Investment cost estimates for gasification-based biofuel production systems

Kristina M. Holmgren

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Summary

This report contains supplementary data for the work presented in Holmgren et al. (2015) in which investment costs for three gasification based biofuel production systems, producing SNG, methanol and Fischer-Tropsch fuels respectively, are estimated and compared to other estimates for similar systems in literature. The paper Holmgren et al. (2015) was also presented at the Conference on Process Integration, Modelling and Optimisation for Energy Saving and Pollution Reduction (PRES 2015) in Kuching, Malaysia, 22-27 August 2015.

Sammanfattning

Denna rapport innehåller bakgrundsdata för arbetet som presenteras i Holmgren et al. (2015). Holmgren et al. (2015) uppskattar investeringskostnader för tre förgasningsbaserade biodrivmedels-produktionssystem, för SNG, metanol och Fischer-Tropsch bränslen, och jämför med kostnader för liknande system som presenterats i litteraturen.

1 Introduction

This report presents details on assumptions and data for equipment cost estimates employed for the investment cost estimates for three biomass based gasification systems producing, synthetic natural gas (SNG), methanol and FT (Fischer-Tropsch) fuels respectively in the study by Holmgren et al. (2015). Process flow diagrams and required inputs and outputs for each of the systems are presented. Even though gasification technologies and syntheses differ between the systems, size has been harmonised based on the gasifier capacity in order to make a fair comparison.

A comparison between estimated investment costs for the SNG, methanol and FT fuel production systems to other cost estimates for similar systems found in the scientific literature and reports is presented in Holmgren et al. (2015) and the data for the compared cases are presented in the current report.

Results and analysis for the investment cost estimates and for the comparison of estimated investment costs to other studies are given in Holmgren et al. (2015) which was presented at the 18th Conference on Process Integration, Modelling and Optimisation for Energy Saving and Pollution Reduction in Kuching, Malaysia, 22-27 August 2015.

Chapter 2 presents the process flow diagrams and the energy balances of the gasification-based systems. Chapter 3 includes detailed data for the process equipment and cost estimates for the individual equipment pieces. In addition to the cost estimates for the stand-alone biofuel systems Table 2 also includes cost estimates for integrated systems. Integrated systems are analysed in Holmgren et al. (submitted b).

2 Description of systems – processes and energy balances

The three analysed biomass gasification systems are illustrated by simplified process schemes in Figure 1. In the SNG case (A) an atmospheric indirect gasifier and adiabatic methanation are employed and the final pressure for the SNG to the grid is 30 bar. More details for the SNG system is found in Heyne et al. (2010). The process scheme for the methanol case (B, blue) is based on Isaksson et al. (2012). The scaled system with determined energy balances is also presented in Holmgren et al. (2014a), with the only difference that the system in the present study has an adjusted heat demand for the air dryer. The liquid phase methanol synthesis does not require CO_2 separation. However, not separating the CO_2 means increased power consumption in the recycle due to the larger compression work for the larger gas volume. A methanol production case with MEA CO_2 separation (same technology as in the SNG case) was therefore also included. For the FT-production system the process scheme and modelling is based on Isaksson et al. (2012) (B) and for the syncrude upgrading part on Johansson et al. (2013) (C). Also in the FT case the heat demand of the air dryer was updated to be the same as in the methanol and SNG cases. Although not depicted in the figure, unreacted gas from the FT synthesis is partly recirculated back to the reformer and unreacted syngas from the methanol synthesis is recycled to the inlet of the reactor at a recycle rate of 2.

All three systems have the same amount of input of biomass (430 MWth LHV @50 % moisture content (MC)) and include onsite process integrated biomass drying. The air dryer used in all three systems of this study has a heat demand of 2600 MJ kg^{-1} evaporated H2O. In the case of the indirect gasifier, (SNG case), the biomass is dried to a MC of 20 % before entering the gasifier and in the methanol and Fischer-Tropsch cases where a pressurised direct oxygen blown circulating fluidised bed gasifier is used the biomass holds an MC of 15 % when entering the gasifier, in all cases this correspond to a gasifier capacity of ~480 MWth LHV of dried biomass.
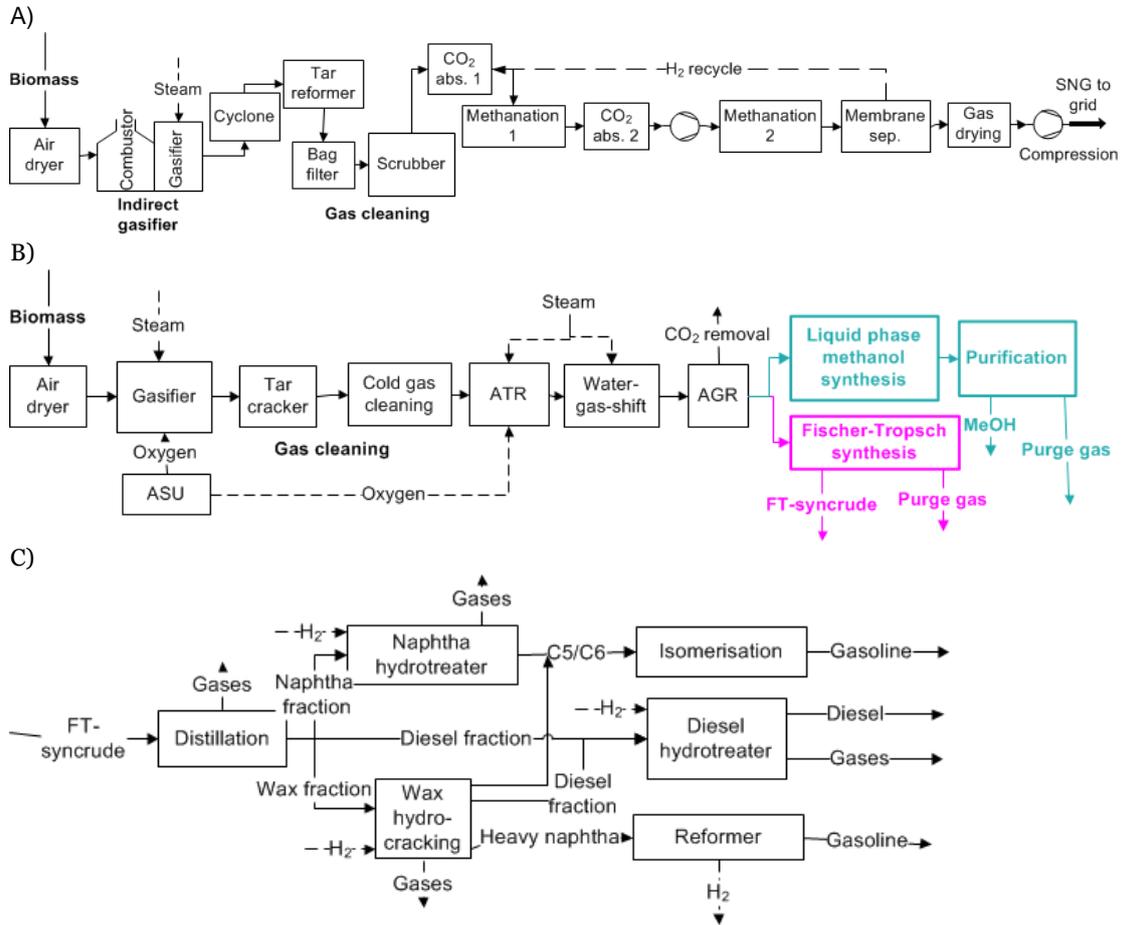
Figure 1   Process flow diagrams of biomass gasification based biofuel production systems; A) SNG production, B) Methanol (blue) and Fischer-Tropsch syncrude (pink) production and C) Upgrading of FT-syncrude.

The energy balance for the upgrading of the FT syncrude to pump fuels was determined by Johansson et al. (2014) (process layout C in Figure 1). Stream data from Aspen simulations for the systems in A and B were used in order to construct grand composite curves (GCCs) for the systems. Split GCCs were used for determining the potential power production in a heat recovery steam cycle (HRSC) utilising the excess heat from the gasification process as conducted in Holmgren et al. (2014b) and Holmgren et al. (2014a) for the SNG and methanol systems respectively.

Table 1   Fuel inputs and product outputs from the gasification based biofuel production systems

	Capacity, dry biomass input to gasifier MW_{LHV} (MC %)	Fuel output MW_{LHV}	Process power demand MW_{el}	Power production HRSC MW_{el}	Net power production MW_{el}
SNG	476 MW[a] (20 %)	299	30.7	42.6	11.9
Methanol without CO_2 separation[b]	485 MW[c] (15 %)	218	39.5	41.1	1.6
Methanol with CO_2 separation[d]	485 MW (15 %)	218	38.6	32.4	-6.2
Fischer-Tropsch incl. upgrading[e]	485 MW (15 %)	172 diesel 33 gasoline	42.6	40.0	-2.6
Fischer-Tropsch excl. upgrading[f]	485 MW (15 %)	172 diesel 33 gasoline	42.6	40.0	-2.6

[a] The energy balance for the SNG case is based on the Aspen simulations by Heyne and Harvey (2014) with adjustments made for the adiabatic methanation instead of the isothermal methanation based on data from Heyne et al. (2010) and changed heating demand for the biomass air dryer to 2600 kJ kg^{-1}H$_2$O evaporated. The biomass feedstock is forest residues with a fuel composition according to Heyne and Harvey (2014). This results in the scaled capacity of 476 MWth $_{LHV}$ of input biomass to the gasifier at 20 % MC.

[b] The energy balance is based on results from Isaksson et al. (2012) with adjustments for the heat demand for the biomass air dryer (same as in SNG case) and updated values for the power demand based on personal communication with Isaksson (2014) and additions to account for the power demand of the air dryer and other auxiliaries.

[c] The biomass fuel composition for the methanol and FT cases is the based on Isaksson et al. (2012) which is slightly different from the one used in the SNG case. Since the fuel is dried to 15 % MC before entering the gasifier the input biomass (LHV) to the gasifier is also slightly different, 485 MW$_{LHV}$ at 15 % MC. The biomass composition in this case is based on Isaksson et al. (2012).

[d] Since the recycle process of unconverted syngas requires compression energy it might be worthwhile to remove CO_2 prior to the methanol synthesis although it is said to not be necessary. In this case the heat demand for a MEA CO_2 separation (with the same assumptions for temperatures and heat demands as Heyne and Harvey (2014)) was added. In addition, it is assumed that the process power demand was reduced by 0.9 MW$_{el}$ for synthesis recirculation.

[e] The energy balance is based on the Aspen simulations by Johansson et al. (2014). In the case including upgrading the fuel yield is based on values given by Johansson et al. (2014). The power demand also includes the estimated power demand for upgrading. The power demand was assumed to be the same for new equipment as for upgrading in existing refinery equipment.

[f] The case excluding upgrading excludes the investment cost for upgrading equipment. The upgrading is assumed to take place in equipment of an existing refinery, as explained by Johansson et al. (2014), hence, power demand for the upgrading is included.

3   Assumptions and methodology for cost estimates

The total plant investment, TPI, is estimated by the following steps:

1. Divide plant into major process areas and sub processes and identify size
2. Identify estimate of base cost for the equipment by: estimate from literature or by consultation with experts
3. Adjust the equipment costs to the size of the design by scaling
4. Adjust the costs to the common time basis using a cost index (i.e. CEPCI) and appropriate exchange rates.
5. Estimate total fixed capital costs using factors (generally necessary to add BOP and indirect costs, (installation costs are generally included in base cost).
6. Add costs for interest during construction

The division of the plant into major plant areas and sub-processes was based on the process flow diagrams (Figure 1). Which size or capacity that is needed for the cost estimates depends on the base cost estimate, but was in most cases be determined from the energy balance, the stream data or other. Process stream data for all three systems are available in the Holmgren et al. (submitted b).

The equipment base costs were estimated by data from literature or by consulting experts. Data for this step is presented in Table 2. In general cost estimates, C_0 for a given component size (capacity) S_0 are given in literature. Scaling is done according to the standard correlation Equation (1).

$$C = C_0 * \left(\frac{S}{S_0}\right)^f \tag{1}$$

Where f is the scaling factor typically ranging between 0.6 and 0.8. Sometimes an upper limit for the size is given and multiple trains (equipment units) could be necessary. In cases where multiple trains are needed equal size of the components are applied and the installed cost of each additional train is estimated to be somewhat less than the cost of the first train. The trained cost, C_m, of a unit is determined by Equation (2) (Liu et al., 2011)

$$C_m = C * n^m \tag{2}$$

where m is the scaling exponent for multiple trains, with a value of 0.9.

In general the scaled component cost estimate needs to be complemented by adding balance of plant costs (BOP) and indirect costs (IC). The BOP is assumed to account for direct costs and includes equipment erection, piping, instrumentation and controls, electrical, utilities, off-sites, buildings (including services) and site preparation. Indirect costs include engineering, head office, start-up and contingencies. In most cases the required addition for BOP and indirect costs are given along with the C_0 cost estimates. All values and comparisons in this study are based on costs and performances of Nth plants[1].

To be comparable, cost estimates need to have a common cost basis, i.e. currency and year. In this study, conversion between currencies is based on annual average values published by the Swedish central bank, Sveriges Riksbank (2014) whereas the updating between years is done by using the Chemical Engineering Plant cost index (CEPCI) (Vatavuk, 2002) according to Equation(3):

$$Cost_{\$Year\,Y} = Cost_{\$Year\,X} \times \frac{CEPCI_{Year\,Y}}{CEPCI_{Year\,X}} \tag{3}$$

The same updating principle could be applied with another of the several indices available but in this study the CEPCI has primarily been used, with a few comparisons to the CERA DCCI index.

The sum of C_m adjusted for BOP and IC gives the TPC. In this study the definition of total plant investment TPI, was based on Kreutz et al. (2008), see Equation (4). The interest during construction, IDC, was set to 5 % of total plant cost, TPC based on Hannula & Kurkela (2013).

$$TPI = TPC * (1 + IDC) \tag{4}$$

3.1 Uncertainty of investment cost estimates

Depending on the purpose and the planning stage there are different are different types of investment cost estimates. The data requirement is quite different between the different estimates types as is described by Peters et al. (2003). This study includes a study estimate (factored estimate) and the accuracy of the resulting estimates is within ±30%.

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[1] An Nth plant is a plant built when the technology is tested in full scale and implemented in several previously commercial scale plants. A first-of-its-kind-plant will have higher investment costs than an Nth plant.

3.2 Equipment cost estimates and data

Equipment cost estimates for the three different gasification systems are given in Table 3. Equation 5 is used for determining the BOP (balance of plant) to be added in some of the cases.

$$BOP(\%) = \frac{0.8867}{\{(\,biomass\;MW_{HHV})^{0.2096}\}}\pi r^2 \qquad (5)$$

The BOP is based on a fit from several studies as stated by Liu et al. (2011).

Table 2   Reference equipment capacities, costs and scaling exponents

	Base cost (original unit)	Base Scale	Scale unit	Scale factor	Scale in present study	BOP (%)	Indirect Cost (%)	Ref.	Note
Biomass pretreatment and drying									
Biomass receive and unload	3.505 M$_{2007}$	198.1	wet biomass input t h^{-1}	0.62	196.7	15.5	32	Black (2013)	a
Biomass storage, preparation and feed into gasifier of 30 bar	2.94 M$_{1999}$	64.6	biomass feed, wet t h^{-1}	0.77	115.7	Eq. 5	32	Liu et al. (2011)	b
Biomass storage, preparation, feeding to atmospheric gasifier	2.028 M$_{1999}$	64.6	biomass feed, wet t h^{-1}	0.77	100.6	Eq. 5	32	Liu et al. (2011)	c
Biomass air dryer	0.72 MSEK$_{2003}$	1.08* 10^6	m^{3}h^{-1}	Eq. 6	5.02*10^6	200	110	Heyne & Harvey, (2014)	d
Gasification island, pressurized CFB gasifier									
GTI gasifier, gas cooling and gas cleaning excl. scrubber	173 M$_{2007}$	483	MW$_{LHV}$ as received biomass	0.5	485	Included	Included	Liu et al. (2011)	e
GTI gasifier, gas cooling and gas cleaning incl. scrubber	208 M$_{2007}$	483	MW$_{LHV}$ as received biomass	0.5	485	Included	Included	Liu et al. (2011)	f
Scrubber	0.418 M€$_{2010}$	200	m^3 h^{-1}	0.6	12101	15	32	Trippe et al. (2011)	g
ASU (including O$_2$ and N$_2$ compression)	82.7 M$_{2007}$	2,202	pure O$_2$ tonne day^{-1}	0.5	1,024 (For FT and MeOH cases)	15.5	27	Liu et al. (2011)	h

	Base cost (original unit)	Base Scale	Scale unit	Scale factor	Scale in present study	BOP (%)	Indirect Cost (%)	Ref.	Note
Gasification island, indirect gasifier									
Indirect gasifier + combustor	8 M€$_{2003}$	8	$MW_{th,LHV}$ (at MC in to gasifier)	0.72	476	200	110	Heyne and Harvey, (2014)	i
Indirect gasifier, slightly pressurized	9.7 M\$$_{2007}$	500	metric t day^{-1}	0.6	2,720	Inst. factor 2.31 + land cost	0.6	Dutta et al. (2011)	j
Indirect gasifier+ combustor	6.7 M€2007	20	MW_{th} input	0.72	476	200	110	Tock et al. (2010)	k
Tar reformer	93.7 M\$$_{2007}$	31,000	syngas flow at exit kmol h^{-1}	0.9	6,180 (SNG)	Included	Included	Liu et al. (2011)	l
Fabric filter	68.8 k\$$_{2002}$	15.6	m^3 s^{-1}	-	74 (SNG)	240	140	Heyne and Harvey (2014)	
Water scrubber	3 M\$$_{2002}$	12.1	m^3 s^{-1}	0.7	43.0 (SNG)	23	50	Hamelinck et al. (2004)	m
Guard bed (Zink bed and guard bed)	2,680 k\$$_{2007}$	2,020	$MW_{HHV,SNG}$	0.67	331 (SNG)	Based on estimates by Arvidsson et al. (2014)		Arvidsson et al., (2014)	n
Water gas shift	3.36 M\$$_{2007}$	815	MW_{LHV} (dried biomass)	0.67	485 (FT, MeOH)	15	32	Liu et al. (2011)	o
CO$_2$ removal									
MEA (in SNG and MeOH cases) (without CO$_2$ storage)	5.19 M€$_{2010}$	.5 kg CO$_{2, sep.}$ S^{-1} 46.7 vol.- % CO$_2$; 3.9 Nm^3 $_{raw\ gas}$ s^{-1}			16.8 kg CO$_{2,sep.}$ s^{-1} 46.7 vol.- % CO$_2$; 18.6 Nm^3 $_{raw\ gas}$ s^{-1}	Based on Heyne and Harvey (2014)		Heyne and Harvey (2014)	p
Gas drying (SNG case)	20 k\$$_{2000}$	1180	m^3 h^{-1}	0.7	34200	Based on Heyne and Harvey (2014)		Heyne and Harvey (2014)	q
H$_2$ membrane (in SNG case)	500 \$$_{2006}$	420	Cost per unit of mounted skid	1	2010	Based on Heyne and Harvey (2014)		Heyne and Harvey (2014)	r
Rectisol (in FT case)	28.8 M\$$_{2007}$	200,000	Syngas feed Nm^3 h^{-1}	0.63	139,200 (FT) (MeOH)	Eq.5	32	Liu et al. (2011)	s
Rectisol compressor (in FT case)	6.31 M\$$_{2007}$	10	Compressor power MW_{el}	0.67	8.1 (FT)	Included	32	Liu et al. (2011)	t

	Base cost (original unit)	Base Scale	Scale unit	Scale factor	Scale in present study	BOP (%)	Indirect Cost (%)	Ref.	Note
HX in Rectisol	52 M$_{2007}$	355	Boiler duty MW$_{th}$	1	67.3 (FT)	Eq. 5	27	Liu et al. (2011)	u
Co$_2$ compression for storage									
CO$_2$ compression for storage	6.31 M$_{2007}$	10	Compressor power MW$_{el}$	0.67	Different for the different cases	Included	32	Liu et al. (2011)	v
Syntheses									
Methanation									
Adiabatic methanation reactor	4381 k€$_{2012}$	175	MW$_{HHV,SNG}$	See note	331	Based on the estimates by Arvidsson et al. (2014)		Arvidsson et al. (2014)	w
Catalyst (initial fill)	845 k€$_{2012}$	175	MW$_{HHV,SNG}$	See note	331	Based on the estimate by Arvidsson et al. (2014)		Arvidsson et al. (2014)	w
FT-synthesis									
FT slurry reactor	246.31 M$_{2011}$	2,420	Amount of produced gasoline and diesel MW	0.75	192	15.50	Included	Goellner et al. (2013)	x
Initial fill of FT catalyst	7.89 M$_{2011}$	2,420	Amount of produced gasoline and diesel, MW	1	192	15.50	Included	Goellner et al. (2013)	y
FT slurry reactor and HXs	13.6 M$_{2007}$	71,360	FT input vol. flow rate, m^{3}h^{-1}	0.75	283,200	Eq. 5	32	Liu et al. (2011)	z
Syngas compressor	6.31 M$_{2007}$	10	Compressor power MW$_{el}$	0.67	70	Included	32	Liu et al. (2011)	
LPMeOH synthesis									
MeOH Synthesis LPMeOH	3.5 M$_{2001}$	87.5	ton MeOH h^{-1}	0.72	39.1	Installation factor 2.1		Hamelinck and Faaij (2002)	aa
Recycle compressor	12.9 M€$_{2002}$	13.2	MW$_{el}$	0.85	4.3	Installation factor 2.1		Hamelinck et al. (2004)	aa
Syngas compressor (before MeOH synthesis)	12.9 M€$_{2002}$	13.2	MW$_{el}$	0.85	10.1	Installation factor 1.86		Hamelinck et al. (2004)	aa

	Base cost (original unit)	Base Scale	Scale unit	Scale factor	Scale in present study	BOP (%)	Indirect Cost (%)	Ref.	Note
MeOH synthesis LPMeOH	81.77 M$2002	10.81	kmol s^{-1} (syngas feed)	0.7	3.96	15	Included	Larson and Tingjin (2003)	aa
Methanol product separation/purification	1.72 M$2002	4.66	kg s^{-1} methanol produced	0.291	10.85	15	Included	Larson and Tingjin (2003)	aa
Final upgrading									
SNG									
Flash tank	13 k€2010	8.85	m3	0.67	42.14	Same factors as used by Heyne and Harvey (2014)		Heyne and Harvey (2014)	
TSA-drying	74 k$2004	21	kg$_{H2O\ adsorbed}$ h^{-1}	0.67	39.15	Same factors as used by Arvidsson et al. (2014)		Arvidsson et al. (2014)	
Drying adsorbent	47 k$2004	21	kg$_{H2O\ adsorbed}$ h^{-1}	0.67	39.15	Same factors as used by Arvidsson et al. (2014)		Arvidsson et al. (2014)	
Final compression	6.31 M$2007	10	MW$_{el}$	0.67	0.65	Included	32	Liu et al. (2011)	
FT upgrading									
HC recovery plant (distillation)	74.15 M$1993	3,190	MW diesel and gasoline	0.65	192	Included	Included	Bechtel (1998a)	bb
Wax hydrocracker	9.3 M$2007	4.1	ton hr^{-1}	0.55	6.44	Eq. 5	32	Liu et al. (2011)	cc
Diesel (distillate) hydrotreater	2.49 M$2007	1.30	ton hr^{-1}	0.6	9.17	Eq. 5	32	Liu et al. (2011)	cc
Naphtha hydrotreater	0.75 M$2007	0.93	ton hr^{-1}	0.65	2.41	Eq. 5	32	Liu et al. (2011)	cc
Naphtha reformer	5.18 M$2007	1.542213	ton hr^{-1}	0.6	2.15	Eq. 5	32	Liu et al. (2011)	cc
C5/C6 isomerization	0.956 M$2007	0.54431	ton hr^{-1}	0.62	0.86	Eq. 5	32	Liu et al. (2011))	cc
MeOH upgrading									
Methanol refining	15.1 M$2001	87.5	ton MeOH h^{-1}	0.7	39.1	Installation factor 2.1		Hamelinck and Faaij (2002)	dd
Heat recovery and integration									
Boiler/steam generator/ ductwork/stack	52 M$2007	355	Boiler duty MW$_{th}$	1	Varying depending on case	Eq.5	27	Liu et al. (2011)	ee

	Base cost (original unit)	Base Scale	Scale unit	Scale factor	Scale in present study	BOP (%)	Indirect Cost (%)	Ref.	Note
Steam cycle (steam turbine, condenser, piping, auxiliaries)	66.7 M$_{2007}$	275	Gross power production MW$_{el}$	1	varying depending on case	15.5	Included	Liu et al. (2011)	ff
Integration cases									
Integration to refinery via molten salt heat transfer			Boiler duty MWth	1	Varying depending on case	18	30	Johansson et al. (2013)	gg
Heat delivery to district heating	See hh								hh
Integration with chemical plant									
Hot water circuits for heat transfer to biomass drying. HXs	32.8 k$_{2000}$	80	m^2 (HX area)	0.68	Specific for each heat exchanger	Factors from Smith (2005) as applied by Hackl and Harvey (2013)		Hackl and Harvey (2013)	ii
Hot water circuits for heat transfer to biomass drying. Piping	Equation (7)	Equation (7)	Pipe diameter DN in mm		Specific for each pipe	Included	Included	Hackl and Harvey (2013)	jj
Piping costs for redistribution of steam	See Hackl & Harvey (2013)		Nominal pipe diameter cm		Specific for each pipe	Included	Included	Hackl and Harvey (2013)	kk
Costs for new HX for increasing LP steam demand at chemical plant	See jj								
Piping costs for redistribution of fuel from cluster	20 MSEK$_{2010}$	2,000	m		1	4,000	Included	Included	Hackl and Harvey (2013)	ll

a) The biomass receive and unload base costs are taken from Black (2013) who gives values for bituminous coal. Based on the assumption in Liu et al. (2011) that the equipment cost scale linearly with bulk density, and that the bulk density for wood chips is 250 kg/m³ according to the BEC, 2014 and that the bulk density of bituminous coal typically is 833 kg/m³, the equipment value should be raised by a factor of 833/250 = 3.33. Further, the assumption by Liu et al. (2011) that the BOP for the values given by NETL (Black, 2010, Black, 2013) are 15.5 % and the indirect costs are 32 % is used. Also Heyne and Harvey (2014) use the cost estimate from Black (2010) but a different method to take indirect and direct costs into account. Applying the same method as Heyne and Harvey (2014) still results in similar TPC values.

b) This is the pretreatment of the feed before feeding into a gasifier operating at 30 bar. The description of the cost estimate is given by Liu et al. (2011) who find the base cost values in Weyerhaeuser (2000) including: conveyor 851 k$_{1999}$, dried wood chip storage 561 k$_{1999}$, feed bin 233 k$_{1999}$, rotary air lock 329 k$_{1999}$ and water cooled feed screw 54 k$_{1999}$. All values were raised by a factor of 1.45 to take into account the higher pressure in the current study. The BOP is estimated by Eq(5). The same data and assumptions can be found in Kreutz et al. (2008).

c) This is based on the same references and includes the same equipment as in b, but no scaling factor since both current and reference system operates at atmospheric conditions.

d) The estimate is based on Heyne and Harvey (2014) who used an original cost given by Johansson et al. (2004) where the following investment cost equation for air dryers is given: $Investment\ cost\ [MSEK] = 0.2 * \frac{(air\ flow\ (hot))^{0.8}}{1000}$, (Eq.6) where the air flow is given in $m^3\ h^{-1}$ and the investment cost includes design, site preparation, bottom slab etc. and conveyors to and from the dryer. Heyne and Harvey (2014) use the method of Smith (2005) for including direct and indirect costs. Note that the scale in present study is divided upon 5 units in the cost calculations.

e) The cost estimates are based on the data and assumptions given in Liu et al. (2011). The gas cleaning units included are; a cyclone and a filter but no a scrubber.

f) The same sources as in (e) are used. The same gas cleaning units as in (e) are included and in addition also a scrubber is included.

g) The cost estimate is based on the original estimate given in Trippe et al. (2011) and the scale for the present study is based on the estimated flows by Johansson et al. (2014)

h) The cost estimate is based on Liu et al. (2011) who state the value 105 M\$$_{2007}$ for the capacity 2202 tonne O_2 day^{-1} including indirect costs. The indirect costs were separated from the base cost in order to obtain a consistent scaling. This estimate was compared to the cost estimates given by Tijmensen et al. (2002) (also used by Hamelinck et al. (2004)) 27.9 M€$_{2002}$ for a stand-alone ASU producing 576 tonne O2 per day, with a scaling factor of 0,75 and an installation factor of 1.3 and 18.1 M€$_{2002}$ for an O_2 compressor using 13.2 MW$_e$ and a scaling factor of 0.85 and an installation factor of 1.86, which resulted in a TPC differing less than 1 % compared to using the estimate from Liu et al. (2011).

i) The estimate is taken from Heyne and Harvey (2014) who based the estimate on data from Güssing gasifier. The value includes the cost for an ejector and a cyclone and the feeding system of biomass into the gasifier. The addition of direct and indirect costs is done according to the method used by Heyne and Harvey (2014) which is based on Smith (2005). Material, pressure and temperature factors used were 1, 1, and 3.1 respectively. Three gasifiers (trains) of equal size were assumed to be needed.

j) This estimate is taken from Dutta et al. (2011) who state that the following equipment is included in the estimate: biomass feeding (excluding storage, pre-treatment, and transport equipment; gasification process reactors (gasification, gas conditioning and combustion); tar reformer; all cyclones and interconnecting ducting; combustion air blower; combustion air pre-heat exchangers; start up equipment; syngas polishing scrubber (final clean up and particulate removal); as discharge system; emergency flare system; process instrumentation; olivine and magnesium oxide makeup system; inert gas system.

k) This estimate was calculated with the same factoring method as used by Heyne and Harvey (2014) with the material, pressure and temperature factors being 1, 1, and 3.1 respectively. Also in this case three gasifier trains were used.

l) This estimate from Liu et al. (2011) was compared to the estimate by Heyne and Harvey (2014) and total plant cost (TPC) differed by less than 5 %.

m) The original cost estimate is taken from Hamelinck et al. (2004) and the data on scaling parameter (physical flow) was estimated by scaling from Johansson et al. (2013). Also Heyne and Harvey (2014) used the cost estimate from Hamelinck et al. (2004) as a basis, but with other factors for adding indirect and direct costs (based on Smith (2005)), which resulted in a very high cost for this equipment which in this study was considered to be inappropriate.

n) The ZnO guard bed is included to remove traces of sulphur. This estimate is taken from Arvidsson et al. (2014) who state that the original source is Brasington et al. (2011). The estimate includes costs for the bed and for initial fill of ZnO absorbent. The factors used for material, pressure and temperature adjustments are: 1, 1.43, 1. Arvidsson et al. (2014) use the method by Smith (2005) to add direct and indirect costs.

o)   This is described as an upstream single stage water gas shift. Note that no water gas shift is included in the SNG case since it was not included in the process scheme by Heyne et al. (2010) and motivated by the fact that steam is added and there will be a simultaneous water gas shift reaction in the methanation reactor. However, Arvidsson et al. (2014) who also uses an adiabatic methanation included a water gas shift reactor.

p)   This includes absorber, stripper, condenser, re-boiler, heat exchangers, and pumps and is entirely based on the work by Heyne and Harvey (2014). Note that MEA CO_2 separation has been added in one of the methanol production cases to the original process layout by Isaksson et al. (2012) since it still might be relevant to remove CO_2 in order to keep compression demand for the recycled gas down.

q)   The gas drying is based on glycol dehydrators. Excessive scaling is applied but is still considered to be valid according to Heyne and Harvey (2014).

r)   According to Heyne and Harvey (2014), the cost estimate is in the higher range of values found in the literature and is originally based on Baker and Lokhandwala (2008). The membrane area determined by the calculation model is increased by a safety factor of 10 % for the cost calculation.

s)   The cost estimate is based on Liu et al. (2011) and the physical flow is based on Johansson et al. (2013). The same base cost estimate was used by Hannula and Kurkela (2013), but with slightly different additions for installation, indirect costs and balance of plant.

t)   The cost estimate is based on Liu et al. (2011) and the estimate for the physical scaling parameter (MW_{el}) is based on Johansson et al. (2013). Note that for all three cases (SNG, methanol and FT-fuels) the same cost estimate for syngas-, SNG- and for CO_2-compressors was used.

u)   The cost estimate is based on Liu et al. (2011) and the physical scaling parameter is based on Johansson et al. (2013).

v)   The cost estimate is based on Liu et al. (2011) and the physical scaling parameter is based on Johansson et al. (2013).

w)   Based on the TREMP technology by Haldor-Topsoe. The base cost estimate for the methanation equipment is the equipment cost considering the factors for material, pressure and temperature (1, 1.4, and 2.54). The methanation equipment is assumed to include reactors, recirculation compressor, hex, and knock-out drum. To the base cost estimates of the methanation equipment and the initial fill of catalyst the factors for direct and indirect costs used by Arvidsson et al. (2014) (which are based on Smith (2005)) are added. According to Arvidsson et al. (2014) the original source for the cost estimates is Brasington et al. (2011).

x)   This cost estimate is included as a comparison and is valid for a LTFT slurry phase reactor using Co catalyst. This estimate renders a very similar value to the estimate below from Liu et al. (2011). It is only included here to show that there are more estimates that result in similar values.

y)   The cost for initial filling of catalyst into the FT reactor is also taken from Goellner et al. (2013) (where also the annual need for refilling catalyst is given). According to Goellner et al. (2013) catalyst losses have been important issue for the project economics and success of for instance the Oryx natural gas based FT fuel production plant. The Oryx plant and the study by Goellner et al. (2013) use the slurry phase reactor.

z)   The technology used in the process setup of this study is a slurry phase reactor with a Co catalyst. The Co-catalyst is more expensive than the alternative Fe-catalyst, but according to Kumar et al. (2009) it has longer life time and higher selectivity and is therefore the preferred choice. According to Dry (2004) the Co-catalyst is used in the slurry phase reactor of the Shell Bintulu plant. At the GTL plant in Qatar, the Oryx uses the Sasol slurry phase reactor over a cobalt catalyst since 2006 similar system is in place at the Escravos plant in Nigeria.
     The cost estimate is based on Liu et al. (2011) and includes heat exchangers needed to keep the reactor at appropriate temperatures.

aa) According to Hamelinck and Faaij (2002) the cost for the liquid phase methanol synthesis (LPMeOH) is only half of the investment cost for the conventional gas phase synthesis. Note that this estimate is for the methanol synthesis reactor only. The estimates for methanol synthesis and methanol purification given by Larson and Tingjin (2003) result in a very similar cost estimate for our case (with capacity of 3.96 kmol/s), if we assume that the estimate by Larson and Tingjin (2003) include the syngas and recycle compressors. In the cost summary we only use the estimates from Larson and Tingjin (2003).

bb) The HC recovery plant is described by Johansson et al. (2014) and includes a distillation plant. In this case the original cost estimate given by Bechtel (1998b) and scaling factors etc. by Bechtel (1998a). These are also the sources that Liu et al. (2011) used. The reason for using the original sources was that the mass flow used for scaling gave results that deviated significantly when compared to the experience and costs of actual refinery equipment (based on information from Karlsson (2014-12-12)). Scaling based on the total amount of diesel and gasoline produced is a less reliable source but we get values more in line with the costs estimated by the refinery expert.

cc) Except for the HC recovery plant the rest of the FT upgrading equipment is based on the cost estimates given by Liu et al. (2011) and the process scheme and physical flows given by Johansson et al. (2013)

dd) The methanol refining includes separation and refining and the methanol has a purity of 99.7 % after the refining.

ee) Boiler/steam generator cost is based on data from Simbeck (2004), (which is also used by Liu et al. (2011)). The cost is valid to the heat transfer duty of the boiler or HRSG and to heat exchange that occurs immediately upstream and/or downstream of the AGR system. The cost for all other heat exchange equipment in our process is included in the cost for the associated main process equipment. The assumed cost scale factor of unity is probably conservative.

ff) The steam turbine and condenser costs are based on estimates found in Liu et al. (2011) who state that they are originally from Woods et al. (2007).

gg) Integration with refinery. High temperature excess heat from the gasification process is transferred to the refinery via a molten salt system. The cost estimate is based on (http://www.ulrichvasudesign.com/cgi-bin/cgiwrap.cgi/econ/econnew.pl) which was used also by Johansson et al. (2013).

hh) The cost estimate for delivering district heating included heat exchangers and piping. The DH system is assumed to have a supply temperature of 90°C and a return temperature of 45°C. The piping costs are based on the same assumption as the hot water circuit used in the integrated methanol case as described by Hackl and Harvey (2013). The estimated pipe diameter was DN300. The estimated length of the pipes was 300 m, i.e. it was assumed that the district heating net and the gasification plant are located close to each other.

ii) The heat exchanger area was estimated for each individual heat exchanger by Hackl and Harvey (2013) according to the method suggested by . The HX investment cost was then estimated base on the values given here. The maximum HX area for one HX was set to 4000 m². The CEPCI-index for heat exchangers was used to update costs from historic values. The additions for direct and indirect costs followed the method by Smith (2005). The cost factors used can be found in Hackl and Harvey (2013). Note that factors for working capital and off-sites were omitted for this case.

jj) Eq. (7) = $-0.0112x^2 + 28.22x + 2707.4$ SEK$_{2007}$/m, and X is district heating pipe diameter DN in mm. This cost function is based on Lindeberg (2007) and valid for district heating pipes in an urban environment. The CEPCI for pipes was used for updating to 2012 values.

kk) See Hackl and Harvey (2013) for a detailed description of the cost estimate methodology for the steam redistribution.

ll) The cost estimate for the fuel redistribution is based on the cost of an existing fuel redistribution pipe at the site. More details are given in Hackl and Harvey (2013).

3.3 Comparison of investment cost estimates

The investment cost estimates for the gasification based biofuel systems achieved from the equipment estimates given in Table 2were compared to estimates of similar systems found in the literature, both scientific and technical reports. In Table 3 data for the investment cost comparison is provided. The included systems are all based on biomass gasification and most systems are based on woody biomass input (with some exception, see comments to Table 3). Several

of the included studies showed cost estimates for many cases (different technical layout, different size etc.), but cases as close in technology and size as possible to the current study were chosen for the comparison. Details are given in Table 3. The cost estimates from the compared systems were scaled based on Equations (1)-(3) where gasifier capacity was considered to be the scaling factor. The scaling factor (f) was set to 0.7, see Equation (1). The gasifier capacity was provided in different units in the studies but was recalculated so as to be given in LHV for the biomass entering the gasifier at the given MC.

Table 3   Data for investment cost comparison

Reference	Case	Values from original study (i.e. unscaled)				Scaled values (on basis gasifier capacity, biomass input LHV)				
		Gasifier capacity MW_{th} LHV input	Investment cost	Cost basis currency, year	Fuel production MW_{fuel} (LHV)	Investment cost, CEPCI, 480 MWth M€$_{2012}$	Investment cost, CERA DCCI, 480 MW$_{th}$ M€$_{2012}$	Fuel production MW_{fuel} (LHV)	Fuel and net power production MWf$_{uel}$+ MW$_{el}$	Investment cost/ output M€$_{2012}$ MW^{-1}
Holmgren et al. (2015) low gasifier estimate[a]	SNG	476	353	M€2012	299	353		299	311	1.1
Holmgren et al. (2015) high gasifier estimate[b]	SNG	476	513	M€2012	299	513		299	311	1.7
Heyne and Harvey (2014)[c]	SNG	100	205.5	M€2010	63.3	670		299	319	2.1
Bio2gas, Möller et al. (2013b)[d]	SNG	376	450	M€2012	200	531		253.2	260	2.0
Norrtorp study, Möller et al. (2013a)[e]	SNG	350	776.1	M€2012	226	963		307.36	250	3.8
Gassner and Marechal (2012)[f]	SNG	100	175.5	M$2006	70.5	476	568.8	335.58	354	1.3
Tunå and Hulteberg (2014)[g]	SNG	400	580	M$2011	266	509		317	333	1.5
Holmgren et al. (2015) no CO_2 separation	MeOH	485	435.2	M€2012	218	435		218	219.6	2.0
Holmgren et al. (2015) MEA CO_2 separation	MeOH	485	502	M€2012	218	502		218	219.6	2.3
Hannula and Kurkela (2013)[h]	MeOH	335	347	M€2010	183	492		265	268.7	1.8
Hamelinck and Faaij (2002), Hamelinck and	MeOH	394	252.4	M$2001	99.3	336.8	622	122.2	179.8	1.9

Reference	Case	Values from original study (i.e. unscaled)				Scaled values (on basis gasifier capacity, biomass input LHV)				
		Gasifier capacity MW_{th} LHV input	Investment cost	Cost basis currency, year	Fuel production MW_{fuel} (LHV)	Investment cost, CEPCI, 480 MWth $M€_{2012}$	Investment cost, CERA DCCI, 480 MWth $M€_{2012}$	Fuel production MW_{fuel} (LHV)	Fuel and net power production MWf_{uel}+ MW_{el}	Investment cost/ output $M€_{2012}$ MW^{-1}
Faaij (2001)[i]										
Norrtorp study, Möller et al. (2013a)[e]	MeOH	350	802.47	M€2012	220	1008		305	246.7	4.1
Tunå and Hulteberg (2014)[g]	MeOH	400	580	M$2011	235	516		285	293	1.8
Holmgren et al. (2015) incl. upgrading[j]	FT	485	515	M€2012	191	515		191	189	2.7
Johansson et al. (2013)[k]	FT	563	540	M€2010	223	532		192	189	2.8
Haarlemmer et al. (2012)[l]	FT	400	946	M€2011	197	1168.4		238.9	231.6	5.0
Liu et al. (2011)[m]	FT	600	724	M$2007	286	540	529	231	265.1	2.0
Hamelinck et al. (2004), Hamelinck et al. (2003)[n]	FT	421	314.6	M€2002	172.7	376	661	199	171.1	2.2
Hannula and Kurkela (2013)[o]	FT	335	370	M€2010	157	525		227.3	248.0	2.1
Tijmensen et al. (2002)[p]	FT	380	387	M$2000	169	530	987	215.7	235	2.3
Swanson et al. (2010)[q]	FT	402	498.3	M$2007	150	492	482	181	200	2.5
van Vliet et al. (2009)[r]	FT	400	390	M$2005	190	434	573	230	263	1.7
Tunå and Hulteberg (2014)[g]	FT	400	760	M$2011	182	676		221	250	2.7

[a] This case is based on the average of two low estimates for the indirect gasifier based on Dutta et al. (2011) and Tock et al. (2010).

[b] This case is based on the cost estimate of the indirect gasifier from Heyne and Harvey (2014).

[c] This estimate is for a system including an indirect gasifier and an isothermal methanation and MEA CO_2 separation. Biomass input to gasifier holds a moisture content of 20 %

[d] The BiO2Gas is a project for a SNG plant in southern Sweden planned by E.ON. The system includes a direct oxygen blown gasifier and the adiabatic TREMP methanation. The biomass requirement is 340 MWth LHV for moisture of 50 % but the biomass is dried to 20 % before entering the gasifier. The values used in this study for comparison is the ones presented as valid for an Nth installation (not first of its kind).

[e] The Norrtorp study is an initial study estimating the cost for potential biomass gasification based fuel production systems at a specific location in Sweden. The study can be considered as somewhat different from the other studies since the systems are not optimized and do not include a heat recovery steam cycle for optimal recovery of excess heat. The gasification technology for both the methanol and SNG case in the Norrtorp study is a high temperature Winkler gasifier based on modified circulating fluidized bed. In the SNG case TREMP methanation is used. In the SNG case compression for utilization in vehicles is included (as compared to the case for the current study where only compression for net distribution is included). Vehicle use requires higher pressures. The methanol synthesis is a gas phase synthesis and CO_2 is removed by Rectisol. In both the SNG and the methanol case the biomass is dried to 10-15 % moisture content and is pelletized The oxygen demand was found to be significant.

[f] This case includes an indirect gasifier. Gassner and Marechal (2012) also included a case with a CFB gasifier but that case was not included in the comparison here.

[g] Tunå & Hulteberg (2014) present investment costs for several biofuel production systems; SNG, methanol; FT; DME and ethanol at a size of 400 MWLHV biomass input. It is unclear from the study at what moisture content but we assumed that it was at dried biomass for gasifier input. In the SNG case an indirect gasifier is used (MILENA-type) whereas in methanol and FT cases a Andritz Carbona type is used (most likely a CFB gasifier). The Andritz Carbona is pressurized and we assumed it is oxygen fed. In the SNG case the TREMP synthesis is applied, whereas the synthesis technology for the cost estimates in the methanol and FT cases are not specified (information on how they have modelled these syntheses is given). Further it is not clearly stated whether FT syncrude upgrading equipment is included but we have assumed it is.

[h] Hannula and Kurkela (2013) included a circulating fluidized bed gasifier, Rectisol CO_2 removal and a gas phase methanol synthesis.

[i] Hamelinck and Faaij (2001) state an initial moisture content of the biomass is 30 % and it is dried to 15 % before entering the gasifier. The gasifier in this case is an IGT pressurized bubbling fluidized bed gasifier and a liquid phase methanol synthesis. It includes a liquid phase methanol synthesis and no CO_2 separation.

[j] The FT case includes slurry phase reactor with a Co catalyst. CO_2 is removed by Rectisol.

[k] The technology choice is based on the process flow diagram by Johansson et al. (2014). The estimated cost for the biomass handling system by Johansson et al. (2013) is very high since it is based on chopped switch grass which has a significantly higher bulk density than wood chips (which is the assumed biomass input by Johansson et al. 2013 and the current study).

[l] The values included are for the case based on an entrained flow gasifier with recycle from Haarlemmer et al. (2012). This study includes FT syncrude upgrading equipment. The biomass is pre-treated at the plant, torrefied and grinded in order to be fed to the gasifier.

[m] In this case chopped switch grass is used in a gasifier at 15 % MC. Fuel upgrading equipment costs are included. The FT synthesis is a slurry phase reactor. It is a setup with a pressurized oxygen blown CFB gasifier and a recycle. Liu et al. (2011) have given the capacity in HHV of dry biomass.

[n] The scenario included from Hamelinck et al. (2004) has a CFB gasifier and slurry phase FT reactor. CO_2 separation by a Selexol unit was included.

[o] Hannula and Kurkela (2013) has a pressurized fluidized bed gasifier, with a Rectisol Co_2 removal and low temperature FT synthesis based on a Co-catalyst. It is a multi-tubular reactor operating at 30 bar and 200°C.

[p] Tijmensen et al. (2002) include several concepts but the one included here is based on an IGT gasifier (CFB) direct oxygen blown with recycle. The FT reactor is a fixed bed reactor.

[q] Swanson et al. 2010 includes gasification of corn stover. Swanson et al. include several cases but the one included here is a low temperature (referring to gasification temperature) case including a fluidized bed gasifier. In this study partial upgrading of the FT syncrude is included. They use a

monoethanolamine-based acid gas removal technology for removing sulphur and CO_2 before the FT synthesis. The FT synthesis has a Co-catalyst and is a fixed bed reactor.

[r] From the study by van Vliet et al. (2009) the PTL 400 scenario with wood pellets as input and a size similar to the current study. The gasifier is of Carbo-V technology, i.e. a two stage gasifier; this type is under development and has the advantage of fuel flexibility and full carbon conversion but the draw-back of being complex to build. van Vliet et al. (2009) describe the development of the Fischer-Tropsch process as having become significantly cheaper due to recent process development. One of the most important developments is the moving from multi-tubular to slurry phase reactor which has reduced construction costs, pressure drop and catalyst consumption by 75 %., increased conversion and reduced maintenance requirements. However, catalyst poisoning is demanding and puts high demands on the reliability of syngas cleaning.

4   Results and Discussion

The resulting investment costs based on the equipment cost estimates and sizes given in Table 2 for the stand-alone gasification based plants are presented in Holmgren et al. (2015). Also the comparison of the estimated investment costs for the gasification based biofuel production systems from literature (Table 3) is presented in Holmgren et al. (2015). Conclusions and discussion from these assessments are also found in Holmgren et al. (2015).

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