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Advanced Gasification Technologies – Review and Benchmarking

Technical assessment and economic analysis

Task 5 Report

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	Name	Signature	Position	Date
Prepared by:	C. Andrea Jordan	Fichtner	Lead Consultant	31/03/2021
Checked by:	John Weatherby	Fichtner	Managing Director	31/03/2021
Checked by:	Andy Cross	AECOM	Project Manager	01/04/2021

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Any enquiries regarding this publication should be sent to us at:
enquiries@beis.gov.uk

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This report has been prepared by Fichtner Consulting Engineers Limited and AECOM. The main contributors were:

- Andrea Jordan, Lead Consultant, Fichtner
- John Weatherby, Managing Director, Fichtner
- Tim Loh, Project Engineer, Fichtner
- David Menmuir, Principal Consultant AECOM
- Andy Cross, Project Manager, AECOM

Additional support was provided by:

- Stephen Othen, Technical Director, Fichtner
- Stephen Florence, Project Engineer, AECOM
- Ian Gilles, Director – Energy, AECOM

The project was managed by the UK Department for Business, Energy and Industrial Strategy (BEIS). The supervisory team consisted of:

- Yannis Dragotis, Project Manager
- Nick Bevan, Senior Engineering Advisor, Industry & CCUS Innovation Team
- Charlotte Powell, Energy Innovation, Head of Team (Bioenergy)
- Katherine Woods, Energy Innovation Programme Lead, Bioenergy Team
- Harry Michalakakis, Senior Energy Research Advisor

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Management Summary

Globally, advanced gasification technologies (AGTs) are recognised as key technologies for the production of hydrogen (H_2), Fischer-Tropsch (FT) fuels, methanol and methane which can be used to substitute fossil derived fuels without the need for changes to fuel handling infrastructure or end user systems. Furthermore, in the UK, AGTs are considered to be potential low or negative carbon routes for sectors of the economy which are known to be difficult to decarbonise such as transport, heat and chemical manufacturing.

Commercial deployment of AGTs has been limited, and many gasification and pyrolysis projects have failed due to technical and/or commercial reasons, as described in the Task 4 report. Given the potential role of AGTs in the transition to net zero, the UK Government requires an evidence-based assessment of the development status of AGTs to inform the development of policies and innovation spending initiatives.

The Department for Business, Energy & Industrial Strategy (BEIS) has commissioned AECOM and Fichtner Consulting Engineers Limited to assess the status of AGTs as part of the overall assessment of the capability and commercial competitiveness of these technologies to decarbonise the heating, industrial and transportation sectors. The previous reports produced as part of this assessment are listed below:

- Task 2: Current status of advanced gasification technologies;
- Task 3: Assessment methodology and counterfactual benchmarks; and
- Task 4: Opportunities and barriers for AGTs.

This report describes the core process systems required for production of H_2, FT fuels and methane using AGTs. It assesses the likely performance of the selected generic AGT plants and provides estimates of capital and operating costs based on information from technology developers and our experience in related sectors. The performance assumptions and estimated costs have been used to derive levelised costs for the products considered and to provide indicative levels of carbon dioxide (CO_2) emissions from these processes.

AGTs

Currently, each AGT plant configuration under development consists of a feedstock preparation plant, feedstock storage and handling systems, a bubbling fluidised bed gasifier (BFB), syngas clean-up systems, a syngas reformer, syngas upgrading systems to convert the syngas into the desired product and product purification systems.

Feedstocks

The feedstocks considered in the study are municipal solid waste (MSW), or commercial waste with a similar composition, and biomass in the form of woodchip or pellets. Pre-treatment of feedstock using conventional mechanical treatment processes is required to prepare feedstock for fuelling the AGT processes. Twenty scenarios based on five products (H_2, synthetic paraffinic kerosene (SPK), FT diesel, methane and methanol) evaluated at two different plant sizes for each feedstock have been assessed in this report.

In 2018 the UK generated a total of around 26 Mtpa of residual MSW, much of which is already contracted to supply existing energy from waste facilities. Provided that sustainability issues can be managed successfully, imported biomass could be used as a feedstock for large scale AGT plants. However, both waste and biomass are finite resources with competing alternative uses, as described in Task 4. While AGTs may be a valuable technology for reducing CO_2 emissions or removing CO_2 from the atmosphere, it is important that feedstock supply limitations are considered in relation to the overall decarbonisation potential of AGTs.

Techno-economic assessment

Technical models were developed to produce generic mass balances for each of the 20 scenarios. The outputs from the mass balances have been used in a cost model to assess operating costs and production rates. Capital costs have been estimated using information provided by technology suppliers combined with internal data available to Fichtner and AECOM. It must be understood that few AGTs have been constructed to date and with the exception of one plant, all of the demonstrator plants to date are smaller than the sizes studied in this report. Consequently, capital cost estimates are likely to vary significantly and this has been modelled by varying capital costs to indicate the sensitivity of this on the product costs. Operating costs have been built up from a combination of supplier information, data from the mass balance models on consumables and residues and experience from similar waste and biomass processes.

The data gathered has been used in the techno-economic models to predict a levelised cost for each product based on the first of a kind (FOAK) AGT configuration evaluated. The models also assessed the impact of carbon capture on the levelised cost of the products. The costs derived for the configurations with carbon capture and storage do not include a cost for export of the CO_2 collected. The results obtained have been compared with the cost of the counterfactuals produced using established methods of production.

These results are presented in Sections 6 to 9 of this report for each product. In all cases, the levelised costs of products from AGTs are higher than traditional ways of generating these products. For AGT configurations that provide a CO_2 saving relative to the counterfactual product, or can operate as CO_2 negative processes with the addition of carbon capture, usage and storage (CCUS), then the difference in the cost of production will be reduced if the cost of emitting CO_2 to the atmosphere increases.

As outlined in the Task 4 report, one of the main risks associated with the development of AGTs is the ability to achieve reliable long-term operation without incurring additional capital costs and excessive operating costs. Consequently, additional systems such as power generation and oxygen separation which increase plant complexity have not been included in our modelling of FOAK plants. Whilst a target plant availability of 85% from the first year of operation has been assumed in deriving the levelised costs of products, this should not be considered as the expected availability for a FOAK plant for the configurations presented. In our experience FOAK plants can exhibit much lower plant availability than this especially in the early years of operation.

Second Generation AGTs

Due to the historical failures of many gasification processes in the UK, we consider it essential that lessons are learnt from these (see Task 4 report). Any new AGTs should be built in a robust manner to have a chance of achieving good availability and this will be reflected in the capital costs. We also consider the complexity of initial projects should be reduced, even if this means initial levelised costs are higher due to lower yields or increased consumables costs.

Unlike solar and wind, where reduced whole life costs have been achieved due to large scale capital cost reductions from mass production of modules in factories, for AGTs we consider product cost reductions are more likely to be achieved by increasing product yields and reduction in consumables costs, particularly electricity costs, by implementing power generation on site from waste heat.

For example, the yield of the hydrocarbon products (transport fuels, methane or methanol) can be significantly increased by using additional imported hydrogen to provide the optimum carbon to hydrogen ratio for synthesis of the products. This has the additional benefit of using the available carbon in the feedstock to generate product rather than CO_2. However, this relies on the availability of H_2 at a suitable cost and with acceptable associated CO_2 emissions. The linking of low carbon H_2 projects with AGTs where yields could be improved and CO_2 emissions reduced by the addition of H_2, is a projected area of development for second generation AGTs.

The results of this report can be used by BEIS to assess the potential for AGTs to contribute to decarbonisation in the UK. Further evaluation of the potential contribution of AGTs could be carried out by a full lifecycle CO_2 assessment to determine the net fossil origin CO_2 savings relative to other methods of obtaining the products under consideration. For AGT configurations that provide a reduction in CO_2 emissions, this can be used to determine a marginal abatement cost of CO_2 using AGTs. This marginal abatement cost of CO_2 can subsequently be compared with the marginal abatement cost of other methods of decarbonising the target sectors of the economy.

Development pathway

Section 3 of this report sets out a potential pathway to develop AGTs and to demonstrate integration and operation of the required technologies. There are different ways in which AGT technology could be developed in relation to feedstock used, products generated, scale of new plants to be developed and financial support mechanisms provided. Information provided in the reports associated with this study can be used to inform decisions on the most appropriate pathway to consider. Feedstock availability, technology development risks and marginal abatement cost of CO_2 may also influence decisions in relation to which feedstocks and products should be targeted.

The current level of technology development for any specific AGT configuration will be a key factor for consideration in determining the scale of any future plant to be developed and the most appropriate mechanisms for providing financial support. If specific technologies have already demonstrated a reasonable level of reliability and performance at a commercial, or near commercial, scale then the next stage of development could be to demonstrate more reliable operation at a similar scale to existing developments. For this type of development, market incentive mechanisms based on product output could be most appropriate.

A development pathway has been proposed based on demonstrating the production of products from MSW at the small scale (100,000 tpa) evaluated in this report. Whilst MSW is a more technically challenging fuel than biomass, if reliable operation can be achieved it brings potential economic advantages due to feedstock costs. A plant size of 100,000 tpa is large enough to demonstrate the performance of the technology at a commercial scale and to allow scale-up to larger sizes without significant risk. The pathway presented relies on the existence of a base technology that has demonstrated reasonable performance at a suitable scale, and our assessment shows that this will not be the case for all products under consideration in this study.

The government's role in the development pathway described may be based around assessment of projects being developed by third parties and working with these organisations to provide appropriate support if robust justification for the support can be provided.

Key steps in developing the technology are firstly for policy makers to assess where AGTs fit best within the overall net zero strategy and how financial support can be provided to assist the development of the technology over the next 18 months. Following this, a well-structured FOAK project could then be set out leading to a construction phase between 2023 and 2026. The plant can then prove the technology and provide confidence to investors allowing a roll out of commercial scale projects from 2030 onwards.

1 Introduction

1.1 Background

Decarbonisation of the heating, industrial and transportation sectors is essential for the transition to a net zero carbon economy. In the UK, advanced gasification technologies (AGTs) have been identified as key enabling technologies for the conversion of biomass and wastes to hydrogen, Fischer-Tropsch (FT) fuels (sustainable aviation fuel (SAF) and diesel), methane and methanol. These energy products which can be synthesised from the syngas produced during gasification are direct substitutes for fossil derived fuels and do not require infrastructural or equipment changes.

Although the importance of AGTs to achieving carbon neutrality is recognised globally, and many of these technologies have been used for the production of fuels from coal for decades, operation on biomass and waste has been limited. Consequently, many of these technologies have only been operated on biomass and waste as large scale demonstrators, as a result commercial deployment of these technologies is at varying stages of development. Given the potential importance of AGTs in the transition to net zero, the UK Government requires evidence-based assessment of the development status of AGTs to inform the development of policies and innovation spending initiatives to incentivise investment and promote wide scale deployment of these technologies by 2030.

The Department for Business, Energy & Industrial Strategy (BEIS) has commissioned AECOM and Fichtner Consulting Engineers Limited to assess the technical feasibility and commercial competitiveness of these technologies to decarbonise the heating, industrial and transportation sectors.

1.2 Objectives

The primary objective of this report has been to carry out a techno-economic assessment of the production of hydrogen, FT fuels, methane and methanol from biomass and waste using AGTs. The assessment was carried out as follows:

1. technical assessment of the key process component systems for fuel production;
2. development of a technical model for assessment of the key inputs to and outputs from each process stage;
3. evaluation of the product yield and process efficiency on an energy basis for a single pass first-of-kind plant (FOAK) not including process recycles;
4. evaluation of the capex and opex for production of each fuel using an in-house cost model;
5. commenting on key technical considerations and process economics for commercialisation;
6. initial evaluation of the decarbonisation potential of AGTs; and

7. identification of the technical and commercial parameters which will be key to development of second generation plants.

1.3 Nomenclature

Table 1: Table of abbreviations

Abbreviation	Meaning
AGT	advanced gasification technologies
Ar	as received
BFB	bubbling fluidised bed
CCS	carbon capture and storage
COS	carbonyl sulphide
CH_4	methane
CO	carbon monoxide
CO_2	carbon dioxide
odt	oven dry tonnes
EfW	energy from waste
EPC	Engineering, procurement and construction
FOAK	first-of-a-kind
FT	Fischer-Tropsch
GCV	gross calorific value
HCN	hydrogen cyanide
H_2	hydrogen
ktpa	thousand tonnes per annum
LCOH	levelised cost of hydrogen
LCOX	levelised cost of product
LG	landfill gas
Mtpa	million tonnes per annum
MSW	municipal solid waste
NCV	net calorific value
NPV	net present value
Nm^3	normal cubic metre
O_2	oxygen
O&M	operating and maintenance
PSA	pressure swing adsorption
RDF	refuse derived fuel
SAF	sustainable aviation fuel
SMR	steam methane reforming
SN	stoichiometric number
SNG	synthetic natural gas
SPK	synthetic paraffinic kerosene

Abbreviation	Meaning
tpa	tonnes per annum
WGS	water gas shift

2 Conclusions

2.1 Feedstock

Biomass and refuse derived fuel (RDF) from residual municipal solid waste (MSW) are the primary feedstocks being considered for fuelling AGTs. Processing of these feedstocks to comply with several critical parameters prior to delivery to the gasifier is essential. The primary criteria are as follows:

1. particle size;
2. quantity of non-fluidisable material;
3. metal content; and
4. moisture

The primary technologies used for processing of feedstock for fuelling AGTs are:

1. shredding;
2. removal of ferrous and non ferrous metals; and
3. drying.

Generally, for large scale commercial operations in the UK, biomass feedstock is pelletised. The processing of feedstock for AGTs is a proven technology and whilst improvements to the processing technology will inevitably be made by various suppliers, innovation of these technologies is not required for supply of feedstock to commercial plants.

As feedstock processing is essential for reliable AGT operation, the capital and operating costs for these systems have been included in the LCOH for each product.

2.2 Technical solution

AGTs for the conversion of biomass and waste to hydrogen (H_2), FT fuels, methane and methanol are currently in development to validate plant operations and the predicted plant performance. Production of each of these products has been demonstrated using bubbling fluidised bed (BFB) gasifiers at plant capacities up to 100,000 tpa. Globally, commercial deployment of these technologies is yet to occur, and the number of technology developers is limited. The largest demonstrator (115,000 tpa) is in operation in Edmonton, Canada and is fuelled by RDF for the production of methanol. The first commercial scale plant (175,000 tpa) is currently under construction in Nevada, USA and will convert RDF to FT fuels. Commissioning of this plant is expected to begin in Q3, 2021. A second commercial scale plant (165,000 tpa) to produce FT fuels from forestry residues is being constructed in Oregon, USA. Plans for another commercial scale plant with a throughput capacity of 200,000 tpa have been announced for Varennes, Canada. Commissioning of the plant is scheduled for 2023.

Overall, the primary design characteristics of the technologies under development, include the following:

1. BFB gasifiers are the primary type of gasifier under development and the preferred feedstock is RDF (for economic reasons).
2. Steam/O_2 mixtures[1] are used as the gasifying agent to produce a raw syngas with the required H_2:CO ratio for upgrading to the relevant product.
3. Syngas treatment technologies are required for removal of a variety of contaminants including acid gases, alkali metals, ammonia (NH_3), carbonyl sulphide (COS), hydrogen cyanide (HCN), hydrogen sulphide (H_2S), particulates and tars. These are the key contaminants produced during gasification of biomass and waste and if not removed, these can reduce process conversion efficiency and plant availability through deposition on process equipment and poisoning of downstream catalysts.
4. Treated syngas from the syngas clean up system is shifted in a water gas shift (WGS) reactor for production of the required H_2:CO ratio prior to syngas upgrading to the desired product.
5. For the production of H_2, pressure swing adsorption (PSA) is the preferred technology for extraction of H_2 from the syngas.
6. Synthesis of FT fuels, methane and methanol is carried out using technologies which are currently used in the synthesis of the same products from syngas from coal.

Our assessment of these technologies clearly shows that due to the very small plant processing capacity, limited number of hours of continuous operation and the need for removal of new and increased concentrations of various contaminants, the primary areas of development and innovation for AGTs are in:

1. process integration at commercial scale;
2. advancement of syngas treatment which can effectively and efficiently reduce key contaminant concentrations to within the required specification; and
3. evolution of syngas upgrading technologies which can efficiently and reliably operate on syngas from biomass and waste.

Overall, the techno-economic assessment carried out is based on the information supplied by a small number of suppliers who to date have only operated demonstration systems which with the exception of one plant are smaller than the smallest scale plant under evaluation. Moreover, many of these plants have been operated in campaigns or for a limited number of

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[1] Steam/O_2 mixtures are currently preferred for the production of H_2 as operation on steam alone incurs additional capex and opex and increases plant complexity

operating hours and have not been operated continuously over several thousand hours. Consequently, many process control, process integration and operational issues are yet to be encountered and assessed. These will need to be resolved before stable long term operation can be achieved. On this basis, it is evident that these technologies are in the early stages of development and significant development of these technologies is necessary before commercial operation is achievable.

2.3 Hydrogen

1. AGTs for the conversion of biomass and waste to high purity H_2 have been demonstrated using RDF, at a maximum feedstock throughput of 780 kg/h (2 MW_{th}).
2. Operation at higher throughputs is technically feasible as the key component systems (feedstock processing, gasification, syngas treatment and CO shifting) required for the production of syngas with a high H_2 content are the same for the other products under consideration.
3. Modelling of a single pass (once through) process for conversion of biomass and waste shows that H_2 yields of 67 kg/odt of waste and 77 kg/odt of biomass could be obtained using first generation AGTs coupled with PSA.
4. Optimisation of plant performance to include key operations such as process recycles and a high efficiency of heat integration will contribute to an increase in product yields.
5. Preliminary assessment on of the capex and levelised cost of hydrogen LCOH for these plants is outlined in Table 2.

Table 2: Summary of the capex and LCOH of FOAK AGTs

Parameters	**Units**	**Plant capacity**			
		Biomass		**MSW**	
Thermal input	MW_{th}	100	643	36	199
Feedstock throughput	tpa	330,000	1,000,000	100,000	550,000
Hydrogen yield	kg/odt	77	78	67	69
Energy efficiency[2]	%	44	40	38	39
CO_2 produced	tpa CO_2	248,000	1,424,500	76,900	432,000
Plant capex	£	263,615,000	842,787,000	148,715,000	431,173,000
Total project capex	£	304,460,000	982,273,000	170,868,000	498,871,000
LCOH	£/kg	7.99	6.50	7.53	3.52

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[2] Energy efficiency here is defined as the efficiency of conversion of energy in the feedstock to energy in the hydrocarbon product.

6. The gasification, syngas treatment and synthesis systems represent approximately 85-90% of the plant costs.

7. The range in the LCOH is to some extent similar to that for the production of H_2 from electrolysis (median value £6.00/kg) but is significantly higher than the cost of production of H_2 using steam methane reforming (SMR) (£1.20/kg) and SMR(£1.70/kg) with carbon capture and storage (CCS).

8. Due to the large differential in feedstock cost, the waste plants have a significantly lower LCOX than the biomass plants. It should be noted that whilst the table above indicates the LCOH for small scale biomass and waste is similar, the small scale biomass plant is three times the size of the small scale waste plant, so benefits from a significant economy of scale.

9. The impact on net zero by AGTs for the production of H_2 could be significant. The following table shows the distribution of carbon emissions in the output streams for a FOAK plant:

Table 3: Carbon emissions from the production of H_2 using AGTs

Parameters	Units	Plant capacity			
		Biomass		MSW	
Rich CO_2 stream	tpa	248,000	1,425,000	76,900	432,000
CO_2 in flue gas	tpa	55,000	288,000	17,000	87,000
C in product	% wt	1	1	1	1
C in CO_2 stream	% wt	81	82	80	82
C in flue gas	% wt	18	17	18	16
C in solid residue	% wt	0	0	1	1
LCOH CO_2 rich stream captured	£/kg	8.21	6.76	7.79	3.81
LCOH CO_2 in flue gas captured	£/kg	9.09	7.25	9.18	4.55

The CO_2 rich stream can be captured using established technologies, compressed and sent via a high pressure pipeline to storage. The LCOH of this, based on costs to the site boundary only with no storage costs are shown in the table above. The table also shows the estimated LCOH if 90% of the CO_2 in the flue gas is also captured using post combustion capture with an amine scrubber.

2.4 Fischer-Tropsch fuels

1. Currently, the production of synthetic paraffinic kerosene (SPK) and FT diesel from biomass and waste using AGTs has only been demonstrated on RDF at a throughput of 170 kg/h (0.5 MW_{th}).

2. Operation at higher feedstock throughputs is technically feasible as the key component systems (feedstock processing, gasification, syngas treatment and CO shifting) required for the production of the raw syngas required are the same for the other products under evaluation.

3. FT systems for the synthesis of liquid fuels from natural gas are currently in operation at production rates of up to 1,900 m^3/day (FT fuels).

4. Modelling of a single pass AGT system for the production of SPK and FT diesel shows the following:

    a. The yields of SPK ranging from 143 l/odt of waste to 181 l/odt of biomass and be obtained; and

    b. yields of FT diesel ranging from 151 l/odt of waste to 190 l/odt for biomass could be achieved in a FOAK AGT.

5. Optimisation of plant performance to include key operations such as process recycles and a high efficiency of heat integration as would be standard in commercial plants would contribute to an increase in product yields.

6. Preliminary assessment on of the capex and the LCOX for these plants is outlined in Table 4.

Table 4: Summary of the capex and LCOX of FOAK AGTs

Parameters	Units	Plant capacity			
		Biomass		MSW	
Thermal input	MW_{th}	100	643	36	199
Feedstock throughput	tpa	330,000	1,000,000	100,000	550,000
SPK yield	kg/odt	172	181	143	151
Energy efficiency (SPK)	%	44	39	37	38
FT diesel yield	kg/odt	181	190	151	159
Energy efficiency (FT diesel)	%	45	41	38	39
CO_2 produced	tpa CO_2	151,800	854,600	47,000	259,300
Plant capex	£	300,162,000	954,706,000	168,512,000	486,563,000
Total project capex	£	346,055,000	1,109,650,000	193,399,000	561,912,000
LCOX SPK	£/kg	4.97	3.83	5.21	2.33
LCOX FT diesel	£/kg	4.61	3.56	4.82	2.17

7. The gasification, syngas and synthesis system represent approximately 85-90% of the plant costs.

8. The LCOX at the plant capacities under consideration for the production of SPK and FT diesel is more than 10 times the cost of production of aviation fuel (median value £0.49/kg) and diesel (£0.65/kg) from the processing of crude oil.

9. Due to the large differential in feedstock cost, the waste plants have a significantly lower LCOX than the biomass plants. It should be noted that whilst the table above indicates the LCOX for small scale biomass and waste is similar, the small scale biomass plant is three times the size of the small scale waste plant, so benefits from a significant economy of scale.

10. The impact on net zero by AGTs for the production of SPK and FT diesel could be significant. The following table shows the distribution of carbon emissions in the output streams for a FOAK plant:

Table 5: Carbon emissions from the production of SPK

Parameters	Units	Plant capacity			
		Biomass		MSW	
Rich CO_2 stream	tpa	151,800	854,600	47,000	259,300
CO_2 in flue gas	tpa	72,000	394,000	23,000	123,000
C in product	% wt	28	28	27	27
C in CO_2 stream	% wt	49	49	49	49
C in flue gas	% wt	23	23	24	23
C in solid residue	% wt	0	0	1	1
LCOX CO_2 rich stream captured	£/kg	5.05	3.92	5.32	2.44
LCOX CO_2 in flue gas captured	£/kg	5.54	4.19	6.11	2.86

Table 6: Carbon emissions from production of FT diesel

Parameters	Units	Plant capacity			
		Biomass		MSW	
Rich CO_2 stream	tpa	151,700	854,200	47,000	259,100
CO_2 in flue gas	tpa	69,000	377,000	22,000	118,000
C in product	% wt	29	29	28	29
C in CO_2 stream	% wt	49	49	48	49
C in flue gas	% wt	22	22	23	22
C in solid residue	% wt	0	0	1	1
LCOX CO_2 rich stream captured	£/kg	4.69	3.65	4.93	2.27
LCOX CO_2 in flue gas captured	£/kg	5.14	3.90	5.66	2.66

The CO_2 rich streams can be captured using established technologies, compressed and sent via a high pressure pipeline to storage. The LCOX of this, based on costs to the site boundary only with no storage costs are shown in the table above. The table also shows the estimated LCOX if 90% of the CO_2 in the flue gas is also captured using post combustion capture with an amine scrubber.

11. By capturing the CO_2, a biomass or waste AGT should become a significant negative contributor as it will generate little CO_2 and the feedstock removes CO_2 from the atmosphere in producing the feedstock. Clearly biomass will remove much more CO_2 as all the feedstock is biogenic.

12. To increase the impact of decarbonisation, rather than capture the CO_2 generated in the process, it will be possible in future plants to add imported hydrogen to the process to convert more of the carbon in the waste streams to product, significantly increasing product yields. This may be a more logical step than capturing the CO_2 generated, which will then need to be stored.

2.5 Methane

1. The production of methane using AGTs has been demonstrated on both biomass and waste at 6 t/h (32 MWth) and 1 t/h (3.5 MWth) respectively for the production of 11,600 tpa methane from biomass and 1,600 tpa methane from waste.

2. Like the other products from AGTs under evaluation here, the key component systems (feedstock processing, gasification, syngas treatment and CO shifting) required for the production of methane have been demonstrated at higher feedstock throughputs.

3. Methanation plants with operating capacities up to 113,000 tpa of methane from syngas from coal are in operation.

4. Modelling of a single pass system indicates that yields of methane ranging from 168 kg/odt of waste to 188 kg/odt of biomass could be achieved for a first generation AGT.

5. Optimisation of plant performance to include key operations such as process recycles and a high efficiency of heat integration as would be standard in commercial plants would contribute to an increase in product yields.

6. Preliminary assessment of the capex and LCOX for the production of methane from first generation AGTs is outlined in Table 7.

Table 7: Summary of the capex and LCOX of FOAK AGTs for the production of methane

Parameters	Units	Plant capacity			
		Biomass		MSW	
Thermal input	MW_{th}	100	643	36	199
Feedstock throughput	tpa	330,000	1,000,000	100,000	550,000
Methane yield	kg/odt	184	188	168	172
Energy efficiency	%	52	47	47	48

Parameters	Units	Plant capacity			
		Biomass		MSW	
CO_2 produced	tpa CO_2	184,500	1,048,200	57,100	317,700
Plant capex	£	253,185,000	810,826,000	143,056,000	415,366,000
Total project capex	£	292,590,000	945,898,000	164,427,000	480,881,000
LCOX Methane	£/kg	3.26	2.66	2.90	1.35

7. The gasification, syngas and synthesis system represent approximately 85-90% of the plant costs.

8. The LCOX at the plant capacities under consideration is significantly higher than the cost of production from landfill gas (median value £0.65/kg), natural gas (£0.28/kg) and for methane produced from anaerobic digestion(£1.10/kg).

9. Due to the large differential in feedstock cost, the waste plants have a significantly lower LCOX than the biomass plants. It should be noted that whilst the table above indicates that the LCOX for small scale biomass and waste is similar, the small scale biomass plant is three times the size of the small scale waste plant, so benefits from a significant economy of scale.

10. The impact on net zero by AGTs for the production of methane could be significant. The following table shows the distribution of carbon emissions in the output streams for a FOAK plant:

Table 8: Carbon emissions from the production of methane using AGTs

Parameters	Units	Plant capacity			
		Biomass		MSW	
Rich CO_2 stream	tpa	184,500	1,048,200	57,100	317,700
CO_2 in flue gas	tpa	47,000	241,000	14,000	68,000
C in product	% wt	26	26	26	27
C in CO_2 stream	% wt	59	60	58	59
C in flue gas	% wt	15	14	14	13
C in solid residue	% wt	0	0	1	1
LCOX CO_2 rich stream captured	£/kg	3.33	2.74	2.99	1.44
LCOX CO_2 in flue gas captured	£/kg	3.65	2.91	3.45	1.68

The CO_2 rich stream can be captured using established technologies, compressed and sent via a high pressure pipeline to storage. The LCOH of this, based on costs to the site boundary only with no storage costs are shown in the table above. The table also shows the estimated LCOX if 90% of the CO2 in the flue gas is also captured using post combustion capture with an amine scrubber.

11. By capturing the CO_2, a biomass or waste AGT should become a significant negative contributor as it will generate little CO_2 and the feedstock removes CO_2 from the atmosphere in producing the feedstock. Clearly biomass will remove much more CO_2 as all the feedstock is biogenic.
12. To increase the impact of decarbonisation, rather than capture the CO_2 generated in the process, it will be possible in future plants to add imported hydrogen to the process to convert more of the carbon in the waste streams to product, significantly increasing product yields. This may be a more logical step than capturing the CO_2 generated, which will then need to be stored.

2.6 Methanol

1. The production of methanol using AGTs has been demonstrated to date on waste at a maximum throughput of 110,000 tpa (by Enerkem at Edmonton, Canada). This is also the highest feedstock throughput rate for a large scale AGT demonstrator.
2. Modelling of the process indicates that yields of methanol ranging from 611 l/odt of waste to 710 l/odt of biomass could be achieved by a first generation AGT plant.
3. The modelled yields are based on a single pass system and indicate the lower end of the range of product output.
4. Suppliers have indicated that initial operations show that optimisation of plant performance by increasing the number of process recycles and supply of additional H_2 will result in increased product yields.
5. Preliminary assessment of the capex and LCOX for the production of methanol from first generation AGTs is outlined in Table 9.

Table 9: Summary of the capex and LCOX of FOAK AGTs for the production of methanol

Parameters	Units	Plant capacity			
		Biomass		MSW	
Thermal input	MW_{th}	100	643	36	199
Feedstock throughput	tpa	330,000	1,000,000	100,000	550,000
Methanol yield	kg/odt	698	710	611	626
Energy efficiency	%	77	69	66	67
CO_2 produced	tpa CO_2	156,200	889,500	48,300	269,700
Plant capex	£	277,579,000	885,536,000	156,272,000	452,325,000
Total project capex	£	320,353,000	1,030,926,000	179,468,000	522,945,000
LCOX Methane	£/kg	1.15	0.93	1.11	0.53

6. The gasification, syngas and synthesis system represent approximately 85-90% of the plant costs.

7. The LCOX of production of methanol at the plant capacities evaluated using AGTs and at the high end of the range it is approximately three times higher than the current cost of production (median value £0.32/kg) using the established SMR of natural gas technology.

8. Due to the large differential in feedstock cost, the waste plants have a significantly lower LCOX than the biomass plants. It should be noted that whilst the table above indicates that the LCOX for small scale biomass and waste is similar, the small scale biomass plant is three times the size of the small scale waste plant, so benefits from a significant economy of scale.

9. The impact on net zero by AGTs for the production of methanol could be significant. The following table shows the distribution of carbon emissions in the output streams for a FOAK plant:

Table 10: **Carbon emissions from the production of methanol using AGTs**

Parameters	Units	Plant capacity			
		Biomass		MSW	
Rich CO_2 stream	tpa	156,200	889,500	48,300	269,700
CO_2 in flue gas	tpa	29,000	140,000	9,000	42,000
C in product	% wt	40	41	40	41
C in CO_2 stream	% wt	50	51	50	50
C in flue gas	% wt	9	8	9	8
C in solid residue	% wt	0	0	1	1
LCOX CO_2 rich stream captured	£/kg	1.17	0.95	1.14	0.56
LCOX CO_2 in flue gas captured	£/kg	1.24	0.99	1.25	0.61

The CO_2 rich stream can be captured using established technologies, compressed and sent via a high pressure pipeline to storage. The LCOX of this, based on costs to the site boundary only with no storage costs are shown in the table above. The table also shows the estimated LCOX if 90% of the CO_2 in the flue gas is also captured using post combustion capture with an amine scrubber.

10. By capturing the CO_2, a biomass or waste AGT should become a significant negative contributor as it will generate little CO_2 and the feedstock removes CO_2 from the atmosphere in producing the feedstock. Clearly biomass will remove much more CO_2 as all the feedstock is biogenic.

11. To increase the impact of decarbonisation, rather than capture the CO_2 generated in the process, it will be possible in future plants to add imported hydrogen to the process to convert more of the carbon in the waste streams to product, significantly increasing

product yields. This may be a more logical step than capturing the CO_2 generated, which will then need to be stored.

2.7 Second generation AGT plants

1. If improvements can be made to first generation designs by making designs modular and reducing some redundancy, there is the potential for capital cost savings. However, these are likely to be small and may be outweighed by modifications needed to improve performance or reliability.
2. It must be recognised that scale is likely to have a significant impact on capital costs as even for modular plant, building larger plants will usually mean a significant reduction in the capital cost per MW or tonne.
3. Construction of larger plants will reduce overall costs. However, entrained flow gasifiers which are the largest gasification technologies in operation require significant innovation before commercial solutions based on this technology can be deployed.
4. Based on our assessment, evolution of advanced techniques and new process development will be able to increase yields by 10% from basic process improvements to as much as 100%, for example by the addition of H_2.
5. Reduction in the carbon footprint of these products will be achieved by improving yields and generating power and oxygen from waste heat. However, the carbon emissions arising from operation on biomass will need to take into account carbon costs in the growing, treatment and transportation of biomass.

3 Development pathway

It is clear from the estimated costs in this report that generating products from waste or biomass using AGTs will not be competitive with alternative ways of generating these products unless the costs of carbon emissions are increased, or the products subsidised. In addition, the current failure of many UK gasification processes has created uncertainty in the investment markets which would need to be overcome.

However, there is significant potential in using AGTs to decarbonise the more difficult sectors such as heat and transport by generating low or zero carbon products such as hydrogen, SAF, diesel, methane or methanol.

The next step in demonstrating where AGTs fit within the overall energy transition to zero carbon is to use the results from this report to allow a fair comparison with alternative means to decarbonise sectors, taking into consideration the overall cost of each process and by carrying out a full life cycle assessment. Technical solutions need to be compared on an equivalent basis and should consider the whole carbon cycle. For AGTs using biomass and waste, this means that the carbon savings generated by using feedstocks which remove carbon from the environment needs to be taken into account, as well as the carbon emissions created in using electricity, oxygen and other consumables used within the process. In addition, carbon emissions from the products need to be assessed. Competing technologies to decarbonise transport and heat need to be assessed in a similar manner. Only then can the full impact of AGTs on net zero be determined to assess its overall potential and to identify which products should be targeted as part of the overall UK net zero strategy.

In terms of large-scale decarbonisation, there will be insufficient waste in the UK to significantly decarbonise sectors such as the gas network or aviation. Whilst in the longer term, biomass is likely to become the key feedstock, in the short-term, economics will strongly favour waste plants due to the revenue from the waste gate fee compared with paying for biomass. Whilst waste is considered a more difficult fuel, we consider that if plants capable of processing waste are developed, it is relatively straightforward to use similar technology to process biomass. It must be recognised that the reverse is not true, as biomass plants have frequently been constructed in a less robust manner as it is perceived that the fuel is better.

We also consider that whilst large scale plants at the scale proposed in this report will be significantly cheaper in terms of product costs and will be required to decarbonise at the scale proposed, building a plant at the size of the small-scale waste plant proposed will be large enough to demonstrate the process, validate process performance and technical viability, before scaling up to the fully commercial sizes.

We therefore propose the following development pathway for AGTs:

- **2021**
  Assess where AGTs fit within overall net zero strategy by carrying out a full life cycle

CO_2 analysis and comparison with alternative methods to decarbonise at a national level.

- **2021/2022**
  Once the product which fits the UK's overall strategy best is decided, select the best technologies to be demonstrated and sites to develop an AGT project. If it is not possible to assess with any accuracy which product is best, it would be possible to build a gasification plant where various slip-streams of the syngas can be used to generate different products. Selection of the technical solution should also be based on an initial evaluation of the syngas treatment and upgrading technologies for cleaning and conversion of the syngas produced.

  It is likely that the study to identify the preferred demonstration plant would be based on using MSW as feedstock at a nominal size of about 100 ktpa. The initial study would then be used to seek a planning consent and an environmental permit.

  A key step will be to decide how the project is to be financed. This could be supported by government grants, early stage funding to de-risk a potential project or product subsidies (e.g. on a similar basis to Contracts for Difference for low carbon electricity generation).
- **2023-2026**
  Plant construction phase, nominally an initial procurement phase of 8 months with a 40 month construction phase.
- **2027**
  Commissioning and testing assumed to take around 12 months. We would caution that the commissioning period for a FOAK plant could be significantly longer.
- **2028-2030**
  Demonstration of performance and availability. This will include detailed assessment of the performance of the syngas clean up and upgrading systems. This will ensure that contaminants are reduced to within the acceptable working range for optimised operation of the process catalysts and that plant performance predictions on product yield are achieved. It would also be possible to test other syngas conversion technologies or process improvements. However, as demonstrating availability will be important to provide credibility for investors, care should be taken to not over complicate the project leading to numerous outages to build additional phases.
- **2029**
  Start to develop larger commercial projects based on results of first plant, with increased focus on biomass.
- **2030 onwards**
  Develop second generation projects at full scale. This will require a combination of product subsidy and taxes on more polluting ways to generate products, such as carbon taxes.

4 Techno-economic assessment

4.1 Introduction

In order to include benchmarking of the full range of technologies that are considered to have high potential to support the pathway to net zero, a range of generic plant configuration types were considered, rather than a number of specific technology providers that all offer a similar type of plant. The range of configurations assessed were as follows:

Table 11: Options for Advanced Gasification Solutions

Product	Biomass		Waste	
	Small-scale	Large-scale	Small-scale	Large-scale
Hydrogen	P	P	P	P
FT fuels	P	P	P	P
Methane	P	P	P	P
Methanol	P	P	P	P

The options highlighted in Table 11 indicate that there are suppliers who have developed solutions for the particular option to a reasonable stage. Suppliers who have provided detailed information to support this work are: Advanced Biofuel Solutions Limited (ABSL); Enerkem; KEW Technology; LanzaTech; PowerHouse Energy Group; Valmet; and Velocys. These suppliers were evaluated in the Task 2 report and contacted to provide more details of their processes. Much of the detail was provided as “commercial in confidence”, on the basis that it would only be used to develop generic models which could be used to assess performance and costs.

Many of the configurations in the above table are similar. For each configuration we have set out a case, based on a combination of supplier information, actual plant operating data and our own experience, to set out generic process solutions and to estimate costs for each of these. To do this, we have sub-divided each case into four discrete process blocks as shown in Figure 1.

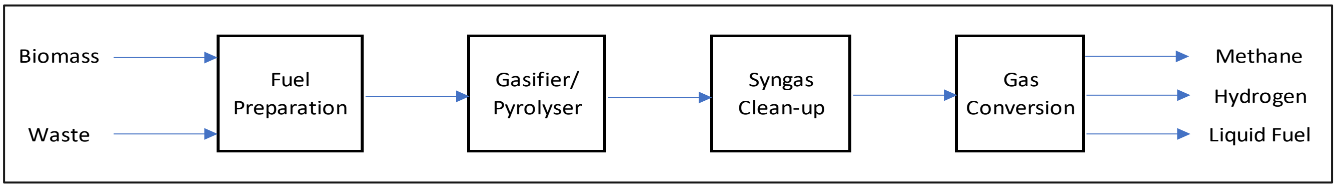

Figure 1: AGT block flow processes

Many of the blocks will be similar whatever the solution considered. For example, if municipal waste is to be converted into a product, it is first processed into RDF and then gasified to make a syngas. These steps will be largely similar whatever the product. The gas clean-up and conversion steps will then be selected to match the required product. Similarly, for a particular product the gas clean-up and conversion steps may be largely the same whether the feedstock is biomass or waste, but the fuel preparation and the gasifier steps may be different.

For each case, a process description is provided which explains the make-up of each of the process blocks. This is based on descriptions and information provided by the main technology suppliers who have made most progress in these areas. Section 5 describes the feedstock preparation which is assumed to be the same independent of the product. Sections 6 to 9 provide process descriptions for each product: H_2; FT fuels, methane and methanol. Whilst the descriptions are intended to be generic, they are based on the specific information provided by suppliers to ensure that there is a sound practical basis for each of the cases. A mass balance has been developed to show the mass flows in and out from the process and to allow consumable requirements, disposal costs and product yields to be estimated.

To allow generic cases to be assessed, these are based on “biomass” using wood chip or pellets and “waste” using MSW. The following typical feedstock properties are used for design purposes.

Table 12: Typical feedstock properties for design basis (as received)

Parameter	Unit	Wood Chip	Wood pellets	MSW
Carbon	% ar	25.50	47.43	26.30
Hydrogen	% ar	3.15	5.86	3.69
Nitrogen	% ar	0.15	0.28	0.77
Oxygen	% ar	20.89	38.85	15.64
Sulphur	% ar	0.01	0.02	0.13
Chlorine	% ar	0.01	0.01	0.96
Ash	% ar	0.30	0.56	17.86
Moisture	% ar	50.00	7.00	34.65
Net Calorific Value (NCV)	MJ/kg	8.14	17.24	9.70

4.2 Plant capacity

Realistic technical and economic assessment of the potential for AGTs to have a significant impact on decarbonisation by 2050 must be based on evaluation of plant with operating capacities which are representative of:

1. the plant capacity under development; and
2. the process capacity required for production of the products under evaluation.

Initial assessment of the process capacity requirements for the products under consideration showed that in 2019, the UK consumed about 12.4 million tonnes (Mt) of aviation fuel[3] and in

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[3] https://www.gov.uk/government/statistics/digest-of-uk-energy-statistics-dukes-2020, dataset tab 3.2

2015, 680 ktpa of hydrogen[4]. The hydrogen was used for existing industrial purposes, not as part of the energy transition away from fossil fuels. The scale of a transition to hydrogen is enormous, with the CCC's analysis of a "Balanced Pathway" predicting low-carbon hydrogen production of 105 TWh (2.7 mtpa)[5] in the UK by 2035 and BEIS predicting production ranging from 250 -460 TWh (6.4 – 11.8 mtpa) by 2050[6].

In 2019, the UK consumed about 878 TWh (GCV basis)[7] of natural gas. Therefore, to have a significant impact on decarbonisation of transport or the gas grid will require very large facilities. For example, to generate 10% of the UK's current hydrogen consumption would require about 1.7 mtpa of MSW or 0.9 mtpa of wood pellets. Moreover, to generate 1% of the aviation fuel the UK used in 2019 would need 2.0 mtpa of MSW or 1.0 mtpa of wood pellets.

Therefore, this report assesses AGT plant sizes which could have a significant impact in the future and for this reason very small plants are not considered. To allow the comparison of plants and their costs, the following sizes have been evaluated:

Small scale MSW

- Small scale MSW, sized at 100,000 tpa of MSW is equivalent to a gasifier thermal capacity of about 36 MW_{th} (NCV basis). Such a plant is at the smaller end of the range of UK energy from waste plants. It is large enough to be seen as a step on the way to proving the technology and developing larger plants. It is also possible that due to local waste arisings, 100,000 tpa is a sensible option for regions with lower waste arisings.

Large scale MSW

- Large scale MSW, sized at about 550,000 tpa of MSW is equivalent to a gasifier thermal capacity of 199 MW_{th}. At the current status of development this would be a multi-stream plant. This is slightly larger than the larger AGT plants currently proposed internationally, but is much smaller than the UK's larger energy from waste plants (the Runcorn plant for example processes almost 1 mtpa of RDF). It should therefore benefit from the economies of scale for a large plant but is still expected to be a manageable size to source waste and for investors, at least once the technology is demonstrated.

Small scale biomass

- Small scale biomass, sized at 330,000 tpa of woodchips is equivalent to a gasifier thermal capacity of 100 MW_{th}. This is a similar size to the larger biomass plants in the UK fired by UK biomass, either wood chip or straw. It is therefore a size which could be

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[4] UK hydrogen production is 26.9 TWh/y - 680ktpa based on GCV of 39.4 MWh/t (2016 study cited by CCC) https://erpuk.org/wp-content/uploads/2016/10/ERP-Hydrogen-report-Oct-2016.pdf, page 8

[5] CCC 6th carbon budget "Balance Pathway" scenario - https://www.theccc.org.uk/wp-content/uploads/2020/12/The-Sixth-Carbon-Budget-The-UKs-path-to-Net-Zero.pdf, pg27 (range of 160-375 TWh/y by 2050, pg 156)

[6] Impact assessment for the sixth carbon budget - https://www.legislation.gov.uk/ukia/2021/18/pdfs/ukia_20210018_en.pdf

[7] https://www.gov.uk/government/statistics/digest-of-uk-energy-statistics-dukes-2020 dataset tab 4.2

scaled up to larger production technically, but should currently be able to source UK biomass if the economics were right.

Large scale biomass

- Large scale biomass, sized at 1m tpa of imported wood pellets is equivalent to a gasifier thermal capacity of 643 MW_{th} which would require multiple streams. This size of plant is included in the report to demonstrate the potential impact and economies of scale. It is larger than could feasibly be expected to attract investment in the current market. It is however smaller than the largest biomass plants in the UK, with the coal conversion plants at Drax and Lynemouth consuming about 5 mtpa of wood pellets annually, and MGT Teesside, which is in construction and is, expected to burn 1.1 mtpa of wood pellets.

For the purposes of the process assessment at the sizes considered, it is assumed that the gasifiers will be bubbling fluidised bed (BFB) systems. Currently, all of the main technology suppliers listed above, have based their systems on the BFB technology. It is possible that suppliers will develop solutions based on larger scale technologies, such as entrained flow gasifiers. There is also more variety available at the very small scale, with suppliers developing small scale pyrolysers as well as other forms of gasifiers. As discussed above, very small systems are not reviewed in this report. The impact of changing the core technology for larger scale plants is discussed in Section 11.

4.3 Mass balances

Generic mass balance models have been derived for each of the fuel production processes considered. As two feedstocks have been assessed producing 5 different products at two different sizes, 20 cases have been modelled. To do this, we have assessed the information provided by suppliers who are developing technologies in this sector together with credible literature sources. As this field is new and there is very little real operating data available from this type of plant, we have generated simple theoretical mass balance models to model the various stages of the process. These models have been adjusted using factors to match the available data from suppliers for syngas composition, yields and consumables. Whilst a detailed mass balance for this type of plant is likely to have around 100 separate streams for each case, as streams are refined and recycled, our models are simpler, focussing on the outputs rather than the internal streams which are required for plant design purposes.

For the fuel preparation, the mass balances are relatively simple and well understood as the feedstock is refined to separate out inert materials and metals and then to dry the feedstock to a level suitable for stable operation of the gasifier. The gasifiers themselves are also relatively similar in concept so the fluidised bed design modelled, has been fluidised by a mixture of steam and oxygen. Once the syngas is produced in the mass balance, the syngas reactions are then considered on a molar basis to allow the various steps of clean-up and conversion to be modelled based on theoretical conversion reactions, adjusted by factors to match the predicted outcomes provided by suppliers. The model refines the product stream to the desired product quality. Waste gas streams from various parts of the process are collected. These are combined to form two streams: a CO_2 rich stream (>90%) which does not require any further treatment prior to compression; and a waste gas stream which consists of all remaining gas streams containing remaining hydrocarbons and trace elements. The model assumes that the waste gas stream is burnt in a thermal oxidiser to meet environmental requirements, assumed to be Industrial Emissions Directive levels. The resulting flue gases are then cooled to recover the energy to be used in the process. In the model it is assume that the flue gases are emitted to atmosphere but in practice the CO_2 in the flue gases could be removed using conventional carbon capture methods.

The model uses different factors according to the size of the plant and the type of feedstock. This allows it to generate results for all 20 scenarios for comparison although the technology suppliers are not considering many of the cases shown at this stage. The results for each model are shown in Sections 6 to 9. The model also assesses the amount of waste products produced such as solid residues together with the amounts of key consumables such as steam and oxygen required.

A full energy balance has not been carried out. For each mass balance the waste energy available from the process was assessed to check whether the overall process is exothermic or whether it will require additional auxiliary fuel to drive the process. The energy which can be recovered in a useful form to generate steam is compared with the amount of energy needed to generate the process steam needed by the process. In all cases considered, there is an excess of heat available. No power generation has been assumed, with all the electricity and oxygen required by the process imported. As discussed in Section 10, as plants become more developed, they will recover more of the available waste heat and may be able to use this to generate electricity to supply at least part of the overall demand or to produce oxygen for the process.

4.4 Capital and operating cost methodology

Capital cost (capex) and operating costs (opex) have been estimated using a combination of supplier information and our independent assessment. It should be recognised that available capital and operating cost data is very limited. There are no similar plants in the UK against which capex and opex can be properly benchmarked. None of the UK's waste or biomass gasification plants established under the Renewables Obligation (RO) scheme make a syngas of good enough quality for conversion to the products. There are a few demonstration plants worldwide (eg GobiGas, Enerkem (at Edmonton), Gussing), but these were built to demonstrate the processes, and cannot be considered to be commercial projects. Accessing accurate and meaningful cost data from these is not easy. In addition, the capital costs for a plant built in the UK under an EPC contract will bear little resemblance to one built in Europe or North America in phases or using a multiple contract based approach with no performance guarantees. There is therefore no available database of costs which can be used to determine the costs for the plants considered.

A number of the technology suppliers have provided cost data for the projects they are developing. This tends to be for specific projects which they have already costed and therefore may not be directly comparable to the cases considered in this report. Providing accurate capital costs for a project requires a high level of project definition, in which each part of the plant must be specified and then costs built up based on the quotations provided. However, as the plant sizes, feedstock and products vary and the plants do not yet exist, it is not possible to get detailed cost data from the market for the cases under consideration.

The approach taken is therefore to take the plant and equipment cost data provided by the suppliers and to normalise the data for scope and size to allow us to assess the cost difference between suppliers and products. We have estimated the fuel preparation capital cost for biomass and RDF using our larger database of such plants and adjusting these for capacity to allow us to estimate the prices for large plants. We have then adjusted the capital costs provided by the suppliers based on thermal capacity of the gasifiers and to ensure that the different costs provided by various suppliers are adjusted for scope as far as possible to make these fair comparisons. All cost data provided has been for the gasification plants at the smaller end of our ranges and we have estimated larger plant costs using sizing factors derived from waste and biomass power projects. This allows us to predict the capital costs for the 20 options being considered. We have also benchmarked the capex against similar sized EfW plants to ensure costs are reasonable.

The capex estimate is for an EPC contract let in 2025 in the North East of the UK. It includes the feedstock preparation plant in each case. None of the capital costs include any power generation, oxygen separation or CO_2 capture and compression equipment[8] as we have assumed this will not be included in the FOAK projects. The cost of these services is included via operating costs.

In all cases we have assumed that the plant will be built under an EPC contract. Once we have estimated the plant costs, project development costs have then been added. There is a much larger database of projects against which to compare this. On the other hand, it must be recognised that project development costs for conventional projects can vary widely depending on planning issues, utility connections, land constraints and the developer’s aims.

The additional project costs reflect the following:

1. land costs;
2. project development costs;
3. project management;
4. planning and permitting;
5. developer’s costs;
6. utility connections;

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[8] The estimated capital costs for CO_2 capture and compression are referenced separately as additional plant costs for each of the products evaluated.

7. construction phase finance;
8. insurance; and
9. contingency.

In our view the likely accuracy of the capital cost estimates provided in the report are -30% to +50%. Whilst this may seem a wide range, cost estimates for current UK EfW plants often vary by 25% when competitive bids for the same project are compared, even though these are based on a fixed scope, a known site and where experience from hundreds of similar projects is available. However, due to the current level of development of the technologies under consideration it is possible that actual plant costs could fall outside of this range.

The approach taken to estimate operating costs is more straightforward:

1. fixed operating costs such as rates, rent and insurance are assessed;
2. staffing costs have been estimated by setting out likely staffing profiles for each case and then applying appropriate salary costs for each grade;
3. consumable costs have been estimated by using the consumption rates calculated in the mass balance model and applying suitable unit costs per tonne for each item;
4. solid and effluent disposal costs are calculated in a similar manner form the predicted material tonnages multiplied by suitable unit costs per tonne;
5. feedstock costs or gate fees are calculated using the tonnages input into the mass balance multiplied by the agreed product rates/gate fees;
6. revenue costs from any streams which can be sold, such as metals, are calculated using the tonnages predicted by the mass balance; and
7. maintenance costs are predicted on an annual average basis using a fixed ratio (2.5%) of the EPC capital cost.

4.5 Carbon capture

The application of CCUS technology to AGTs provides an opportunity for negative CO_2 emissions from a several of the proposed configurations. The requirement for CCUS deployment may be essential for some waste-fired configurations which, due to their efficiency of conversion, may otherwise have higher associated fossil origin CO_2 emissions than the direct use of fossil fuels.

The application of CCUS to the various AGT configurations has been considered to illustrate the contribution of CCUS to the LCOX for all of the end products. Based on the mass balances described in Section 4.3 there are two waste gas streams that contain CO_2 that can be captured and sequestered for all configurations, as follows:

1. a CO_2-rich stream from the WGS reactor within the syngas upgrading stage; and
2. flue gas from combustion of the tail gas in the final syngas upgrading stage and other waste gas streams.

CO_2 capture from these two gas streams has been considered separately, as the relative volumes of CO_2 and cost of capture differs significantly. Given the biogenic carbon content of the feedstocks, it may be considered uneconomic and / or unnecessary to capture the CO_2 from both gas streams for all of the configurations. The LCOX for capture of the rich CO2 stream only and capture of CO_2 from both the rich and flue gas streams have been calculated.

The assessment of the cost of carbon capture for the various AGT configurations is based on the following assumptions:

1. The CO_2-rich stream from the WGS reactor and synthesis system will be separated using a physical solvent process such as Rectisol or Selexol, to provide a pipeline quality CO_2 stream directly from the WGS stage with little additional clean-up. The capex and opex for capture of this CO_2 is therefore based on the cost of compression of this CO_2 stream to the required export pressure. These costs are derived from public-domain costs for CO_2 compressors and internal cost database information, with appropriate scaling factors applied.
2. CO_2 in the flue gas can be captured using a chemical solvent-based process such as monoethanolamine (MEA). Capex and opex for capture of this CO_2 include the key component systems for an amine plant (flue gas pre-treatment, absorber, stripper, solvent reclaiming), as well as drying, conditioning and compression of the captured CO_2 to export pressure are estimated. These costs are derived from various benchmarking studies for such carbon capture technology, including the 2018 Wood study undertaken on behalf of BEIS, and internal cost database information, with appropriate scaling factors applied.

As noted in the Task 3 Methodology report, these costs include plant and equipment up to the battery limit of the AGT plant, i.e. they do not include the cost of transport and storage of the captured CO_2 off-site.

4.6 Techno-Economic model

The mass balance data and capital and operating cost estimates have been used to derive an equivalent Levelised Cost of End Product (LCOX) for each of the end products and plant configurations considered. Generally, the LCOX provides an indication of the unit cost of production over the full life of a project, including capex, opex and financing costs. It is important to note that the LCOX in this report only includes plant related costs and does not take into account any additional investment related to national or regional infrastructure outside the plant boundary that specific technologies may require, for example transport and storage infrastructure for disposal of captured CO_2.

In this assessment, the LCOX is expressed as the cost per kg of end product produced throughout the design life of the plant, based on the whole life capital and operating costs. The LCOX derived can be compared with the cost of the same end products produced by other conventional or low carbon means.

The basis of this calculation and the assumptions used are described in detail in the “Methodology for Techno-Economic Assessment of Advanced Gasification Technologies” report prepared as part of Task 3 of this study. Counterfactual benchmark costs for comparison with the derived LCOX are provided in Appendix A of that report.

Fundamental to the development of the LCOX is the calculation of the Net Present Value (NPV) for the total lifetime costs of developing and operating the plant, and the total quantity of end product produced. These are derived from the following equations:

Equation 1 NPV of Total Costs

Equation 2 Discounted Sum of End Product

Equation 3 Levelised Cost of End Product

For each plant configuration, the NPV, sum of end product and levelised cost have been derived using a Microsoft Excel calculation.

Reference should be made to the Task 3 Methodology report for a full understanding of all of the assumptions used in these calculations and their sources. A summary of the key assumptions is provided in Table 13.

Table 13: Main assumptions used in the LCOX calculation

Parameter	Value
Design life	25 years
Base year for all costs	2025
Discount rate	7.8%
Deflator rate[1]	2%
Plant availability[2]	85%
Feedstock cost	
MSW (gate fee)	(£112/t)
Woodchip	£61/t
Biomass pellets	£152/t

Source: AECOM / Fichtner Task 3 report "Methodology for Techno-Economic Assessment of AGTs"

[1] Deflator rate is the inflation rate used to adjust prices from different sources to a consistent basis.

[2] The assumed availability is a target availability for next generation AGT plants to provide levelised cost figures that are representative of the operation of AGTs as established technologies within a low

carbon energy system. This should not be considered as the initial expected availability for a FOAK plant for the configurations considered in this study.

The total sum of end product produced by the plant is derived from the yield assumed in the mass balance for each configuration as described in the following sections of this report.

5 Feedstock processing

The two types of feedstock considered in this report, biomass and MSW, have to be prepared to meet the requirements of the gasifier suppliers. As all the suppliers who have provided information for this stage of the project are developing projects using bubbling fluidised beds, the fuel requirements are broadly similar. As biomass differs significantly in character from MSW, the fuel preparation systems required are quite different. This section describes these systems.

For a fluidised bed to function well, the fuel must be prepared to adjust its critical parameters as follows:

- Particle size. Fluidised beds function by blowing the fluidising medium (oxygen and steam in this case) through a bed of solid particles. If enough fluidising medium is used, the bed particles fluidise. As this is dependent on the size and density of the bed particles, the size of the feedstock entering the fluidised beds needs to be controlled, normally by screening and shredding.
- Amount of non-fluidisable material. Biomass and waste contain material which does not gasify and which is dense. In biomass this is typically stones and possibly some metal. In waste there is much more of this material, such as stones, bricks, metal objects, glass etc. For the fluidised bed to operate properly, the amount of this material added has to be controlled carefully and the bed designed to allow whatever is fed to the bed to be able to leave it. If this does not happen, the bed will de-fluidise and no longer function. The amount of non-fluidisable or large dense material needs to be controlled, normally by using screens and density separators.
- Metal. Metal will not fluidise easily as it is dense and inert. It acts as non-fluidisable material in a bed and it must be possible to remove metal particles fed into a bed. In addition, material such as wire easily catches on parts of the structure or on other pieces of wire and can build large structures in the bed, often called “birds’ nests” which then leads to defluidisation. As much metal as possible should be removed before the bed, and if this is done it has a value as a recyclable material. Metal is removed either by magnets if ferrous or by eddy current separators if non-ferrous.
- Moisture. If feedstock is too wet, it can cause problems with bed temperature control. More fundamentally too much moisture reduces the efficiency of the gasification process. For AGTs, feedstock moisture is normally limited to no greater than 15% by weight. This is achieved by driers installed in the fuel processing plant.

It is recognised that feedstocks can be procured pre-treated. Many plants processing waste will utilise an RDF produced from MSW and commercial waste by a fuel supplier. However, in doing so the gate fee paid to the AGT operator will reduce compared with the gate fee for MSW. Also, in terms of the overall impact on decarbonisation, the fuel preparation stage needs to be taken into account to try to ensure than when systems are compared, they are compared on as like-for-like basis with as similar envelopes as possible. Therefore, the approach taken in this report has been to consider the residual MSW in its raw form and to include the capital and operating costs as part of the cost of the AGT plant, although in reality some projects may elect to procure pre-treated RDF from suppliers instead.

All feedstock preparation systems are considered to be commercially proven with many facilities in operation. As the feedstock is prepared for a fluidised bed, of which there are many operational examples, we do not consider there are any innovative techniques to be demonstrated because the fluidised bed is part of an AGT plant. Preparation of a good quality fuel is a key factor in achieving good availability for a fluidised bed.

5.1 Biomass preparation

5.1.1 Wood chips

Virgin wood is normally supplied in various forms. Commonly wood used for energy processes is the off-cuts from saw mills or timber production so the material will be supplied in a chipped form. Wood can also be supplied as “round wood” which can be stored outside and then chipped as required. In addition, virgin wood can be supplied in the form of forestry trimming containing small branches and the like. Sawdust can also make up part of the fuel mix. Virgin wood should contain few contaminants, although if not harvested or stored well it can be very wet and may contain soil and stones entrained during the harvesting and transport process.

A typical wood preparation plant will therefore consist of a reception area where delivery vehicles deliver the feedstock, an outside store if round wood is to be used and an internal storage area for chips or sawdust. Round wood would be chipped on-site using a chipper. Material would be moved by a combination of mobile shovels and walking floors so that woodchip is adequately mixed and fed onto conveyors without the need for manual intervention through the night. Magnets would be provided to remove any ferrous metal objects which have become mixed with the feedstock. The fuel would be screened prior to feeding to the drier, with oversize either removed from site or sent to the chipper.

The reception and storage area would be enclosed as would all feed conveyors. Fire protection and detection equipment would be included to the Environmental Permit and insurer’s requirements.

The material would then be dried using a conventional wood chip drier using process steam as the drying medium. Except for start-up, steam would be generated from waste heat from the AGT process. The dried fuel would then be sent to a temporary holding store before being supplied to the gasifier.

5.1.2 Wood pellets

For larger plants wood pellets are the more likely feedstock. Wood pellets would most likely be sourced from abroad and be delivered by sea to dedicated unloading facilities. Depending on the location of the AGT plant, the pellets could be conveyed directly to the plant if it is adjacent to the port, or they would be transported by train to the plant.

For the purposes of this report, it is assumed that pellets are transported by rail to the AGT plant. Automatic rail unloading systems are used to remove the pellets and transport them to large storage silos. Pellets are then extracted from the silos and transported to the gasifier on demand. As wood pellets undergo strict quality control to produce the pellets, which preclude the inclusion of metal objects and which mean the wood is shredded and dried to around 7% in the pellet making process, no on-site separation, drying or treatment is required. Extensive fire detection, protection and suppression systems are provided to ensure the pellets do not overheat in transport or storage.

Examples of such wood pellet storage systems are in operation at Drax and Lynemouth coal conversion plants.

5.2 MSW preparation

The MSW feedstock proposed is a mixture of either municipal waste or commercial waste similar in character. The material is the residual solid waste left over from households or commercial properties after removal of source segregated recyclables.

Material is delivered either by refuse collection vehicles or in bulkers to the MSW preparation facility on site where it is unloaded inside the reception building. The building also provides adequate storage, typically about 4 days full load operation, to cater for delivery interruptions or holidays.

The material is then fed into a material recycling facility to prepare suitable fuel for the gasifier and to remove any materials with value. This facility is likely to consist of:

1. bag splitters to split any refuse sacks and to coarsely shred the waste;
2. magnets to separate metal objects – typically these are located on various conveyors to remove as much metal as possible;
3. eddy current separators to extract non-ferrous metal;
4. density separators to remove coarse, sense objects such as glass, bricks and other inert materials;
5. secondary shredders to shred the remaining material to the size required by the gasifier; and
6. a conventional drier using process steam as the drying medium -except for start-up, steam would be generated from waste heat from the AGT process.

The dried waste material is then transported by conveyor to a buffer store before being sent to the gasifier.

All operations will be carried out inside and the equipment is set up to operate automatically with manual supervision and monitoring. The equipment proposed is all conventional and is in use in numerous facilities. An extensive fire detection, protection and suppression system will be installed to protect the plant from fires.

6 Hydrogen

6.1 Overall technical solution

6.1.1 Gasification

Feedstock processed as described in section 5, is conveyed via lock hoppers to a BFB gasifier where the feedstock is heated to temperatures ranging from 700-800°C using a mixture of superheated steam and oxygen. The oxy-steam mixture acts as both the gasifying agent and the fluidising medium. The syngas produced is then discharged at temperatures of up to 750°C to a multistage syngas clean-up system for removal of a range of contaminants which can deactivate the catalysts used in the syngas upgrading process and or impact on the quality of the H_2 produced. Steam for the gasification process is raised using heat recovered from downstream process reactions.

Residual ash and char from the process are continuously discharged from the base of the gasifier.

To date, the process for the production of hydrogen using AGT has been developed for operation at pressures from -2 mbar(g) to up to 50 mbar(g). Currently, the largest H_2 producing AGT demonstrator plant has been demonstrated on RDF and has a design throughput of approximately 5,800 tpa (2 MW_{th}) which is approximately 6% of the small-scale waste to H_2 plant evaluated in this study.

6.1.2 Syngas treatment

Treatment of syngas prior to upgrading for the production of H_2 is carried out using a range of systems including:

1. plasma conversion;
2. high efficiency cyclones for collection of entrained particulates and droplets of hydrocarbons;
3. wet scrubbing for the removal of entrained particulate matter, tars and NH_3;
4. activated carbon beds for adsorption of metals;
5. hydrolysis reactors for the reduction of COS, HCN and unsaturated hydrocarbons; and
6. amine scrubber for the removal of CO_2 and some acid gases.

As wet scrubbing reduces the syngas temperature, syngas treatment is generally carried out at temperatures ranging from 150 – 250°C and at can occur at pressures up to 50 bar(a) depending on the downstream process requirements. As some of the reactions occurring during gas clean-up are exothermic the heat produced is recovered and used in the overall heat recovery system to generate process steam. Effluent from the clean-up process is treated and some of it is recycled to the plant for syngas scrubbing.

6.1.3 Syngas reforming

Following gas clean up, the treated syngas is then discharged to a multistage CO shift reactor (reformer) to maximise conversion of CO in the syngas to H_2 using the WGSreaction.

Generally, as in other syngas reforming processes the syngas is split into two where one stream is delivered to the reformer and the other bypasses the reformer. The stream entering the reformer is mixed with medium pressure steam to increase the ratio of steam:CO to provide the optimal operating temperature for reaction.

Reforming of the syngas is carried out at temperatures ranging from 200°C (low temperature shift) to 500°C (high temperature shift) and is catalysed by cobalt-molybdenum catalysts in low temperature shift reactors and chromium or copper promoted iron oxide catalysts in high temperature shift reactors.

Prior to extraction of H_2 from the shifted syngas, CO_2 in the mixture of gases is captured using a solvent based process.

6.1.4 Pressure swing adsorption

Syngas from the CO shift reactor is then sent to the PSA unit where approximately 70% of the H_2 in the syngas is recovered from the syngas with a purity of 98% or more as required by BS ISO 14687:2019[9]. Generally, adsorption occurs at pressures ranging from 10 - 40 bar(g) and temperatures from ambient to 60 °C. Desorption of the gas retained by the PSA unit is then carried out and the tail gas stream collected can then be combusted without any support fuel for power generation.

The H_2 produced is stored onsite in storage tanks before delivery to the offtaker.

6.1.5 Process residues handling

The CO_2 captured from the shifted syngas has a high purity (>90%) and can be cleaned and processed through liquefaction to food grade quality, supplied for industrial gases or sequestered.

Process residues from the gasifier and the syngas treatment system are continuously discharged from the plant and are collected for disposal offsite. Effluent from the wet scrubber systems is treated onsite in a wastewater treatment plant.

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[9] BS ISO 14687:2019 - Hydrogen fuel quality — Product specification. Part: All applications except proton exchange membrane (PEM) fuel cell for road vehicles

Any waste gases are collected from various stages of the process and sent to a thermal oxidiser[10] to oxidise all the hydrocarbons. The flue gas is then cooled using the heat to produce process steam and cleaned to ensure any emissions are below limits set by the IED.

6.2 Plant description

AGT plants for the production of hydrogen consist of the following key system components:

1. feedstock storage and delivery system;
2. BFB gasifier;
3. ash and char handling system;
4. multistage syngas clean up system;
5. syngas compression;
6. CO shift reactor;
7. pressure swing adsorption;
8. CO_2 scrubbing and compression;
9. thermal oxidiser for waste gases;
10. heat recovery systems;
11. storage tanks for products, oxygen and nitrogen;
12. auxiliary systems to provide effluent treatment, start-up boiler, cooling, compressed air etc; and
13. stack.

The estimated plant area for the four thermal sizes modelled is shown in Table 14.

Table 14: Projected AGT plant area for modelled thermal capacity

	Plant size			
	MSW		Biomass	
	100,000 tpa	550,000 tpa	330,000 tpa	1,000,000 tpa
Plant area (m^2)	25,000	100,000	70,000	300,000

6.3 Process model assumptions

A process model was derived to estimate the key process inputs to and outputs from oxy-steam gasification of woodchip and RDF for the production of hydrogen. The model was developed

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[10] Complies with the Industrial Equipment Directive (IED) 20/75/EU, 2-second 850°C requirement

based on a detailed evaluation of process data from several technology providers coupled with our experience in the assessment of a large number of gasification systems and technologies. The model is based on a single pass first-of-a-kind plant in which all of the feedstock delivered to the gasifier is used to produce hydrogen and does not consider the following:

1. number of process recycles; and
2. use of any excess energy for power generation.

In addition to the assumptions described in Section 4.3, the key assumptions are as follows:

1. 100% conversion of feedstock to syngas, ash and char;
2. The primary syngas components considered were CO, H_2, CO_2, H_2O, O_2, N_2, CH_4, NH_3, HCl, H_2S;
3. equilibrium gas composition in the gasifier;
4. benzene and naphthalene were used as the model component for tars;
5. the gasifier was assumed to operate at between 700-800° at pressures higher than atmospheric;
6. 100% conversion of CO in a single pass;
7. 70% efficiency of capture of H_2 from the shifted syngas;
8. the estimated NCV of H_2 was 94 MJ/kg (as the hydrogen is 98.5% pure);
9. electricity consumption is based on data provided by suppliers;
10. the main consumables have been derived from the model based on the process requirements; and
11. operation at the maximum feedstock throughput for 85% of the hours in a year (7,446 h)

The results from the model for the two biomass plant sizes considered are outlined in Table 15 and illustrated in Section 6.4.2.

6.4 Biomass to hydrogen

6.4.1 Process model

Table 15: Biomass to H_2 process model

Model Parameter	**Units**	**Model outputs**	
		330,000 tpa	**1,000,000 tpa**
Process streams	---	1	1 [11]
Feedstock throughput	t/h	44	134
Thermal input rate	MW_{th}	100	643
	MWh_{th}/year	744,600	4,788,000
Syngas output from gasifier	t/h	341,300	1,831,000
Syngas output from reformer	t/h	242,100	1,364,000
Hydrogen yield	kg/odt	77	78
Hydrogen production	tpa	12,600	72,600
Electricity consumption	MWe	15	82
	MWh/year	107,100	604,500
Plant availability	%	85	85
Energy efficiency	%	44	40
Main consumables	tpa natural gas	130	836
	tpa oxygen	65,900	372,000
	tpa steam	314,000	817,800
	tpa water	214,100	1,209,000
	tpa chemicals	16,500	93,000
Key material output streams	tpa CO_2 [12]	248,400	1,424,500
	tpa flue gas	492,600	2,805,200
	tpa flue gas CO_2	55,000	288,000
	tpa bottom ash and char	716	5,600
	tpa effluent	231,000	1,302,000

6.4.2 Mass balance

The mass balances derived from the process models are illustrated in Figure 2 and Figure 3 overleaf.

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[11] Currently, the largest BFB gasifier has a thermal capacity of 100 MW. The proposal here of a single stream is based on the potential for development of gasifiers (for conversion of biomass and waste) with a significantly higher thermal capacity as the technology matures.

[12] CO_2 from the CO_2 rich stream extracted from the shifted syngas

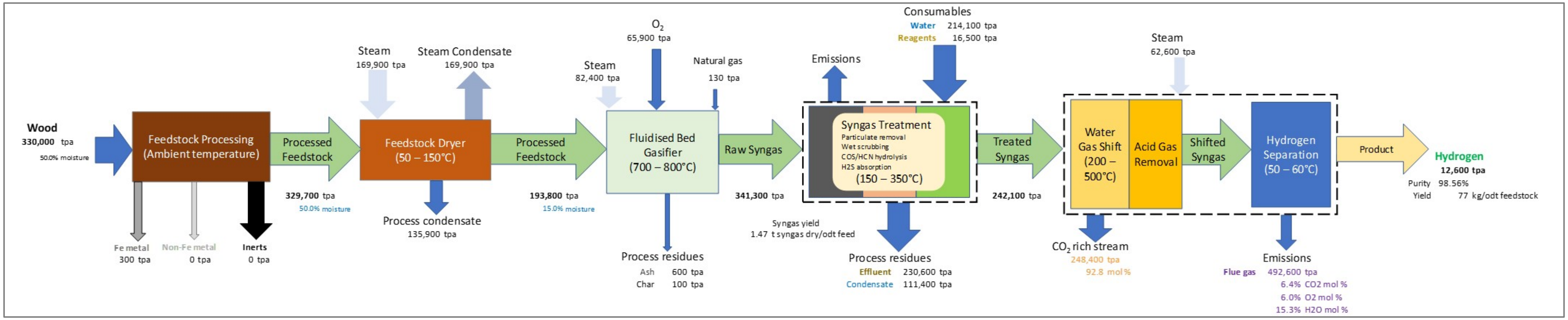

Figure 2: Modelled mass balance for 330,000 tpa AGT biomass to H_2 plant

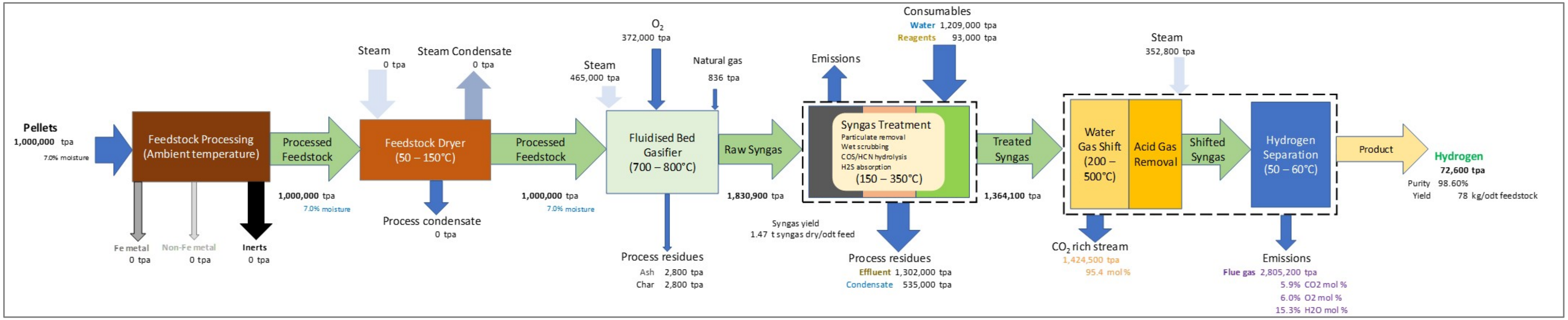

Figure 3: Modelled mass balance for 1,000,000 tpa AGT biomass to H_2 plant

6.4.3 Capital cost estimate

The estimated capital costs for FOAK biomass to H_2 plants are outlined in Table 16.

Table 16: Estimated capital cost of FOAK biomass to H_2 plants

Item	Cost, £	
	330,000 tpa	**1,000,000 tpa**
Project Development Costs		
Land	1,347,500	5,775,000
Consultancy Services	2,636,100	8,427,900
Planning & Other Regulatory	5,422,700	17,622,500
Developers Costs	18,453,000	58,995,100
Start-up Costs	9,226,500	29,497,500
Utility connection	3,759,000	19,168,100
Sub-total development costs	**40,844,800**	**139,486,100**
EPC costs		
Fuel preparation and storage	14,102,900	78,096,100
Gasifier and synthesis plant	249,512,000	764,691,000
EPC cost	**263,614,900**	**842,787,100**
Total Project Costs	**304,460,000**	**982,273,000**

6.4.4 Operating cost estimate

The estimated operating costs for FOAK biomass to H_2 plants are outlined in Table 17.

Table 17: Estimated annual operating costs for FOAK biomass to H_2 plants

Item	Basis	Total Cost, £ pa	
		330,000 tpa	**1,000,000 tpa**
Fixed Costs			
Labour		3,016,100	4,820,200
Administration & Other Overheads		7,611,500	24,556,800
Sub-total Fixed Costs		**10,627,600**	**29,377,000**
Feedstock		**20,130,000**	**151,783,200**
Consumables			
Electricity		13,916,400	78,016,000
Oxygen		6,589,000	37,200,000
Natural gas	GCV	233,800	1,500,900
Water		69,400	348,000
Chemicals		164,700	930,000
Catalysts		303,600	921,500
Sub-total Consumables		**21,276,900**	**118,916,400**
Disposal costs			

Item	Basis	Total Cost, £ pa	
		330,000 tpa	**1,000,000 tpa**
Metal revenue		-11,700	0
Non Fe metal revenue		0	0
Landfill waste	Active waste	0	0
Coarse ash disposal and rejected inerts	Inert	5,800	27,900
Fine ash disposal	Hazardous	23,100	449,600
Effluent		352,800	1,992,100
Sub-total Disposal costs		**370,000**	**2,469,600**
Maintenance	2.5% of capex/year	**7,611,500**	**24,556,800**
Total Opex		**60,016,000**	**327,103,000**

6.4.5 Capital and operating costs estimate with carbon capture

The additional capital and operating costs for carbon capture equipment for FOAK biomass to H_2 plants are listed in Table 18.

Table 18: Estimated capital and operating costs of carbon capture plant for FOAK biomass to H_2 plants

Parameters	330,000 tpa	1,000,000 tpa
Capital costs	**Cost, £**	
CO_2 rich stream capture cost	3,470,000	19,047,000
CO_2 in flue gas capture cost	59,082,000	172,052,000
Total capture costs	62,552,000	191,099,000
Operating costs	**Cost, £ pa**	
CO_2 rich stream capture	2,161,000	15,218,000
CO_2 in flue gas capture	4,060,000	14,454,000

6.5 Waste to hydrogen

6.5.1 Process model

Table 19: Waste to H_2 process model

Model Parameter	Units	Model outputs	
		100,000 tpa	550,000 tpa
Process streams	---	1	2
Feedstock throughput	t/h	13	74
Thermal input rate	MW_{th}	36	199
	MWh_{th}/year	268,100	1,481,800
Syngas output from gasifier	t/h	121,000	665,900
Syngas output from reformer	t/h	75,000	413,700

Model Parameter	Units	Model outputs	
		100,000 tpa	550,000 tpa
Hydrogen yield	kg/odt	67	69
Hydrogen production	tpa	3,900	22,000
Electricity consumption	MWe	5	29
	MWh/year	37,800	208,100
Plant availability	%	85	85
Energy efficiency	%	38	39
Main consumables	tpa natural gas	235	1,295
	tpa oxygen	29,100	160,000
	tpa steam	70,500	387,700
	tpa water	75,700	416,100
	tpa chemicals	5,300	32,000
Key material output streams	tpa CO_2[12]	76,900	432,000
	tpa flue gas	152,500	850,700
	tpa flue gas CO_2	17,000	87,000
	tpa bottom ash and char	10,700	58,900
	tpa effluent	81,500	448,100

6.5.2 Mass balance

The mass balances derived from the process models are illustrated in Figure 4 and Figure 5 on the following pages.

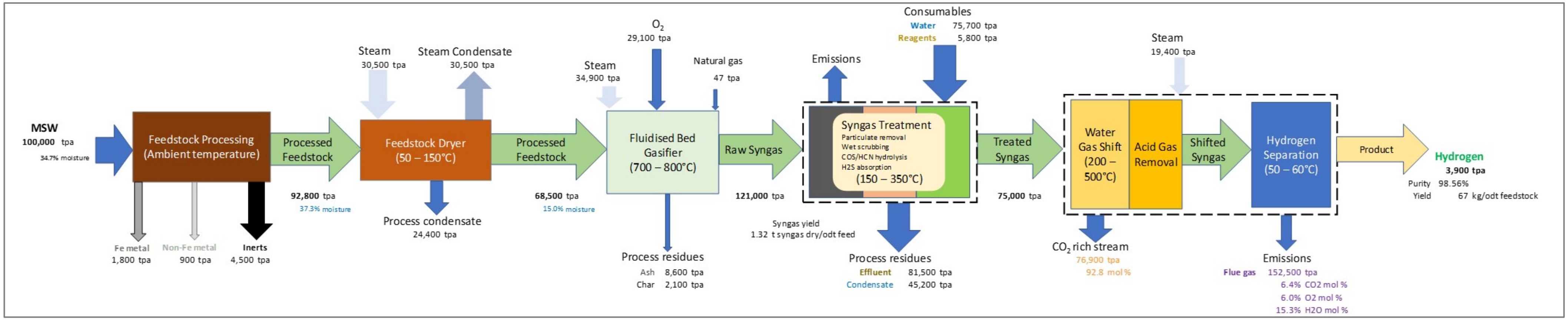

Figure 4: Modelled mass balance for 100,000 tpa AGT MSW to H_2 plant

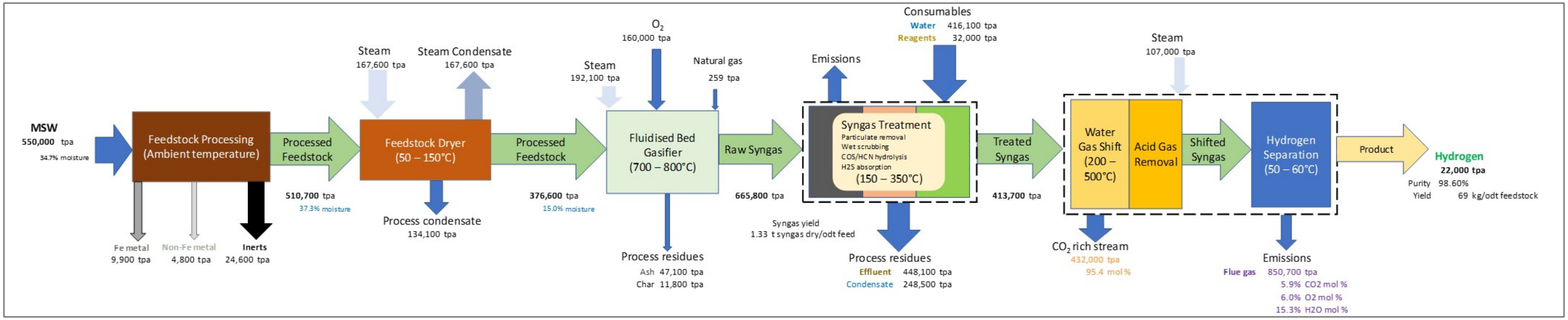

Figure 5: Modelled mass balance for 550,000 tpa AGT MSW to H_2 plant

6.5.3 Capital cost estimate

The estimated capital costs for FOAK MSW to H_2 plants are outlined in Table 20.

Table 20: Estimated capital cost for FOAK MSW to H_2 plants

Item	Cost, £	
	100,000 tpa	550,000 tpa
Project Development Costs		
Land	481,300	1,925,000
Consultancy Services	1,487,200	4,311,700
Planning & Other Regulatory	3,035,600	8,914,400
Developers Costs	10,410,100	30,182,100
Start-up Costs	5,205,000	15,091,000
Utility connection	1,533,500	7,274,100
Sub-total development costs	**22,152,700**	**67,698,300**
EPC costs		
Fuel preparation and storage	13,549,100	53,015,700
Gasifier and synthesis plant	135,165,900	378,157,000
EPC cost	**148,715,000**	**431,172,700**
Total Project Costs	**170,868,000**	**498,871,000**

6.5.4 Operating cost estimate

The estimated operating costs for FOAK MSW to H_2 plants are outlined in Table 21.

Table 21: Estimated annual operating costs for FOAK MSW to H_2 plants

Item	Basis	Total Cost, £ pa	
		100,000 tpa	550,000 tpa
Fixed Costs			
Labour		2,455,600	5,100,900
Administration & Other Overheads		4,271,700	12,471,800
Sub-total Fixed Costs		**6,727,300**	**17,572,700**
Feedstock		**-11,200,000**	**-61,600,000**
Consumables			
Electricity		5,098,200	28,040,100
Oxygen		2,910,000	16,004,900
Natural gas	Gross calorific value (GCV)	84,400	464,400
Water		30,700	126,000
Chemicals		58,200	320,100
Catalysts		162,600	455,300
Sub-total Consumables		**8,344,100**	**45,410,800**

Item	Basis	Total Cost, £ pa	
		100,000 tpa	550,000 tpa
Disposal costs			
Fe- metal revenue		-80,000	-440,000
Non-Fe metal revenue		-135,900	-747,500
Landfill Waste	Active waste	536,500	2,950,600
Coarse ash disposal + Inerts	Inert	85,700	471,200
Fine ash disposal	Hazardous	345,200	1,898,400
Effluent		124,700	685,600
Sub-total Disposal costs		**876,200**	**4,818,300**
Maintenance	2.5% of capex/year	**4,271,700**	**12,471,800**
Total Opex		**9,019,000**	**18,674,000**

6.5.5 Capital and operating costs estimate with carbon capture

The additional capital and operating costs for carbon capture equipment for FOAK biomass to H_2 plants are listed in Table 22.

Table 22: Estimated capital and operating costs for carbon capture plant for FOAK waste to H_2 plants

Parameters	100,000 tpa	550,000 tpa
Capital costs	**Cost, £**	
CO_2 rich stream capture cost	1,545,000	8,699,000
CO_2 in flue gas capture cost	29,236,000	84,090,000
Total capture costs	30,781,000	92,789,000
Operating costs	**Cost, £ pa**	
CO_2 rich stream capture	769,000	4,943,000
CO_2 in flue gas capture	1,880,000	6,161,000

6.6 Levelised cost of production

Comparison of the LCOX for H_2 production using AGTs is illustrated in

Figure 6 for all of the plant sizes evaluated. Overall, the LCOX for hydrogen is significantly higher than for production from SMR and SMR with CCS. However, it is comparable to some extent with the production of H_2 from electrolysis. The counterfactual product costs are discussed in detail in the Task 3 report[13]. Figures 7 and 8 show the impact of including carbon capture.

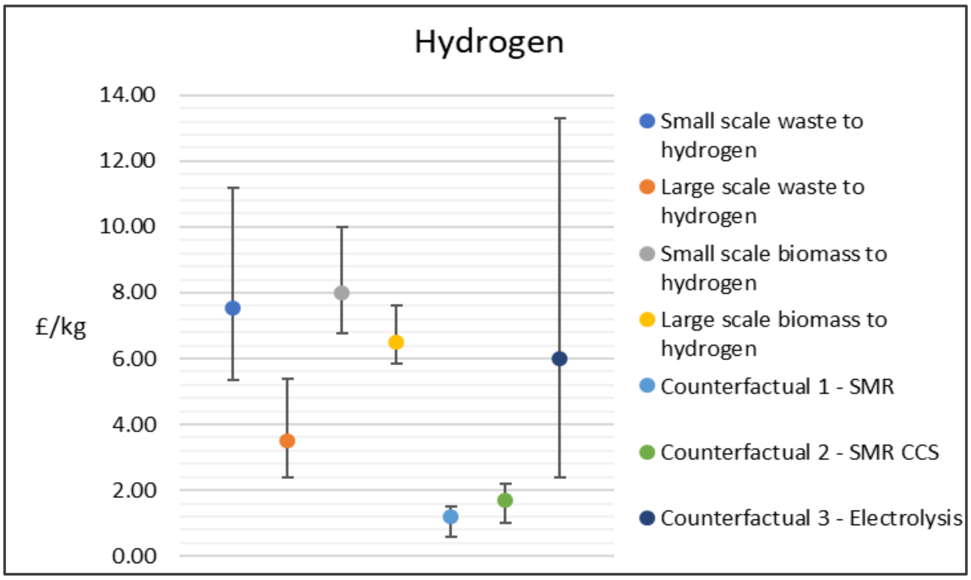

Figure 6: LCOH for biomass and waste using AGTs

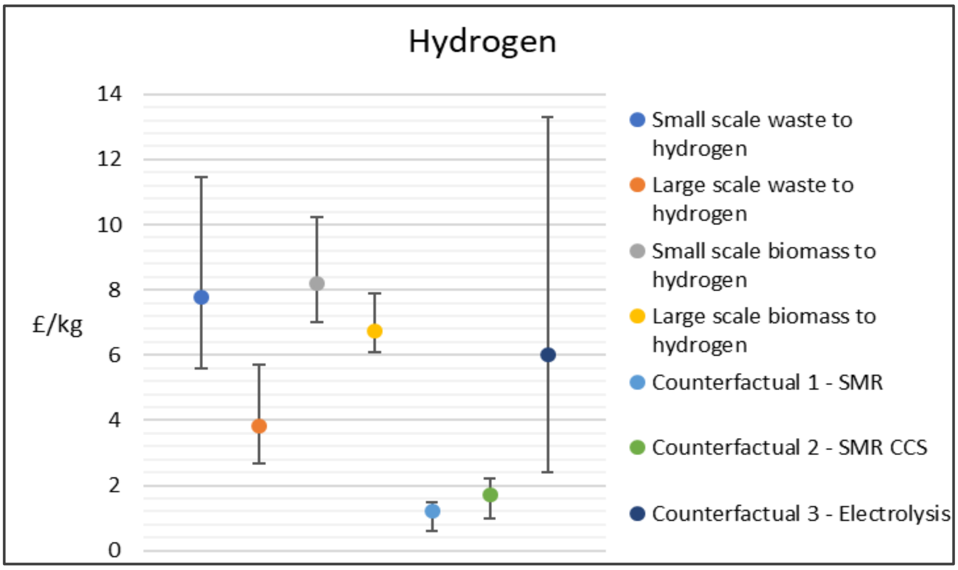

Figure 7: LCOH for biomass and waste with capture of the rich CO_2 stream

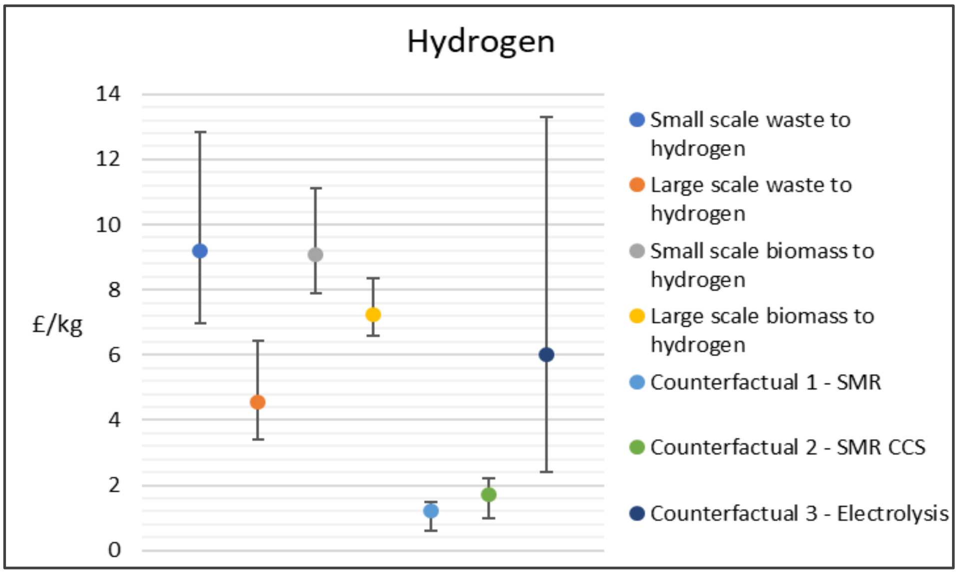

Figure 8: **LCOH for biomass and waste with capture of the CO_2 from the rich and flue gas streams**

Given, that commercial deployment of this technology has not yet occurred, the derived values reflect the uncertainty in plant capex and opex in the absence of front-end engineering design (FEED) studies and long term plant performance and operating costs.

The economic impact of the cost incurred to purchase biomass (as compared to the revenue earned from gate fees for waste) is highlighted by the lower LCOH for large scale waste than for large scale biomass. Whilst it may appear from the graph that the small scale biomass and waste plants are comparable, it must be remembered that the biomass plants are about 3 times larger than the waste plants in both the small and large scale sizes. Therefore, the biomass plants have a significant benefit of scale. As would be expected due to the large differential in feedstock cost, the waste plants have a significantly lower LCOH than the biomass plants. Evaluation of the split of the cost of production by capex, opex and feedstock cost shows the impact of the cost of feedstock on the LCOH[13] (Table 23).

Table 23: **LCOX breakdown for H_2**

	Hydrogen				
			LCOX		
Feedstock	Plant size (tpa)	LCOX (£/kg)	Capex (£/kg)	Opex (£/kg)	Feedstock cost (£/kg)
MSW	100,000	7.53	4.99	5.70	-3.15
	550,000	3.52	2.58	4.01	-3.08
Biomass	330,000	7.99	2.75	3.48	1.76
	1,000,000	6.50	1.54	2.66	2.30
Rich CO_2 stream capture					
MSW	100,000	7.79	5.04	5.91	-3.15
	550,000	3.81	2.63	4.26	-3.08

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[13] Details of the calculation of the LCOX are described in detail in the Task 3 report: Assessment methodology and counterfactual benchmarks.

	Hydrogen				
Biomass	330,000	8.21	2.78	3.67	1.76
	1,000,000	6.76	1.57	2.89	2.30
Rich CO_2 and flue gas stream CO_2 capture					
MSW	100,000	9.18	5.89	6.44	-3.15
	550,000	4.55	3.06	4.56	-3.08
Biomass	330,000	9.09	3.32	4.02	1.76
	1,000,000	7.25	1.84	3.11	2.30

6.7 Hydrogen output

The H_2 produced will comply with BS ISO 14687:2019[14].

6.8 Overall assessment

AGTs for the conversion of biomass and waste to high purity H_2 are currently under development. To date, the technology has only been demonstrated on RDF, at a maximum feedstock throughput of 780 kg/h (2 MW_{th}). Given that the key component systems (feedstock processing, gasification, syngas treatment and CO shifting) required for the production of syngas with a high hydrogen content are the same for the other products under consideration and that these systems have been demonstrated for production at higher throughputs this suggests that operation at higher throughputs for the production of H_2 is technically feasible.

Modelling of the process indicates that H_2 yields of 67 kg/odt of waste and 77 kg/odt of biomass could be obtained using AGTs coupled with PSA. The modelled yields are based on a single pass system and indicate the lower end of the range of product output. Optimisation of plant performance to include use of adsorbents which have a very low affinity for H_2, recycling of the desorbed tail gas stream to maximise recovery of H_2 from the syngas and a high efficiency of heat integration will contribute to an increase in product yields.

Preliminary assessment on the capex for these plants suggests that for the plant sizes considered the capex could range from £312M at the smaller end of the range to £1 billion for a large scale biomass to hydrogen plant. Similarly, the capex for a waste to hydrogen plant could be on the range £175M to £511M. The gasification, syngas and synthesis system represent approximately 85-90% of the plant costs.

The LCOH without carbon capture at the plant sizes under consideration has been calculated to range from £3.50 - £8/kg using AGTs and although the range in cost is to some extent similar to that for the production of H_2 from electrolysis it is significantly higher than the cost of production of H_2 using SMR and SMR with CCS.

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[14] BS ISO 14687:2019 - Hydrogen fuel quality — Product specification. Part: All applications except proton exchange membrane (PEM) fuel cell for road vehicles

7 Fischer-Tropsch Fuels

7.1 Overall technical solution

7.1.1 Gasification

Processed feedstock for FT fuel production using AGTs undergoes gasification in a BFB gasifier at temperatures ranging from 750 - 800°C using a mixture of oxygen and superheated steam. The oxygen and steam provide not only the gasifying medium but are also the fluidising agents which maintain the fuel bed in a fluidised state at a stable temperature. The syngas produced is then subjected to further treatment in a hydrocarbon reformer at temperatures between 1,200 – 1,400°C to increase the ratio of H_2:CO from 0.8:1 to > 1.

The high temperature syngas is then discharged to a multistage syngas clean-up system for the removal of a range of contaminants which can deactivate the catalysts used in the CO shift reactor and in the FT process.

Residual ash and char from the process are continuously discharged from the base of the gasifier.

Gasification for the production of syngas for FT fuels production has been carried out at pressures up to 2 bar(g). To date the largest gasification system which has been demonstrated for the production of FT fuels from biomass and waste has a design throughput of approximately 170 kg/h (0.5 MW_{th}) which is approximately 1% of the small- scale waste to FT fuels plant evaluated in this study. A 175,000 tpa plant under construction in the USA, is currently scheduled to start commissioning in 2021.

7.1.2 Syngas treatment

Treatment of syngas prior to upgrading for the production of FT fuels is carried out using a range of systems including:

1. high efficiency cyclones for collection of entrained particulates and droplets of hydrocarbons;
2. wet scrubbing for the removal of entrained particulate matter, tars and ammonia;
3. activated carbon beds for adsorption of metals;
4. hydrolysis reactors for the reduction of COS, HCN and unsaturated hydrocarbons; and
5. amine scrubber for the removal of CO_2 and some acid gases.

As wet scrubbing reduces the syngas temperature, syngas treatment is generally carried out at temperatures ranging from 150 – 250°C and at pressures up to 50 bar(a). Heat is recovered from the gas clean-up and used to produce process steam in the overall plant heat recovery system. Effluent from the clean-up process is treated and some of it is recycled to the plant for syngas scrubbing.

7.1.3 Syngas reforming

Following gas clean up, the treated syngas is then discharged to a multistage CO shift reactor (reformer) to adjust the H_2:CO molar ratio from approximately 1 to 2 by steam reforming, using the WGS reaction.

Generally, as in other syngas reforming processes, the syngas is split into two where one stream is delivered to the reformer and the other bypasses the reformer. The stream entering the reformer is mixed with medium pressure steam to increase the ratio of steam:CO and to provide the optimal operating temperature for reaction.

Reforming of the syngas is carried out at temperatures between 450 – 500°C using superheated steam at around 30 bar(g) over metal sulphide catalysts. The syngas produced contains a mixture of H_2 and CO with a SN close to the optimal value of 2.

7.1.4 FT Synthesis

In the FT process, H_2 and CO in the syngas are catalytically converted into a complex mixture of olefins, paraffins and other compounds such as alcohols and aldehydes. The primary factors influencing product yield and FT liquid chemical composition are the syngas composition, the catalyst used and the operating temperature. The objective is to preferentially produce C8 to C18 paraffins as these compounds contain very low concentrations of sulphur, heavy metals and aromatics which result in a significant reduction in emissions during combustion.

Typically, each multi-reactor train is made up of 2 - 4 reactors which are arranged such that each may be taken out of service periodically for regeneration of the catalyst. Proprietary cobalt-based catalysts have been found to be the most effective in the production of the range of hydrocarbons required for synthesis of SPK whereas iron-based catalysts are preferentially used for production of FT diesel.

FT synthesis is highly exothermic and FT reactors include extensive heat exchange systems and heat integration to ensure that isothermal conditions are maintained in each reactor. The heat transfer rate is strictly controlled and the heat recovered is used to produce medium pressure steam for the process.

The mixture of FT liquids produced is then sent to a hydrocracking unit for upgrading to SPK or FT diesel.

7.1.5 Hydrocracking

From the FT reactor the liquids produced are sent to a hydrocracking unit where they undergo heating indirectly with steam in a hydrocracking unit at temperatures up to 400°C and pressures up to 60 bar(g). Hydrocracking produces a mixture of shorter chain hydrocarbons, water and unreacted H_2 which is known as syncrude (synthetic crude oil).

The syncrude is then cooled, condensed in a series of separators followed by fractionation to produce a mixture of liquid hydrocarbons which are similar in composition to SPK or to FT diesel. The SPK or diesel produced is cooled and sent for storage.

The remaining gases are condensed and are separated into three streams, which are:

1. naphtha (C5-C9) hydrocarbons;
2. fractionator water; and
3. offgas.

The naphtha is a secondary product which can be collected for sale or refluxed back into the process. The offgas is recycled to the fractionator for reprocessing.

7.1.6 Process residues handling

The CO_2 extracted during syngas treatment has a high purity stream (> 90%) and can be cleaned for supply for industrial gases or for storage.

Process residues from the gasifier and the syngas treatment system are continuously discharged from the plant and are collected for disposal offsite.

The output from the FT reactor includes FT tail gas (unreacted syngas and other gases), steam and FT water which contains several dissolved hydrocarbons. Most of the FT tail gas is compressed by the FT recycle compressor and recycled to the FT reactor. Any waste gases are collected from various stages of the process and sent to a thermal oxidiser to oxidise all the hydrocarbons. The flue gas is then cooled using the heat to produce process steam and cleaned to ensure any emissions are below limits set by the IED.

The FT water and effluent from the wet scrubber systems are treated onsite in a wastewater treatment plant.

7.2 Plant description

AGTs for the production of FT diesel and SPK from biomass and waste consist of the following key system components:

1. feedstock storage and delivery system;
2. BFB gasifier;
3. multistage syngas clean up system;
4. ash and char handling system;
5. syngas compression;
6. CO shift reactor;

7. CO_2 scrubbing and compression;
8. FT reactor;
9. cobalt catalyst;
10. hydrocracker;
11. hydro-isomerisation reactor;
12. hydrotreater;
13. thermal oxidiser for waste gases;
14. heat recovery systems;
15. storage tanks for products, oxygen and nitrogen;
16. auxiliary systems to provide effluent treatment, start-up boiler, cooling, compressed air etc; and
17. stack.

The estimated plant area for the four thermal sizes modelled is shown in Table 24.

Table 24: Projected AGT plant area for modelled thermal capacity

	Plant size			
	MSW		Biomass	
	100,000 tpa	550,000 tpa	330,000 tpa	1,000,000 tpa
Plant area (m^2)	25,000	100,000	70,0000	300,000

7.3 Process model assumptions

A process model was derived to estimate the key process inputs to and outputs from oxy-steam gasification of woodchip and RDF for the production of hydrogen. The model was developed based on a detailed evaluation of process data from several technology providers coupled with our experience in the assessment of a large number of gasification systems and technologies. The model is based on a single pass first-of-a-kind plant in which all of the feedstock delivered to the gasifier is used to produce FT fuels and does not consider the following:

1. number of process recycles; and
2. use of any excess energy for power generation.

In addition to the assumptions described in Section 4.3, the key assumptions are as follows:

1. 100% conversion of feedstock to syngas, ash and char;

2. The primary syngas components considered were CO, H_2, CO_2, H_2O, O_2, N_2, CH_4, NH_3, HCl, H_2S;
3. equilibrium gas composition in the gasifier;
4. benzene and naphthalene were used as the model component for tars;
5. the gasifier was assumed to operate at between 700-800°C;
6. 100% conversion of CO in a single pass;
7. H_2 is not recovered from the syngas for use in hydrocracking downstream of the FT reactor;
8. the assumed NCV of SPK is 43 MJ/kg;
9. the assumed NCV of FT diesel is 43 MJ/kg;
10. electricity consumption is based on data provided by suppliers;
11. the main consumables have been derived from the model based on the process requirements; and
12. operation at the maximum feedstock throughput for 85% of the hours in a year (7,446 h)

The results from the model for the key inputs and outputs for the two biomass plant sizes considered for the production of SPK and FT diesel are outlined in Table 25, Table 26 and illustrated in Section 7.4.2.

7.4 Biomass to FT fuels

7.4.1 Process model

Table 25: Biomass to SPK/SAF process model

Model Parameter	Units	Model outputs	
		330,000 tpa	1,000,000 tpa
Process streams	---	1	1[11]
Feedstock throughput	t/h	44	134
Thermal input rate	MW_{th}	100	643
	MW_{th}/year	744,600	4,787,778
Syngas output from gasifier	tpa	341,300	1,830,900
Syngas output from reformer	tpa	242,200	1,364,100
SPK yield	l/odt	172	181
SPK production	tpa	21,300	126,100
FT naphtha yield	l/odt	52	48
FT naphtha	tpa	5,800	30,300

Model Parameter	Units	Model outputs	
		330,000 tpa	1,000,000 tpa
Energy efficiency	%	44	39
Electricity consumption	MWe	15	82
	MWh/year	107,100	604,500
Plant availability	%	85	85
Main consumables	tpa natural gas	650	4,180
	tpa oxygen	65,900	372,000
	tpa steam	268,700	557,500
	tpa water	214,100	1,209,000
	tpa chemicals	16,500	93,000
Key material output streams	tpa CO_2[12]	151,800	854,600
	tpa flue gas	476,500	2,604,800
	tpa flue gas CO_2	72,000	394,000
	tpa bottom ash and char	716	5,580
	tpa effluent	230,600	1,302,000

Table 26: Biomass to diesel process model

Model Parameter	Units	Model outputs	
		330,000 tpa	1,000,000 tpa
Feedstock throughput	t/h	44	134
Thermal input rate	MW_{th}	100	643
	MW_{th}/year	744,600	4,787,778
Syngas output from gasifier	tpa	341,300	1,830,900
Syngas output from reformer	tpa	242,200	1,364,100
FT diesel yield	l/odt	181	190
FT diesel production	tpa	23,000	135,900
FT naphtha yield	l/odt	46	42
FT naphtha	tpa	5,200	26,500
Efficiency of conversion	%	45.4	40.8
Electricity consumption	MWe	15	82
	MWh/year	107,100	604,500
Plant availability	%	85	85
Main consumables	tpa natural gas	650	4,180
	tpa oxygen	65,890	372,000
	tpa steam	268,700	557,500
	tpa water	214,100	1,209,000
	tpa chemicals	16,500	93,000
Key material output streams	tpa CO_2[12]	151,700	854,200
	tpa flue gas	455,300	2,485,600

Model Parameter	**Units**	**Model outputs**	
		330,000 tpa	**1,000,000 tpa**
	tpa flue gas CO_2	69,000	377,000
	tpa bottom ash and char	716	5,600
	tpa effluent	230,600	1,302,000

7.4.2 Mass balance

The mass balances derived from the process models are illustrated in Figure 9, Figure 10, Figure 11 and Figure 12 overleaf.

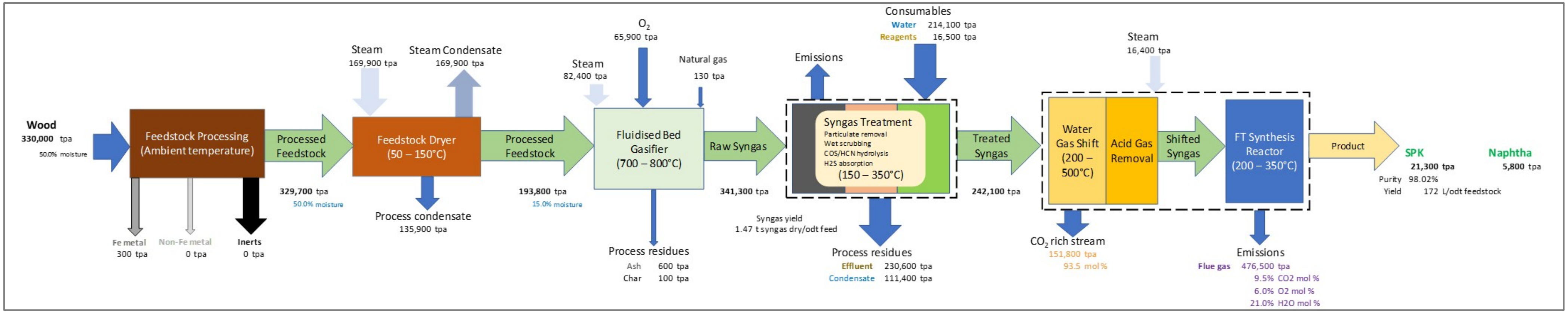

Figure 9: **Modelled mass balance for 330,000 tpa AGT biomass to SPK plant**

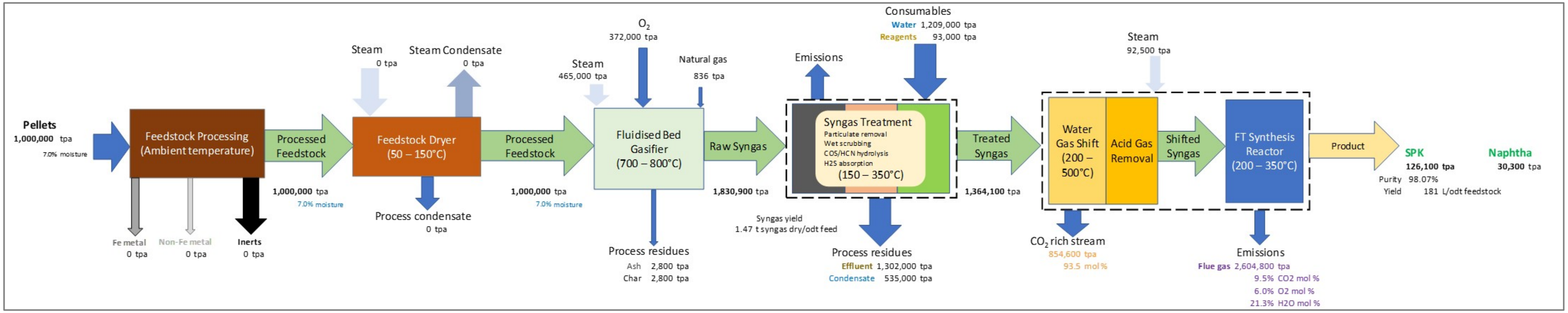

Figure 10: **Modelled mass balance for 1,000,000 tpa AGT biomass to SPK plant**

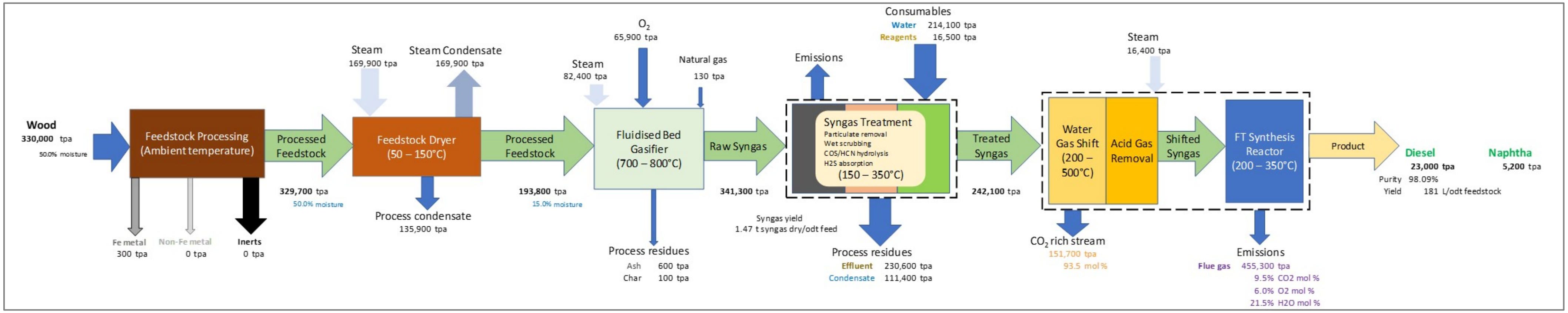

Figure 11: Modelled mass balance for 330,000 tpa AGT biomass to diesel plant

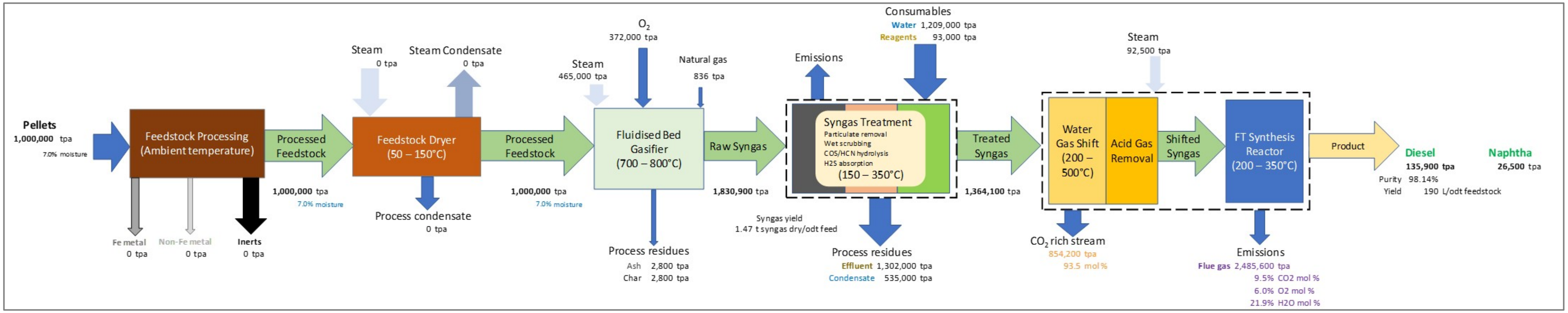

Figure 12: Modelled mass balance for 1,000,000 tpa AGT biomass to diesel plant

7.4.3 Capital cost estimate

The capital costs for FOAK biomass to SPK/FT diesel plants are outlined in Table 27. The costs for an SPK or diesel plant are considered to be similar.

Table 27: Estimated capital cost of FOAK biomass to SPK/FT diesel plants

Item	Cost, £	
	330,000 tpa	**1,000,000 tpa**
Project Development Costs		
Land	1,347,500	5,775,000
Consultancy Services	3,001,600	9,547,100
Planning & Other Regulatory	6,158,000	19,874,300
Developers Costs	21,011,300	66,829,400
Start-up Costs	10,505,700	33,414,700
Utility connection	3,868,700	19,503,900
Sub-total development costs	**45,892,800**	**154,944,400**
EPC costs		
Fuel preparation and storage	14,096,200	78,602,900
Gasifier and synthesis plant	286,065,900	876,102,600
EPC cost	300,162,100	954,705,500
Total Project Costs	**346,055,000**	**1,109,650,000**

The additional costs for installing carbon capture equipment are estimated as £AA (information to be provided in the updated version of the report)

7.4.4 Operating cost estimate

The operating costs for a FOAK biomass to SPK/FT diesel plant are shown in Table 28.

Table 28: Estimated annual operating costs for FOAK biomass to SPK/FT diesel plants

Item	Basis	Total Cost, £ pa	
		330,000 tpa	**1,000,000 tpa**
Fixed Costs			
Labour		3,016,100	4,820,200
Administration & Other Overheads		8,651,400	27,741,200
Sub-total Fixed Costs		**11,667,500**	**32,561,400**
Feedstock		**20,130,000**	**151,783,200**
Consumables			
Electricity		13,916,400	78,016,000
Oxygen		6,589,000	37,200,000
Natural gas	GCV	233,800	1,500,900
Water		69,400	348,000

Item	**Basis**	**Total Cost, £ pa**	
		330,000 tpa	**1,000,000 tpa**
Chemicals		164,700	930,000
Catalysts		303,600	921,500
Sub-total Consumables		**21,276,900**	**118,916,400**
Disposal costs			
Fe-Metal revenue		-11,700	0
Non-Fe Metal revenue		0	0
Naphtha revenue (diesel)		-1,678,800	-8,615,500
Naphtha revenue (SPK)		-1,885,400	-9,858,100
Landfill Waste	Active waste	0	0
Coarse ash disposal + Rejected inerts	Inert	5,800	27,900
Fine ash disposal	Hazardous	23,100	449,600
Effluent		352,800	1,992,100
Sub-total Disposal costs (diesel)		**-1,308,800**	**-6,145,900**
Sub-total Disposal costs (SPK)		**-1,515,400**	**-7,388,500**
Maintenance	2.5% of capex/year	**8,651,400**	**27,741,200**
Total Opex (diesel)		**60,417,000**	**324,856,000**
Total Opex (SPK)		**60,210,000**	**323,614,000**

7.4.5 Capital and operating costs estimate with carbon capture

The additional capital and operating costs for carbon capture equipment for FOAK biomass to SPK and FT diesel plants are listed in Table 29.

Table 29: Estimated capital and operating costs for carbon capture plant for FOAK biomass to SPK and FT diesel plants

Parameters	**330,000 tpa**		**1,000,000 tpa**	
	SPK	FT diesel	SPK	FT diesel
Capital costs	**Cost, £**			
CO_2 rich stream capture cost	2,462,000	2,461,000	13,962,000	13,956,000
CO_2 in flue gas capture cost	53,386,000	53,213,000	158,206,000	157,691,000
Total capture costs	55,848,000	55,674,000	172,168,000	171,647,000
Operating costs	**Cost, £ pa**			
CO_2 rich stream capture	1,396,000	1,395,000	9,353,000	9,348,000
CO_2 in flue gas capture	3,780,000	3,767,000	14,181,000	14,125,000

7.5 Waste to FT fuels

7.5.1 Process model

Table 30: Waste to SPK/SAF process model

Model Parameter	Units	Model outputs	
		100,000 tpa	550,000 tpa
Process streams	---	1	2
Feedstock throughput	t/h	13	74
Thermal input rate	MW_{th}	36	199
	MW_{th}/year	268,100	1,481,800
Syngas output from gasifier	tpa	121,000	665,900
Syngas output from reformer	tpa	75,000	413,700
SPK yield	L/odt	143	151
SPK production	tpa	6,200	36,300
FT naphtha yield	l/odt	48	46
FT naphtha	tpa	1,900	9,900
Electricity consumption	MWe	5	29
	MWh/year	37,800	208,100
Plant availability	%	85	85
Energy efficiency	%	37	38
Main consumables	tpa natural gas	235	1,295
	tpa oxygen	29,100	160,100
	tpa steam	70,500	387,700
	tpa water	75,700	416,100
	tpa chemicals	5,280	32,000
Key material output streams	tpa CO_2[12]	47,000	259,300
	tpa flue gas	151,900	814,000
	tpa flue gas CO_2	23,000	123,000
	tpa bottom ash and char	716	5,600
	tpa effluent	81,500	448,100

Table 31: Waste to diesel process model

Model Parameter	Units	Model outputs	
		100,000 tpa	550,000 tpa
Process streams	---	1	2
Feedstock throughput	t/h	13	74
Thermal input rate	MW_{th}	36	199
	MW_{th}/year	268,100	1,481,800
Syngas output from gasifier	t/h	121,000	665,900
Syngas output from reformer	t/h	75,000	413,700

Model Parameter	**Units**	**Model outputs**	
		100,000 tpa	**550,000 tpa**
FT diesel yield	L/odt	151	159
FT diesel production	tpa	6,800	39,200
FT naphtha yield	l/odt	46	42
FT naphtha	tpa	1,700	8,800
Electricity consumption	MWe	5	29
	MWh/year	37,800	208,100
Plant availability	%	85	85
Energy efficiency	%	38	39
Main consumables	tpa natural gas	235	1,295
	tpa oxygen	29,100	160,000
	tpa steam	70,500	387,700
	tpa water	75,700	416,100
	tpa chemicals	5,280	32,000
Key material output streams	tpa CO_2[12]	47,000	259,100
	tpa flue gas	145,300	777,900
	tpa flue gas CO_2	22,000	118,000
	tpa bottom ash and char	716	5,600
	tpa effluent	81,500	448,100

7.5.2 Mass balance

The mass balances derived from the process models are illustrated in Figure 13, Figure 14, Figure 15, and Figure 16.

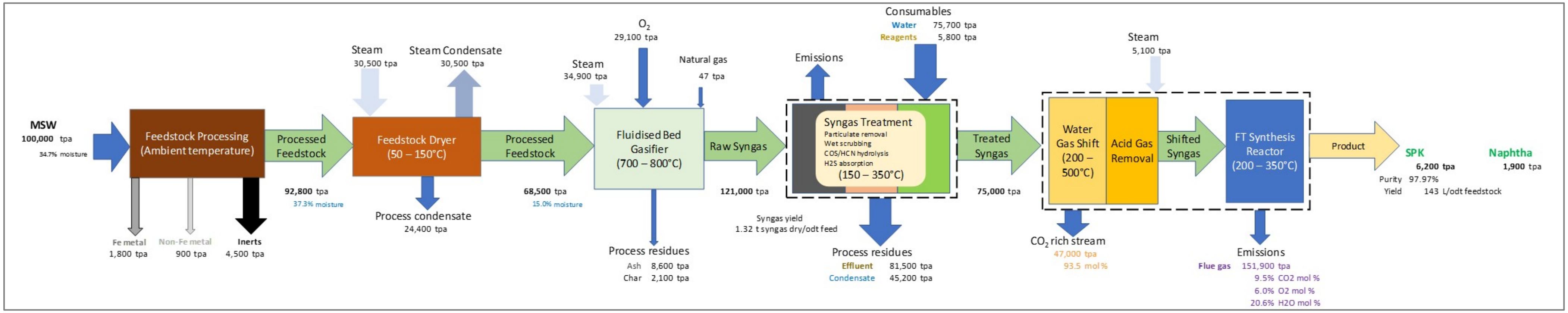

Figure 13: Modelled mass balance for 100,000 tpa AGT MSW to SPK plant

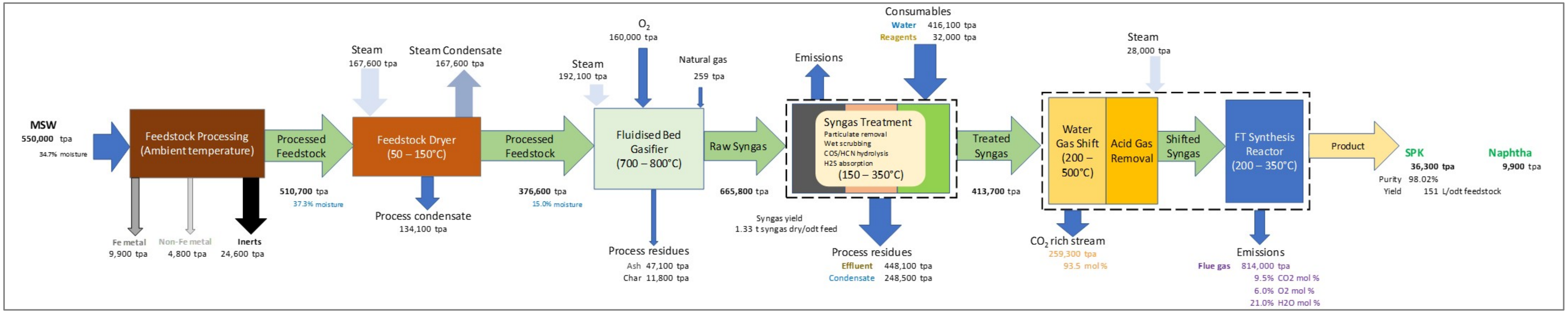

Figure 14: Modelled mass balance for 550,000 tpa AGT MSW to SPK plant

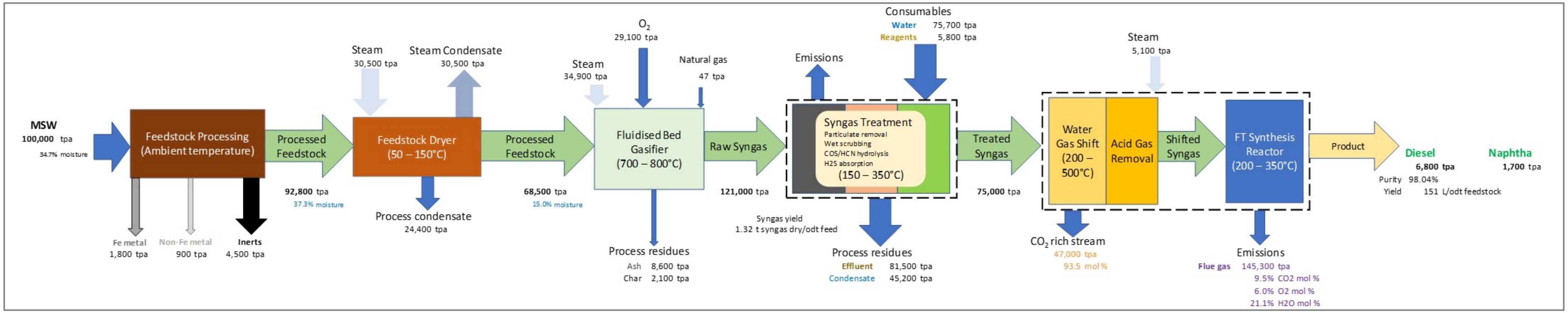

Figure 15: Modelled mass balance for 100,000 tpa AGT MSW to diesel plant

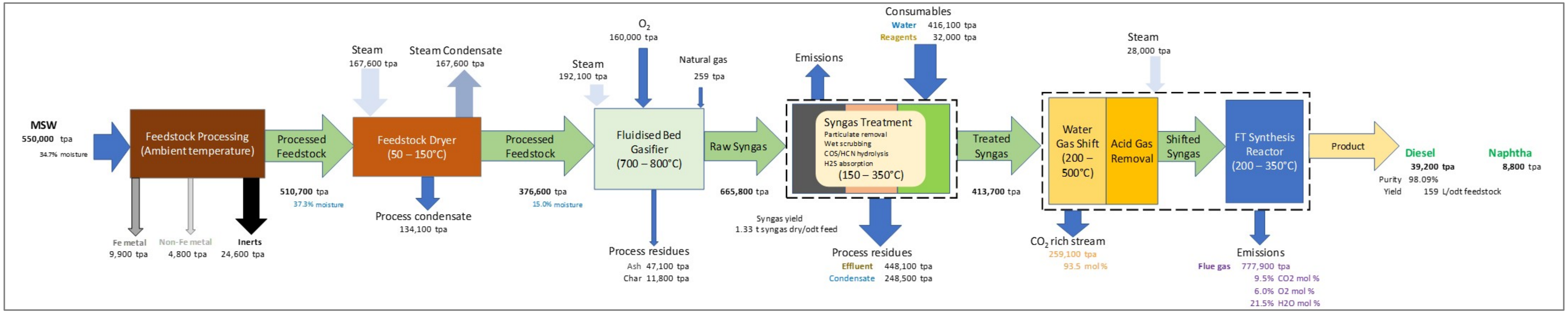

Figure 16: Modelled mass balance for 550,000 tpa AGT MSW to diesel plant

7.5.3 Capital cost estimate

The capital costs for FOAK MSW to SPK/FT diesel plants are outlined in Table 32.

Table 32: Estimated capital cost for FOAK MSW to SPK/FT diesel plants

Item	Cost, £	
	100,000 tpa	550,000 tpa
Project Development Costs		
Land	481,300	1,925,000
Consultancy Services	1,685,100	4,865,600
Planning & Other Regulatory	3,434,000	10,028,900
Developers Costs	11,795,800	34,059,400
Start-up Costs	5,897,900	17,029,700
Utility connection	1,592,900	7,440,300
Sub-total development costs	**24,887,000**	**75,348,900**
EPC costs		
Fuel preparation and storage	13,549,100	53,015,700
Gasifier and synthesis plant	154,962,800	433,547,100
EPC cost	**168,511,900**	**486,562,800**
Total Project Costs	**193,399,000**	**561,912,000**

7.5.4 Operating cost estimate

The estimated operating costs for FOAK MSW to SPK/FT diesel plants are outlined in Table 33.

Table 33: Estimated annual operating costs for FOAK MSW to SPK/FT diesel plants

Item	Basis	Total Cost, £ pa	
		100,000 tpa	550,000 tpa
Fixed Costs			
Labour		2,455,600	5,100,900
Administration & Other Overheads		4,835,000	14,047,800
Sub-total Fixed Costs		**7,290,600**	**19,148,700**
Feedstock		**-11,200,000**	**-61,600,000**
Consumables			
Electricity		5,098,200	28,040,100
Oxygen		2,910,000	16,004,900
Natural gas		84,400	464,400
Water		30,700	126,000
Chemicals		58,200	320,100
Catalysts		162,600	455,300
Sub-total Consumables		**8,344,100**	**45,410,800**

Item	Basis	Total Cost, £ pa	
		100,000 tpa	550,000 tpa
Disposal costs			
Fe- metal revenue		-80,000	-440,000
Non-Fe metal revenue		-135,900	-747,500
Naphtha revenue (diesel)		-563,100	-2,867,900
Naphtha revenue (SPK)		-622,800	-3,220,900
Landfill Waste	Active waste	536,500	2,950,600
Coarse ash disposal + Inerts	Inert	85,700	471,200
Fine ash disposal	Hazardous	345,200	1,898,400
Effluent		124,700	685,600
Sub-total Disposal costs (diesel)		**313,100**	**1,950,400**
Sub-total Disposal costs (SPK)		**253,400**	**1,597,400**
Maintenance	2.5% of capex/year	**4,835,000**	**14,047,800**
Total Opex (diesel)		**9,583,000**	**18,958,000**
Total Opex (SPK)		**9,523,000**	**18,605,000**

7.5.5 Capital and operating costs estimate with carbon capture

The additional capital and operating costs for carbon capture equipment for FOAK MSW to SPK/FT diesel plants are listed in Table 34.

Table 34: Estimated capital and operating costs for carbon capture plant for FOAK MSW to SPK/FT diesel plants

Parameters	100,000 tpa		550,000 tpa	
	SPK	FT diesel	SPK	FT diesel
Capital costs	**Cost, £**			
CO_2 rich stream capture cost	1,108,000	1,108,000	6,223,000	6,220,000
CO_2 in flue gas capture cost	26,475,000	26,389,000	77,490,000	77,239,000
Total capture costs	27,583,000	27,497,000	83,713,000	83,459,000
Operating costs	**Cost, £ pa**			
CO_2 rich stream capture	526,000	525,000	3,119,000	3,117,000
CO_2 in flue gas capture	1,745,000	1,739,000	5,964,000	5,942,000

7.6 Levelised cost of production

Comparison of the levelised cost of production of SPK and FT diesel from AGTs and crude oil is illustrated in Figure 17 and Figure 18. Figures 19, 20, 21 and 22 show the impact of integrating carbon capture.

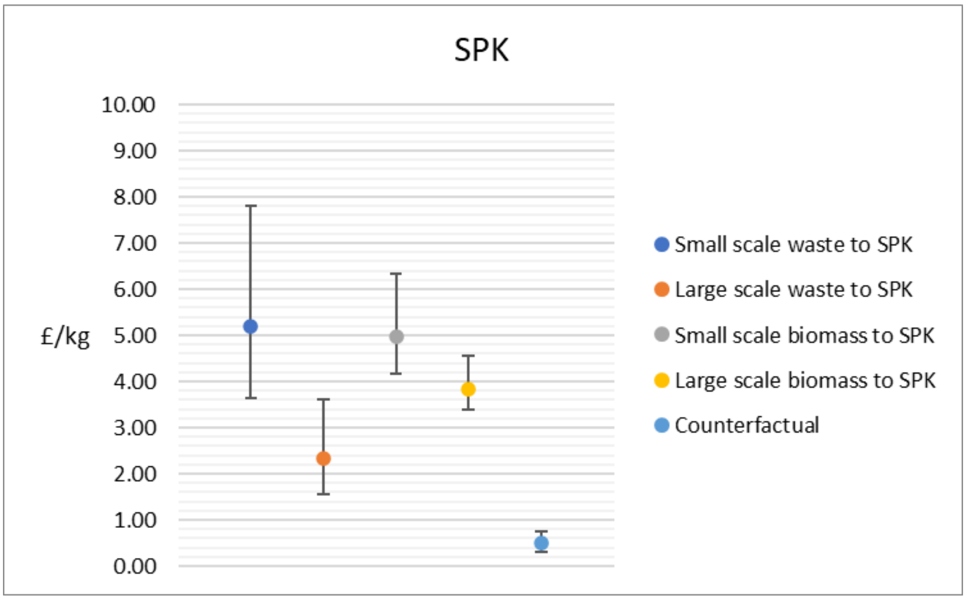

Figure 17: Levelised cost of SPK from biomass and waste using AGTs

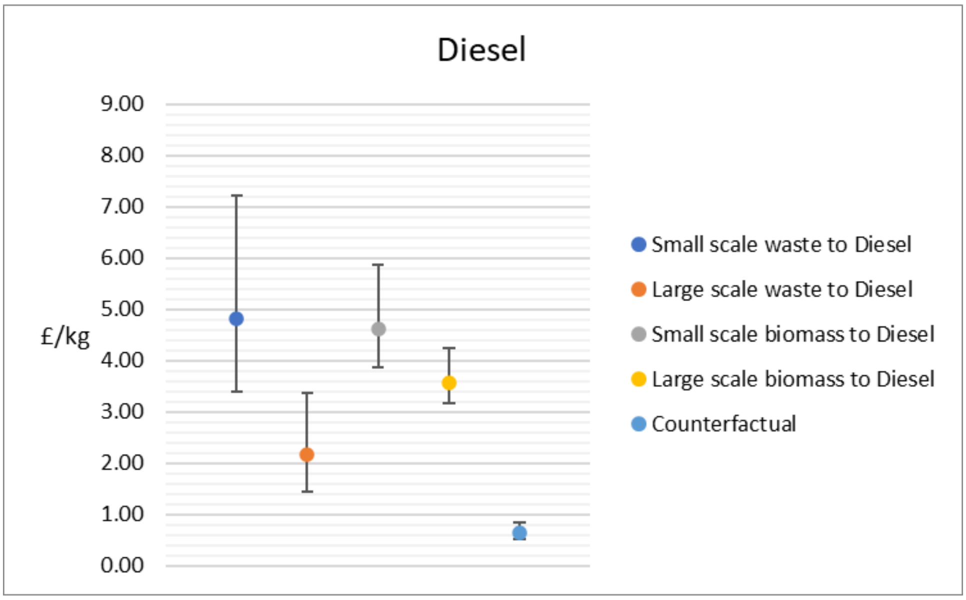

Figure 18: Levelised cost of FT diesel from biomass and waste using AGTs

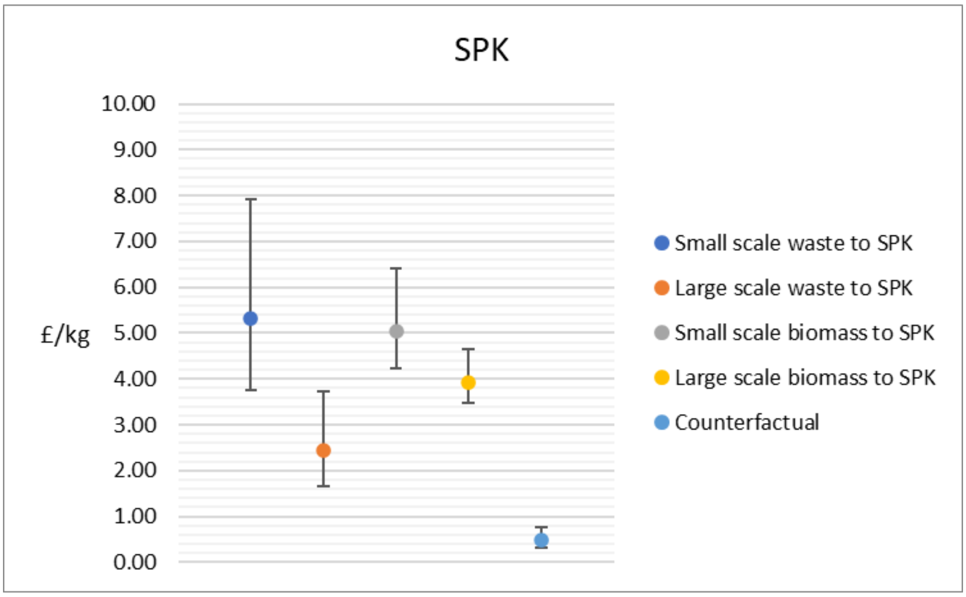

Figure 19: **Levelised cost of SPK from biomass and waste with capture of the rich CO_2 stream**

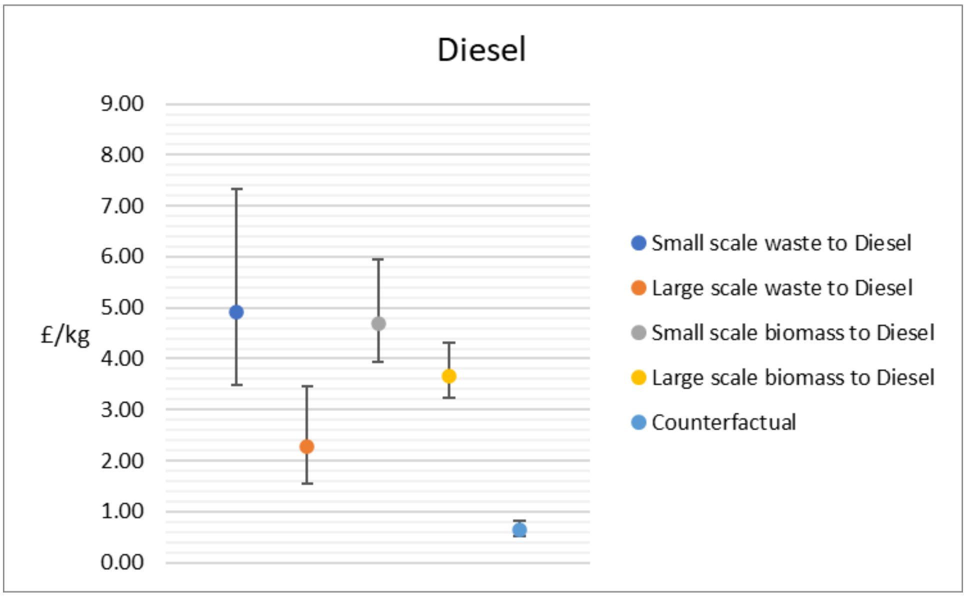

Figure 20: **Levelised cost of FT diesel from biomass and waste with capture of the rich CO_2 stream**

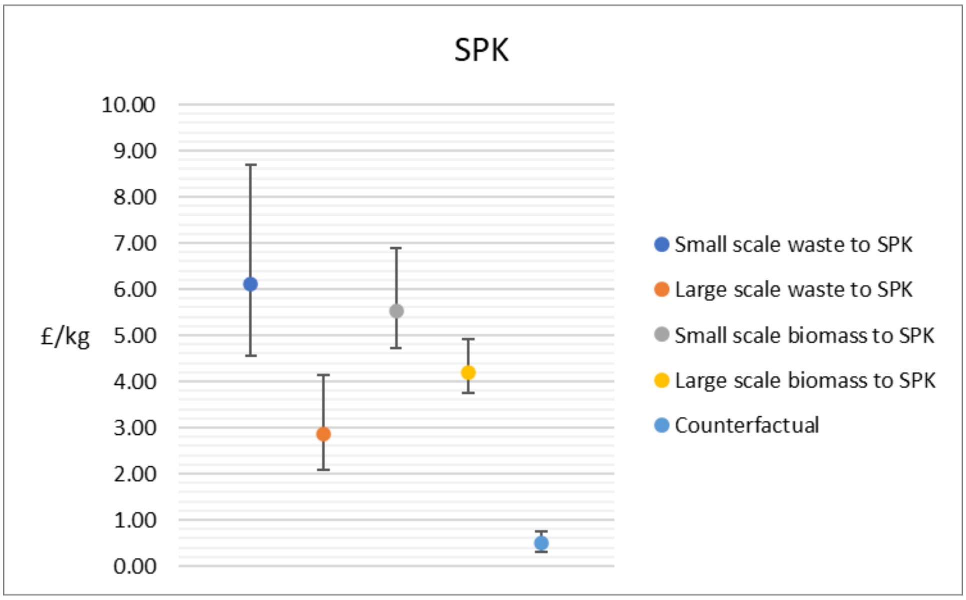

Figure 21: **Levelised cost of SPK from biomass and waste with capture of the CO_2 from the rich and flue gas streams**

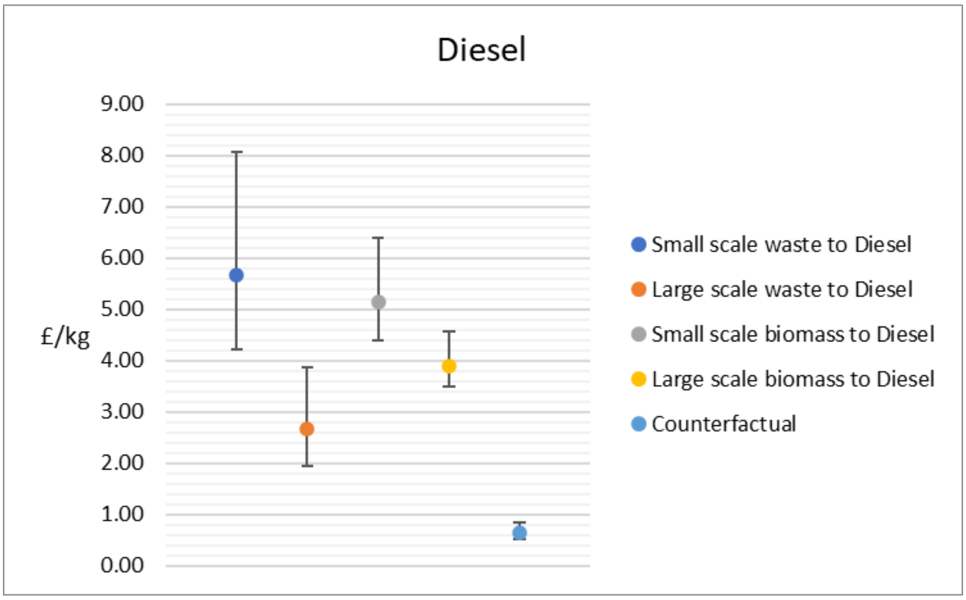

Figure 22: **Levelised cost of FT diesel from biomass and waste with capture of the CO_2 from the rich and flue gas streams**

Overall, the LCOX for these fuels using FT technology is significantly higher than for production from crude oil. Given that commercial deployment of this technology has not yet occurred, the calculated values reflect the uncertainty in plant capex and opex in the absence of FEED studies and long term plant performance and cost details.

The economic impact of the cost of biomass on the levelised cost of production is illustrated by the lower LCOX for the production of SPK and FT diesel using large scale waste when compared to large scale biomass. Whilst it may appear from the graph that the smalls scale biomass and waste plants are comparable, it must be remembered that the biomass plants are about 3 times larger than the waste plants in both the small and large scale sizes. Therefore, the biomass plants have a significant benefit of scale. As would be expected due to the large differential in feedstock cost, the waste plants have a significantly lower LCOX than the biomass plants.

Table 35 and Table 36 show that although the capex and opex of the small-scale biomass plants benefit from economies of scale, the cost of feedstock increases the cost of fuel production.

Table 35: LCOX breakdown for SPK

SPK					
Feedstock	Plant size (tpa)	LCOX (£/kg)	LCOX		
			Capex (£/kg)	Opex (£/kg)	Feedstock cost (£/kg)
MSW	100,000	5.21	3.53	3.65	-1.97
	550,000	2.33	1.76	2.43	-1.86
Biomass	330,000	4.97	1.86	2.07	1.04
	1,000,000	3.83	1.00	1.50	1.32
Rich CO_2 stream capture					
MSW	100,000	5.32	3.55	3.74	-1.97
	550,000	2.44	1.78	2.52	-1.86
Biomass	330,000	5.05	1.87	2.14	1.04
	1,000,000	3.92	1.02	1.58	1.32
Rich CO_2 and flue gas stream capture					
MSW	100,000	6.11	4.03	4.05	-1.97
	550,000	2.86	2.03	2.70	-1.86
Biomass	330,000	5.54	2.15	2.34	1.04
	1,000,000	4.19	1.16	1.70	1.32

Table 36: LCOX breakdown for FT Diesel

FT Diesel					
			LCOX		
Feedstock	Plant size (tpa)	LCOX (£/kg)	Capex (£/kg)	Opex (£/kg)	Feedstock cost (£/kg)
MSW	100,000	4.82	3.26	3.38	-1.82
	550,000	2.17	1.63	2.26	-1.73
Biomass	330,000	4.61	1.72	1.93	0.96
	1,000,000	3.56	0.93	1.40	1.23
Rich CO_2 stream capture					
MSW	100,000	4.93	3.28	3.47	-1.82
	550,000	2.27	1.65	2.35	-1.73
Biomass	330,000	4.69	1.73	2.00	0.96
	1,000,000	3.65	0.94	1.48	1.23
Rich CO_2 and flue gas stream CO_2 capture					
MSW	100,000	5.66	3.73	3.75	-1.82
	550,000	2.66	1.88	2.51	-1.73
Biomass	330,000	5.14	2.00	2.18	0.96
	1,000,000	3.90	1.08	1.59	1.23

7.7 FT fuels output

SPK is a complex and variable mixture which needs to meet a performance rather than a compositional specification. Our review of analyses carried out on samples from AGTs using FT technology for the production of SPK shows that the product is compliant with the requirements of ASTM D7566[15] and blending with jet fuel up to a maximum of 50% on a volume basis would be permitted.

Globally, blended jet fuel must also comply with ASTM D1655[16] and DEF STAN 91-91 to ensure that fuel can be safely used to fuel passenger aircraft.

Initial data on FT diesel from AGTs shows that the diesel produced is compliant with ASTM D975[17] and EN 15940:2016 + A1:2018 + AC:2019[18]

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[15] ASTM D7566 - Standard specification for aviation fuel containing synthesised hydrocarbons

[16] ASTM D1655 - Standard Specification for Aviation Turbine Fuels

[17] ASTM D975 - Standard Specification for Diesel Fuel

[18] EN 15940:2016 + A1:2018 + AC:2019 - Automotive fuels. Paraffinic diesel fuel from synthesis or hydrotreatment. Requirements and test methods

7.8 Overall assessment

Currently, the production of SPK and FT diesel from biomass and waste using AGTs has only been demonstrated on RDF at a throughput of 170 kg/h (0.5 MW_{th}). Given that the key component systems (feedstock processing, gasification, syngas treatment and CO shifting) required for the production of syngas with a high hydrogen content are the same for the other products under consideration and that these systems have been demonstrated for production at higher throughputs this indicates the potential for operation at higher feedstock throughputs. In addition, based on the experience gained from operation of FT systems on natural gas, plants with a product output of up to 1,900 m^3/day of FT liquids are in operation.

Modelling of the process indicates that yields of SPK ranging from 143 l/odt of waste to 181 l/odt of biomass and from 151 l/odt of waste to 190 l/odt for biomass could be achieved. The modelled yields are based on a single pass system and indicate the lower end of the range of product output. Optimisation of plant performance to include key operations such as process recycles (recovery of unconverted syngas and recirculation to the FT reactor to maximise the efficiency of conversion to product) and a high efficiency of heat integration as would be standard in commercial plants would contribute to an increase in product yields.

Preliminary assessment on the capex for these plants suggests that for the plant sizes considered the capex could range from £354M at the smaller end of the range to £1.1 billion for a large scale biomass to SPK/FT diesel plant. Similarly, the capex for a waste to SPK/FT diesel plant could be on the range £198M to £576M. The gasification, syngas and synthesis system represent approximately 85-90% of the plant costs.

The levelised cost of production without carbon capture at the plant capacities under consideration ranges from £2.17 - £4.82/kg using AGTs and at the high end of the range it is as much as nine times the cost of production from crude oil.

8 Methane

8.0 Overall technical solution

8.0.1 Gasification

After processing of the feedstock, it is then conveyed via lock hoppers to the BFB gasifier where the feedstock is heated to temperatures ranging from 700-800°C using a mixture of superheated steam and oxygen. The oxy-steam mixture acts as both the gasifying agent and the fluidising medium. Syngas at temperatures of up to 750°C is then discharged to a multistage syngas clean-up system for the removal of a range of contaminants which can deactivate the catalysts used in the syngas upgrading process and or impact on the quality of the methane produced. The steam is raised using heat recovered from downstream process reactions.

Residual ash and char from the process are continuously discharged from the base of the gasifier.

To date the process for the production of methane has been developed for operation at atmospheric or slightly lower (-2 to -10 mbar(g)) pressure. Currently, the largest demonstrator plant for the production of methane is fuelled by forestry residues and has a design throughput of approximately 47,000 tpa (32 MW_{th}). This plant is marginally lower than the small scale waste to methane plant evaluated in this study but is 14% smaller than the small scale biomass to methane plant considered here.

8.0.2 Syngas treatment

Treatment of syngas prior to upgrading for the production of methane is carried out using a range of systems including:

1. plasma conversion;
2. high efficiency cyclones for collection of entrained particulates and droplets of hydrocarbons;
3. wet scrubbing for the removal of entrained particulate matter, tars and ammonia;
4. activated carbon beds for adsorption of metals;
5. hydrolysis reactors for the reduction of COS, HCN and unsaturated hydrocarbons; and
6. amine scrubber for the removal of CO_2 and some acid gases.

As wet scrubbing reduces the syngas temperature, syngas treatment is generally carried out at temperatures ranging from 150 – 250°C and at pressures up to 50 bar(a). As some of the reactions occurring during gas clean-up are exothermic the heat produced is recovered and used in the overall heat recovery system to generate process steam. Effluent from the clean-up process is treated and some of it is recycled to the plant for syngas scrubbing.

8.0.3 Syngas reforming

Following gas clean up, the treated syngas is then discharged to a multistage CO shift reactor (reformer) to adjust the H_2:CO molar ratio from approximately 1 to 3 by steam reforming, using the WGS reaction.

Generally, as in other syngas reforming processes the syngas is split into two where one stream is delivered to the reformer and the other bypasses the reformer. The stream entering the reformer is mixed with medium pressure steam to increase the ratio of steam:CO and to provide the optimal operating temperature for reaction.

Reforming of the syngas is carried out at temperatures ranging from 200°C (low temperature shift) to 500°C (high temperature shift) and is catalysed by cobalt-molybdenum catalysts in low temperature shift reactors and chromium or copper promoted iron oxide catalysts in high temperature shift reactors. The syngas produced contains a mixture of H_2 and carbon oxides with a stoichiometric number (SN) close to the optimal value of 3.

8.0.4 Methanation

Syngas from the CO shift reactor is then compressed to the operating pressure required for methanation. In the methanator the mixture of H_2, CO and CO_2 in the syngas is reacted over proprietary catalysts including mixtures of nickel, iron and ruthenium. The optimal temperature range for methanation is between 250-350°C and at pressures between 15 – 30 bar(g), however the process is highly exothermic, and process operating temperatures must be tightly controlled to maximise on the operating life of the catalyst. Consequently, the heat produced is recovered to produce medium pressure steam for use in the process.

The raw methane produced is then treated for the removal of CO_2 followed by drying before being sent to buffer storage tanks prior to injection into the gas grid.

8.0.5 Process residues handling

The CO_2 extracted from the methane has a high purity (> 90%) and can be cleaned and processed through liquefaction to food grade quality or as supply for industrial gases or stored.

Process residues from the gasifier and the syngas treatment system are continuously discharged from the plant and are collected for disposal offsite. Effluent from the wet scrubber systems is treated onsite in a wastewater treatment plant.

Any waste gases are collected from various stages of the process and sent to a thermal oxidiser to oxidise all the hydrocarbons. The flue gas is then cooled using the heat to produce process steam and cleaned to ensure any emissions are below limits set by the IED.

8.1 Plant description

AGT plants for the production of methane consist of the following key system components:

1. feedstock storage and delivery system;
2. BFB gasifier;
3. ash and char handling system;
4. multistage syngas clean up system;
5. syngas compression;
6. CO shift reactor;
7. methanators;
8. CO_2 scrubbing and compression;
9. thermal oxidiser for waste gases;
10. heat recovery systems;
11. storage tanks for products, oxygen and nitrogen;;
12. auxiliary systems including effluent treatment, start-up boiler, cooling, compressed air etc; and
13. stack

The estimated plant area for the four thermal sizes modelled is shown in Table 37.

Table 37: Projected AGT plant area for modelled thermal capacity

	Plant size			
	MSW		Biomass	
	100,000 tpa	550,000 tpa	330,000 tpa	1,000,000 tpa
Plant area (m^2)	25,000	100,000	70,0000	300,000

8.2 Process model assumptions

A process model was derived to estimate the key process inputs to and outputs from oxy-steam gasification of woodchip and RDF for the production of methane. The model was developed based on a detailed evaluation of process data from several technology providers coupled with our experience in the assessment of a large number of gasification systems and technologies. The model is based on a single pass first-of-a-kind plant in which all of the feedstock delivered to the gasifier is used to produce methane and does not consider the following:

1. number of process recycles; and
2. use of any excess energy for power generation.

In addition to the assumptions described in Section 4.3, the key assumptions are as follows:

1. 100% conversion of feedstock to syngas, ash and char;
2. The primary syngas components considered were CO, H_2, CO_2, H_2O, O_2, N_2, CH_4, NH_3, HCl, H_2S;
3. equilibrium gas composition in the gasifier;
4. benzene and naphthalene were used as the model component for tars;
5. the gasifier was assumed to operate at between 700-800°C at pressures greater than atmospheric;
6. the assumed NCV of methane is 46 MJ/kg;
7. electricity consumption is based on data provided by suppliers;
8. the main consumables have been derived from the model based on the process requirements; and
9. operation at the maximum feedstock throughput for 85% of the hours in a year (7,446 h)

The results from the model for the key inputs and outputs for the two biomass plant sizes considered are outlined in Table 38 and illustrated in Section 8.3.2.

8.3 Biomass to methane

8.3.1 Process model

Table 38: Biomass to methane process model

Model Parameter	Units	330,000 tpa	1,000,000 tpa
Process streams	---	1	1[11]
Feedstock throughput	t/h	44	134
Thermal input rate	MW_{th}	100	643
	MW_{th}/year	744,600	4,788,000
Syngas output from gasifier	tpa	341,300	1,831,000
Syngas output from reformer	tpa	242,100	1,364,000
Methane yield	kg/odt	184	188
Methane production	tpa	30,300	175,200
Energy efficiency	%	52	47
Electricity consumption	MWe	15	82
	MWh/year	107,000	604,500
Plant availability	%	85	85

Model Parameter	Units	330,000 tpa	1,000,000 tpa
Main consumables	tpa natural gas	650	4,180
	tpa oxygen	65,900	372,000
	tpa steam	284,100	644,200
	tpa water	214,100	1,209,000
	tpa chemicals	16,500	93,000
Key material output streams	tpa CO_2[12]	184,500	1,048,200
	tpa flue gas	315,400	1,743,100
	tpa flue gas CO_2	47,000	241,000
	tpa bottom ash and char	716	5,600
	m^3/year effluent	230,600	1,302,000

8.3.2 Mass balance

The mass balances derived from the process models are illustrated in Figure 23 and Figure 24 overleaf.

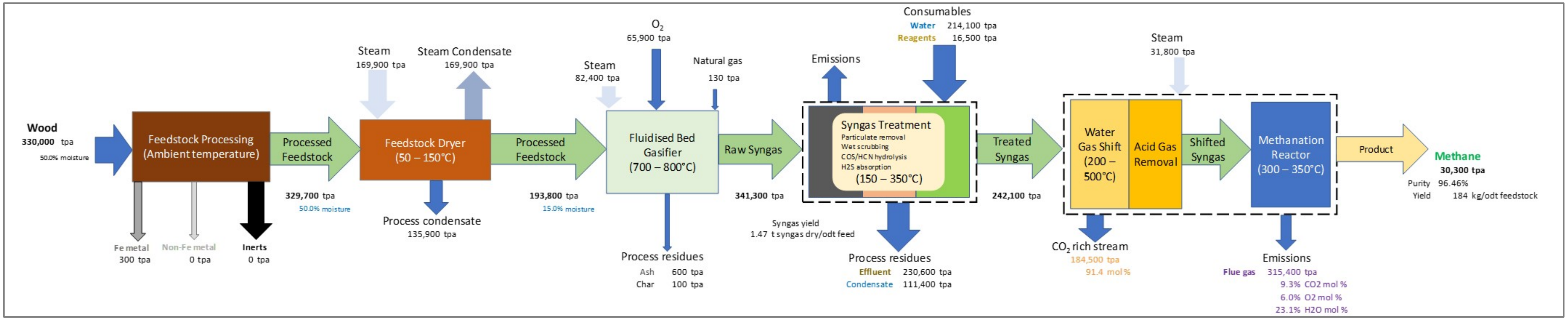

Figure 23: Modelled mass balance for 330,000 tpa AGT biomass to methane plant

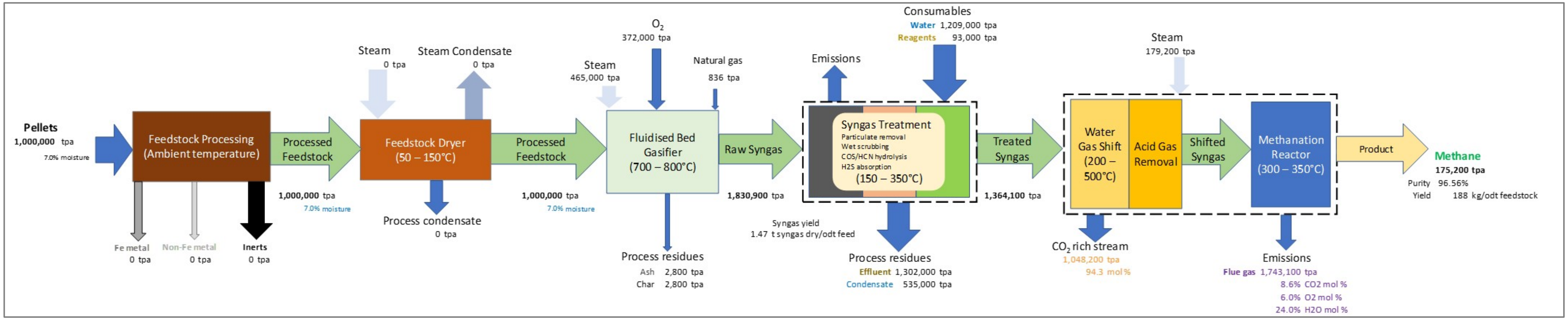

Figure 24: Modelled mass balance for 1,000,000 tpa AGT biomass to methane plant

8.3.3 Capital cost estimate

The estimated capital costs for FOAK biomass to methane plants are outlined in Table 39.

Table 39: Estimated capital costs for FOAK biomass to methane plants

Item	Cost, £	
	330,000 tpa	**1,000,000 tpa**
Project Development Costs		
Land	1,347,500	5,775,000
Consultancy Services	2,531,900	8,108,300
Planning & Other Regulatory	5,212,800	16,979,400
Developers Costs	17,723,000	56,757,800
Start-up Costs	8,861,500	28,378,900
Utility connection	3,727,700	19,072,300
Sub-total development costs	**39,404,400**	**135,071,700**
EPC costs		
Fuel preparation and storage	14,096,000	78,603,100
Gasifier and synthesis plant	239,089,200	732,222,800
EPC cost	253,185,200	810,825,900
Total Project Costs	**292,590,000**	**945,898,000**

The additional costs for installing carbon capture equipment are estimated as £AA (information to be provided in the updated version of the report).

8.3.4 Operating cost estimate

The estimated operating costs for FOAK biomass to methane plants are outlined in Table 40.

Table 40: Estimated annual operating costs for FOAK biomass to methane plants

Item	Basis	Total Cost, £ pa	
		330,000 tpa	**1,000,000 tpa**
Fixed Costs			
Labour		3,016,100	4,820,200
Administration & Other Overheads		7,314,700	23,647,400
Sub-total Fixed Costs		**10,330,800**	**28,467,600**
Feedstock		**20,130,000**	**151,783,200**
Consumables			
Electricity		13,916,400	78,016,000
Oxygen		6,589,000	37,200,000
Natural gas	GCV	233,800	1,500,900
Water		69,400	348,000
Chemicals		164,700	930,000

Item	Basis	Total Cost, £ pa	
		330,000 tpa	**1,000,000 tpa**
Catalysts		303,600	921,500
Sub-total Consumables		**21,276,900**	**118,916,400**
Disposal costs			
Fe- metal revenue		-11,700	0
Non-Fe metal revenue		0	0
Landfill Waste	Active waste	0	0
Coarse ash disposal + inerts	Inert	5,800	27,900
Fine ash disposal	Hazardous	23,100	449,600
Effluent		352,800	1,992,100
Sub-total Disposal costs		**370,000**	**2,469,600**
Maintenance	2.5% of capex/year	**7,314,700**	**23,647,400**
Total Opex		**59,422,000**	**325,284,000**

8.3.5 Capital and operating costs estimate with carbon capture

The additional capital and operating costs for carbon capture equipment for FOAK MSW to methane plants are listed in Table 41.

Table 42: Estimated capital and operating costs for carbon capture plant for FOAK biomass to methane plants

Parameters	330,000 tpa	1,000,000 tpa
Capital costs	**Cost, £**	
CO_2 rich stream capture cost	2,819,000	15,724,000
CO_2 in flue gas capture cost	51,509,000	147,208,000
Total capture costs	54,328,000	162,932,000
Operating costs	**Cost, £ pa**	
CO_2 rich stream capture	1,656,000	11,347,000
CO_2 in flue gas capture	3,621,000	12,729,000

8.4 Waste to methane

8.4.1 Process model

Table 43: Waste to methane process model

Model Parameter	Units	100,000 tpa	550,000 tpa
Process streams	---	1	2
Feedstock throughput	t/h	13	74
Thermal input rate	MW_{th}	36	199
	MW_{th}/year	268,100	1,481,800

Model Parameter	Units	100,000 tpa	550,000 tpa
Syngas output from gasifier	kg/h	121,000	665,900
Syngas output from reformer	kg/h	75,000	414,000
Methane yield	kg/odt	168	172
Methane production	tpa	9,800	55,100
Energy efficiency	%	47	48
Electricity consumption	MWe	5	29
	MWh/year	37,800	208,100
Plant availability	%	85	85
Main consumables	tpa natural gas	235	1,295
	tpa oxygen	29,100	160,100
	tpa steam	75,200	414,000
	tpa water	75,700	416,100
	tpa chemicals	5,800	32,000
Key material output streams	tpa CO_2[12]	57,100	317,700
	tpa flue gas	89,700	482,300
	tpa flue gas CO_2	14,000	68,000
	tpa bottom ash and char	10,700	58,900
	m^3/year effluent	81,700	448,100

8.4.2 Mass balance

The mass balances derived from the process models are illustrated in Figure 25 and Figure 26 overleaf.

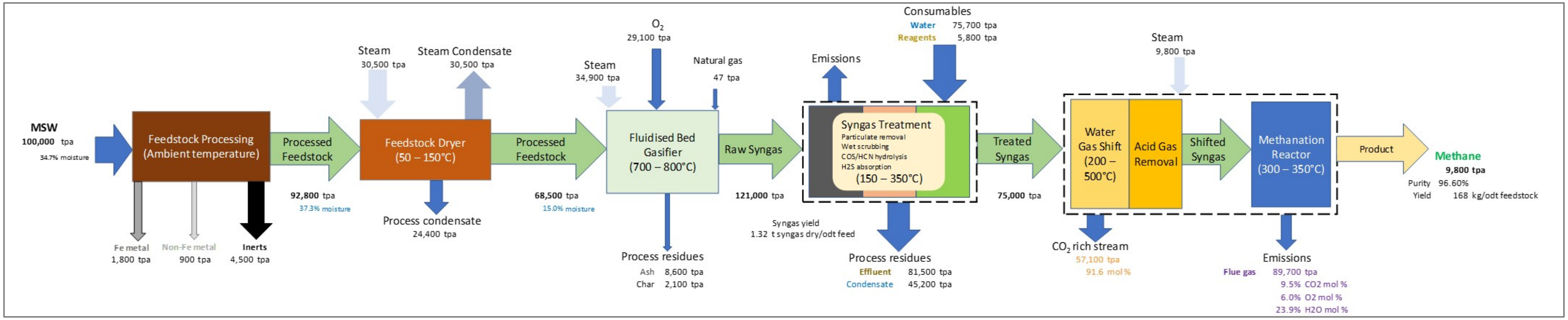

Figure 25: Modelled mass balance for 100,000 tpa AGT MSW to methane plant

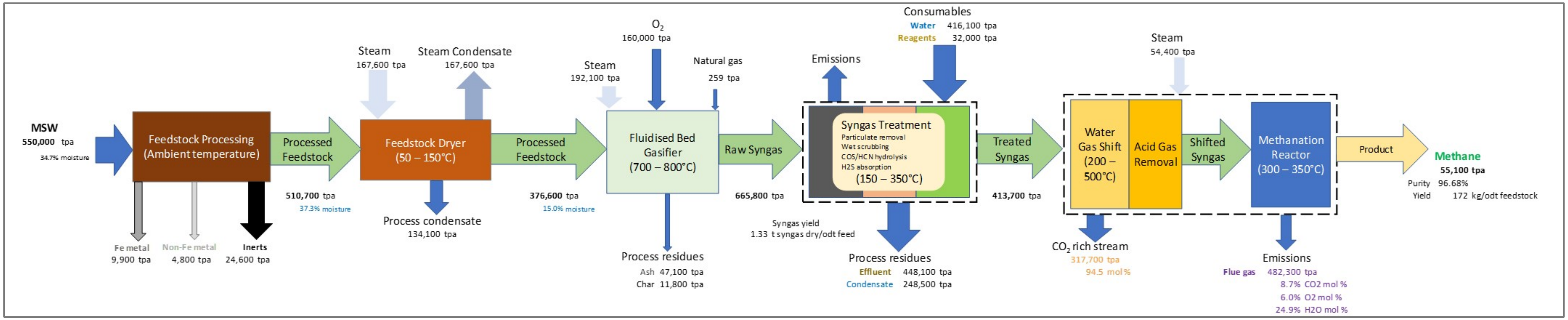

Figure 26: Modelled mass balance for 550,000 tpa AGT MSW to methane plant

8.4.3 Capital cost estimate

The estimated capital costs for FOAK MSW to methane plants are outlined in Table 44.

Table 44: Estimated capital costs for FOAK MSW to methane plants

Item	Cost, £	
Project Development Costs	**100,000 tpa**	**550,000 tpa**
Land	481,300	1,925,000
Consultancy Services	1,430,600	4,153,700
Planning & Other Regulatory	2,921,800	8,596,400
Developers Costs	10,013,900	29,075,600
Start-up Costs	5,006,900	14,537,800
Utility connection	1,516,600	7,226,700
Sub-total development costs	**21,371,100**	**65,515,200**
EPC costs		
Fuel preparation and storage	13,552,100	53,017,100
Gasifier and synthesis plant	129,503,600	362,348,500
EPC cost	143,055,700	415,365,600
Total Project Costs	**164,427,000**	**480,881,000**

8.4.4 Operating cost estimate

The estimated operating costs for FOAK MSW to methane plants are outlined in Table 45.

Table 45: Estimated annual operating costs for FOAK MSW to methane plants

Item	Basis	Total Cost, £ pa	
		100,000 tpa	**550,000 tpa**
Fixed Costs			
Labour		2,455,600	5,100,900
Administration & Other Overheads		4,110,700	12,022,000
Sub-total Fixed Costs		**6,566,300**	**17,122,900**
Feedstock		**-11,200,000**	**-61,600,000**
Consumables			
Electricity		5,098,200	28,040,100
Oxygen		2,910,000	16,004,900
Natural gas	GCV	84,400	464,400
Water		30,700	126,000
Chemicals		58,200	320,100
Catalysts		162,600	455,300
Sub-total Consumables		**8,344,100**	**45,410,800**
Disposal costs			
Fe- metal revenue		-80,000	-440,000

Item	Basis	Total Cost, £ pa	
		100,000 tpa	550,000 tpa
Non-Fe metal revenue		-135,900	-747,500
Landfill Waste	Active waste	536,500	2,950,600
Coarse ash disposal + Inerts	Inert	85,700	471,200
Fine ash disposal	Hazardous	345,200	1,898,400
Effluent		124,700	685,600
Sub-total Disposal costs		**876,200**	**4,818,300**
Maintenance	2.5% of capex/year	**4,110,700**	**12,022,000**
Total Opex		**8,697,000**	**17,774,000**

8.4.5 Capital and operating costs estimate with carbon capture

The additional capital and operating costs for carbon capture equipment for FOAK MSW to methane plants are listed in Table 46 .

Table 46: Estimated capital and operating costs for carbon capture plant for FOAK MSW to methane plants

Parameters	100,000 tpa	550,000 tpa
Capital costs	**Cost, £**	
CO_2 rich stream capture cost	1,262,000	7,106,000
CO_2 in flue gas capture cost	24,447,000	68,627,000
Total capture costs	25,709,000	75,733,000
Operating costs	**Cost, £ pa**	
CO_2 rich stream capture	608,000	3,738,000
CO_2 in flue gas capture	1,606,000	5,136,000

8.5 Levelised cost of production

Figure 27 illustrates the levelised cost of methane for all of the plant sizes considered. Our assessment shows that the LCOX for methane from both biomass and waste using AGTs is significantly higher than the cost of production from natural gas (NG), landfill gas (LG) and anaerobic digestion (AD) which are the established sources for methane. Figures 28 and 29 show the impact of integrating carbon capture.

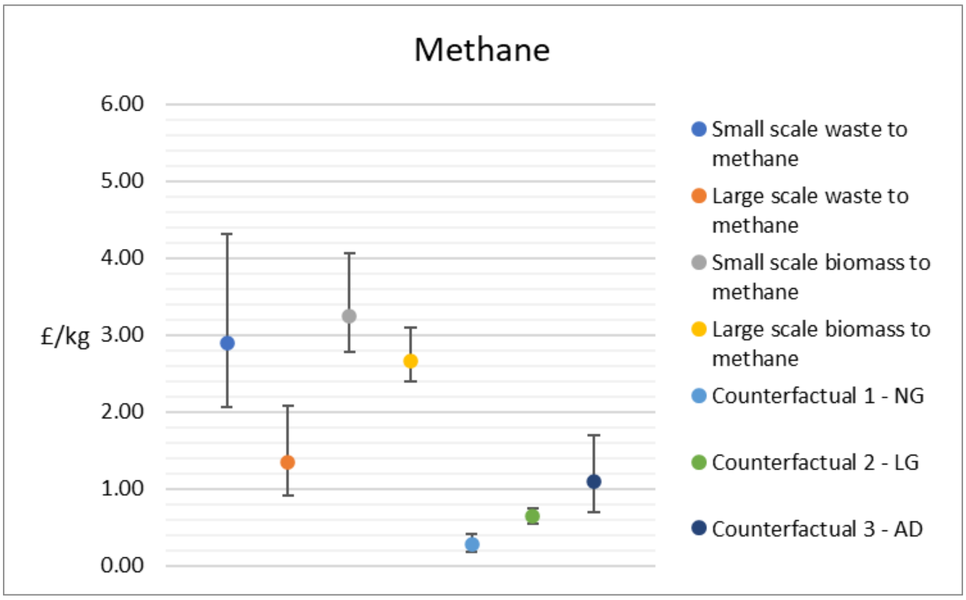

Figure 27: Levelised cost of production of methane from biomass and waste using AGTs

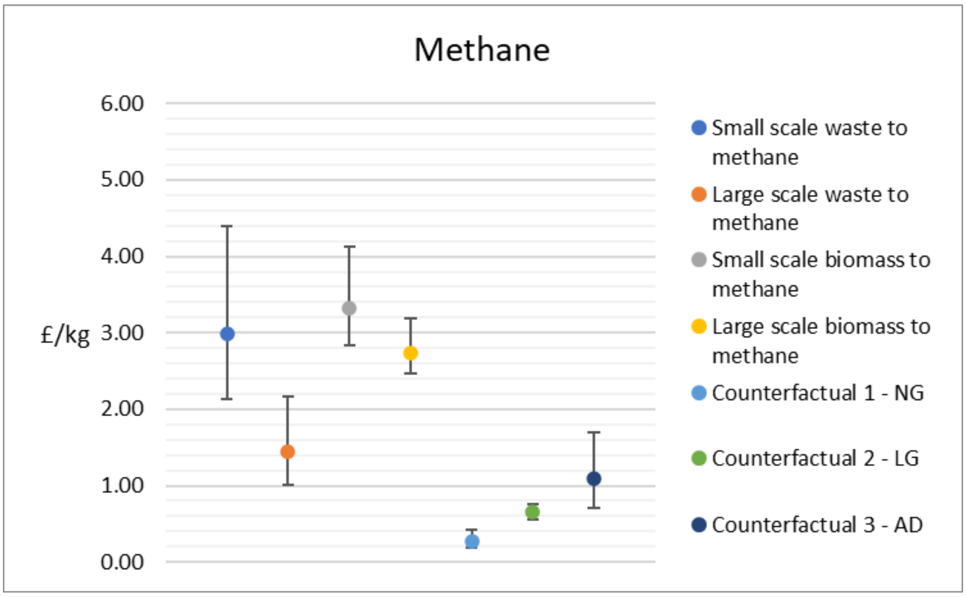

Figure 28: Levelised cost of production of methane from biomass and waste with capture of the rich CO_2 stream

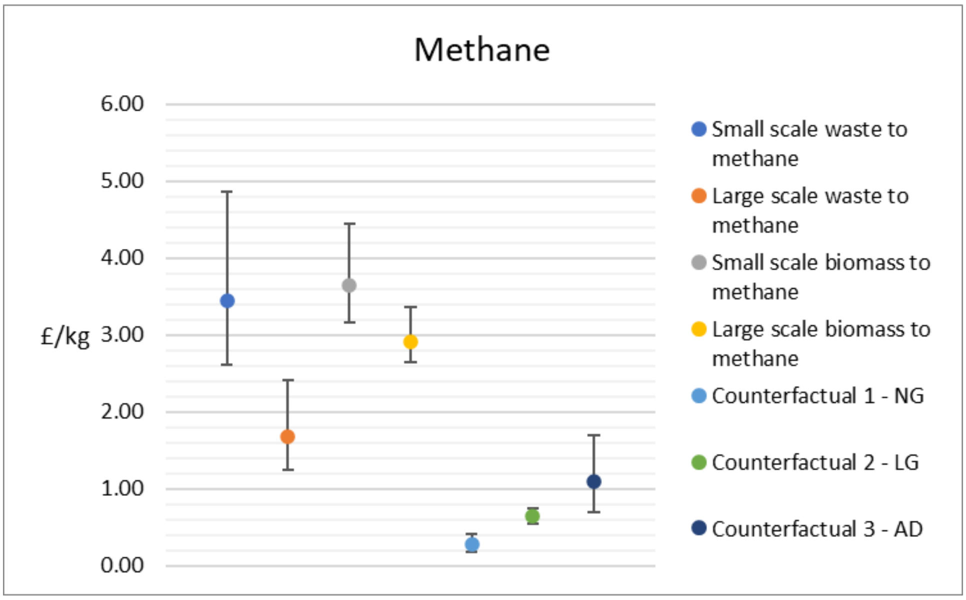

Figure 29: Levelised cost of production of methane from biomass and waste with capture of the CO_2 from the rich and flue gas streams

As with the other products being evaluated in this study the LCOX values derived indicate the uncertainty in plant capex and opex in the absence of FEED studies and long term plant performance and cost details.

The economic impact of the cost of biomass on the levelised cost of production is illustrated by the lower LCOX for the production of methane using large-scale waste when compared to large-scale biomass. Whilst it may appear from the graph that the smalls scale biomass and waste plants are comparable, it must be remembered that the biomass plants are about 3 times larger than the waste plants in both the small and large scale sizes. Therefore, the biomass plants have a significant benefit of scale. As would be expected due to the large differential in feedstock cost, the waste plants have a significantly lower LCOX than the biomass plants.

Evaluation of the split of the cost of production by capex, opex and feedstock cost showing the impact of the cost of feedstock on fuel production is outlined in Table 47.

Table 47: LCOX breakdown for methane

Methane					
Feedstock	Plant size (tpa)	LCOX (£/kg)	LCOX		
			Capex (£/kg)	Opex (£/kg)	Feedstock cost (£/kg)
MSW	100,000	2.90	1.92	2.24	-1.26
	550,000	1.35	0.99	1.58	-1.23
Biomass	330,000	3.26	1.10	1.43	0.73
	1,000,000	2.66	0.62	1.09	0.95
Rich CO_2 stream capture					

Methane					
MSW	100,000	2.99	1.94	2.31	-1.26
	550,000	1.44	1.01	1.66	-1.23
Biomass	330,000	3.33	1.11	1.49	0.73
	1,000,000	2.74	0.63	1.16	0.95
Rich CO_2 and flue gas stream CO_2 capture					
MSW	100,000	3.45	2.22	2.49	-1.26
	550,000	1.68	1.15	1.76	-1.23
Biomass	330,000	3.65	1.30	1.62	0.73
	1,000,000	2.91	0.72	1.24	0.95

8.5 Methane output

Methanation of syngas from the gasification of biomass or waste produces BioSNG which has to be compliant with EN 16723-1:2017[19], the European Standard for injection into the natural gas grid.

8.7 Overall assessment

The production of methane using AGTs has been demonstrated on both biomass and waste at 6 t/h (32 MW_{th}) and 1 t/h (3.5 MW_{th}) respectively for the production of 11,600 tpa methane from biomass and 1,600 tpa from waste. Like the other products from AGTs under evaluation, the key component systems (feedstock processing, gasification, syngas treatment and CO shifting) required for the production of methane have been demonstrated at higher feedstock throughputs. Moreover, methanation plants with production capacities of up to 113,000 tpa of methane from coal are in operation.

Modelling of the process indicates that yields of methane ranging from 168 kg/odt of waste to 188 kg/odt of biomass could be achieved. The modelled yields are based on a single pass system and indicate the lower end of the range of product output. Optimisation of plant performance to include key operations such as process recycles (recovery of unconverted syngas and recirculation within the methanator to maximise the efficiency of conversion of syngas to methane) and a high efficiency of heat integration as would be standard in commercial plants would contribute to an increase in product yields.

Preliminary assessment on the capex for these plants suggests that for the plant sizes considered the capex could range from £300M at the smaller end of the range to £969M for a large scale biomass to methane plant. Similarly, the capex for a waste to methane plant could be on the range £169M to £493M. The gasification, syngas and synthesis system represent approximately 85-90% of the plant costs.

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[19] EN 16723-1:2017 - Natural gas and biomethane for use in transport and biomethane for injection in the natural gas network

The levelised cost of production without carbon capture at the plant throughputs under consideration ranges from £1.35 - £3.26/kg using AGTs and is significantly higher than the cost of production from natural gas, landfill gas (LG) and anaerobic digestion.

9 Methanol

9.1 Overall technical solution

9.1.1 Gasification

The processed feedstock then undergoes gasification in a BFB gasifier at temperatures ranging from 700 – 800°C and at pressures up to 15 bar(a) using a mixture of oxygen and steam as the gasifying agent.

As the syngas leaves the fluidised bed, initial conversion of some of the entrained particulates and tars is carried out in the upper section of the gasifier using a mixture of superheated steam and oxygen. The syngas at temperatures of up to 1,300°C is then discharged to a multisystem syngas treatment system for the removal of a range of contaminants which can poison the catalysts used in the syngas upgrading process and or impact on the quality of the methanol produced.

Residual ash and char from the process are continuously discharged from the gasifier.

To date the maximum design feedstock throughput for commercial scale AGT demonstrator plants for the production of methanol is 115,000 tpa. This plant is of a similar size to the small scale waste to methanol plant evaluated in this study.

9.1.2 Syngas treatment

Treatment of syngas prior to upgrading for the production of methanol is carried out primarily using a range of systems including:

1. high efficiency cyclones for collection of entrained particulates and droplets of hydrocarbons;
2. wet scrubbing for the removal of entrained particulate matter, tars and ammonia;
3. activated carbon beds for adsorption of metals;
4. hydrolysis reactors for the reduction of COS and HCN; and
5. acid gas removal unit for the removal of CO_2, acid gases and tars such as benzene and naphthalene.

As wet scrubbing reduces the syngas temperature, syngas treatment is generally carried out at temperatures ranging from 150 – 250°C and at pressures up to 50 bar(a). As some of the reactions occurring during gas clean up are exothermic the heat produced is recovered to preheat boiler feedwater. Treated effluent from the process is returned to the plant for syngas scrubbing.

Any waste gases are collected from various stages of the process and sent to a thermal oxidiser to oxidise all the hydrocarbons. The flue gas is then cooled using the heat to produce process steam and cleaned to ensure any emissions are below limits set by the Industrial Emissions Directive.

9.1.3 Syngas reforming

Following gas clean up, the treated syngas is then discharged to a multistage CO shift reactor to adjust the H_2:CO ratio from approximately 1 to 2.2 by steam reforming, using the WGS reaction. Generally, as in other syngas reforming processes the syngas is split into two where one stream is delivered to the reformer and the other bypasses the reformer. The stream entering the reformer is mixed with medium pressure steam to increase the ratio of steam:CO and to provide the optimal operating temperature for reaction.

Reforming of the syngas is carried out at temperatures ranging from 200°C (low temperature shift) to 500°C (high temperature shift) and is catalysed by cobalt-molybdenum catalysts in low temperature shift reactors and chromium or copper promoted iron oxide catalysts in high temperature shift reactors. The syngas produced contains a mixture of H_2 and carbon oxides with a stoichiometric number (SN) close to the optimal value of 2.2.

9.1.4 Methanol synthesis

The reformed syngas from the CO shift reactor is then compressed to the required operating pressure before delivery to the methanol synthesis loop. In the synthesis loop the mixture of H_2, CO and CO_2 in the syngas is heated to between 200 - 320°C at 40 – 120 bar(g) in the presence of copper, zinc oxide or alumina-based catalysts to produce methanol.

The two primary processes occurring are illustrated by the equations below.

Equation 1

$$CO + 2H_2 \rightleftarrows CH_3OH$$

Equation 2

$$CO_2 + 3H_2 \rightleftarrows CH_3OH + H_2O$$

The heat produced by the reaction is recovered by heat exchangers in the synthesis loop to produce medium pressure steam. The sensible heat of the hot methanol vapour is also recovered to produce steam and to heat the incoming syngas.

The raw methanol produced is then condensed, washed and any unconverted gases are recycled to the synthesis loop. The methanol is filtered to remove any entrained particulates before discharge to the raw methanol storage tank. The raw methanol is then distilled to remove water, hydrocarbons and any other soluble contaminants. The distilled methanol is then stored on site prior to delivery to the offtaker. Distillation produces methanol with a purity exceeding 98 %.

9.1.5 Process residues handling

Process residues from the gasifier and the syngas treatment system are continuously discharged from the plant and are collected for disposal offsite. Effluent from the wet scrubber systems is treated onsite in a wastewater treatment plant with clean effluent recycled to the scrubber.

Any waste gases are collected from various stages of the process and sent to a thermal oxidiser to oxidise all the hydrocarbons. The flue gas is then cooled using the heat to produce process steam and cleaned to ensure any emissions are below limits set by the Industrial Emissions Directive.

9.2 Plant description

AGT plants for the production of methanol consist of the following key system components:

1. feedstock storage and delivery system;
2. BFB gasifier;
3. a boiler;
4. multistage syngas clean up system;
5. methanol synthesis loop;
6. CO_2 scrubbing and compression;
7. ash handling system;
8. thermal oxidiser for waste gases;
9. heat recovery systems;
10. storage tanks for products, oxygen and nitrogen;
11. auxiliary systems to provide effluent treatment, start-up boiler, cooling, compressed air etc; and
12. stack

The estimated plant area for the four thermal sizes modelled is shown in Table 48.

Table 48: Projected AGT plant area for modelled thermal capacity

	Plant size			
	MSW		Biomass	
	100,000 tpa	550,000 tpa	330,000 tpa	1,000,000 tpa
Plant area (m^2)	25,000	100,000	70,0000	300,000

Source: AGT suppliers and Fichtner Consulting Engineers

9.3 Process model assumptions

A process model was derived to estimate the key process inputs to and outputs from oxy-steam gasification of woodchip and RDF for the production of hydrogen. The model was developed based on a detailed evaluation of process data from several technology providers coupled with our experience in the assessment of a large number of gasification systems and technologies. The model is based on a single pass first-of-a-kind plant in which all of the feedstock delivered to the gasifier is used to produce methanol and does not consider the following:

1. number of process recycles;
2. addition of H_2 to improve methanol yield; and
3. use of any excess energy for power generation.

In addition to the assumptions described in Section 4.3, the key assumptions are as follows:

1. 100% conversion of feedstock to syngas, ash and char;
2. The primary syngas components considered were CO, H_2, CO_2, H_2O, O_2, N_2, CH_4, NH_3, HCl, H_2S;
3. equilibrium gas composition in the gasifier;
4. benzene and naphthalene were used as the model component for tars;
5. the gasifier was assumed to operate at 1200°C at pressures higher than atmospheric;
6. the assumed NCV of methanol is 23 MJ/kg;
7. electricity consumption is based on data provided by suppliers;
8. the main consumables have been derived from the model based on the process requirements; and
9. operation at the maximum feedstock throughput for 85% of the hours in a year (7,446 h)

The results from the model for the key inputs and outputs for the two biomass plant sizes considered are outlined in Table 49 and illustrated in Section 9.4.2.

9.4 Biomass to methanol

9.4.1 Process model

Table 49: Biomass to methanol process model

Model Parameter	Units	Model outputs	
		330,000 tpa	1,000,000 tpa
Process trains	---	1	1[11]
Feedstock throughput	t/h	44	134
Thermal input rate	MW_{th}	100	643
	MW_{th}/year	744,600	4,787,800
Syngas output from gasifier	tpa	351,200	1,886,200
Syngas output from reformer	tpa	250,500	1,411,300
Methanol yield	L/odt	698	710
Methanol production	tpa	91,000	523,200
Electricity consumption	MWe	15	82
	MWh/year	107,100	604,500
Plant availability	%	85	85
Energy efficiency	%	77	69
Main consumables	tpa natural gas	650	4,180
	tpa oxygen	75,800	427,800
	tpa steam	281,200	627,700
	tpa water	214,100	1,209,000
	tpa chemicals	16,500	93,000
Key material output streams	tpa CO_2[12]	156,200	889,500
	tpa flue gas	102,200	502,500
	tpa flue gas CO_2	29,000	140,000
	tpa bottom ash, char and incombustibles	716	5,600
	m^3/year effluent	230,600	1,302,000

9.4.2 Mass balance

The mass balances derived from the process models are illustrated in the figures overleaf.

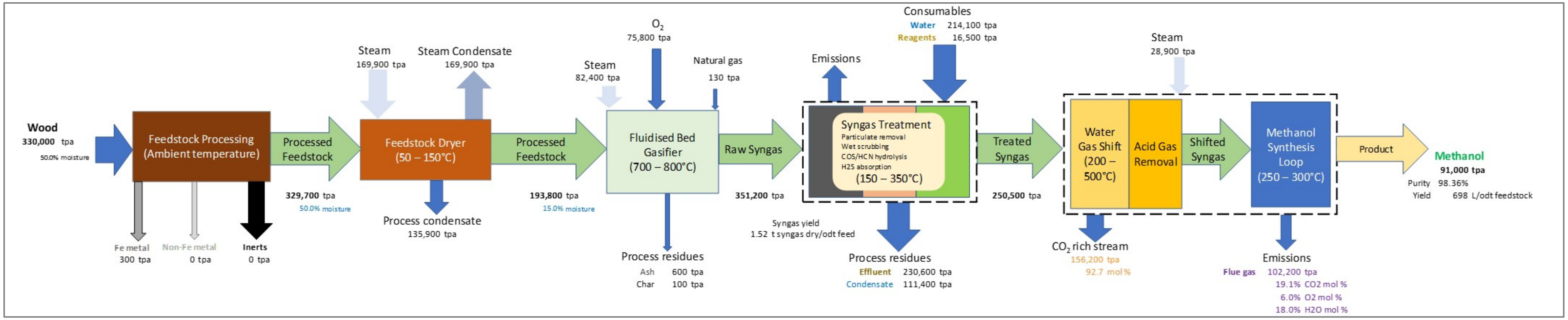

Figure 30: Modelled mass balance for 330,000 tpa AGT biomass to methanol plant

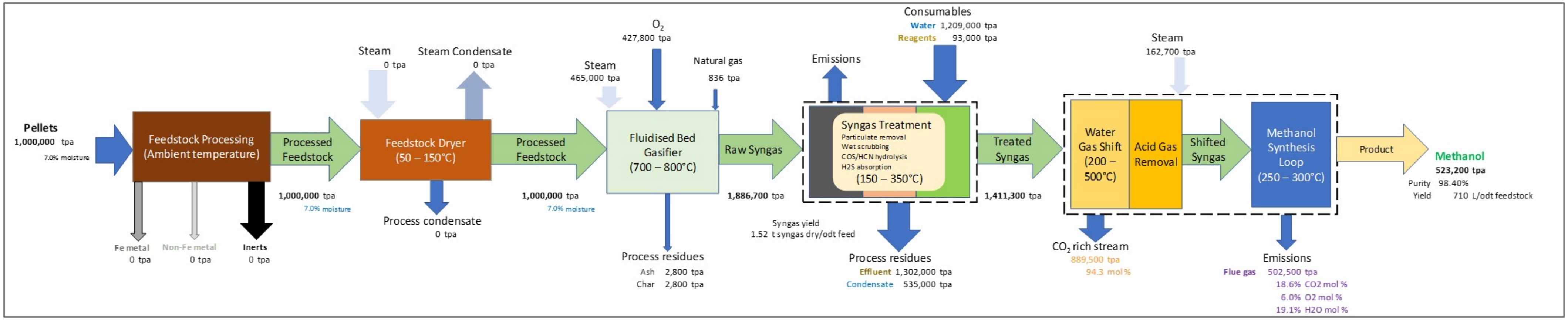

Figure 31: Modelled mass balance for 1,000,000 tpa AGT biomass to methanol plant

9.4.3 Capital cost estimate

The estimated capital costs for FOAK biomass to methanol plants are outlined in Table 50.

Table 50: Estimated capital costs for FOAK biomass to methanol plants

Item	Cost, £	
	330,000 tpa	**1,000,000 tpa**
Project Development Costs		
Land	1,347,500	5,775,000
Consultancy Services	2,775,800	8,855,400
Planning & Other Regulatory	5,703,600	18,482,600
Developers Costs	19,430,500	61,987,500
Start-up Costs	9,715,300	30,993,800
Utility connection	3,800,900	19,296,400
Sub-total development costs	**42,773,600**	**145,390,700**
EPC costs		
Fuel preparation and storage	14,096,400	78,603,900
Gasifier and synthesis plant	263,482,600	806,931,900
EPC cost	**277,579,000**	**885,535,800**
Total Project Costs	**320,353,000**	**1,030,927,000**

The additional costs for installing carbon capture equipment are estimated as £AA (information to be provided in the updated version of the report).

9.4.4 Operating cost estimate

The estimated operating costs for FOAK biomass to methanol plants are outlined in Table 51.

Table 51: Estimated annual operating costs for FOAK biomass to methanol plants

Item	Basis	Total Cost, £ pa	
		330,000 tpa	**1,000,000 tpa**
Fixed Costs			
Labour		3,016,100	4,820,200
Administration & Other Overheads		8,008,800	25,773,200
Sub-total Fixed Costs		**11,024,900**	**30,593,400**
Feedstock	tpa	**20,130,000**	**151,783,200**
Consumables			
Electricity		13,916,400	78,016,000
Oxygen		7,577,400	42,780,000
Natural gas	GCV	233,800	1,500,900
Water		69,400	348,000
Chemicals		164,700	930,000

Item	Basis	Total Cost, £ pa	
		330,000 tpa	1,000,000 tpa
Catalysts		303,600	921,500
Sub-total Consumables		**22,265,300**	**124,496,400**
Disposal costs			
Fe- metal revenue		-11,700	0
Non-Fe metal revenue		0	0
Landfill Waste	Active waste	0	0
Coarse ash disposal + Inerts	Inert	5,800	27,900
Fine ash disposal	Hazardous	23,100	449,600
Effluent		352,800	1,992,100
Sub-total Disposal costs		**370,000**	**2,469,600**
Maintenance	2.5% of capex/year	**8,008,800**	**25,773,200**
Total Opex		**61,799,000**	**335,116,000**

9.4.5 Capital and operating costs estimate with carbon capture

The additional capital and operating costs for carbon capture equipment for FOAK biomass to methanol plants are listed in Table 52.

Table 53: Estimated capital and operating costs for carbon capture plant for FOAK biomass to methanol plants

Parameters	330,000 tpa	1,000,000 tpa
Capital costs	**Cost, £**	
CO_2 rich stream capture cost	2,511,000	14,344,000
CO_2 in flue gas capture cost	33,658,000	91,216,000
Total capture costs	36,169,000	105,560,000
Operating costs	**Cost, £ pa**	
CO_2 rich stream capture	1,431,000	9,716,000
CO_2 in flue gas capture	2,388,000	7,778,000

9.5 Waste to methanol

9.5.1 Process model

Table 54: Waste to methanol process model

Model Parameter	Units	Model outputs	
		100,000 tpa	550, 000 tpa
Process streams	---	1	2
Feedstock throughput	t/h	13	74
Thermal input rate	MW_{th}	36	199

Model Parameter	**Units**	**Model outputs**	
		100,000 tpa	**550, 000 tpa**
	MW_{th}/year	268,100	1,481,800
Syngas output from gasifier	tpa	125,300	689,900
Syngas output from reformer	tpa	77,600	428,000
Methanol yield	L/odt	611	626
Methanol production	tpa	28,200	158,700
Electricity consumption	MWe	5	29
	MWh/year	37,800	208,100
Plant availability	%	85	85
Energy efficiency	%	66	67
Main consumables	tpa natural gas	235	1,295
	tpa oxygen	33,500	184,100
	tpa steam	74,300	409,000
	tpa water	75,700	416,100
	tpa chemicals	5,800	32,000
Key material output streams	tpa CO_2[12]	48,300	269,700
	tpa flue gas	31,600	152,400
	tpa flue gas CO_2	9,000	42,000
	tpa bottom ash and incombustibles	8,600	47,100
	m^3/year effluent	81,500	448,100

9.5.2 Mass balance

The mass balances derived from the process models are illustrated in Figure 32 and Figure 33 overleaf.

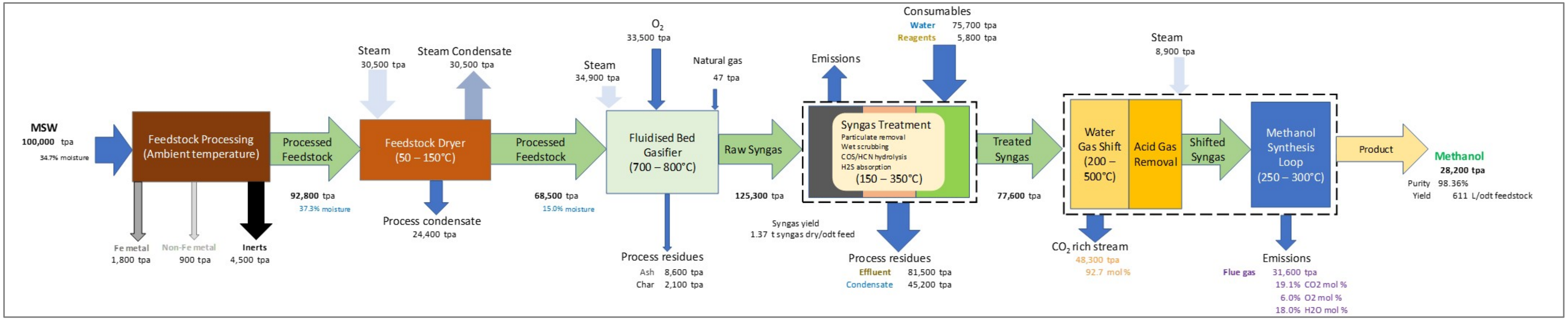

Figure 32: Modelled mass balance for 100,000 tpa AGT MSW to methanol plant

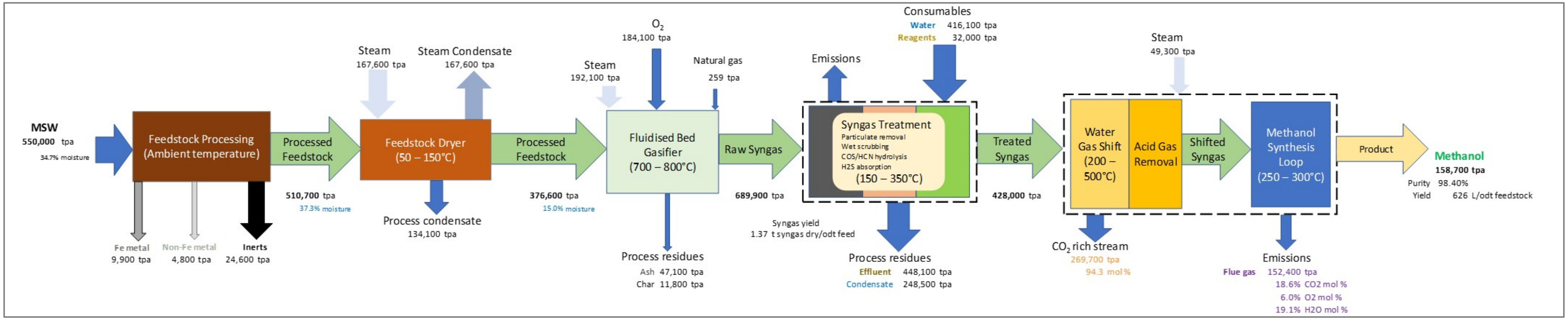

Figure 33: Modelled mass balance for 550,000 tpa AGT MSW to methanol plant

9.5.3 Capital cost estimate

The estimated capital costs for FOAK MSW to methanol plants are outlined in Table 55.

Table 55: Estimated capital costs for FOAK MSW to methanol plants

Item	Cost, £	
	100,000 tpa	550,000 tpa
Project Development Costs		
Land	481,300	1,925,000
Consultancy Services	1,562,700	4,523,200
Planning & Other Regulatory	3,187,700	9,340,000
Developers Costs	10,939,000	31,662,700
Start-up Costs	5,469,500	15,831,400
Utility connection	1,556,200	7,337,500
Sub-total development costs	**23,196,400**	**70,619,800**
EPC costs		
Fuel preparation and storage	13,552,100	53,017,100
Gasifier and synthesis plant	142,719,600	399,907,700
EPC cost	**156,271,700**	**452,924,800**
Total Project Costs	**179,468,000**	**523,545,000**

9.5.4 Operating cost estimate

The estimated operating costs for FOAK MSW to methanol plants are outlined in Table 56.

Table 56: Estimated annual operating costs for FOAK MSW to methanol plants

Item	Basis	Total Cost, £ pa	
		100,000 tpa	550,000 tpa
Fixed Costs			
Labour		2,455,600	5,100,900
Administration & Other Overheads		4,486,700	13,073,600
Sub-total Fixed Costs		**6,942,300**	**18,174,500**
Feedstock		**-11,200,000**	**-61,600,000**
Consumables			
Electricity		5,098,200	28,040,100
Oxygen		3,346,500	18,405,600
Natural gas	GCV	84,400	464,400
Water		30,700	126,000
Chemicals		58,200	320,100
Catalysts		162,600	455,300
Sub-total Consumables		**8,780,600**	**47,811,500**

Item	Basis	Total Cost, £ pa	
		100,000 tpa	550,000 tpa
Disposal costs			
Fe-Metal revenue		-80,000	-440,000
Non-Fe Metal revenue		-135,900	-747,500
Landfill Waste	Active waste	536,500	2,950,600
Coarse ash disposal + Inerts	Inert	85,700	471,200
Fine ash disposal	Hazardous	345,200	1,898,400
Effluent		124,700	685,600
Sub-total Disposal costs		**876,200**	**4,818,300**
Maintenance	2.5% of capex/year	**4,486,700**	**13,073,600**
Total Opex		**9,886,000**	**22,278,000**

9.5.5 Capital and operating costs estimate with carbon capture

The additional capital and operating costs for carbon capture equipment for FOAK MSW to methanol plants are listed in Table 57.

Table 58: Estimated capital and operating costs for carbon capture plant for FOAK MSW to methanol plants

Parameters	100,000 tpa	550,000 tpa
Capital costs	**Cost, £**	
CO_2 rich stream capture cost	1,129,000	6,385,000
CO_2 in flue gas capture cost	16,656,000	44,582,000
Total capture costs	17,785,000	50,967,000
Operating costs	**Cost, £ pa**	
CO_2 rich stream capture	537,000	3,230,000
CO_2 in flue gas capture	1,127,000	3,387,000

9.6 Levelised cost of production

Assessment of the capex, opex and feedstock costs for all of the plant sizes evaluated shows that the levelised cost of production of methanol from biomass and waste using AGTs could be as much as three times higher than the cost of production of methanol from SMR which is the primary source of production globally. Whilst it may appear from the graph that the small scale biomass and waste plants are comparable, it must be remembered that the biomass plants are about three times larger than the waste plants in both the small and large scale sizes. Therefore, the biomass plants have a significant benefit of scale.

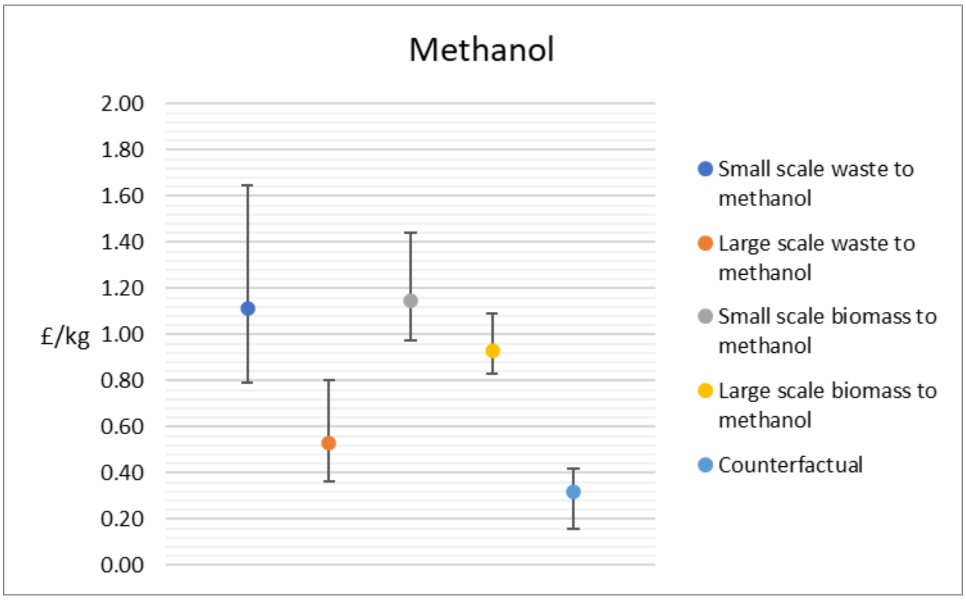

Figure 34: Levelised cost of production of methanol from biomass and waste using AGTs

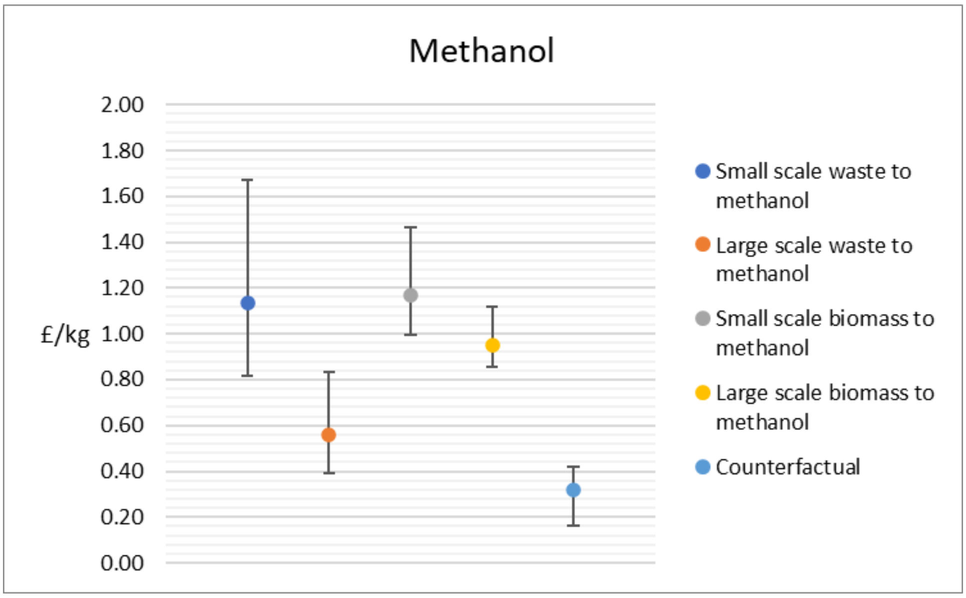

Figure 35: Levelised cost of production of methanol from biomass and waste with capture of the rich CO_2 stream

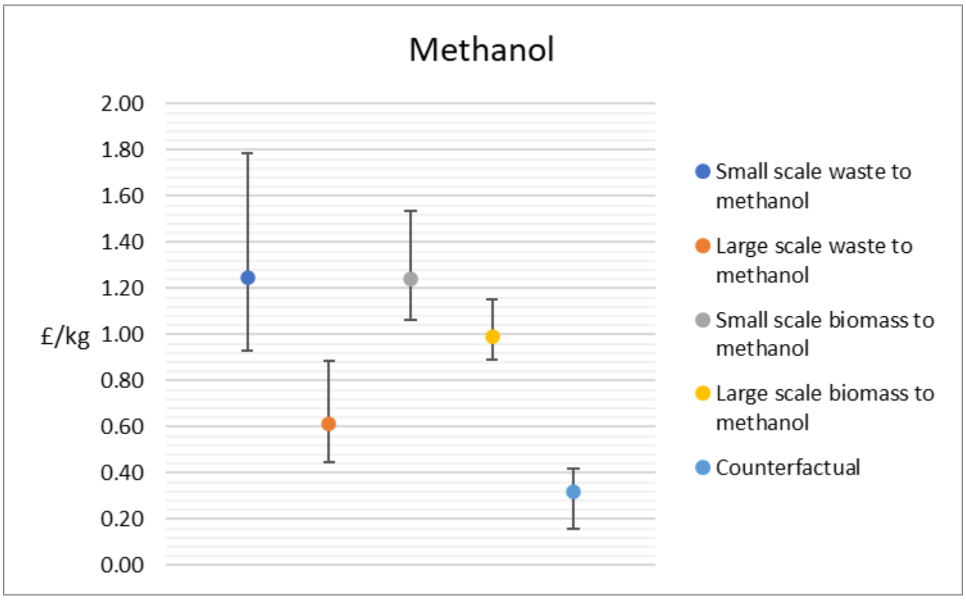

Figure 36: Levelised cost of production of methanol from biomass and waste with capture of the CO_2 from the rich and flue gas streams

As would be expected due to the large differential in feedstock cost, the waste plants have a significantly lower LCOX than the biomass plants.

As with the other products being evaluated in this study the LCOX values derived indicate the uncertainty in plant capex and opex as commercial plants using AGT for the production of methanol have not yet been built. The high level of variability in the predicted LCOX for methanol from several feedstocks at varying levels of production is clearly illustrated in Figure 34. The LCOX values derived here for large scale waste and biomass are illustrated by the yellow, red and green triangles. The high cost of wood pellets in the UK is the primary contributor to the very high LCOX for the large scale AGT biomass when compared to the production cost of wood shown in Figure 37.

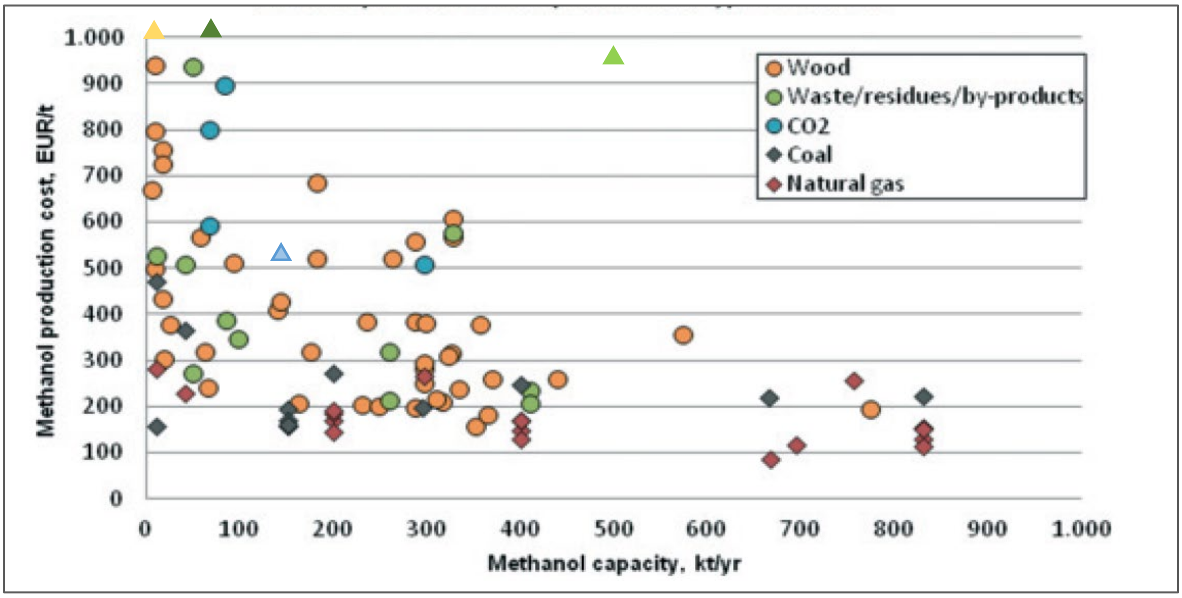

Figure 37: Production costs at varying capacities for methanol from several feedstocks

Source: IEA-ETSAP and IRENA Technology Brief 108 – 2013

Table 59: LCOX breakdown for methanol

Methanol					
Feedstock	Plant size (tpa)	LCOX (£/kg)	LCOX		
			Capex (£/kg)	Opex (£/kg)	Feedstock cost (£/kg)
MSW	100,000	1.11	0.73	0.82	-0.44
	550,000	0.53	0.38	0.58	-0.43
Biomass	330,000	1.15	0.40	0.50	0.24
	1,000,000	0.93	0.22	0.39	0.32
	Rich CO_2 stream capture				
MSW	100,000	1.14	0.73	0.84	-0.44
	550,000	0.56	0.38	0.60	-0.43
Biomass	330,000	1.17	0.40	0.52	0.24
	1,000,000	0.95	0.23	0.41	0.32
	Rich CO_2 and flue gas stream CO_2 capture				
MSW	100,000	1.25	0.80	0.89	-0.44
	550,000	0.61	0.41	0.63	-0.43
Biomass	330,000	1.24	0.45	0.55	0.24
	1,000,000	0.99	0.25	0.42	0.32

As shown in Table 59 and for the other products from AGTs, although the capex and opex of the small-scale biomass plant benefits from economies of scale over the small scale waste plant, the cost of biomass will be an important contributor to the cost of production.

9.7 Methanol output

The outputs from the technical model and our review of the process plant data to date from large scale demonstrators show that methanol produced from the gasification of biomass and wastes followed by syngas treatment and upgrading can produce methanol with a purity of 98 w/w% or greater. Moreover, provided that a multisystem approach to syngas treatment is part of the plant process configuration, contaminant concentrations can be reduced to the levels such that the methanol produced is compliant with the International Methanol Producers and Consumers Association (IMPCA) Methanol Reference Specifications (2015).

9.8 Overall assessment

The production of methanol using AGTs has been demonstrated to date on waste at a maximum throughput of 110,000 tpa This is also the highest feedstock throughput rate for a large scale AGT demonstrator.

Modelling of the process indicates that yields of methanol ranging from 611 l/odt of waste to 710 l/odt of biomass could be achieved. The modelled yields are based on a single pass system and indicate the lower end of the range of product output. Suppliers have also indicated that initial operations show that optimisation of plant performance by increasing the number of process recycles (recovery of unconverted syngas and recirculation within the methanol synthesis loop to maximise the efficiency of conversion of syngas to methanol) and supply of additional hydrogen has resulted in increased product yields. Yields of up to 1200 l/odt of waste have been predicted.

Preliminary assessment on the capex for these plants suggests that for the plant sizes considered the capex could range from £328M at the smaller end of the range to £1.1 billion for a large scale biomass to methanol plant. Similarly, the capex for a waste to methanol plant could be on the range £184M to £536M. The gasification, syngas and synthesis system represent approximately 85-90% of the plant costs.

The levelised cost of production without carbon capture at the plant sizes evaluated ranges from £0.53 - £1.15/kg using AGTs and at the high end of the range it is more than three times higher than the current cost of production using the established SMR of natural gas technology.

10 Second generation AGT plants

10.1 Introduction

The plants described in the previous sections are considered to be first generation plants aimed at demonstrating the fundamentals of the process. The plants have been simplified to reduce complexity, numbers of interacting components and to ensure the best possible chance of proving that it is possible to produce the required product reliably.

In doing so, it is recognised that the projects will not operate as economically as if they were fully optimised. Future plants will need to operate in a commercial world and second generation plants will have to concentrate on reducing the costs of production. This can be achieved in a number of ways:

1. reduction of capital costs;
2. increase of yields;
3. reduction of consumable costs; and
4. reduction of disposal costs.

In addition to reducing costs, much focus will be placed on improving the process to reduce carbon emissions.

10.2 Technology

Whilst BFB gasifiers are the primary type of gasifier being developed for the production of biofuels from biomass and wastes, entrained flow gasifiers which are currently the largest gasifiers in commercial operation are also being considered. Entrained flow systems are quite common for large scale coal gasifiers with the largest entrained flow coal gasifier having a thermal capacity of 1,000 MW.

Although several initial studies have indicated that similar or higher efficiencies of conversion of biomass and waste can be achieved with entrained flow gasifiers, the chemical composition of these feedstocks varies considerably. In addition, these feedstocks exhibit lower NCVs and ash fusion temperatures when compared with coals. Consequently, significant innovation is required before entrained flow gasification at a similar scale to coal conversion plants becomes technically and commercially feasible.

Currently the majority of development is based on using fluidised beds, which is why this has been the focus of this report. It is possible that in the future technology will switch to entrained flow in the long term, but even if this is the case, the majority of the process steps will be very similar to those discussed in this report.

10.3 Capital costs

Comparisons are frequently made in renewable terms with the costs of solar and wind power generation. The cost of this has fallen dramatically over recent years. However, it is important to realise that there is no direct comparison between solar/wind generation and process plants like AGTs. The large savings in capital costs for solar and wind farms have been made because the solar panels and wind turbines are modular, with many units making up large facilities. The individual units can be made in factories where mass production processes have brought capital costs down rapidly. By comparison, AGTs tend to be built as large, independent pieces of equipment which are fabricated for specific projects. Whilst common designs and fabrication techniques can bring costs down, this is not significant when compared with mass production. AGTs are comparable with conventional biomass or energy from waste plants in design and scale. Historically the cost of these plants has risen rather than fallen due to reduced competition and increased plant costs as EPC contractors have failed and left the market.

Therefore, whilst we consider there will be some potential cost savings from first generation designs if improvements can be made, designs made modular and some redundancy avoided, this will be small and may be outweighed by other modifications needed to improve performance or reliability.

However, scale is likely to have a significant impact on capital costs. Even for modular plant, building larger plants will usually mean a significant reduction in the capital cost per MW or tonne. This can be seen in the comparisons between the smaller and larger plants proposed.

Another potential saving could be to adopt different technical solutions allowing larger plants to be built. Whilst modularity can bring cost savings due to shared designs, building single units larger rather than providing smaller modular units normally reduces overall costs. One large plant is almost always cheaper than two units with the same overall capacity. As fluidised beds can be scaled up relatively easily and there are very large fluidised beds in operation, it is possible that second generation plants will become larger, with larger units to reduce capital costs. Any potential for capital costs reduction using large scale entrained flow gasification will be dependent on the process changes required to ensure a high efficiency of conversion of biomass and waste to syngas.

10.4 Yield

Whilst lower capital costs per MW generated have reduced the cost of renewables generation from solar or wind, we foresee the most effective way to reduce production costs for products from AGTs is to increase their yield. This has an additional benefit that the carbon emissions will reduce as more of the carbon in the feedstock is turned into product.

Some of this will be an inevitable progression. Once first generation plants have been built and the performance monitored, this will allow modifications to be made to boost yield. Once the process is proven, it will be worthwhile to increase its complexity, for example by increasing the amount of recycle loops or by targeting losses in the system and improving the performance of individual plant units. New catalysts can be developed and tested to improve the reaction efficiencies and to optimise the processes.

In addition, advanced techniques will be used to improve yields. One example of this is the approach of Enerkem. It built a first generation demonstration plant in Edmonton, Canada, to demonstrate the production of methanol and ethanol from RDF. It is now planning to build plants in Rotterdam and Varennes, Quebec, both of which will use hydrogen to maximise the production of methanol. If methanol is produced from waste alone, the balance of carbon to hydrogen in the feedstock means that once syngas is made, it is beneficial to then use WGS reaction to produce more hydrogen by reacting some of the CO. This leads to a yield of about 610 l/odt of RDF. However, if imported hydrogen is available, all the CO in the syngas can be converted to methanol, increasing the yield to closer to 1,200 l/odt. This also reduces the amount of CO_2 generated. At Varennes, the intention is to use green electricity from Canada's hydro-electric system to generate hydrogen for the process by electrolysis.

Therefore, the expectation should be that advanced techniques and new process development will be able to increase yields by 10% from basic process improvements to as much as 100%, for example by the addition of hydrogen.

10.5 Consumable costs

It can be seen from the economic assessments that the AGT processes have high consumable costs, chiefly from the use of imported power and oxygen. These also contribute to the carbon footprint due to the need to generate the power or oxygen, normally using some fossil fuels.

Our assumption, which matches most of the plants in development, is that most FOAK plants will focus on the product and not the generation of power. Most technology suppliers recover large amounts of waste heat in the process and use this to produce process steam. Our models, matched by those of most suppliers, show that there is plenty of excess heat to produce the steam required. The remainder of the waste heat is wasted usually requiring cooling systems. In future plants we foresee that much more attention will be taken to recover as much available heat as possible and to use this to generate some or all of the electricity requirements for the plant. Some of the processes will have an excess of waste energy, allowing some electricity to be exported and to generate the oxygen requirements for the process. Other processes which optimise the yield, such as Enerkem using imported hydrogen, will have less available waste heat.

Therefore, using the waste heat to supply all or some of the plant's requirements for electricity and oxygen will significantly reduce the operating costs of the plant and also reduce carbon emissions further.

10.6 Disposal costs

Operating costs can be further reduced by minimising disposal costs. There are two principal sources of these, the solid ash streams from the process and effluent.

As the gasification process is made more efficient, the solid material will contain less carbon and is likely to be classified as inert material. Such material can be recycled as secondary aggregate. Therefore, disposal costs for the solid material can be reduced by minimising or eliminating the production of hazardous solid residues from the gasifier and by reducing disposal costs for inert material by finding secondary aggregate consumers.

Effluent costs from gasification processes can be very significant if these contain tars and other hazardous material which require expensive refinement. For the purpose of assessing first generation plants we have assumed that the effluent streams will undergo a basic waste water treatment process on site to reach acceptable discharge levels, but the aqueous streams will require secondary treatment off-site. As more data becomes available from operating AGTs, the gasification process will be refined to make it more efficient and to reduce the amount of tars and long-chain hydrocarbons produced. As well as improving the overall efficiency of the process, this will reduce the amount of effluent treatment required. In the long term it is likely that better on-site effluent treatment systems will be used to minimise the cost of off-site disposal.

10.7 Carbon emissions

As the main reason to promote AGTs is to contribute to reducing carbon emissions, particularly for sectors hardest to affect such as transport and heat, much focus will go into improving second generation plants to further reduce carbon emissions. As mentioned above, improving yields and generating power and oxygen from waste heat will contribute strongly to this.

In addition, more focus is likely to be placed on reducing the carbon footprint of the feedstocks. Whilst biomass is considered to be 100% biogenic, there are carbon costs in growing, treating and transporting the biomass. In particular, importing biomass from abroad in pellet form means there is a significant cost in producing the pellets and also to achieve the sanitising requirements needed to import biomass. In the longer term, more biomass may be sourced from the UK, reducing the carbon cost of pelletising and transport.

By comparison, residual MSW is considered to be about 50% biogenic by energy, with the non-biogenic material such as plastics sourced from fossil fuels. With time it is likely that the amount of plastic in residual waste will reduce, either because we use less plastics or because plastic recycling expands.

In this report we have estimated where the carbon in the feedstock ends up. This is primarily in four streams:

1. a very small percentage remains in the solid residues streams as carbon;

2. some of the carbon is retained in the product in varying quantities dependent on the product and the yield;
3. some of the carbon is separated in the syngas conversion process as a CO_2 rich stream which can be pressurised and used or captured as required; and
4. the remaining carbon is emitted as CO_2 in the flue gas.

Depending on the level of carbon capture required and whether solutions are developed to use or store any CO_2 which is captured, the CO_2 rich stream is readily available with compression and possibly further clean-up, once pressure and quality requirements are set. The CO_2 in the flue gas can be captured using conventional carbon capture methods such as amine-based solvents if high carbon capture levels are required. In principle, a second generation AGT plant with carbon capture fitted to the flue gas stream and using biomass, provides a strongly carbon negative process as the feedstock used extracts CO_2 from the atmosphere and virtually all the carbon in the feedstock is then captured for use or storage.