor volume increases close to three times (0.5 m^3 to 1.6 m^3) as the WHSV is decreased from 2610 ml/h*g_{cat} to 873.3 ml/h*g_{cat}. The reactor is designed similar to a commercial fixed bed FT reactor, with a single tube diameter of 0.05 m for a catalyst particle size ranging from 1.5 to 2 mm [30]. The syncrude leaves the reactor in two separate streams. One stream is in liquid state, consisting primarily of waxes (C_{24+}), hot condensate (C_{12}-C_{24}), cold condensate (C_5-C_{12}) and aqueous product (reaction water, alcohols,

Table 2
Biomass gasification - mass & energy parameter.

Input	m^* (kg/h)
Biomass	1000
Steam	660
Oxygen	230
Output	**m^* (kg/h)**
Syngas	1350
Char	20
Water	515.2
Syngas composition	**Vol%**
X_{H2}	47 ± 2
X_{CO}	22 ± 3
X_{CO2}	27 ± 1.6
X_{CH4}	4 ± 0.5
MW_{syngas}	19.6 kg/kmol
ρ_{syngas}	0.79 kg/m^3
$E_{i-biomass}$	4.9 MW
$E_{o-syngas}$	3.9 MW
$\eta_{gasification}$	79.5%
Syngas composition after CO_2 & CH_4 removal	
X_{H2} (vol%)	68
X_{CO} (vol%)	32
$m^*_{FT-syngas}$	488.5 kg/h

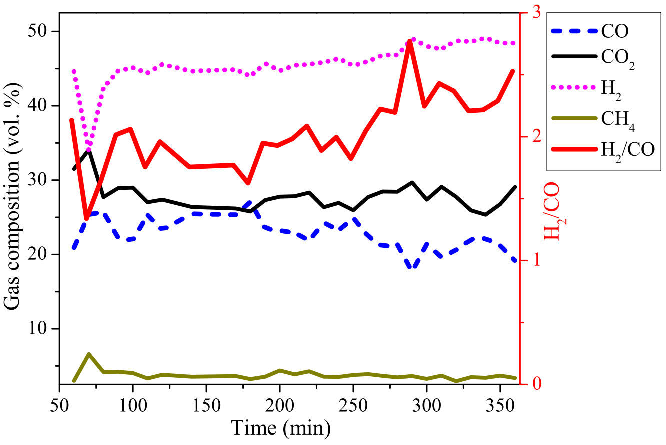

Fig. 2. Variation in the syngas composition with time; SBR = 0.8, ER = 0.8 [22].

carboxylic acids and dissolved paraffins and olefins). The second stream is gas phase products consisting of unconverted syngas and lighter gas phase hydrocarbons. A fraction of the exit gas products are used for the generation of steam in a gas fired steam boiler and the balance gases are used in an IC engine to generate electricity for in-house power application and for supply to the grid. The wax products are separated and directed to a hydrocracker where they are converted into high quality liquid transportation fuel. The mass flow rates of C_{6+} and the gas phase hydrocarbons, which include the unconverted hydrocarbons are evaluated as shown in Eqns. (7) and (8). The total energy output of the liquid fuels, expressed in MW, is evaluated assuming a calorific value of 40 MJ/kg. Fig. 4 shows the overall process layout for the conversion of biomass to liquid fuel via oxy-steam gasification and FT synthesis.

$$m_{cat}\ (kg) = \frac{\dot{m}_{syngas} \times 1000}{\rho_{syngas} \times WHSV} \tag{6}$$

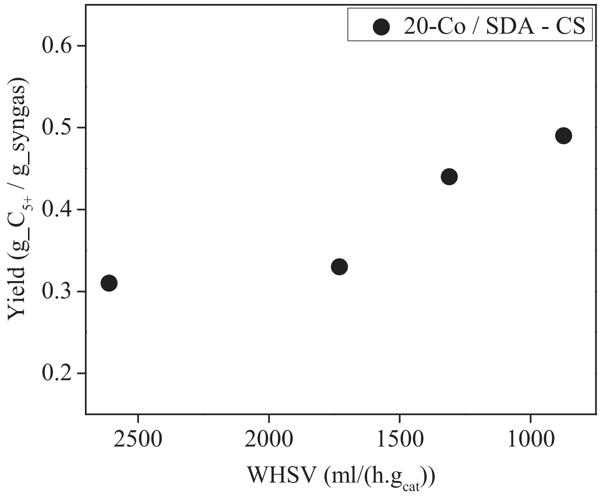

Fig. 3. Variation of C_{6+} yields with space velocity. (*Note that the space velocity is presented in the decreasing order*) [15].

Table 3
FT reaction - mass & energy related parameters; T = 503 K, P = 3 MPa.

Parameter	Case 1	Case 2	Case 3	Case 4
WHSV (ml/h*g_{cat})	2610	1730	1310	873.3
m_{cat} (kg)	451.0	680.4	898.6	1347.9
Reactor volume (m^3)	0.5	0.8	1.1	1.6
ξ_{CO} (%)	42.9	51.2	65.6	73.1
ξ_{H2} (%)	54.7	61.4	66.1	80.6
S_{CH4} (wt.%)	14.6	13.9	12.8	11.4
S_{C2-C5} (wt.%)	8.6	7.9	6.9	5.3
S_{C6+} (wt.%)	75.7	75.6	77.9	78.8
S_{CO2} (wt.%)	1.10	2.49	2.39	4.58
$\dot{m}_{C6+}$ (kg/h)	136.8	161.2	215.0	239.4
$\dot{m}_{fuel-gas}$ (kg/h)	351.7	327.3	273.5	249.2
Calorific Value $_{Fuel-gas}$ (MJ/kg)	27.2	28.3	31.3	32.7
$E_{liquid-fuel-output}$ (MW)	1.5	1.8	2.4	2.7
$E_{fuel-gas-output}$ (MW)	2.7	2.6	2.5	2.2
η_{FT-BTL} (% of biomass)	30.6	36.0	48.0	53.5

$$\dot{m}_{C6+}\ (kg/h) = \frac{S_{C6+} \times \xi_{CO} \times \dot{m}_{CO-in}}{10000} \tag{7}$$

$$\dot{m}_{fuel-gas}(kg/h) = \frac{(100 - S_{C6+}) \times \xi_{CO} \times \dot{m}_{CO-in}}{10000} + \dot{m}_{CO-out} + \dot{m}_{H2-out} \tag{8}$$

3. Methodology

This section describes the approach for determining the liquid fuel cost based on the annualized life cycle (ALC) of the systems involved. The ALC is evaluated by taking into account the annualized capital cost, annual fuel cost and annualized current value of all the future costs that include, operation and maintenance cost, equipment replacement cost, and electricity cost. The ALC of the liquid fuel is determined by dividing the ALC of the system with the annual liquid product output from the BTL plant. Eqns. (9) and (10) show the expressions for the ALC calculations. Here, C_{cap}, C_{op}, C_m, C_{ele} are the annualized capital cost, annual operation cost, annual maintenance cost, and the annual electricity cost respectively. CRF is the capital recovery factor, *d* and *n* are the discount factor and the system life. All costs are evaluated in Indian Rupee (INR) and the

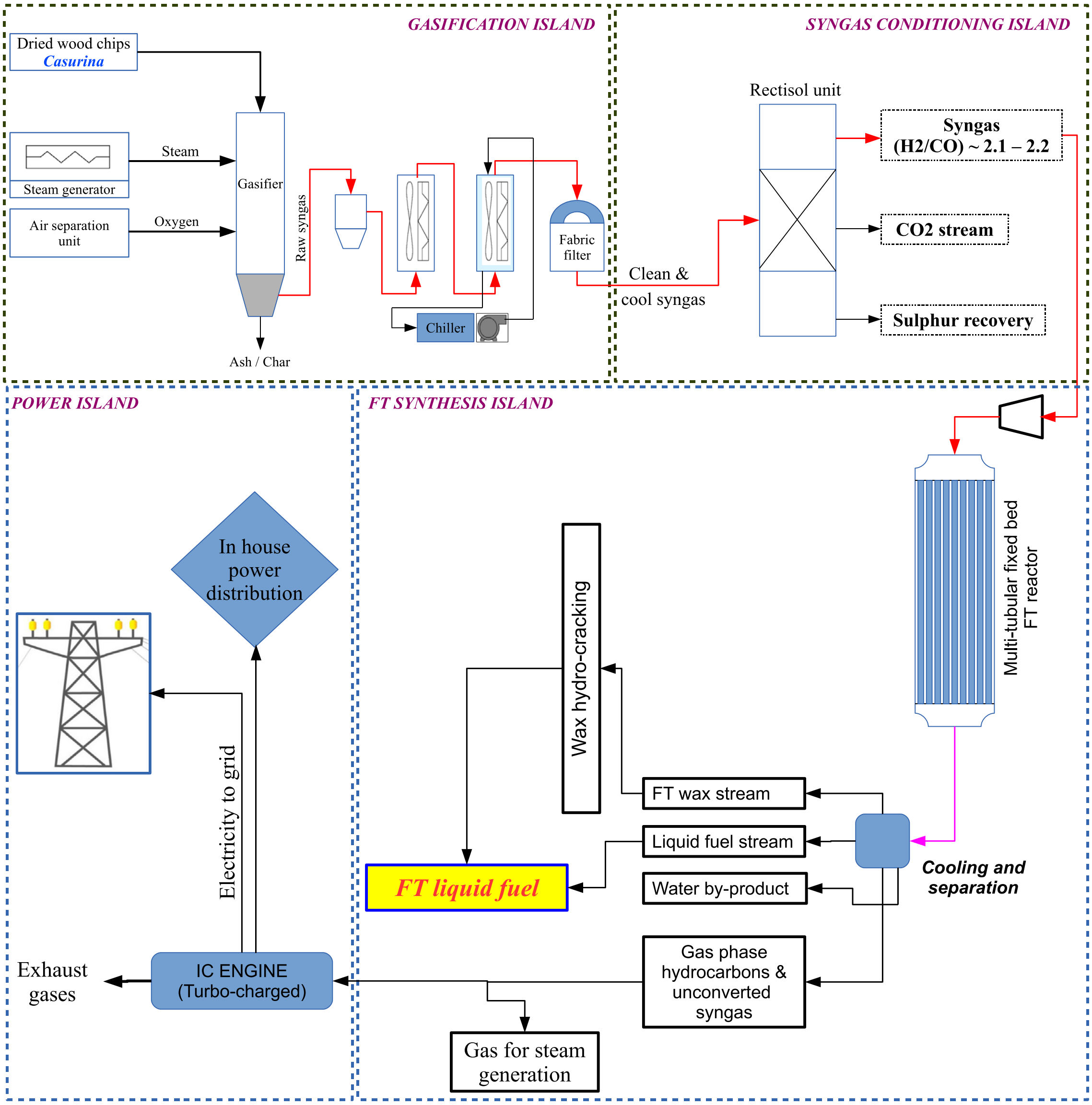

Fig. 4. Overall process layout for BTL via oxy-steam gasification and FT synthesis.

Table 4
Assumptions for economic analysis.

Component	Parameter
Grid electricity cost (INR/kWh)	6
Discount factor (%)	12
Annual maintenance cost (INR)	5% of capital cost
Total plant manpower	19
Per head manpower cost (INR/h)	100
Daily operational hour (h)	24
Annual operation (days)	300
1 US \$ = INR (Indian Rupees)	66.77 (October 06, 2016)

conversion to US dollar (USD) at this time is INR 66.77/USD. The assumptions involved in the economic analysis are listed in Table 4. The costs reported in this techno-economic analysis are considered for the baseline year of 2015–2016.

$$ALC = \frac{(C_{cap} \times CRF) + C_{op} + C_m + C_{ele}}{\left(\dot{m}_{C6+} \times 300 \times 24 \times 1000 \big/ \rho_{liq}\right)} \tag{9}$$

$$CRF = \frac{d \times (1+d)^n}{(1+d)^n - 1} \tag{10}$$

The total operating cost, which is the costs related to the total manpower required for various plant activities, is INR 1900/h. The annual maintenance cost is assumed as 5% of the total capital cost [31]. The total operation and maintenance (O&M) cost are generally the expenses incurred in the servicing and repair of associated equipments, manpower cost, facility planning and management and services cost. During the duration of plant operation, the costs associated with these elements may change, affecting the product cost. Electricity from the grid is obtained at INR 6/kWh [32] and the plant is operated for 7200 h annually. The discount factor, which is required to evaluate the net present value of the various operating sub-systems, is assumed to be 12%. The associated capital costs for each of the components are specified in Table 5. All the costs are expressed in Indian rupees (INR) and the capital cost were obtained from literature sources, equipment vendor quotes, industry experts and from the estimates of earlier work [31] [33]. Since the FT

Table 5
Estimated installed capital, operation and maintenance cost.

Unit	Capacity (kg/h)	Power use (kW_e)	Capital cost (INR)	Annual O&M cost (INR)	Life (years)
ASU	250	20	9750000	1927500	10
Steam generator	750	510	1495000	2234750	5
Gasifier	1000	100	30000000	5100000	15
CO_2 separator	820	50	22750000	3297500	10
FT reactor[a]	500	250	15000000	4350000	10
FT reactor[b]	500	250	22500000	4725000	10
FT reactor[c]	500	280	30000000	8700000	10
FT reactor[d]	500	300	45000000	9450000	10
Engine	0.75 MW_e	–	24000000	2220000	10

[a] WHSV = 2610 ml/(h*g_{cat}).
[b] WHSV = 1730 ml/(h*g_{cat}).
[c] WHSV = 1310 ml/(h*g_{cat}).
[d] WHSV = 873.3 ml/(h*g_{cat}).

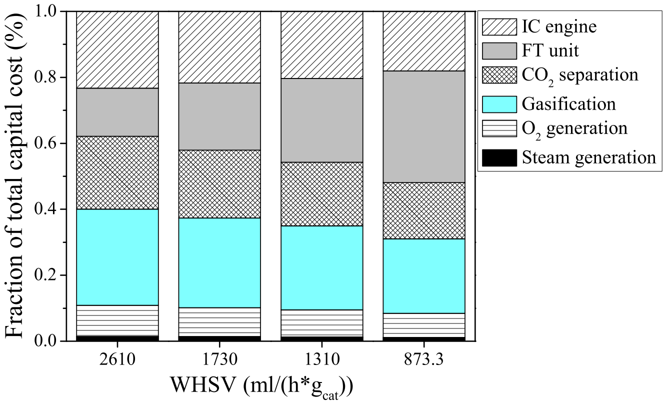

Fig. 5. Varying capital cost fraction of individual BTL components as a function of decreasing syngas WHSV in the FT unit.

reactor volume increases as the WHSV decreases from 2760 ml/(h*g_{cat}) to 873.3 ml/(h*g_{cat}), the fraction of FT associated capital cost increases from 15% to 34% of the total capital cost. Fig. 5 shows the increase in the fraction of the capital cost required for the FT section with decreasing weight hourly space velocity. For higher space velocities, the maximum capital cost is required for installation of oxy-steam gasifier along with the gas cleaning and cooling components. The gas cleaning and cooling unit, which comprises of a cyclone separator, water scrubber, water-chiller and a fabric filter, is a synchronized and customized accessory along with the oxy-steam gasification unit.

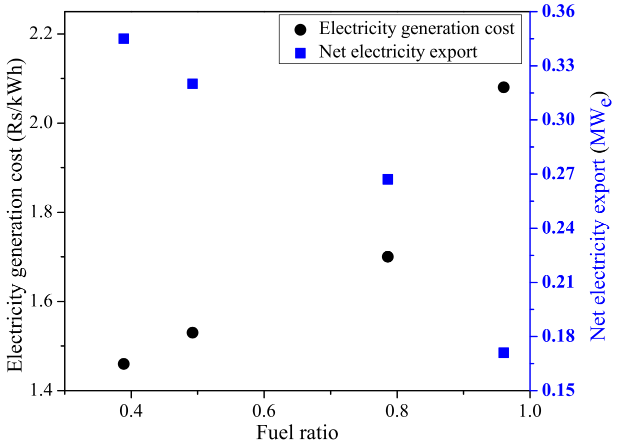

Fig. 6. Effect of increasing fuel-ratio on the net-grid electricity export and on the electricity generation cost.

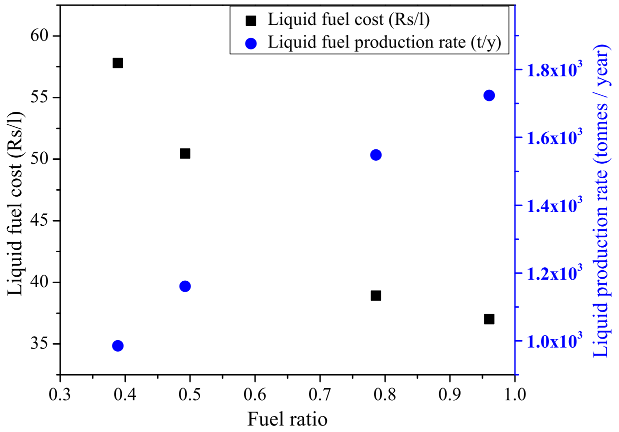

Fig. 7. Effect of increasing fuel-ratio on annual production rate and the liquid fuel cost.

4. Results and discussion

This section describes the impact of conversion and yield on the BTL economics for the scale described above. As shown in Table 2, following the cleaning and cooling process, the 1000 kg/h biomass gasifier yields 488.5 kg/h of FT suitable syngas with H2:CO = 2.1:1. Moreover, the hydrocarbon spectrum distribution with the combustion synthesized cobalt based catalysts yields mixed fractions of waxes and middle distillates with nearly 60 wt% of higher hydrocarbons in the form of waxes [23]. By decreasing the WHSV in the FT reactor, the fuel ratio increases from 0.4 to 1.0, indicating a reduced fraction of gas phase hydrocarbons in the product stream. The BTL system is designed such that a part of the fuel-gas energy is used for generating steam for the gasification reaction and the rest is converted to electricity in an internal combustion engine. The engine is a turbocharged producer gas fuelled engine with a rated power output of 750 kW_e and an efficiency of 28% [34,35]. A fraction of the generated electricity is used for in-house applications that include the air separation unit (ASU), gasifier, CO_2 separator and the FT system, and the rest is exported to the grid at the rate of INR 6/kWh. Decreasing fractions of fuel-gas ratio reduces the net electricity export to the grid. As a result, the cost of electricity per kilo watt hour (kWh) increases and also decreases the gross cash inflow emerging from the sale of electricity as a product. Fig. 6 shows a decreasing trend in the net grid electricity export with the increasing fuel ratio and a consequent increase in the effective unit electricity cost. The total electricity available for export to the grid decreases from 0.35 MW_e to 0.17 MW_e as the fuel ratio increases from 0.39 to 0.96.

As the WHSV is reduced from 2610 ml/(h·g$_{cat}$) to 873.3 ml/(h·g$_{cat}$), the CO conversion increases from 43% to 73.1% and the hydrocarbon yield increases from 0.28 kg$_{C6+}$/kg$_{syngas}$, to 0.49 kg$_{C6+}$/kg$_{syngas}$ (as shown in Fig. 3), with a simultaneous decrease in the fuel ratio. Fig. 7 shows a decrease in the liquid fuel cost with increasing fuel ratio. The unit price of the liquid fuel decreases from INR 57.80/l to INR 37.01/l, as the fuel ratio increases from 0.39 to 0.96. Equivalently, the liquid fuel production increases from 1000 tonnes/year to 1800 tonnes/year. The analysis distinctly evidences that for a once-through FT reactor system, market competitive liquid fuel prices can be achieved only with CO conversions greater than 55–60% and when the electricity is produced as a co-product that contributes to the overall reduced liquid fuel prices.

A possible way to determine the volume of profitability of a BTL plant, is to estimate the internal rate of return (IRR) for the varying fuel ratio scenarios. A project with a significantly higher IRR value compared to other reviewed options, indicates a higher rate of growth that a project is expected to return. The present diesel cost in India ranges from INR 45/litre to INR 56/litre. For the FT operating condition with fuel ratios of 0.39 and 0.49, the estimated liquid fuel cost are INR 57.8/litre and INR 50.4/litre, respectively. Under these conditions, no profits can be obtained by the sale of generated liquid fuels. However, the volume of net profits generated is mainly due to the revenues earned from the export of electricity to the grid at INR 6/kWh. Even under these conditions, the IRR are 27% and 32% for the BTL plant with the fuel ratio of 0.39 and 0.49, respectively, which is a clear indication of electricity as major co-product of the BTL plant. The net investment of INR 68000/ton (fuel ratio = 0.4) and INR 58050/ton (fuel ratio = 0.5) of liquid fuel has a payback period of 4.5–5 years. On the other hand, the net investment of INR 45000/ton (fuel ratio = 0.78) and INR 41310/ton (fuel ratio = 0.96) of liquid fuel has a payback period of 3–3.5 years, with an IRR of 43.17% and 45.06%, considering a liquid fuel selling price of INR 52.1/l and the export electricity unit price of INR 6/kWh. These parameters clearly indicate that a BTL plant with the CO conversion greater than 60% (fuel ratio >0.8) results in an economical and sustainable production of liquid fuels.

Fig. 8 shows the variation in the internal rate of return for varying liquid fuel selling price. Ideally, if the IRR of the contemplated BTL scenario is greater than demanded rate of return on the total investment, that framework is desirable. In Fig. 8, for fuel ratio 0.39 and 0.49, the liquid fuel selling price needs to be greater than INR 40/litre for an IRR >20%. Above all, under these conditions, the positive revenues for the BTL plants are earned entirely from the sales of generated electricity, instead of liquid fuel. For fuel ratio of 0.96, the IRR increases by 19.7% with a 45% increase in the liquid fuel selling price. Similarly, for the fuel ratio of 0.78, the IRR increases by 15.3% when the liquid fuel selling price increases from its production cost of INR 38.86/litre to INR 52.14/litre.

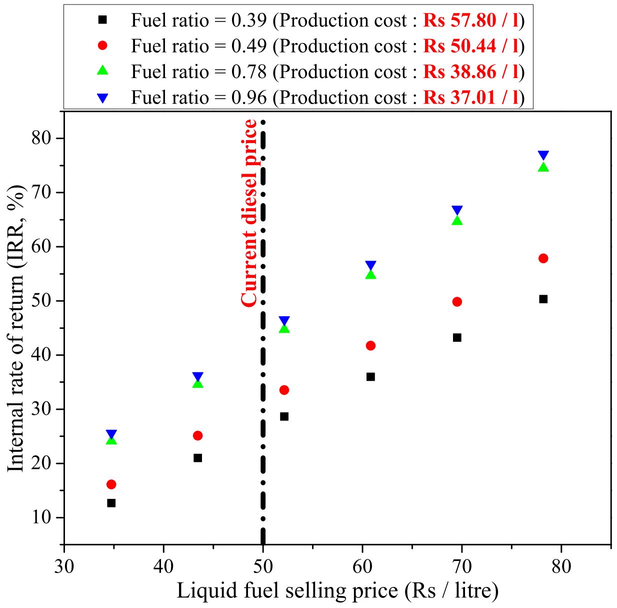

Fig. 8. Effect of the variation in the liquid fuel cost on the IRR.

4.1. Sensitivity analysis for cost parameters

Biomass being the raw material, its variation in the cost is likely to affect the liquid fuel cost. Foreseeing biomass cost is difficult since it relies on assorted factors, such as local supply chain, resource availability, processing costs, land availability, deforestation risks, simultaneous competitive uses and the sustainability touchstone [5,23]. Each of these crucial factors influence the biomass cost directly, or indirectly. Above all, the sustainability of biomass based power plants is steadily bound to the long term availability of biomass feedstock and its low costs. If agro residues or industrial wastes are employed as a source of biomass, which otherwise incur disposal costs, the feedstock costs may be reduced. Similarly, for biomass produced on site, the transportation cost of biomass can be reduced substantially to further reduce the unit costs of biomass. Fig. 9 shows the effect of variation in the cost of liquid fuel with varying biomass cost. The base cost of biomass is considered to be INR 4/kg and, a two fold increase in the biomass cost increases the fuel cost by 40%–45%. For a fuel ratio of 0.96, a doubling increase in the biomass cost, elevates the cost of liquid fuel from INR 37.01/l to INR 50.4/l. Similarly, for a fuel ratio of 0.4, a two fold increase in the biomass cost, increases the liquid fuel cost from INR 57.80/l to INR 86.5/l. The analysis clearly indicates that a sustainable and cost effective source of biomass is absolutely essential for continual and propitious operation of the BTL plant. The effect of variation in the capital cost on the product cost is shown in Fig. 10. For a fuel ratio of 0.96, a two fold increase in the total capital cost increases the liquid fuel cost from INR 37.41/l to INR 41.02/l, signifying a slender 9.5% increase in the product cost. A two fold increase in the biomass cost has a more severe consequences on the liquid fuel cost compared to a proportional increase in the capital costs.

Identical to the biomass cost, the operation and maintenance (O&M) cost is subject to variation. Fig. 11 shows the effect of

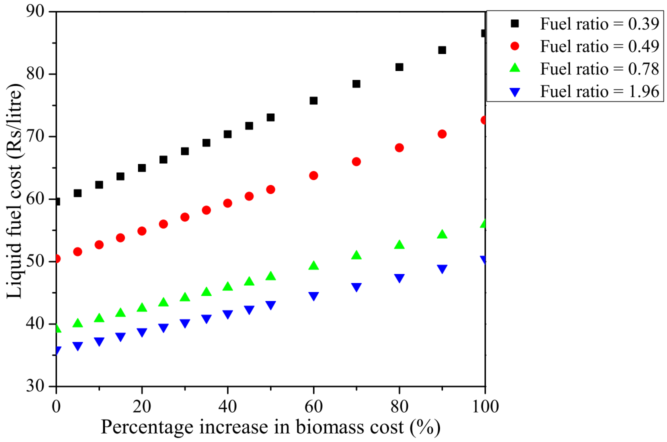

Fig. 9. Liquid fuel cost variation with variation in the biomass cost.

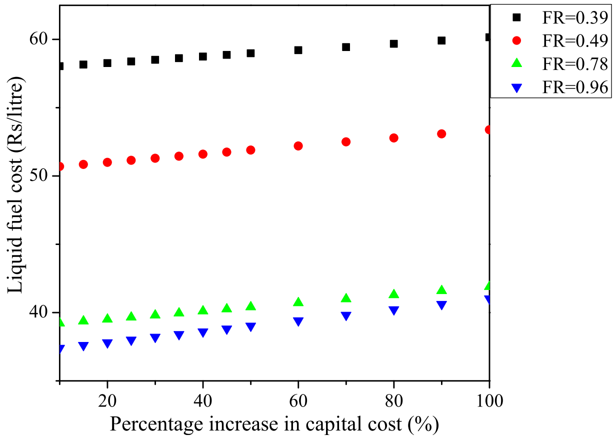

Fig. 10. Liquid fuel cost variation with variation in the capital cost.

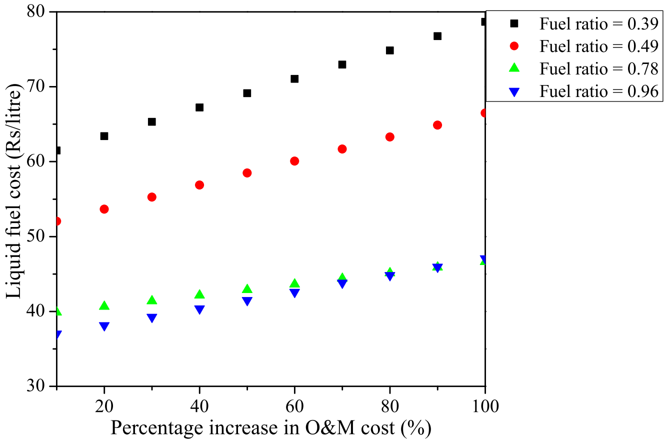

Fig. 11. Liquid fuel cost variation with variation in the operation and maintenance cost.

variation in the O&M costs on the liquid fuel costs. On an average, the cost of liquid fuel increases by 10%–14% with 50% increase in the O&M cost. Clearly, the percentage increase in the fuel cost with O&M costs is not as severe as that caused by the increase in biomass cost.

5. Conclusions

The extensive techno-economic analysis established that the thermochemical gasification of biomass route to produce liquid fuels via Fischer Tropsch synthesis can be commercially and sustainably established and sustainably operated even at biomass throughput levels of 1000 kg/h. The four different conversion scales of FT reactors considered for the analysis, i.e. a CO conversion ranging from 43% to 73% suggested that for a market competitive liquid fuel generation, the CO conversion should be maintained above 60%. It was shown that the liquid fuel cost for CO conversion of 65.6% and 73.1% yielded liquid fuels priced at INR 39.1/litre and INR 35.9/litre. Electricity is a major co-product in the BTL system. The estimates of the payback period for the return on gross investment showed that the FT based BTL facility has a payback period ranging from 5 to 3 years depending on the extent of syngas conversion, particularly for a once through fixed bed FT reactor system. The payback period ranged from 4.5 to 5 years for a fuel ratio of 0.5 (net investment of INR 58050/ton of liquid fuel) and from 3 to 3.5 years for a fuel ratio of 0.96 (net investment of INR 41310/ton of liquid fuel). Moreover, since a considerable fraction of the FT products remain in the gas-phase, it was observed that the combustion of the fuel gas in an IC engine can be used to generate electricity, for export to the grid at INR 6/kWh. Another major factor affecting the liquid fuel cost, is the price of biomass. The cost analysis has indicated that a 50% increase in the biomass cost results in a 15–18% increase in the liquid fuel cost. For a sustainable and long-term operation of the BTL plant, a consistent and economic source of biomass is absolutely crucial.

In contrast to all other studies and industrial explorations in BTL that have always used million tonnes per year scenario leading to an impression that at throughput less than this value, the technology is economically unsustainable, this study shows that even at a fraction of this throughput it is economically sustainable. This conclusion acquires greater importance in view of biomass supply chain management at large throughputs, especially for developing nations.

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