Programme Area: Bioenergy

Project: TEA Biomass Pre-processing

Title: Review and Benchmarking Report

Abstract:

This report assesses the current techno-economic status of biomass pre-processing, combustion and gasification technologies used to transform forestry and perennial energy crop feedstocks into heat, power and syngas. The report uses a common review framework to assess all technologies, including technology descriptions, status, impacts of feedstock choice and options for improvement. Shorter descriptions are also provided for a further set of typically less mature conversion and pre-processing technology options, as part of a horizon scanning exercise.

This report enables the reader to understand which technologies are nearest to/furthest from being commercialised, which are cheapest on a stand-alone basis, and the impact that different feedstock characteristics have on the operation and costs of the technologies. The report presents the techno-economic data gathered on these technologies for use in the modelling of different bioenergy supply chains later in the TEAB project, highlighting any data gaps and where uncertainty ranges are highest.

Context:

The techno-economic project will provide a greater understanding of the options available to modify or improve the physical and chemical characteristics of different types of UK-derived 2nd generation energy biomass feedstocks, that may otherwise reduce the cost-effective performance of conversion technologies.

BI2011

Techno-Economic Assessment of Biomass Pre-Processing (TEABPP)

Deliverable 1

Review and benchmarking report

Version 5.0

The TEABPP Consortium

For the Energy Technologies Institute

4th July 2016

Title	Deliverable 1: Review and benchmarking report	
Client	Energy Technologies Institute	
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Contents

1 Executive Summary

The **Techno-Economic Assessment of Biomass Pre-Processing** (TEABPP) project aims to compare the costs, performance and emissions of biomass supply chain configurations with and without pre-processing, and with and without conversion plant improvements. The primary objective of the Project is to establish optimal system designs for different scales, feedstock types and end uses, highlighting areas of the supply chain with greatest potential for improvement. This will develop the ETI's understanding of those pre-processing activities that do or do not benefit the overall levelised cost of energy of the supply chain.

There are two objectives of Work Package 1 (WP1) of the project: firstly, to carry out a review of the pre-processing, combustion and gasification conversion technologies used to transform forestry and perennial energy crop feedstocks into heat, power and syngas; and secondly, to gather techno-economic data of sufficient quality to be used in the initial chain cost results and down-selection process (in WP2) and for population of the gPROMS databases for the process modelling (in WP3).

This document reports on the review and data gathering activity in WP1. It provides the groundwork for the subsequent model development, by identifying key factors, metrics and issues for parameterisation. This will enable the ETI to understand which technologies are nearest to/furthest from being commercialised, which are cheapest on a stand-alone basis, and the impact that different feedstock characteristics have on the operation and costs of the technologies. The report also highlights any data gaps and where uncertainty ranges are highest.

The report uses a common review framework to assess all technologies, including technology descriptions, status, impacts of feedstock choice and options for improvement. Shorter descriptions are also provided for a further set of typically less mature conversion and pre-processing technology options, as part of a horizon scanning exercise. The report concludes with a benchmarking analysis, which compares the technology status, as well as the levelised costs of producing hot water, electricity, combined heat & power (and syngas, if applicable) from each conversion technology and the levelised costs for each pre-processing technology.

The development status of each technology was assessed according to its Technology Readiness Level (TRL). Figure 1 shows that all the biomass combustion technologies assessed are fully commercial (TRL 9), along with updraft and downdraft gasification. Gasifier designs that are suitable for larger plants are either at the demonstration or first commercial plant stage (TRL 6-8). Downstream technologies to use the syngas produced are commercial (TRL 8-9), having been adapted from natural gas applications.

Biomass pre-processing technologies vary significantly in their development status (Figure 2), with many already commercialised, particularly those involving physical resizing, drying and densification. A number of thermochemical pre-processing technologies sit at TRL 8, with first commercial plants recently built. Pre-processing technologies at earlier stages of development could provide opportunities for cost reduction across the value chain - this will be assessed further through the chain calculations in WP2 and detailed process modelling in WP3.

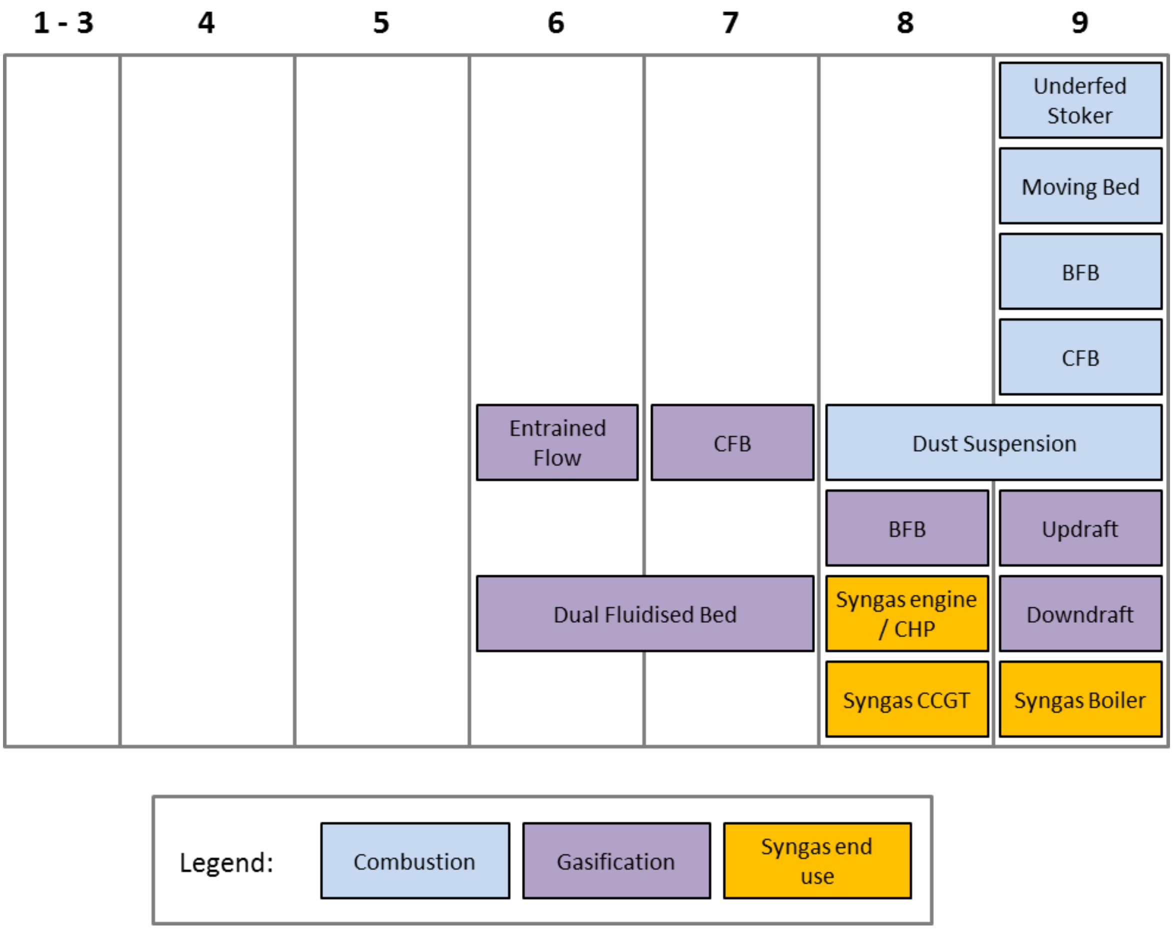

Figure 1: Current TRLs of combustion, gasification and syngas conversion technologies from Sections 3 and 4

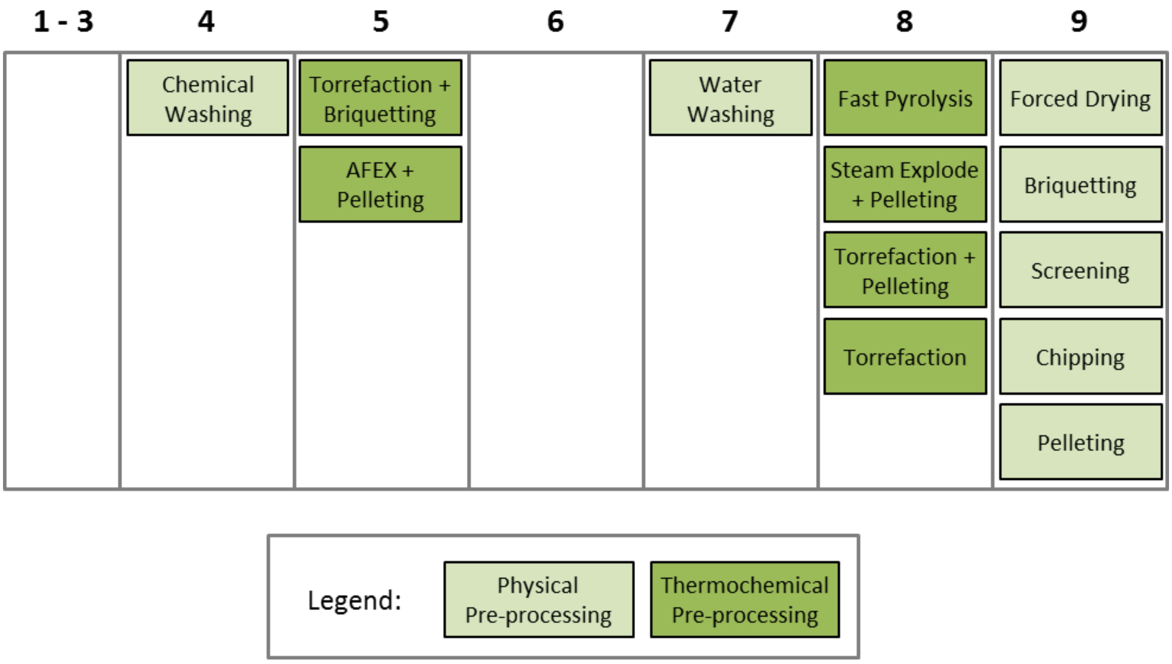

Figure 2: Current TRLs of pre-processing technologies from Section 5

Figure 3 shows that the majority of technologies covered in the horizon scanning exercise are at a much earlier stage of development (TRL 4-5) – these also lack data as to their likely costs at commercial scale. The grouping of syngas clean-up technologies, pyrolysis, ORC and Stirling engines is however closer to commercialisation (TRL 6-9), but publically available data is very limited, given only one or two developers are working on each technology. The horizon scanning technologies are therefore written up to give a “watching brief” on potential technologies for inclusion in future modelling exercises and if more data were to become publically available.

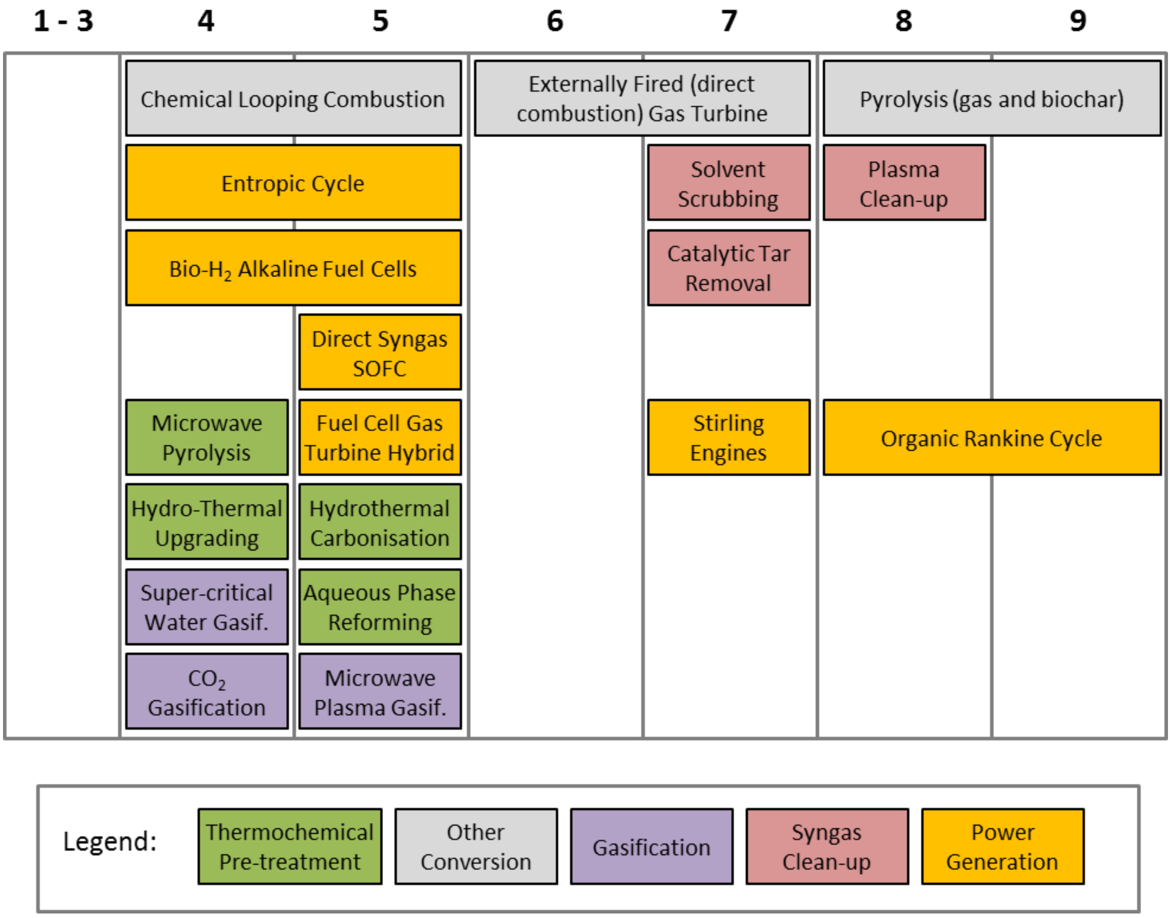

Figure 3: Current TRLs of Horizon Scanning technologies from Section 6

Table 1 below sets out for each technology the current TRL, the expected number of years to reach mass commercial deployment (TRL 9) and the current range of expected plant scales that would be commercially viable/technical feasible (were the technology commercialised). Justifications for TRL progressions can be found in the relevant technology chapters, which involved gauging the level of interest in each technology based on the number of developers, industrial scale-up plans, applications or feedstocks in/out of scope (such as biofuels or wastes), and the opportunity for faster development via cross-over from other more mature technologies that share similarities in components or plant design. For each conversion technology, Table 1 provides a summary of the end use applications analysed, and for the pre-processing technologies, a very high-level summary of some of the key benefits of the intermediate product. Table 1 also summarises the degree of confidence in the techno-economic data gathered in WP1.

Table 1: Summary of conversion and pre-processing technology TRL development, commercial scales (given in MW input), applications/benefits and data quality

Technology	Current TRL	Years to TRL 9	Commercial MW_{in} scale	Applications/ Benefits	Data quality
Conversion technologies – combustion					
Underfeed stoker	9	-	0.2-2.9	Heat	High
Moving bed	9	-	0.5-190	Heat, CHP, power	High
Bubbling Fluidised bed	9	-	28-417	CHP, power	High
Circulating Fluidised Bed	9	-	193-1,300	Power	High
Dust Suspension	8-9	<5	602-1,740	Power	Medium
Conversion technologies – gasification					
Updraft	9	-	1-15	Syngas	High
Downdraft	9	-	0.05-2	Syngas	High
Bubbling Fluidised Bed	8	10	2-20	Syngas	High
Circulating Fluidised Bed	7	10-20	15-100	Syngas	Medium
Dual Fluidised Bed	6-7	15-20	15-100	Syngas	Medium
Entrained Flow	6	15-20	100-2,000	Syngas	Medium
Syngas Boiler	9	-	0.05-52	Heat	High
Syngas Engine/CHP	8	<5	0.4-50	Power, CHP	High
Syngas CCGT	8	5-10	50-800	Power	High
Pre-processing technologies					
Water Washing	7	10	1-182	↓ alkali, ash	Low
Chemical Washing	4	~ 20+	1-182	↓ alkali, ash	Low
Belt Drying	9	-	3-114	↓ moisture	High
Drum Drying	9	-	3-68	↓ moisture	High
Briquetting	9	-	1-34	↑ density	High
Screening	9	-	34-159	↓ rejects	High
Chipping	9	-	1-150	↓ size	High
Pelleting	9	-	2-387	↑ density	High
Pyrolysis	8	10	4-205	→ liquid	Medium
Torrefaction	8	5	33-106	↑ LHV	High
Torrefaction **+** pelleting	8	5	33-106	↑ density, LHV	High
Torrefaction + briquetting	5	10	33-106	↑ density, LHV	High
Steam Explosion + pelleting	8	5	16-301	↑ density	High
AFEX **+** pelleting	5	10-15	4-19	↑ density	Medium

A similar table can be produced for the horizon scanning technologies, as shown in Table 2, but focusing on only the TRL development, as there is very little information available on the potential commercial scales of these technologies. Note that in both Table 1 and Table 2, the symbol "~" indicates a more approximate estimate has been made, due to a greater than average level of uncertainty regarding the specific technology's TRL progression, in general due to a lack of developers.

Those technologies at TRL 5 or below may take up to 20 years or more to reach TRL 9, except in those cases where there is a combination of two more mature technologies (e.g. torrefaction + briquetting), or the technology is relying on uptake from other sectors (e.g. use of AFEX in biofuels and biochemicals pre-treatment).

Table 2: TRL progression assumptions for the Horizon scanning technologies

Technology	Current TRL	Years to TRL 9	Rationale
Lower TRL			
Microwave Plasma Gasification	5	15	Two developers. Could piggy-back off standard plasma gasification once microwave resonator proven
Super-critical Water Gasification	4	~ 25+	No developers. Several academic groups still interested. Wet feedstock
CO_2 Gasification	4	~ 25+	No developers. But recent academic research, at only one group. Focus may be on char gasification, not biomass
Chemical Looping **Combustion**	4	25	No developers, but several academic groups, and uses standard CFB equipment. Depends on coal CCS development and competing tech
Chemical Looping **Gasification**	3	~ 35	No developers and little academic activity
Entropic Cycle	4-5	~ 20+	No developers and little sign of academic activity. May not be developed
Direct Syngas SOFCs	5	15-20	One developer, few academics, but US DOE and industry is driving a program of work in coal
Bio-H_2 Alkaline Fuel Cells	4-5	20	One main developer looking at wastes/biomass angle. WGS presents additional challenge. But small modules lends themselves to roll-out
Fuel cell Gas Turbine Hybrid	5	15-20	One developer, few academics, but US DOE and industry is driving a program of work in coal
Aqueous Phase Reforming	5	Not applicable	One developer, major industry presence. Liquid sugars at present, unlikely to apply to heat and power. For biofuels/chemicals only
Hydro-Thermal Upgrading	4	~ 30+	No developer, industry pulled out long time ago. Wet biomass. May not be developed, too similar to pyrolysis
Microwave Pyrolysis	4	25	Multiple developers for wastes, but only academics for biomass. Not likely to piggy-back, but substitute conventional fast pyrolysis
Hydrothermal Carbonisation	5	20-25	Two developers, a few academics, but little info. Wet feedstock
Higher TRL			
Plasma Clean-up	8	<5	Three developers. Extensive deployment of clean-up torches already, although very waste focused
Solvent Scrubbing	7	5-10	One academic developer, but strong industrial partner with demos, and multiple plans at scale
Catalytic Tar Removal	7	5-10	Two developers and strong industrial partner. Very waste focused. One planned plant
Pyrolysis (gas and biochar)	8-9	5+	Significant history. Industry currently not engaged in developing new projects, though there is interest related to biochar soil developers
Externally Fired Gas Turbine	6-7	10	Two developers. Standard components being used and variety of relevant feedstocks
Stirling Engine	7	5-10	Multiple developers and examples, including some on biomass. CHP gaining attention
Organic Rankine Cycle	8-9	<5	Several developers and examples, including on biomass

Figure 4, Figure 5, Figure 6, Figure 7 and Figure 8 show the levelised cost of energy (LCOE) for the main output of each of the technologies assessed (in the given intermediate or end use applications). These costs all assume a common traded biomass feedstock price of £14.7/MWh (NB: this value will be higher and vary in WP2 and WP3 to reflect actual Miscanthus, SRC willow and SRF production costs) and a 10% annual discount rate across their technical lifetimes, with a co-product heat price of £32.3/MWh$_{th}$ assumed for CHP applications.

Note that these plants are at a variety of scales, due to the different range of applicable scales for each technology – for full details of the Base Case scales used, please see Section 7.2. Appropriate syngas clean-up costs have also been incorporated so that the clean syngas costs shown in Figure 8 are suitable for use in a gas turbine. These LCOE values cannot simply be combined from the charts (e.g. pelleting + BFB combustion), since this LCOE benchmarking in WP1 is only for stand-alone plants, with no consideration for the rest of the value chain. To calculate the LCOE for full chains, including multiplicative efficiency losses, plus storage and logistics costs, the Excel tool in Deliverable 2 is required.

Figure 4 shows that those pre-processing technologies that add the fewest costs to the underlying cost of the biomass are the simplest (chipping and screening); whereas those with the highest LCOE either have relatively low efficiency (e.g. pyrolysis), high capital costs or a relatively low availability for a new technology based on seasonal feedstock (e.g. AFEX + pelleting). For information on the steps included in each of the technologies shown in the following graphs, see Table 3. Note that several of the technologies duplicate components, e.g. drying is included in multiple technologies, so these benchmarking costs cannot simply be added together.

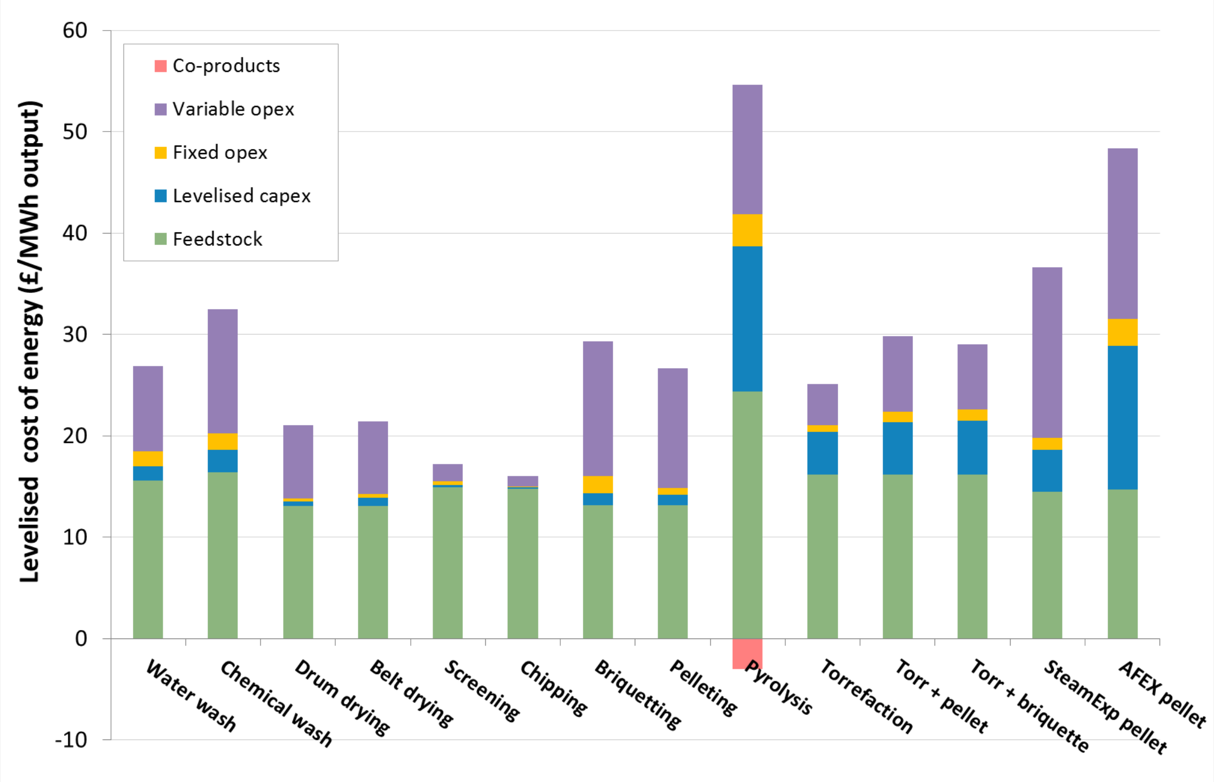

Figure 4: LCOE benchmarking for the TEABPP pre-processing technologies

Scale plays a large role in determining the LCOE of the heating technologies, as shown in Figure 5. As a result, underfed stokers have the highest levelised Capex. Downdraft and updraft gasifiers have significant maintenance and labour cost assumptions, whereas moving bed combustion has the lowest heating cost due to high efficiencies and plant automation.

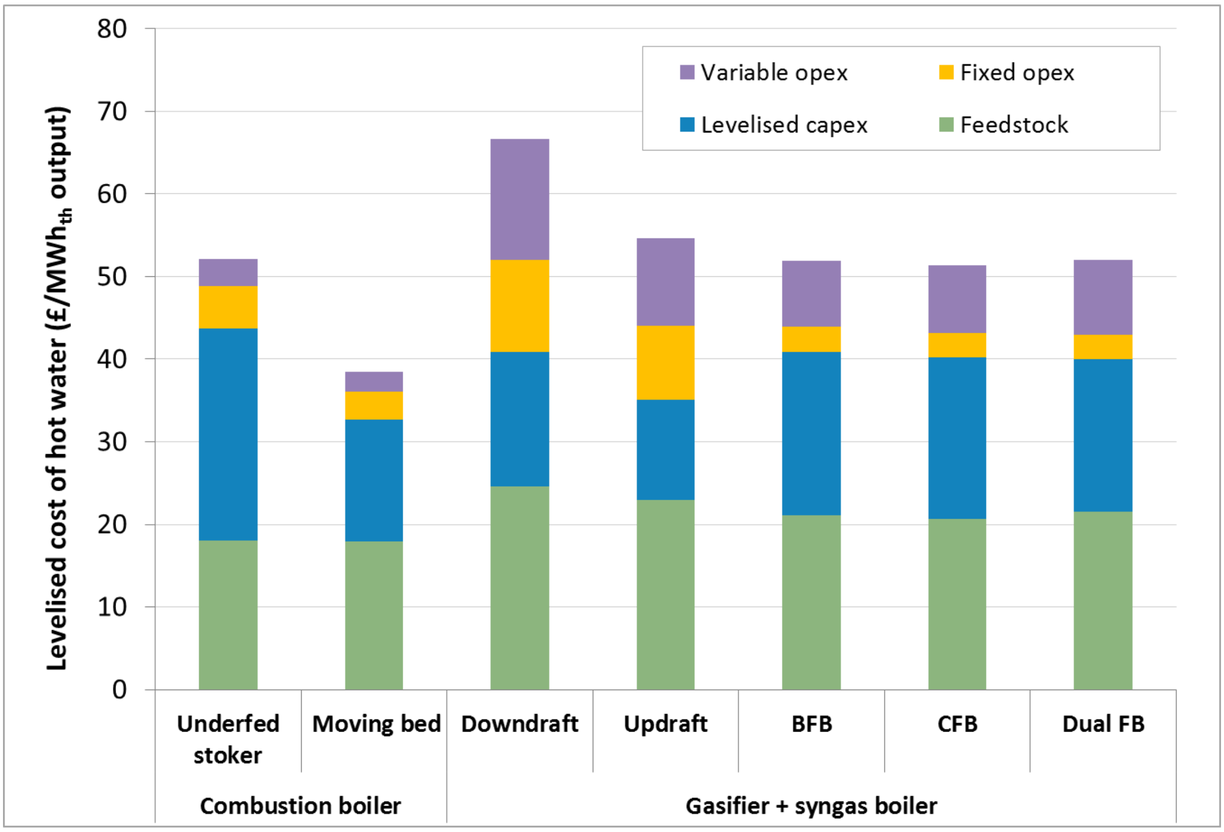

Figure 5: LCOE benchmarking for the TEABPP heating conversion technologies

Figure 6 shows the breakdown of CHP costs, with co-product heat revenues contributing to lower the net total LCOE values. Moving bed combustion + steam CHP systems are particularly expensive due to their small scale and low electrical efficiency. The fluidised bed gasifier + syngas CHP systems have the highest efficiencies, but BFB combustion + steam CHP is cheapest overall due to lower Capex and Opex.

Figure 7 again shows that the smallest technologies have the highest levelised Capex, with the operating costs of the gasifier + syngas engine systems adding significantly to overall LCOE. Gasifier + CCGT systems are able to achieve high efficiencies (reaching 40% on an LHV basis), with very large dust combustion systems able to achieve the lowest overall costs.

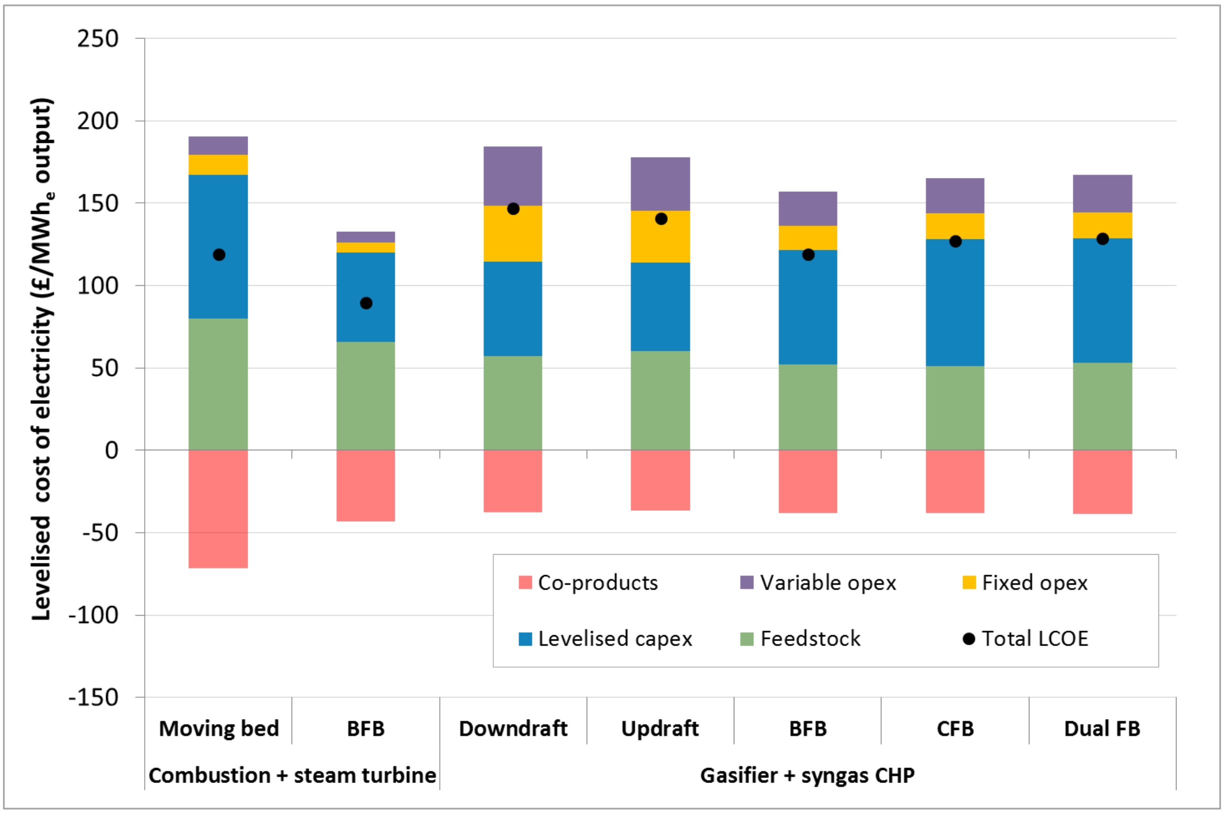

Figure 6: LCOE benchmarking for the TEABPP CHP conversion technologies (dots show net LCOE)

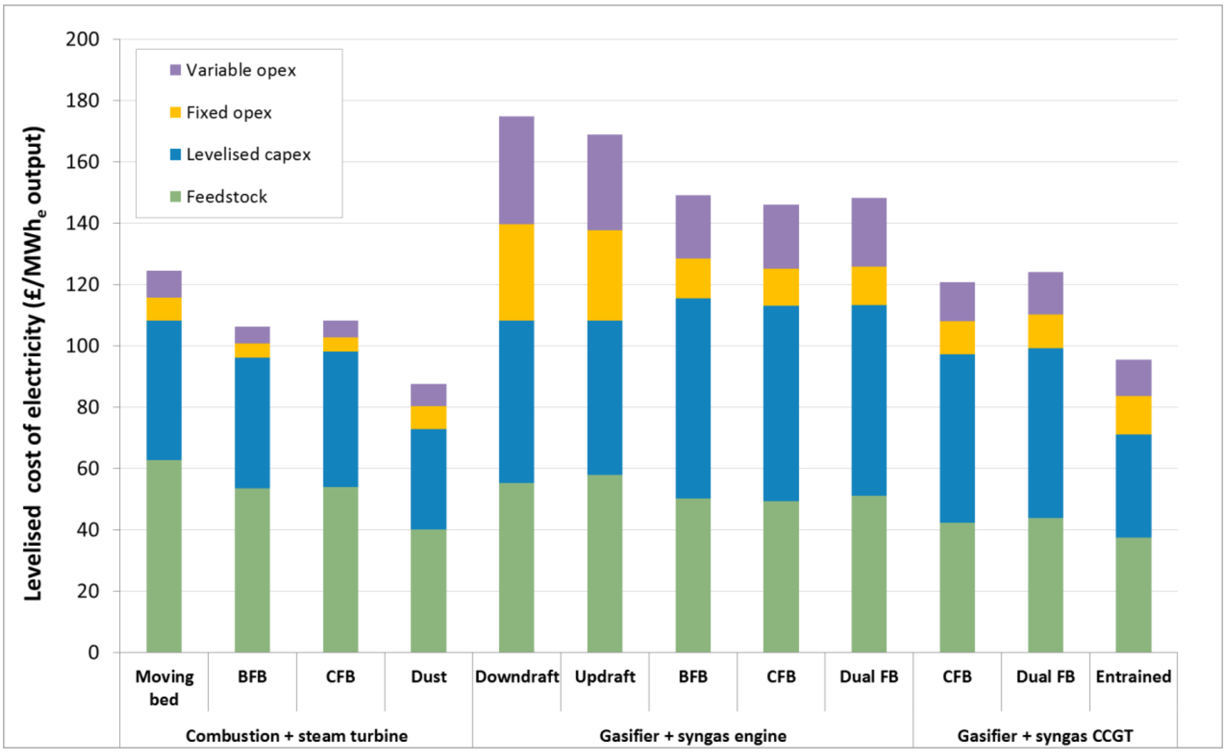

Figure 7: LCOE benchmarking for the TEABPP power conversion technologies

Finally, Figure 8 provides the LCOE for cleaned syngas from each of the gasification technologies, including sufficient clean-up costs to reach gas turbine specifications (which can result in an efficiency loss of 5% to 20% from raw syngas). The operating costs of fixed bed systems add significantly to their overall LCOE, whereas the largest scale entrained flow gasifiers have the lowest Capex and highest efficiency.

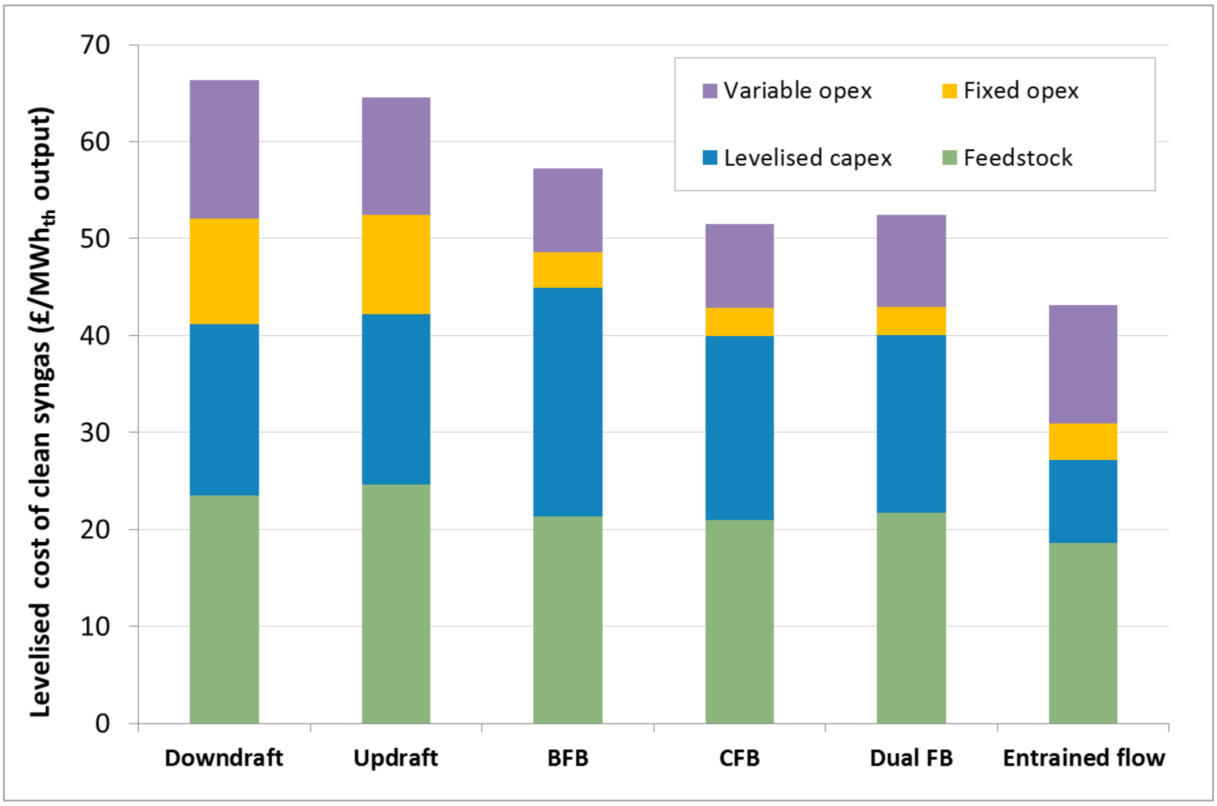

Figure 8: LCOE benchmarking for the TEABPP gasification technologies producing clean syngas

Table 3: Component steps in each of the technologies presented

Technology	**1**	**2**	**3**	**4**	**5**	**6**	**7**
Pre-processing technologies							
Water washing	Screening to remove stones	Chipping	Magnetic screening	Water washing	Filtering		
Chemical washing	Screening to remove stones	Chipping	Magnetic screening	Water washing	Chemical washing	Filtering	
Drum drying	Chipping	Screening	Drying				
Belt drying	Chipping	Screening	Drying				
Briquetting	Chipping	Screening	Drying	Press	Cooling		
Screening	Chipping	Screening					
Chipping	Chipping						
Pelleting	Chipping	Screening	Drying	Grinding	Conditioning	Pelleting	Cooling
Pyrolysis	Drying	Grinding	Fast pyrolysis	Char separation	Liquid recovery		
Torrefaction	Chipping	Screening	Drying	Torrefaction	Cooling		
Torrefaction + pelleting	Chipping	Screening	Drying	Torrefaction	Pelleting	Cooling	
Torrefaction + briquetting	Chipping	Screening	Drying	Torrefaction	Press	Cooling	
Steam explosion + pelleting	Screening	Chipping	Drying	Steam Explosion	Pelleting	Cooling	
AFEX + pelleting	Screening	Chipping	Drying	AFEX	Pelleting	Cooling	
Conversion technologies – combustion							
Underfed stoker	Boiler						
Moving bed	Boiler						
BFB	Boiler						
CFB	Boiler						
Dust suspension	Grinding	Boiler					
Conversion technologies – gasification							
Downdraft	Gasifier	Syngas clean-up					
Updraft	Gasifier	Syngas clean-up					
BFB	Drying	Gasifier	Syngas clean-up				
CFB	Drying	Gasifier	Syngas clean-up				
Dual fluidised bed	Drying	Gasifier	Syngas clean-up				
Entrained flow	Drying	Grinding	Gasifier	Syngas clean-up			

2 Introduction

2.1 Deliverable Objectives

There are two objectives of Work Package 1 (WP1) in the Techno-Economic Assessment of Biomass Pre-Processing (TEABPP) project: firstly, to carry out a review of the pre-processing, combustion and gasification conversion technologies used to transform forestry and perennial energy crop feedstocks into heat, power and syngas; and secondly, to gather techno-economic data of sufficient quality to be used in the initial chain cost results (in WP2) and for population of the gPROMS databases for the process modelling (in WP3).

This document reports on the review and data gathering activity in WP1 and covers all those technologies identified in the Contract. It provides the groundwork for the subsequent model development, by identifying key factors, metrics and issues for parameterisation. This report is to enable the ETI to understand which technologies are nearest/furthest to being commercialised, which are cheapest on a stand-alone basis and the impact that different feedstock characteristics have on the operation and costs of the technologies. The report will also highlight any data gaps and where uncertainty ranges are highest.

We have used a common review framework to assess all 33 technologies covered in this report. For each combustion (Section 3) and gasification (Section 4) conversion technology we have provided:

- A short technology description and diagram
- Development status and timescales
- The impact of different feedstock parameters on operation and cost
- Available options for improvement

For each pre-processing technology (Section 5), we have provided:

- A short technology description and diagram
- Development status and timescales
- The impact of different feedstock parameters on operation and cost
- A review of strengths, weakness, opportunities and barriers (SWOB analysis)

As part of a Horizon scanning exercise (Section 6), we have provided short technology descriptions, development status and timescales for a further set of typically less mature conversion and pre-processing technology options (identified via a literature search and brainstorming sessions).

The report concludes with a benchmarking analysis (Section 7), which calculates Levelised Cost Of Energy (LCOE) in £/MWh for each technology at its Base Case scale. These are used to compare the costs of the various options for producing hot water, electricity, CHP and syngas, plus all of the pre-processing technologies, on a common basis. Summary tables are also provided with current technology readiness level, timescales to reach commercialisation, the range of commercial plant scales and the key benefits of the pre-processed biomass forms.

This Introduction section provides a description of the project scope, the key methodology decisions taken when carrying out the landscaping.

2.2 Project scope

The conversion technology scope is limited to current data in 2015 GBP for new build, dedicated biomass conversion plants making syngas, heat, power or heat and power, with an input thermal capacity greater than 200kWth. Plants incorporating Carbon Capture and Storage or producing liquid biofuels are also out of scope.

We have not created techno-economic spreadsheets and write-up chapters for the following technologies, for the reasons stated below:

- **BFB combustion → heat:** We have provided the required information for BFB combustion → power and BFB combustion → CHP. However, there is no data available on BFB combustion → heat. We are aware that BFB combustion has been used in the past for raising heat with coal as a fuel and for hot gas generation, but we are not aware of any systems using BFBs fired on biomass for heating hot water. This is because BFB combustor scales are far too large for most hot water heating applications.
- **CFB combustion → heat** and **→ CHP:** We have provided the required information for CFB combustion → power but, bearing in mind these are for very large installations, we do not believe there is any application for such large installations raising hot water (which rules out both heat and CHP applications). There may be a use for pass-out steam or steam extraction for process heat applications, but not for hot water and again not in such large installations.
- **Underfeed stoker combustion → power** and **→ CHP:** Underfeed stokers are used up to about 2 MW_{th} hot water. Use of the same device albeit raising steam for power or CHP generation would not be cost effective as the efficiency would be extremely low.
- **Log burner → heat:** Ruled out of scope early on based on scale, as the only systems available have thermal capacities up to 50 kW_{th} input. We are not aware of any automatically fed log burners that are >200 kW_{th}.
- **Fluidised bed drying:** We sent enquiries to a number of dryer manufacturers, but we were not able to obtain a quotation for a fluidised bed dryer using biomass. One of the drying equipment suppliers, who provided quotes for both belt drying and drum drying, advised that they would not offer fluidised bed dryers for biomass, because the alternative technologies were much more cost effective. Their fluidised bed dryer would be only typically be used for sludges and for fine pharmaceuticals.
- **Blending**. We looked into blending during D1 and gathered some cost evidence – but the majority of the costs do not relate to technology, rather relate to the costs of storage and handling, which are inherently logistics steps (part of D2, not D1). The minimal costs of blending therefore did not justify creating a whole new technology datasheet – we are instead proposing to allow blending within the WP3 modelling upon the user selecting multiple feedstocks, and the model will add the necessary incremental costs of an extra hopper and conveyor, treating blending as a mathematical function to combine the costs, emissions and biomass parameters of the two feedstocks into new distributions. We note that academic research in the area of biomass feedstock blending is still fairly early, and linear assumptions about mixing proportions are still commonly used, since the impacts of blended material on the conversion technology is not well known[1]. The only cost estimates

we have seen relate to blending onsite at the conversion or pre-processing plant, with different feed lines being integrated[2]. Offsite blending could be as simple as alternate truck deliveries unloading onto a pile or into a storage facility.

Note that for the pre-processing plant system boundaries, there are often multiple steps costed into a single plant. For example, “torrefaction + pelleting” requires grinding or chipping, drying, torrefaction and pelleting. The choice was made to not separate each plant component, as the process modelling will be complex enough as it is, without adding multiple new steps in each supply chain. The data provided is therefore for a whole plant, with “as received” biomass entering the system boundary, and an “intermediate product” leaving the system boundary. The costs and efficiencies (and any biomass characteristic limits) of all the component steps are incorporated within the plant data.

For gasification, we have taken the approach throughout this report and the data collection to gather information on the gasifier and the downstream syngas boiler/engine/CCGT separately. The final LCOE system costs in the benchmarking section are presented as fully integrated systems, whereby efficiencies are multiplied and costs added together from the two components. This approach is different to that of combustion, whereby e.g. the boiler and steam turbine are presented together. We took this approach for gasification for the following reasons:

- Integrated gasification systems tend in fact at some point to be decoupled into the syngas production and clean-up and the end-use technologies, usually with quite different vendors for each component (the only exception we are aware of is GE, who can provide both parts).
- Furthermore, the technology development trajectories are different, with boiler companies, engine companies and turbine companies focussing on developing technologies that are flexible in the composition of syngas they use (whether coal, oil or biomass based), rather than trying to develop "integrated solutions" from biomass to end use. Because these syngas boilers/engine/CCGTs are adaptations of existing natural gas based technologies, they tend to be quite high TRL and are not limiting the deployment of integrated solutions. The biomass gasifier is the rate limiting component.

We have therefore considered the gasifiers and syngas end uses separately, but included additional clean-up Capex and Opex for the cases where syngas is coupled with engines or turbines to account for the higher levels of clean-up required; these clean-up costs are also included in the LCOE calculations for these technologies with syngas as the final application (as turbine/pipeline purities also need to be met). Syngas clean-up costs are avoided for integrated heating boiler applications.

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[1] From INL (2014) “Feedstock Supply System Design and Economics for Conversion of Lignocellulosic Biomass to Hydrocarbon Fuels”. Available at: http://www5vip.inl.gov/technicalpublications/Documents/6038147.pdf
What still needs to be determined is if blended feedstocks will behave like a single feedstock in a conversion facility. The testing of blends is currently underway for several conversion technologies to determine feedstock behaviour from front-end through finished blend-stock. More research is required to understand the behaviour blended feedstocks will have on overall fuel conversion. Even though it may be possible to blend to specification as measured by composition and physical properties, an additional challenge of the blended feedstock approach is to have the blended feedstock actually perform as well as or better than a singular feedstock in the conversion process. Better understanding of the interactions of blend-stocks in the conversion process will require additional research and development focus to better inform blended feedstock development.

[2] INL (2014) “Feedstock Supply System Design and Economics for Conversion of Lignocellulosic Biomass to Hydrocarbon Fuels: Conversion Pathway: Fast Pyrolysis and Hydrotreating Bio-oil Pathway”, Available at: http://www5vip.inl.gov/technicalpublications/Documents/6038147.pdf

2.3 Technology Readiness Level (TRL) definitions

The development status, expressed in current, of all the technologies covered in this report is summarised in the following pages. This status is expressed using the current Technology Readiness Level (TRL), as defined by the European Commission under Horizon 2020.

TRL is a relative measure, first introduced by NASA, of the maturity of evolving technologies on a scale of 1 to 9. TRL 1 corresponds to basic research on a new invention or concept, TRL 5 to pilot scale testing, whilst TRL 9 corresponds to mass deployment of a fully commercialised technology. The definitions of each TRL are given in Table 4 as given by Horizon 2020.

Table 4: TRL definitions[3]

TRL	Plant stage	Definition
1	Basic research	Principles postulated and observed but no experimental proof available
2	Technology formulation	Concept and application have been formulated
3	Applied research	First laboratory tests completed; proof of concept
4	Small scale prototype	Built in a laboratory environment ("ugly" prototype)
5	Large scale prototype	Tested in intended environment
6	Prototype system	Tested in intended environment close to expected performance
7	Demonstration system	Operating in operational environment at pre-commercial scale
8	First of a kind commercial system	Manufacturing issues solved
9	Full commercial application	Technology available for consumers

We note that TRL definitions are not necessarily set by plant capacity, because different markets operate at different scales. Heat and CHP market applications can be an order of magnitude (or more) smaller than power market applications and it is still unclear at what scale syngas will be utilised downstream. Therefore, what might be a small demonstration plant for the use of a particular technology in generating power, could potentially count as a first commercial plant if used in another application. The MW output of a plant is therefore only a guide to the level of commercialisation, particularly for pre-processing technologies.

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[3] Schild, P. (2013) “Horizon 2020: Calls-Overview”, Available at: http://ec.europa.eu/research/conferences/2013/energy_infoday/pdf/session_3_summary_of_the_calls_open_in_2014_-_philippe_schild.pdf

2.4 Biomass feedstocks

There are five biomass feedstocks in scope:

- Miscanthus
- Short Rotation Coppice (SRC) willow
- Coniferous Short Rotation Forestry (SRF)
- Deciduous Short Rotation Forestry (SRF)
- Imported Long Rotation Forestry (LRF) wood pellets

For use in the project modelling, these feedstocks require characterisation by a number of properties, including proximate and ultimate analysis, moisture and ash contents, heating values, density and physical form. The necessary data to characterise the different feedstocks will be provided by the ETI's Characterisation of Biomass Project, but in this section we provide an overview of the likely values for some of the main biomass properties and discuss the implications that different properties can have for supply chains and downstream conversion plants.

For clarification, agricultural residues and waste feedstocks are not in scope.

2.4.1 Overview

Biomass properties of interest to this project include: calorific value, moisture content, ash content, chlorine, nitrogen and sulphur content and alkali metals content. Each of these affects the technical feasibility and cost of biomass conversion technologies. Furthermore, these properties are subject to considerable variability both within and between species (see table below) and hence there is a need to understand the effect of variation of these properties on technology operation.

Table 5: Indicative values of "green" biomass properties[4,5,6,7,8]

Property	Units	Miscanthus	SRC willow	SRF conif.	SRF decid.	Imported LRF pellet
Moisture content (green biomass)	% (wet basis)	15-30	45-60	40-60	35-55	6-10
LHV (green basis)	GJ/t	10-13	10	7-9	8-10	16-18
Ash content	Mass % (wet basis)	4	1	1-2	1.5	0.25-2
Cl content	Mass % (wet basis)	0.11	trace	0.02	0.007	0.02
N content	Mass % (wet basis)	0.35	0.3	0.04	0.4	0.2
S content	Mass % (wet basis)	0.1	0.03	0.01	0.01	0.01
Alkali metals	kg/GJ (alkali index)	0.33[9]	0.14-0.17[9]	0.034	0.067 (oak) – 0.22 (poplar)	0.03

Each of these components have impacts on the pre-processing and conversion technologies, typically manifesting themselves as increases in Capex or Opex, or reductions in efficiency, as the component percentages increase. We have explored the sensitivity of appropriate key performance measures to variations in these biomass properties.

An important part of our analysis is the definition of a reference "as received" feedstock, against which each technology evaluation and quantification is performed. This is not the same as a "green" (i.e. immediately harvested) feedstock, since most feedstocks will be subject to a degree of seasoning before they arrive at a pre-processing or conversion plant. As expected, the “as received” values in Table 6 (used for the analysis in WP1) are for drier feedstocks and hence higher energy contents than the “green” literature sources given in Table 5.

Table 6: Values for as-received biomass used in WP1 analysis

Property	Units	Miscanthus	SRC willow	SRF conif.	SRF decid.	Imported LRF pellet
Form	-	Bale	Chip	Chip	Chip	Pellet
Moisture content (green biomass)	% (wet basis)	16	30	30	30	8
LHV (wet basis)	GJ/t (wet)	13.9	12.1	12.3	12.3	17
LHV (dry basis)	GJ/t	17	18.4	18.6	18.6	18.7
Density (wet basis)	t (wet)/m^3	0.120	0.240	0.142	0.142	0.600

2.4.2 Implications of different properties

The general implications of variations in key properties are outlined below. We do not focus on form, but where it affects a particular technology this will be pointed out in the relevant section. In general, particle size distribution is a key fuel property for any combustor or gasifier, and hence the plant designer will ensure that this physical characteristic of the fuel will be maintained at all times, thereby avoiding bridging issues in hoppers and conveyors, and ensuring correct flow and residence time for efficient conversion (and minimal carbon losses).

2.4.2.1 Lower Heating Value

Lower Heating Value (LHV) is the principle measure of the “as is” calorific value of the biomass, measured on a wet basis. Lower values against a baseline will result in increased Capex per unit of energy output, increased Opex and potentially reduced efficiency. Some technologies are also characterised by a lower limit on LHV below which operation is not feasible, or at least not recommended.

2.4.2.2 Moisture content

None of SRC Willow, LRF or SRF are particularly friable, primarily because of their high moisture content. The primary difference between the different feedstocks is the amount of moisture level in the biomass as it is received. With the exception of Miscanthus, the moisture content when harvested will be in the range of 45 to 60%. Miscanthus contains a narrower range of moisture levels when harvested in the spring, ranging from 14% to 23%, with wider limits at other times, and is a similar crop to straw.

Increasing moisture content also serves to reduce the LHV of the biomass and reduce the efficiency of the thermal conversion technologies from a baseline, because more of the energy content of the biomass is used to evaporate water rather than participate in the primary conversion process. Hence the major sensitivities considered in the analysis relate technology efficiency to biomass moisture content. We assume a heat of vaporisation of 2260 kJ/kg water, with an efficiency of 0.85. This counts as a parasitic heat lead on the process.

2.4.2.3 Ash content

The ash content of biomass varies considerably. The ash **composition** itself has different implications on system cost; these are considered in the sub-sections and section below. We assume that in addition to these, the total ash content directly affects the waste disposal cost (variable Opex) primarily; i.e. there is a solid waste effect. We have assumed an ash disposal cost of £120/tonne to relate ash content to variable Opex in our sensitivity analysis (based on handling plus landfill charges).

More generally, biomass materials have significant levels of inorganic matter as impurities. Many of the practical problems encountered with the combustion of biomass are therefore associated with the nature and behaviour of the biomass ash. Ashes from biomass have, in general, relatively low fusion temperatures and have relatively high levels of the alkali metals, characteristically containing high levels of potassium and other alkaline earth metals such as sodium, calcium and magnesium and other non-alkali elements such as silicon and phosphorus. These compounds can vaporise or react with other elements as they pass though the boiler, condensing onto surfaces and forming sticky deposits on metal and refractory surfaces[10].

Potassium is the most common alkali metal in herbaceous biomass fuels and is responsible for much of the fouling and corrosion found in biomass boilers. Alkaline earth materials such as calcium form more stable compounds and are less volatile. This explains in part why woody materials, with a high concentration of calcium in the ash, pose fewer problems than herbaceous materials such as straws and grasses whose ash contains higher concentrations of alkali material.

The ash melting temperature can be an issue in some technologies; this is affected by the ash composition. The ash melting temperature is not a simple function of any one parameter, but results from the elemental composition of the materials fed to the combustor or gasifier[11]. According to Jenkins et al.[12] the alkali index (in units of kg/GJ) can be used to determine in a combustion context if slagging or fouling are likely:

$$\frac{1}{Q} Y_f^a \left(Y_{K2O}^a + Y_{Na2O3}^a \right)$$

Here, Q is the (higher) heating value of the fuel on a dry basis, Y_f^a is the dry mass fraction of ash in the fuel, and Y_{K2O}^a and Y_{Na2O3}^a are the respective mass fractions of K_2O and Na_2O_3 in the ash.

2.4.2.4 Cl, S, N mitigation costs

Most biomass materials have relatively low total sulphur contents, generally less than 0.5% w/w and the sulphur generally remains in the ashes. The chlorine content of biomass materials varies significantly; typically from 0.05 to 0.5% w/w. Despite the likelihood of some of these remaining in the ash, to avoid overoptimistic cost assumptions, it is assumed that conversion of biomass containing these impurities will lead to gas clean-up requirements: with Cl leading to HCl formation, S leading to SO_x and N leading to NO_x, and that the variable Opex of these clean-up operations are (a) proportional to the impurity content and (b) well defined by the reagent requirements. In the case of the products of Cl and S processing we assume the principal reagent to be lime (£95/t[13]) and in the case of N to be urea (£190/t[14]).

These are then used to evaluate the sensitivity of Opex against increasing impurity concentrations, by assuming that the additional opex against the base value is given by the additional mitigation reagent required on a stoichiometric molar basis (1:1 S to lime, 1 to 0.5 Cl to lime, 1 to 2/3 N to urea based on assumption of equal NO/NO_2 proportion).

2.4.2.5 Alkali metals content: Opex and other implications

The principal alkali metals present in biomass are sodium (Na) and potassium (K). They are problematic in thermal conversion processes such as gasification and combustion for three reasons: they can cause fouling of the thermal conversion and downstream equipment; they can cause corrosion and they can change the ash melting temperature and therefore lead to slagging.

As mentioned above, the TEABPP project generally treats contaminants in two ways:

- Linear variable cost: ash, S, Cl, N content are treated with linear Opex relations which are derived by disposal cost for ash and treatment cost for SO_x, HCl, NO_x arising from thermal processes.
- Thresholds (constraints): other key constituents are treated as hard constraints due to e.g. warranties or operational feasibility problems.

For certain contaminants such as alkali metals, we need a slightly more sophisticated approach. We have developed two approaches:

- Evaluating the cost per MWh to bring the feedstock into spec by blending it with a “purer”, more expensive feedstock, either on-site with the thermal process or off-site. This would involve additional Capex for blending equipment and for the higher grade feedstock. With the right data, the cost calculation is straightforward.
- Evaluate the increased Opex that would be implied by increasing fouling, corrosion etc. This does require the development of justifiable assumptions which are described below.

For our cost sensitivities, we have focussed on the latter.

According to Fahmi et al[15]: “The alkali index (kg K_2O and Na_2O per GJ energy) can be used to predict performance in a thermochemical setting (Jenkins et al., 1998). With an alkali index above 0.17 kg/GJ, fouling is probable, and above 0.34 kg/GJ, it is almost certain.”

High feedstock mineral content can be mitigated to a certain extent by using newer alloys to construct components that can minimize and withstand some corrosion, and controlling the temperature of the reaction (Jenkins et al., 1998; Fahmi et al., 2008).

From IRENA’s cost analysis[16], it is indicated that slagging/fouling/corrosion are a significant part of maintenance costs – of the order of 50%. Note it says “Operations and maintenance costs make a significant contribution to the LCOE of biomass plants and typically account for between 9% and 20% of the LCOE for biomass power plants.”

Therefore we have proposed a simple model as follows:

- BaseOpex = the baseline Opex
- NaKthreshold = 0.17 kg/GJ when expressed as amount of K2O and Na_2O that would be associated with 1 GJ of energy. At this level we are still at BaseOpex associated with the technology
- At 2xNaKthreshold = 0.34 kg/GJ, at which Opex = multiplier * BaseOpex

- So, Opex = MAX (BaseOpex, BaseOpex + BaseOpex*((NaK – Nakthreshold) /0.17)*(multiplier-1))

Performing sensitivity analysis by looking at different types of biomass and their alkali metals content, the process Opex and Capex, a multiplier of 1.15 seems suitable, i.e. the Opex is 15% higher at an alkali index of 0.34 (corresponding to around 0.5% by mass of 30% moisture content biomass with LHV of 12.5 GJ/t).

Although originally derived from coal, the alkali metals index has been used for biomass fuels as an indicator of operational problems, and its use in the TEABPP project is considered valid, given the level of detail of the analysis. Examples of such analyses are below.

Table 7: Further data on alkali metals content and potential operability problems from slagging[17]

			Total Alkali content			
Feedstock	**Btu/lb (dry)**	**Ash %**	**% in Ash**	**lb/ton**	**lb/MMBtu**	**Slagging risk**
Pine Chips	8,550	0.70%	3.0%	0.4	0.07	Minimal Slagging
White Oak	8,165	0.40%	31.8%	2.3	0.14	
Bagasse - washed	8,229	1.70%	12.3%	4.2	0.25	Threshold 0.4 lb/MMBtu
Hybrid Poplar	8,178	1.90%	19.8%	7.5	0.46	Probable Slagging
Urban Wood Waste	8,174	6.00%	6.2%	7.4	0.46	
"Clean" Tree Trimmings	8,144	3.60%	16.5%	11.9	0.73	
Almond Shells	7,580	3.50%	21.1%	14.8	0.97	Certain Slagging
Refuse Derived Fuel	5,473	9.50%	9.2%	17.5	1.60	
Switch Grass	7,741	10.10%	15.1%	30.5	1.97	
Wheat Straw-average	7,978	5.10%	31.5%	32.1	2.00	
Wheat Straw-high alkali	7,167	11.00%	36.4%	80.0	5.59	
Rice Straw	6,486	18.70%	13.3%	49.7	3.80	

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[5] Practical on-farm renewable energy (http://www.dardni.gov.uk/4-heat-from-biomass.pdf)

[6] Wood as Fuel, Forestry Commission, (http://www.forestry.gov.uk/pdf/FC-BEC-InfoSheet-Wood-as-Fuel-TechSupp.pdf/$FILE/FC-BEC-InfoSheet-Wood-as-Fuel-TechSupp.pdf)

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[8] Wood fuels handbook, AEBIOM, http://www.aebiom.org/IMG/pdf/WOOD_FUELS_HANDBOOK_BTC_EN.pdf

[9] Saddawi A., Jones J. M., Williams A., and Le Coeur C. "Commodity Fuels from Biomass through Pretreatment and Torrefaction: Effects of Mineral Content on Torrefied Fuel Characteristics and Quality" Energy and Fuels dx.doi.org/10.1021/ef2016649

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[17] Thomas R. Miles, Thomas R. Miles Jr., Larry L. Baxter, Bryan M. Jenkins, Laurance L. Oden. Alkali Slagging Problems with Biomass Fuels, First Biomass Conference of the Americas: Energy, Environment, Agriculture, and Industry, Volume 1, 1993.

2.5 Syngas quality

In addition to valuable products such as combustible gases (carbon monoxide, hydrogen and methane) and inert gases (carbon dioxide, water vapour, nitrogen), various gaseous 'impurities' are also formed during biomass gasification. Although the impurities are highly dependent on the nature of the biomass feedstock, they can be categorized as follows:

- Low and high molecular weight organic compounds ('tars'), which cause pipe clogging, soot formation during combustion and blinding of ceramic filters
- Particles such as char, ash and gasification additives (e.g. sand, limestone or dolomite), which cause plugging and abrasion of downstream equipment
- Nitrogen containing impurities (ammonia and hydrogen cyanide) that convert into problematic NO_x during combustion
- Sulphur containing impurities (hydrogen sulphide) that release sulphur dioxide during combustion
- Other impurities such as hydrogen chloride, some vapour phase metals and alkali metals, which cause gas turbine and heat exchangers corrosion and poison catalysts of the catalytic tar crackers

In order to decide what to do with the impurities, it is first important to decide what the end-use device is for the produced fuel gas. If syngas is to be used to generate electricity and/or heat (or be sent down a pipeline), it potentially needs to go through a clean-up process in order to meet the generator equipment specifications, as these contaminants above can prevent syngas from meeting environmental emission regulations, or can damage the downstream equipment.

In particular, the presence of tars in syngas has been one of the major challenges faced in the use of syngas for industrial applications. Tar consists mainly of polycyclic aromatic components with a high boiling point. An efficient removal of these products is essential because tar condenses on moving parts, like gas turbines, which leads to blockage, damage and interruption in operation. For closed combustion purposes (e.g. heating boilers) it may not be necessary to eliminate 'tars' from the produced fuel gas. This is especially so as the tars actually add to the energy value of the fuel gas. In internal combustion (IC) engine applications, acidic compounds such as phenols and cresols need to be reduced as they cause corrosion of the engine parts. However, some light 'tars' such as benzene and toluene are not considered harmful to the engine, especially as small quantities are found in gasoline. It may be important to note that of the IC engines, the high speed and 'new generation' high performance engines are less robust than the low speed and 'old' engines.

2.5.1 Allowable tar limits

For gas engines, gas turbines and fuel cells, the allowable tar levels are typically 0.05 g/Nm^3, 0.005 g/Nm^3 and 0.001 g/Nm^3 respectively. In order to attain these low tar levels it may be necessary to clean and condition the produced fuel gas before use.

For turbines, ORNL state[18]:

"The major areas of difficulty for materials performance in applications involving syngas are likely to be those associated with the specific characteristics of the combustion environment, especially the type and amount of minor species present in the fuel, viz:

- the increased content of water vapour in the combustion gas, resulting not only from the composition of the fuel gas but also from water scrubbing during clean-up;
- the sulphur level, depending on the level of clean up and the efficiency of the process; and
- the particulate content, which is expected to be very low from clean-up scrubbing processes, but will vary depending on the functioning of those processes with the possibility of very high levels being attainable (albeit for relatively short times) during plant upset conditions.

In each case, the value of these parameters is expected to be higher than for turbines fired with natural gas, and could have an important influence on component performance."

Contaminants aside, a range of syngas compositions have been used in IGCC turbines, as demonstrated in Table 8.

Table 8: Range of syngas compositions in IGCC applications, note that A-G reflect different operating IGCCs[19]

Syngas	A	B	C	D	E	F	G
H_2 (%)	24.8	37.2	35.4	34.4	8.6	61.9	22.7
CO (%)	39.5	46.6	45	35.1	26.2	26.2	30.6
CH_4 (%)	1.5	0.1	0	0.3	8.2	6.9	0.2
CO_2 (%)	9.3	13.3	17.1	30	14	2.8	5.6
N_2+AR (%)	2.3	2.5	2.1	0.2	42.5	1.8	1.1
H_2O	22.7	0.3	0.4	0	0	0	39.8
Pre diluent LHV							
BTU/ft^3	209	253	242	210	183	317	163
kJ/m^3	8224	9962	9528	8274	7191	12492	6403
GT Temp °C	330	372	121	96	204	38	200
H_2/CO Ratio	0.63	0.79	0.98	0.98	0.33	2.36	0.74
Diluent	Steam	N_2	N_2/Steam	Steam	-	Steam	Moisture
Equiv. LHV post diluent							
BTU/ft^3	150	118	113	198	-	200	-
kJ/m^3	5910	4649	4452	7801	-	7880	-

As a result of these considerations, we have assumed that in the gasification technology description sections, the gasification produces raw syngas. However, in the whole chain analysis in D2 and the benchmarking analysis in this report, we include the cleanup costs and efficiency penalties because these now must take account of the end use technologies. This way, the gasification technologies can be compared on a common basis without a bias brought about by the end use.

The syngas end-use technologies are modelled "as is", i.e. the syngas composition is within the required specification.

2.5.2 Assumptions

Following gasification, in the benchmarking analysis we assume that the syngas used for downstream applications is always subjected to an appropriate series of clean-up steps (typically particulate removal, wet scrubbing and adsorption) to mitigate tars and other contaminants, and therefore meets the specifications of standard equipment (e.g. engines and turbines). This also includes removal of alkali metals which can take place post gasification if required (e.g. if high and not removed through biomass pre-processing). However, the sensitivities and performance indicators reported for each gasification technology are for the production of raw syngas, which allows each technology to be compared prior to consideration of the end use. This is complemented by the inclusion of the syngas clean-up costs and efficiency penalties in the benchmarking analysis.

Where possible, we try to estimate the potential for reduced gas treatment and poorer quality syngas/feedstock on the syngas-using equipment or on the post-gasification clean-up in section 4.7 onwards, either through increased Capex due to more expensive materials or higher Opex through more cleaning/parts replacement.

Clean-up constitutes two different elements: contaminant removal and tar destruction; these sometimes take place in the same equipment and sometimes separately.

2.5.3 Contaminant removal

This involves removal of lower concentration contaminants present in raw syngas (e.g. hydrogen sulphide, ammonia and hydrogen chloride as well as various forms of trace metals). Typical clean-up and conditioning processes include cyclone and filters for bulk particulates removal; wet scrubbing to remove fine particulates, ammonia and chlorides/sulphides; solid absorbents for mercury and trace heavy metal removal, and acid gas removal where necessary. Alkali metals will be progressively removed throughout these steps, with the most part removed in the first steps.

2.5.4 Tar reduction

Present tar reduction or destruction methods can be divided into three main processes: mechanical methods, self-modification and thermal cracking. More advanced technologies for syngas clean-up using plasma and catalysts are described in the Horizon-Scanning section.

2.5.4.1 Mechanical method

The mechanical methods include: use of filters (fabric, ceramic, baffle, etc.), cyclone separator, rotational particle separator, scrubbers, etc. Although the main purpose of this method is to remove fly ash or particles from the produced fuel gas, the removal of tar is also very good. Water scrubbers have been known to reduce tar levels to 20mg/m^3 – 40mg/m^3 and particulate levels down to 10mg/m^3 - 20mg/m^3. However wet scrubbers are expensive equipment, which generate further environmental problems, as it is even more difficult to dispose of the scrubbing medium (e.g. waste water or oil) after the process. There is also the issue of knowing the particle size distribution in order to be able to select the correct cleaning system. Another disadvantage of this method is the fact that by removing the tar from the fuel gas, the energy value of the fuel gas is reduced significantly. The disadvantages of this method have been largely overcome by the oil scrubbing approach of the OLGA technology, which is discussed in Section 6.3.1.2.

2.5.4.2 Self-modification

The self-modification method promotes optimising gasification conditions to minimise tar production, rather than trying to destroy the tars after production, so this method should be more efficient than the other tar reduction or destruction method – although is still unlikely to reach the strictest syngas purities required by turbines/pipelines. Raising the temperature (promoting secondary cracking), raising the equivalence ratio and using more steam have been found to be effective methods of reducing tar output. Using a suitable low tar producing gasifier (e.g. downdraft or entrained flow), can also be a means of ensuring a low tar gas is obtained.

2.5.4.3 Thermal cracking

Thermal cracking is another suggestion for decreasing tar yield. This method involves heating up raw fuel gas (obtained from gasification) at temperatures higher than 1000°C at reasonable residence times. It has been found that at temperatures higher than 1250°C at a residence time of 0.5s in a fluidized bed, tar yield is significantly reduced. At this temperature tar yield was found to be about 0.05g/Nm^3 (considered the maximum limit for engine applications). A further increase in temperature to 1290°C reduced tar yield to about 0.015g/Nm^3 in an updraft gasifier. CHOREN's Carbo-V EF technology operated at 1200-1400°C, and was acquired by Linde in 2012. Linde has demonstrated the technology at a plant in Freiburg and in 2013 licensed it to Finnish Forest Btl Oy, who had planned to build a large biofuel plant under NER300[20].

Under high temperature conditions, the refractory unsubstituted aromatics can be destroyed without the use of a catalyst. However thermal cracking involves high temperatures which bring about high fixed and operating costs. It can also produce more problematic products such as soot, acid gases and volatile alkali metals.

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[18] Gibbons, T.B. and Wright I.G. "A review of materials for gas turbines firing syngas fuels" 2009 ORNL http://info.ornl.gov/sites/publications/files/pub15496.pdf

[19] Meher Homji et al., "Gas turbine fuels: system design, combustion and reliability", Proc 39th Turbomachinery Symposium, 2010.

[20] The Linde Group (2013) "Linde and Forest BtL sign licensing agreement for Carbo-V technology". Available at: http://www.linde.pl/pl/news_and_media/press_releases/news_130125.html

2.6 Standard Assumptions

A number of standard cost, thermodynamic and related assumptions are made in our calculation sheets as part of WP1, and these assumptions are discussed in this section.

We assume that the individual technologies are implemented as greenfield, standalone plants, with investment costs at current overnight (2015) GBP rates. Where we have obtained other cost data (currencies, years), we use plant cost indices and foreign exchange rates to correct them to present dates.

We distinguish total installed cost (TIC) for procurement, installation and commissioning of equipment, from the total investment cost (TinvC). The latter includes TIC plus the following elements:

- Operations and utilities – includes balance of plant, provision of water, electricity, gas, cabling between motor control centre and motors etc.
- Civils and land. This requirement will vary depending on pre-processing technology and conversion technology, and local conditions. No land lease is assumed
- Project development includes design, construction management, project risk and engineers' fee – activities often covered by an engineering procurement construction (EPC) contractor
- Contingency, to cover project overruns

The TinvC is annualised over the technical lifetime of the technology using a discount rate of 10%. All of the Capex figures presented in the technology chapters further below use TinvC, not TIC. The availabilities of the technologies are based on real-world data in some cases and engineering assumptions in others, and also depends on the process complexity.

We split out Opex into fixed and variable elements as follows:

- Fixed Opex
    - Insurance (1% of total investment cost)
    - Maintenance (parts). Usually taken as a % of TIC; depends on complexity of equipment and operations
    - Maintenance (labour). Usually taken as a % of TIC; depends on complexity of equipment and operations
- Variable Opex
    - Operations labour. This varies widely depending on shift patterns/level of plant complexity and automation. For example 1MW boilers can be mainly left unattended, whereas most pre-processing plants are labour intensive
    - Inputs of energy/functional materials/chemicals/reagents, such as:
        - Binder
        - Ammonia (£410/t[21])
        - Air
        - Oxygen

        - Steam
        - Nitrogen
        - Lime/limestone (£95/t based on internet searches)
        - Urea (£190/t[22])
        - Activated carbon
        - Ammonium acetate
        - Hydrochloric acid
        - Natural gas (£0.03/kWh)
        - Diesel/red diesel (£55/MWh)
        - Heavy fuel oil (£40/MWh), used for conversion plant start-up, with typically 1-2% of biomass fuel input by energy consumed
        - Electricity (£100/MWh)
        - Water input
    - Treatment/disposal of wastes, such as:
        - Waste water treatment
        - Inerts/rejects disposal
        - Ash disposal of bottom ash and fly ash (£120/t)

For each of the conversion plants and pre-processing plants, given the Base Case feedstocks and plant availabilities, we have also gathered data on the annual outputs of:

- CO_2 emissions
- CO emissions
- NO_x emissions
- SO_x emissions
- PM emissions
- Heavy metal emissions

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[21] USDA-IL Dept of Ag Market News Service, Springfield, IL October 2015 http://www.ams.usda.gov/mnreports/gx_gr210.txt

[22] Indexmundi Commodity price data, Available at: http://www.indexmundi.com/commodities/?commodity=urea

3 Conversion technologies – combustion

3.1 Underfed Stoker

3.1.1 Technology description

Underfed stokers offer an operationally safe and relatively cheap technology for small and medium-scale systems requiring heat energy. The biomass fuel is burned to produce typically low or medium temperature hot water heat (LTHW / MTHW) at a maximum flow temperature of c. 90°C (LTHW) or 120°C (MTHW). This can be used for process or space heating (as well as cooling via absorption chillers).

Biomass is fed into the combustion chamber by augers from below a grate and is transported upwards onto the grate. The potential for burn-back along the stoker auger is high, typically requiring it to be emptied on boiler shut down. An air gap and rotary valve arrangement is often installed between the stoker auger (feeding into the boiler) and the transporter auger (extracting directly from the fuel store). In the case of pneumatic delivery systems (typically only suitable for wood pellets in smaller quantity / output systems) a day hopper would be used to feed the stoker auger. Water dousing is common in the event of over-temperature in the auger, controlled via a bimetallic strip or basic wax seal.

The stoker auger delivers fuel into a retort, a transition piece between the feed system and the grate. Forcing increasing amounts of fuel through the auger tube creates a “fuel plug” preventing the intrusion of air via the fuel store. Underfed boilers combust the fuel on a grate situated either inside or outside of the retort. External grates are more common in modern underfed boilers due to their increased flexibility during operation and simpler ash extraction options. As the fuel fills the retort, it spills over and spreads across the grate, where it is subject to primary air and radiant heat[23].

In smaller units (c. 500kW and less), the retort is typically located centrally in the combustion chamber. In larger units, the retort is positioned closer to the supply end of the fuel system. Fuel is shunted along the fixed grate by new feedstock from the stoker auger: fuel feed rate is regulated to ensure sufficient time to achieve maximum burnout on the grate (varies with moisture content).

Remaining ash and char solids continue to be pushed until reaching the edge of the grate (or other opening) where it drops into an ash collection area. From here the ash can either be manually removed from the furnace or, more commonly, transported via another auger into an ash storage vessel to aid simple waste disposal for the operator.

Secondary air is provided above the grate to encourage complete combustion. This is typically provided by a separate, dedicated fan unit in larger output boilers but smaller units may utilise one fan with modulating flap valves to direct air around the furnace as required. A lambda sensor is often situated within the flue gas stream to continually monitor O_2 levels and regulate primary and secondary air flows to improve combustion conditions and target complete burnout.

The combustion zone usually has moderate levels of refractory lining. The volume of water in the boiler is typically low in order to produce a fast response to heating load demands[24]. Water cooled walls and flue gas recirculation are two methods that can be used to regulate temperatures in larger output plant.

Hot combustion gases may navigate a number of baffles to increase residence time in the furnace before exiting via the heat exchangers. Typically, the water moves through the shell side with hot gases passing within the tubes. Automatic tube cleaning is now considered standard practice in all but the most basic models with compressed air (via separate compressor unit) or physical agitation (via hollow core auger or similar). The method of physical (as opposed to pneumatic) tube cleaning also has a potential benefit of acting as vortex shedders to increasing turbulence, residence time and ultimately heat transfer of flue gases within the tubes; however, care must be taken to avoid fouling on the agitator itself.

Numerous heat exchanger arrangements are available including multiple passes and baffles to improve heat transfer; exchangers are typically horizontally or vertically aligned. Vertical designs have an advantage in that gravity may aid the removal of any fouling but this arrangement typically requires a greater boiler footprint.

Before products of combustion exit via the flue, most boiler units incorporate a multi-cyclone to limit fly ash and particulate matter emissions. The fly ash is collected by gravity in a holding vessel or combined with the bottom ash via auger.

Most OEMs recommend a regular plant shut down and maintenance routine to ensure all ash is removed from the combustion areas and that there is no build-up of clinker deposits within the combustion chamber. Most heat exchangers are supplied with suitable tools for rodding – this is especially important in horizontally mounted heat exchanger units.

Underfed Stoker applications

All underfed stoker applications are for heating purposes. There is no call for power only generation in the size range operated by underfed stoker boilers. There are very few steam-raising underfed stoker fired boilers available and these are for process heat only, not for steam turbine or steam engine applications. This also rules out CHP applications as a likely use of underfed stoker technology.

As stated in the next section, although the inherent design of underfed boilers limits output to c. 6 MW_{th}, typically above c. 1 MW_{th} the moving bed boilers are more appropriate (financially and operationally).

3.1.2 Development status and timescales

Underfed stoker technology has been developed over more than 100 years; its operation is therefore well understood, robust and reliable. Having evolved from coal firing boiler systems developed in the early 1900's, this technology has been adapted to suit certain biomass feedstocks and has been proven over the last 20 years to offer an operationally safe and relatively cheap technology for small and medium-scale systems requiring heat energy. The current status is therefore TRL 9.

Review of the principal boiler supply companies finds an advertised maximum output of 2 MW_{th} for standard "off-the-shelf" units. There are references in literature of a nominal max boiler capacity of 6 MW_{th} but units at this range are not common and considered bespoke plant. It is expected that timescales from design finalisation to commercial operation can be within six months for the lower size range plant and within a year in the majority of larger projects.

In the UK, the introduction of the non-domestic renewable heat incentive (RHI) in 2011 has sparked a significant surge in demand for boilers in the sub-1 MW_{th} range. Many European boiler manufacturers (in particular, Austrian manufacturers) have established strong installed bases at commercial and small-industrial sites across the country, with the wider biomass market developing successfully across the supply chain (from fuel supply to installers).

Due in part to RHI tariff structures, the most significant area of growth has been the <200kW$_{th}$ domestic and small commercial boiler market. This size is outside of the scope of the report, but it should be noted that underfed stoker systems are commonplace in this market due in part to their relatively simple construction, small footprint and responsiveness to change in heating demand.

Within the 400 – 2,000kW$_{th}$ band, underfed stokers and moving grate boilers compete for market share. Many of the major OEMs appear to place greater emphasis on their moving bed technology and many do not offer underfed options above 500kW$_{th}$. The advantages in robustness and suitability for a wider variety of fuels in moving bed technology have led to an acceleration in development compared to underfed units of similar scale. This increased development has brought costs down over the last 10 years and there is often little difference between installed costs for either technology.

There is still a strong market presence for underfed stoker boilers below 500kWth and, providing that a fuel supply agreement including a consistent quality fuel specifications (such as ENPlus A1 wood pellets) is in place, the advantages of a simpler construction will continue to make it a viable option to supply process and space heating.

3.1.3 Impact of different feedstock parameters on operation and cost

Emission limits

To be eligible to receive payment under the UK RHI scheme, all applicants must be using boiler equipment that is certified to discharge within published emission limits for NO_x and PM_{10} (now 150g/GJ and 30g/GJ respectively). For non-bespoke units, equipment manufacturers can complete type emissions testing for a single unit and this will be applicable for all further units installed. However, a unique certificate is required for each boiler model across the available size range and for each feedstock combusted; e.g. an emissions certificate is valid for all 300kW$_{th}$ underfed boilers burning wood pellets to ENPlus A1 Standard.

The maximum moisture content of the fuel used during testing must also be stated on the certificate and any fuel combusted above this moisture content would not be considered to have certified emissions eligible for RHI payments. For less homogenous fuels such as wood chips, plant operators must take care to ensure the fuel that is combusted within the furnace is within the specification approved on the emissions certificate.

Commercially operating plants often incorporate an integrated multi-cyclone unit to remove fly ash and particulates from the flue gases and it is unusual for any additional flue gas treatment to be included. The strict fuel specifications and lower output capacities typically negate the requirement for additional emission control equipment (including chemical additions) for all but the lowest permitted cases.

Moisture Content

Strict feedstock moisture content limits are typically specified by the manufacturers of underfed stoker boilers. This can be for emission certification reasons as described above but also to ensure combustion performance is maintained at suitable levels. Often boiler warranties are dependent on the sole usage of feedstock deemed as suitable by the OEM.

A relatively small grate footprint when compared to moving bed technologies means that underfed stokers are typically less accommodating to wet fuel. The reduced combustion area limits the degree of drying and can lead to incomplete burnout and lower temperature product gases. These factors increase the likelihood of corrosive condensation within the heat exchanger tubes and ultimately either increase the maintenance requirements or reduce boiler life.

Preheating the primary combustion air may increase the drying ability of underfed boilers and upper limits of moisture content can be up to 35% in these cases. However, consideration within the boiler design must account for the reduction of cooling effect to the grate, increasing its vulnerability, especially if the bed of fuel or ash is insufficient to protect the grate from radiant heat in the furnace.

For similar reasons, fuel that is too dry (<8%) may burn too strongly and the resultant intense heat (particularly for larger output units) may cause damage within the furnace. The addition of flue gas recirculation within the flow of secondary combustion air can help to control the refractory temperatures but does less to protect the grate as described above.

Size Limits

An advantage of underfed stoker boilers is their good partial-load behaviour and their simple load control[25]. Load changes can often be achieved more easily and quickly than in grate combustion plants because the fuel supply can be controlled more easily and the fuel mass in the furnace is comparatively low. The small fuel inventory reduces the need for slumber mode as it is simple enough to burn out fuel on the grate and reignite when required.

However, these advantages mean high sensitivity to fuel input. A relatively small grate footprint when compared to moving bed technologies means that underfed stokers are typically less accommodating to non-homogeneous fuel: the fuel particle size must be consistent otherwise the small, intense combustion zone may be easily disrupted.

Boiler suppliers typically specify fuel size limits as per the Onorm or EU standards at under G35 (35mm). Oversize pieces may lead to blockages or bridging of fuel in the fuel feed system and also may not complete burn out before falling off the edge of the grate. Conversely, large amounts of dust / undersize material may also inhibit the flow of primary air or fall through the grate uncombusted.

Ash Limits and Chemical Composition

Maximum ash content is specified by the boiler manufacturer with underfed stokers typically requiring biomass fuels with low ash content (virgin, woody biomass). Ash-rich biomass fuels, such as Miscanthus, require increasingly efficient ash removal systems. Sintered or melted ash particles covering the upper surface of the fuel bed can cause unstable combustion conditions when the fuel and the air break through the ash-covered surface[25].

If the combustion process temperatures are not strictly regulated and the temperature of the surrounding refractory exceeds c. 900°C, the inert fractions within the fuel become sticky and collect or condense on the nearest, porous and cooler surface. This is often the adjacent refractory. These deposits, typically called slagging, may corrode the refractory with which it is in contact but, more typically, may cause damage during ash removal as part of maintenance operations.

Major problems associated with slagging within the heat exchangers are the reduction of heat transfer from the boiler and the restriction of gas flows by the deposits, causing channels of higher velocity, higher temperature and erosion. As fouling rates increase, it becomes more difficult to remove deposits via conventional means such as sootblowing so plant shut downs for cleaning become more frequent and the costs of cleaning and lost generation become more significant.

Sensitivity Analysis

The cost curve in Figure 9 corresponds to the total investment cost per unit of output capacity for an underfed heat boiler. It has been derived from quotes provided by suppliers (2009-2013) for all-in overnight EPC costs.

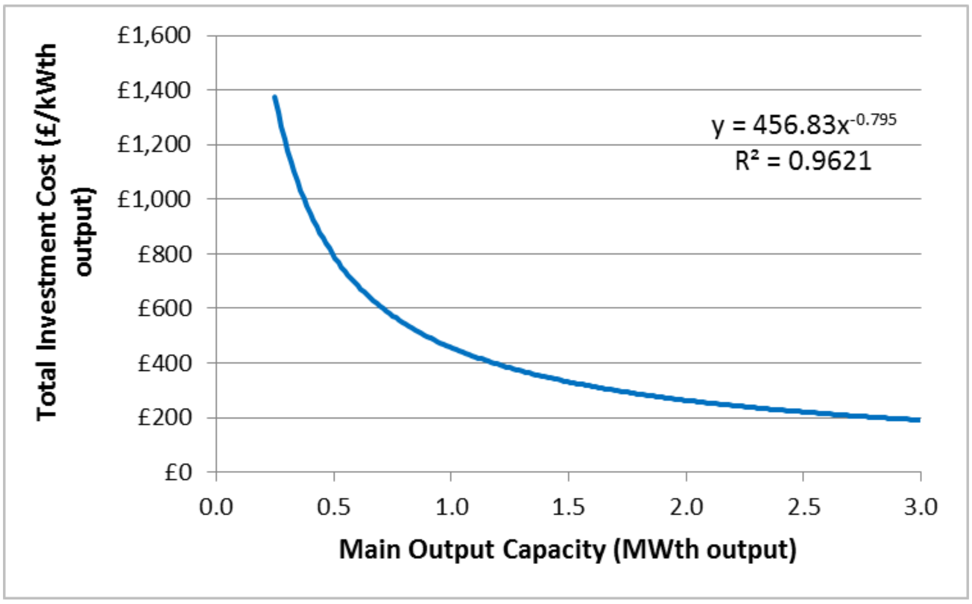

Figure 9: Underfed (heat) total investment cost vs. heat output capacity (derived from seven supplier quotes (2009 - 2013) and B&V data for all-in overnight EPC costs).

Total operational costs for underfed heat technologies includes fixed costs (insurance, maintenance parts and labour) and variable costs (operations labour, additional fuel, chemical dosing). For LTHW/MTHW boilers it is considered that water use is negligible and that the operating thermal output range does not require the addition of reagents. Figure 10 shows the relationship between annual total operational cost for an underfed boiler per unit of energy output and the main output capacity rating.

Except where the parameter of interest is shown on the x axis, the curves in Figure 10 to Figure 14 are all created using the following base values: Main output capacity = 0.4 MW_{th}; station capacity factor of 50%; feedstock moisture content (WB) = 35%; feedstock ash content (DB) = 1.5%; feedstock nitrogen content (DB) = 0.35%; feedstock sulphur content (DB) = 0.05%; and a feedstock chlorine content (DB) = 0.04%.

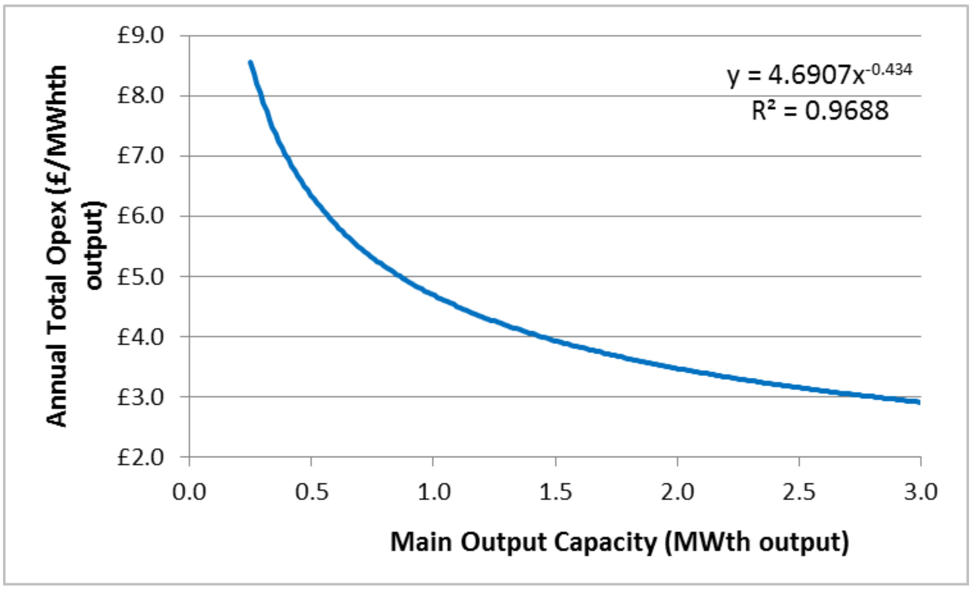

Figure 10: Underfed (heat) annual operating cost per unit of energy output vs. heat output capacity (derived from B&V industry data)

Figure 11 below highlights the change in overall underfed boiler efficiency (LHV) over varying feedstock moisture content. The efficiency is calculated as heat energy output over fuel energy input. The curve is considered to be a representative average across the underfed potential rated output range.

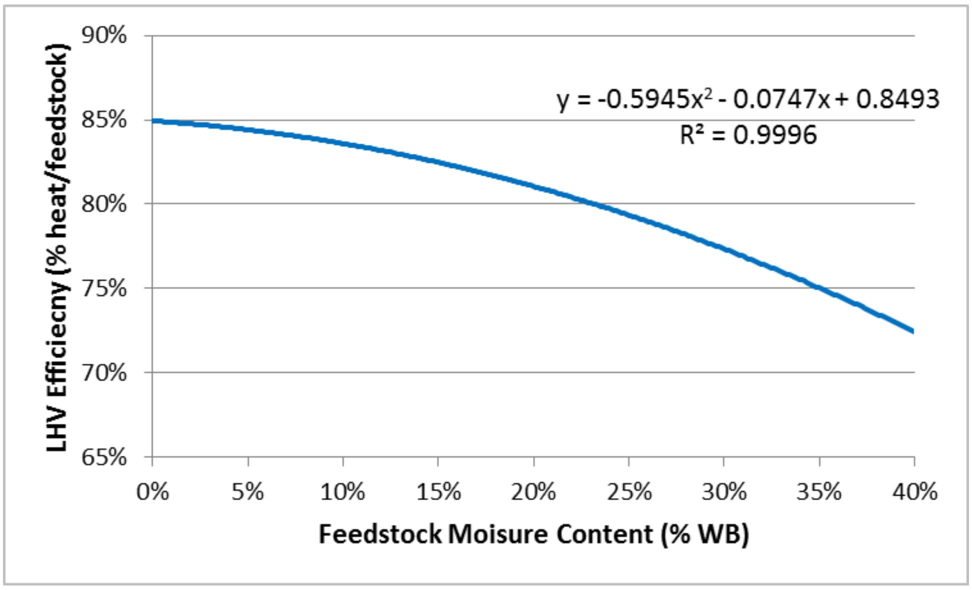

Figure 11: Underfed (heat) efficiency vs. feedstock moisture content (derived from OEM data provided by B&V, considering a representative average across the technology potential rated output range)

The total operational costs for underfed heat technologies (as previously described for Figure 10) are related to feedstock moisture content as shown in Figure 12 below.

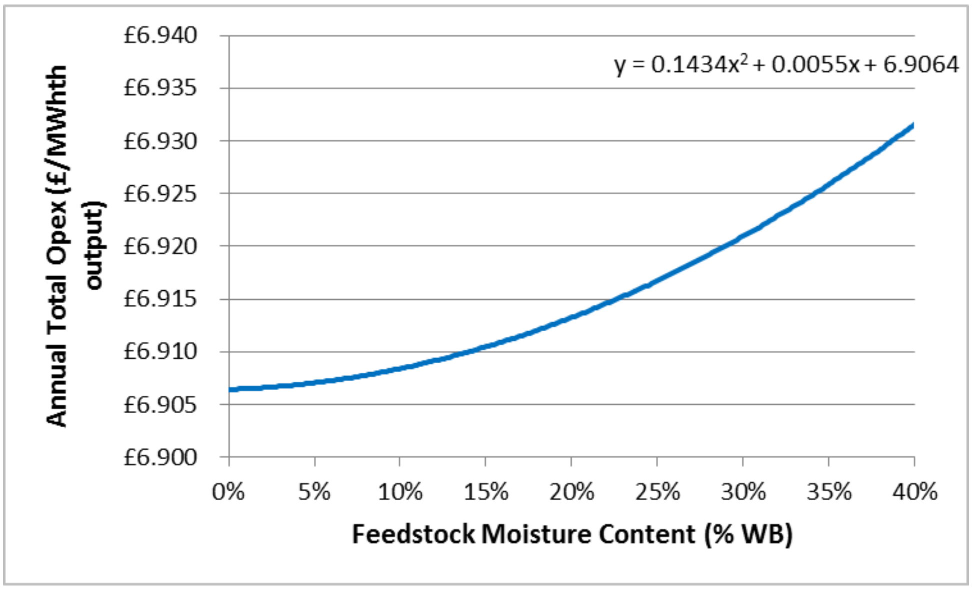

Figure 12: Underfed (heat) annual operating cost per unit of energy output vs. feedstock moisture content (derived from B&V industrial data).

Annual operational costs for underfed boilers vary with the percentage of ash present within the feedstock, as shown in Figure 13 below.

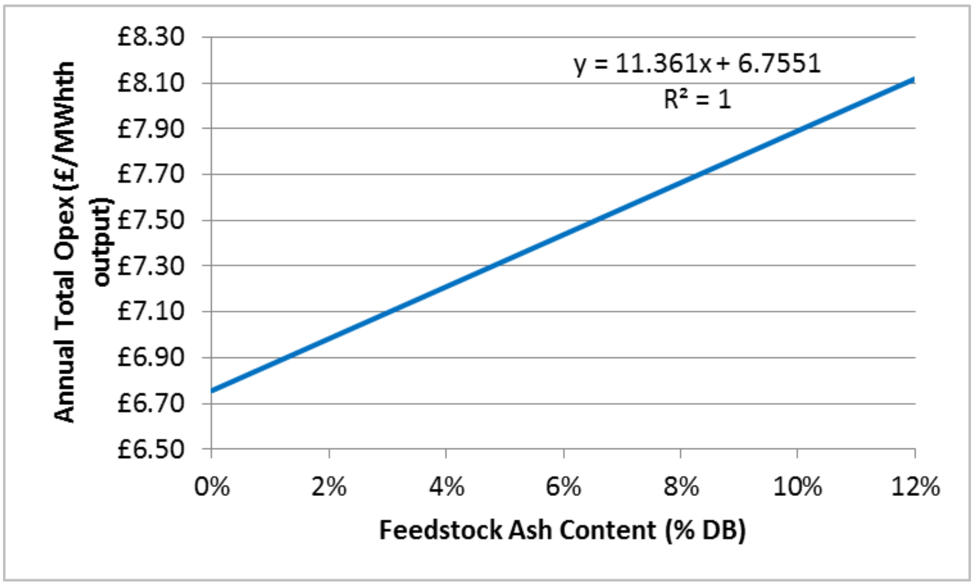

Figure 13: Underfed (heat) annual operating cost per unit of energy output vs. feedstock ash content (derived from B&V industrial data).

As described above, it is considered that reagent use is negligible across the output range available for heat only boilers, although high alkali metals will lead to increased maintenance. Figure 14 below displays the relationship between the alkali index (kg K_2O and Na_2O per GJ energy) and the annual total operational cost for an underfed per unit of energy output.

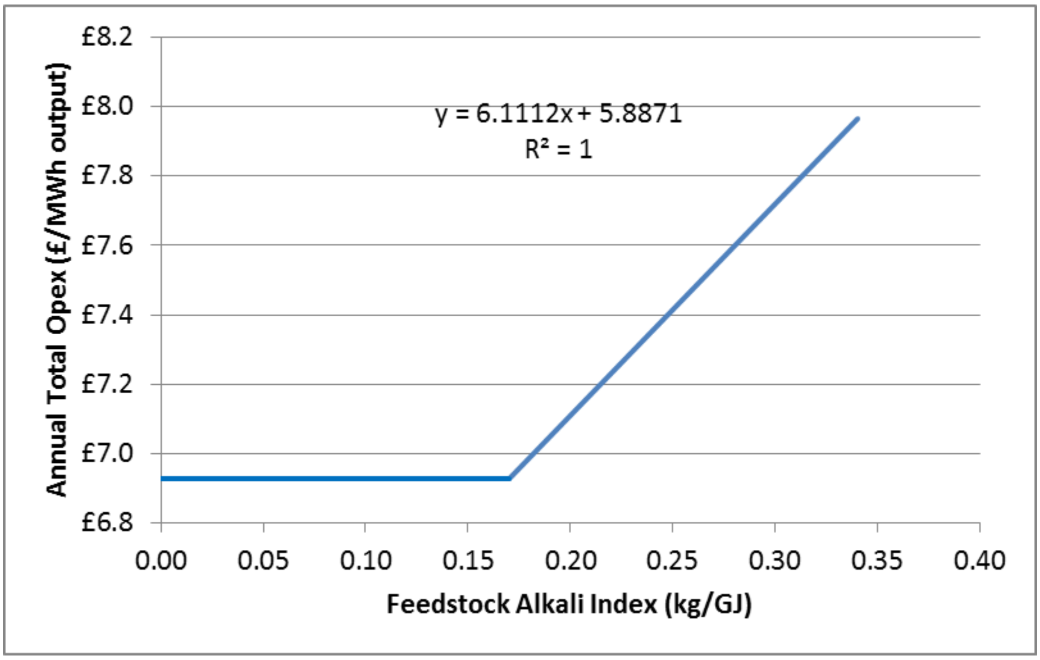

Figure 14: Underfed (heat) annual operating cost per unit of energy output vs. feedstock alkali index (derived from formula based on [26] and section 2.4.2.5).

3.1.4 Available options for improvement

Some of the more significant improvements in underfed stoker boiler design are listed in Table 9 and summarised below.

Table 9: Improvements Analysis of Underfed Stokers

Issue arising due to biomass characteristic	Options to ameliorate	Evaluation of effect	Long Term Improvements
Poor bed conditions from non-homogeneous feedstock	Larger grate area; Multiple fuel input points; Improved fuel preparation; Variable fuel feed rates.	Increased control over residence time will improve burnout of fuel and allow drying of higher moisture fuels; Consistent bed depth will reduce hot / cool spots on the grate.	Reduced grate wear; Increased tolerance of fuel specs; Scaling up options; Reduced emissions and environmental impact.
Slagging in furnace from biomass ash and chemical composition	Separate zones for primary and secondary air supply. Controlled modulation of air flows; Furnace turbulators (vortex fans and or baffles); Control of airflow to	Optimum oxygen supply to handle changing combustion conditions; Improved mixing and residence time to burn off volatiles.	Increased tolerance of fuel specs; Reduced emissions and environmental impact; Improved efficiency; Improved availability; Reduced maintenance Opex.

	regulate furnace temperature and ensure flow consistency.		
Grate damage from dry fuels	Zoning of primary air across grate; Recirculation of inert flue gases (if primary air is preheated); Water cooled grate.	Water cooled grate most expensive but provides best availability.	Reduced maintenance Opex; Improved availability.
Fouling of heat exchanger tubes from biomass ash and chemical composition.	Automated soot blowing; Plate cleaning; Turbulator coils within the tubes.	Sootblowing cost prohibitive for smaller boilers; Movable turbulator coils improve residence time (heat transfer) and physically agitate deposits to prevent deposits (vertical HX).	Reduced maintenance Opex; Improved availability; Improved efficiency.

Underfed stoker boilers are often considered the relatively lower cost option compared to competing technologies within the size range. As discussed in the previous sections, this is often in concession to its fuel and ash handling abilities and, therefore, any design additions or modifications to improve operational improvement must consider the increases in equipment costs to retain competitiveness.

Grate Design

A well-designed and well-controlled grate significantly aids distribution of the fuel over the entire grate surface area. The movement of fuel over the grate should be to be as consistent as possible in order to keep the bed of embers calm and homogeneous. This reduces the likelihood of low fuel levels, exposing sections of the grate to higher temperatures, or deeper fuel levels, causing unburnt fuel in the less exposed areas. In more advanced underfed boilers, primary air supply can be regulated across the grate to provide greater flow to the deepest sections of burning fuel and less where the ash burnout occurs. Incorrect air flow may cause slagging, higher fly-ash amounts and may increase the excess oxygen needed for a complete combustion, resulting in increased dry gas heat loss and reduced boiler efficiency[25].

Combustion Air

To achieve higher combustion efficiencies, underfed designs can achieve improved combustion by separating the zones of primary and secondary combustion. Because of the low turbulence necessary for a calm bed of fuel embers on the grate, the mixing of air and flue gas in the primary combustion chamber is not typically sufficient. Therefore, the secondary combustion zone should be designed to create a mixture of flue gas and air that is as complete as possible. Improved mixing quality can lower the amount of excess oxygen necessary for complete combustion and therefore improve boiler efficiency. Mixing can be encouraged by increasing velocities of both gases through geometry of the furnace and the injection of secondary air via high speed nozzles that are well distributed over the cross-section of this channel. Some boiler designs incorporate dedicated fan units to create a vortex within the combustion chamber; this increased turbulence is reported to improve mixing.

On smaller and/or more basic underfed boiler designs where above-grate combustion is less effective the potential for overheating and subsequent slag formation on the grate is higher and some boiler manufacturers have included a water cooling circuit within the grate as a preventative measure[24].

Slag Formation

Ash deposition occurs when the combustion process temperatures are not strictly regulated. Underfed stoker boilers can address this problem in two ways. The first way is by constantly monitoring the furnace temperatures and implementing control of air and exhaust gas recirculation rates to slow the combustion process. Secondly, increasing the water cooling within the furnace can help to lower the combustion temperatures (particularly in the case of drier fuels) and reduce the potential for fouling and structural damage.

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[23] Solid Fuel/Biomass Energy Systems Component & System Guide, volume 45; Hurst Boiler And Welding Co., INC.,http://www.hurstboiler.com/documents/component-system-guide.pdf

[24] Palmer, D., Tubby, I., Hogan, G. and Rolls, W. (2011). Biomass heating: a guide to medium scale wood chip and wood pellet systems. Biomass Energy Centre, Forest Research, Farnham.

[25] The handbook of biomass combustion and co-firing (2008), edited by Sjaak van Loo and Jaap Koppejan, Earthscan London.

[26] Fahmi R., Bridgwater A. V., Darvell L. I., Jones J. M., Yates N., Thain S., et al (2007). The effect of alkali metals on combustion and pyrolysis of Lolium and Festuca grasses, switchgrass and willow. Fuel 86, 1560–1569 10.1016/j.fuel.2006.11.030.

3.2 Moving Bed

3.2.1 Technology description

Moving bed biomass boilers are versatile plant capable of successful operation in a range of applications and capacities. Available on the market in a wide range of construction designs, they are commonly referred to as “grate” systems and occasionally generalised as “stoker” boilers to differentiate from fluidised bed systems:

- **A chain grate** comprises a continuous chain type belt constructed from multiple links connected with pins. The chains are mounted on sprockets at the front and rear of the boiler. Fuel is fed onto the grate by gravity from a hopper at the front of the boiler and the grate transports the fuel away from the feeder. Primary combustion air is fed through the grate and acts as a coolant for the grate. Issues may arise when it is necessary to preheat the combustion air so that wet fuels may be burned. Preheating the air means that its cooling effect is limited and the grate is more vulnerable, especially if the bed of fuel or ash is insufficient to protect the grate from radiant heat in the furnace. The fuel dries out at the front end of the grate and is considered fully burned by the time it reaches the back of the boiler. The ash falls off the end and discharges into a waste collection system. As a high proportion of the combustible content of wood and other biomass fuels burns in the form of volatile compounds, it is necessary that a large proportion of the combustion air requirement is added above the fuel in stoker and fluidised bed boilers as 'overfire air'. The use of preheated combustion air will reduce the time required for fuel drying prior to ignition and may be essential to spreader stoker combustion systems.
- **A travelling grate** is similar to a chain grate but differs in that grids or inserts of cast iron are mounted on carrier bars to form the grate, and are driven by an endless chain. This means that the drive is not effected through the grate so that allowance for expansion can be more easily made and individual grids may be replaced without disturbing the drive. The fuel can be gravity fed as per the grate stoker or may use a mechanical mechanism for sprinkling the fuel across the grate: the action of distributing the fuel over the grate is referred to as a ‘spreader stoker’.
- **A reciprocating/stepping grate** comprises a number of sets of grate bars that move to and fro in sequence. The grate is inclined downwards so that the movement of the grate causes the fuel to move downwards. Primary air passes up through the gaps between the grate bars. Ash is discharged from the end of the grate but some particles tend to get through the grate and may be separately collected below the grate.
- **A water cooled grate** is a type of reciprocating/stepping grate in which the grate bars are maintained at the required working temperature by passing water through the grate. The passage of water through the grate reduces the damage to the grate that may be caused by excessive air, temperature or by too much radiant heat.
- **Vibrating grates** are considered the most commonly used grate in the larger scale, new applications. Compared to traveling and chain grate stokers, vibrating grates require substantially less maintenance and have lower excess air requirements, which improve boiler efficiency and emissions.

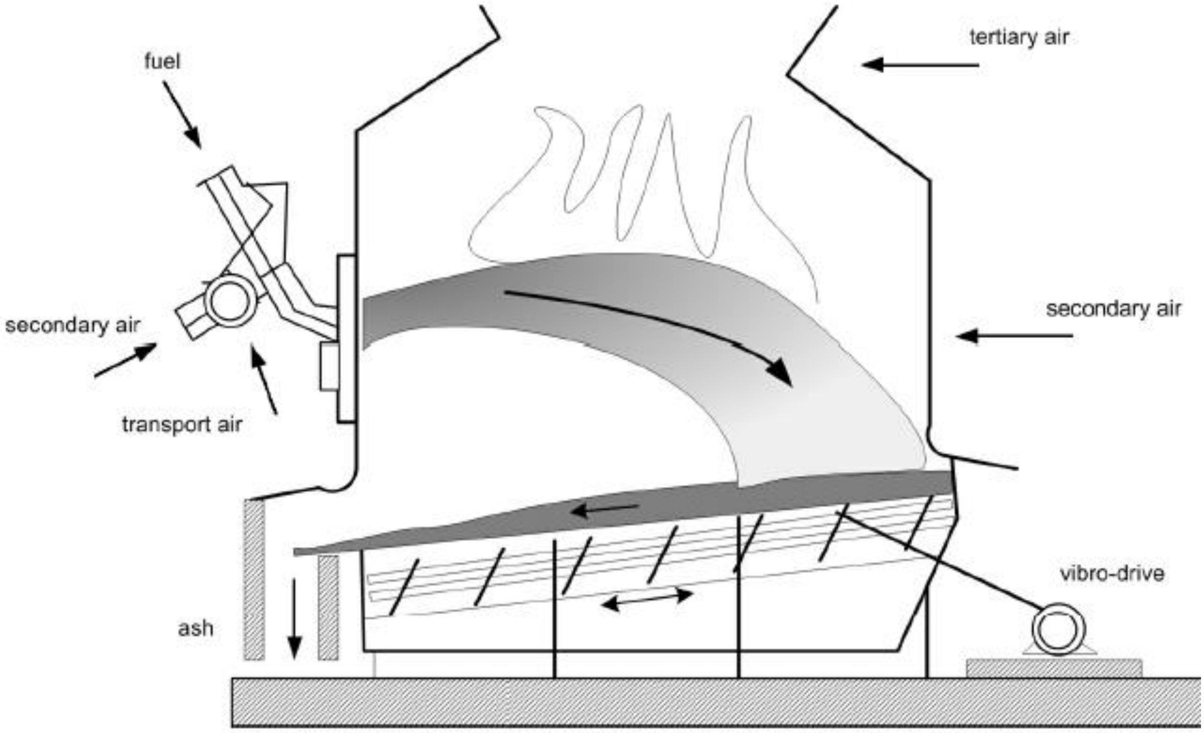

Figure 15: Vibrating Grate Stoker, derived from[27]

Moving Bed Power

Moving bed biomass combustion systems for power raise steam from the combustion gases created in the furnace and operate in tandem with a steam turbine. All moving bed boiler designs are compatible with power generation and are generally limited to plant outputs not greater than 50 MW_e due to their inability to feed large quantities of fuel onto a grate. For power only application the bed would be integrated with a condensing steam turbine with a steam cycle focused on high efficiency electricity generation.

A modern steam cycle power plant using biomass will have an overall electrical generation efficiency in the range of c.20-40% depending upon scale (LHV basis). Furthermore, biomass plants commonly have lower efficiencies than modern fossil plants. The efficiency is lower because of the smaller scale of the plants and the higher moisture content of the biomass fuel compared to fossil fuel.

Moving Bed CHP

The primary use for the application of a moving bed is for power generation; if the design requires the use of CHP then it is because of the need to combine the output with a suitable heat requirement. Steam turbines used in CHP systems are usually either back-pressure systems or extraction-condensing systems.

With a steam cycle power plant, heat at sufficiently high temperature to be useful for district heating or process demands can only be taken at the cost of power output and the electrical efficiency may drop to c. 15% (steam flow dependant).

CHP plants are likely to be smaller, using simpler steam cycles and less extreme steam conditions than conventional electricity generating plant. The emphasis will be more on achieving high plant availability and low construction cost rather than small gains in the efficiency of electricity generation. Even though the CHP electrical efficiency is low, the effective overall system efficiency of steam turbine systems is generally high because of the recovered heat and typically ranges from 55-70%.

Moving Bed Heat

This biomass technology is used for integration into an existing steam or hot water process demand. Systems are typically limited in their thermal output, up to circa 10 MW_{th}, since larger units would generally operate in CHP mode. Moving bed heating systems are typically preferred to underfeed stoker systems at larger scales. The overall thermal boiler conversion efficiency is typically 70-85% (based upon LHV), with the lower efficiencies realised when using high moisture content fuels and without the use of heat recovery systems on the boilers (e.g., air heaters or economisers).

3.2.2 Development status and timescales

Moving bed boilers are a well-developed technology and have evolved over more than 100 years. Consequently, their operation is well understood, robust and reliable. Moving bed technology has been successfully utilised in a range of output mediums and capacities capable of consistent production of hot water and low grade steam for commercial and industrial process heating or high grade steam for power generation via a steam turbine generator. Mass commercial deployment places this technology at TRL 9.

Typically all moving bed applications are limited to plant outputs not greater than 50 MW_e due to their inability to effectively feed larger quantities of fuel onto a grate.

Unit outputs for heat only applications beyond 10 MW_{th} thermal rating are available but typically not normally economic or required for the provision of heat-only demands. Above this size range, it is common for operators to switch to moving bed plant raising high-grade steam suitable for electricity production and then extract the heat energy at a suitable grade in a CHP arrangement. CHP size limits are under 50 MW_e to allow for economic supply of sufficient thermal energy supply.

At the minimum heat output range, the introduction of the UK non-domestic renewable heat incentive (RHI) in 2011 incentivised the demand for boilers below 1 MW_{th} thermal output. Many European boiler manufacturers (in particular Austrian manufacturers) have a mature moving bed boiler product to service the commercial and small-industrial sectors over the range of 0.4 MW_{th} to 5 MW_{th} thermal output. This technology competes with underfed stoker boiler technology at the lower output range.

The range of timescales to develop a boiler system through the overall construction period to full operation would be in the order of 2 to 2.5 years for the largest scale systems and for the smaller heat only system this would reduce to 6 to 9 months.

3.2.3 Impact of different feedstock parameters on operation and cost

General feedstock considerations

The majority of biomass materials of industrial interest have key chemical characteristics which have an influence on the high temperature corrosion processes that any biomass combustion system must be designed to withstand, as described in section 2.4.2.

Consequently, suppliers of standard biomass boilers specify acceptable fuels for use in their boilers and establish a maintenance regime to deal with any fouling issues. The majority of boilers are specified for virgin woodchip or wood pellets whilst systems able to tolerate herbaceous fuels such as Miscanthus are much less common for heat only applications, particularly at the smaller scale (<1 MW_{th} heat output).

Specific feedstock considerations

In Europe, experience with combustion of straw (considered to be similar to Miscanthus in terms of combustion properties) has shown that feedstock type variation cannot readily be implemented. Experience has shown that to fire straw effectively, a specifically designed stoker incorporating a “slagging superheater” has to be employed. In this design, a layer of molten slag is continuously built up on the heat transfer surface of the boiler, which mitigates corrosion concerns. This design would not be suitable for combustion of woody biomass because sufficient slag would not be generated, and temperatures within the furnace would exceed design limits for the moving bed. To fire both woody biomass and herbaceous biomass such as Miscanthus on a conventional moving bed grate would require significantly lower steam temperatures (c. 450°C or less). It is anticipated that such moving bed boilers would be much less efficient than wood-only systems.

Herbaceous biomass fuels such as Miscanthus typically contain relatively high quantities of alkali (i.e., potassium and phosphorous). This elevated alkali content significantly increases the corrosion and slagging potential within combustion systems. Also Miscanthus has a higher chlorine content that the other intended fuels listed above and this will also increase corrosion on the internal parts of the boiler system.

The handling and combustion of Miscanthus at a facility designed to primarily fire woody biomass fuels is likely to require the installation of dedicated material handling and storage equipment for Miscanthus fuels and likely to alter the design and operation of combustion systems to allow the firing of both woody fuels and Miscanthus.

It is anticipated the Miscanthus fuels would be delivered to the biomass facility in the form of bales or pellets. If Miscanthus is provided in baled form, the baled biomass must be unloaded by forklift or specially designed equipment rather than dumped in receiving hoppers. Baled biomass must be stored in covered stacks rather than a reclaimable pile. The footprint of baled biomass stacks is greater than that of a biomass pile with similar energy content. Baled biomass must be conveyed by specially designed bale conveyors rather than conventional conveyor systems typically employed woody biomass.

If Miscanthus is provided in pellet form, conventional receiving equipment and conveyors may be suitable for the feedstock. However, Miscanthus pellets are likely to require separate storage systems (e.g., silos) with dedicated systems for pellet reclaim and delivery to the boiler.

The moving bed boiler requires the feedstock to be sized. Depending on the manufacturer, the maximum size of the fuel may range from 75 to 150mm but more normally the maximum size would be 75mm in larger units. Heat only systems are suited to smaller outputs and often have a more constrained fuel specification dictated by the boiler supplier. This is to ensure that the boiler operation and performance is not compromised, and is acceptable for RHI accreditation (see underfed stoker section for further details for RHI details).

Nitrogen oxide emissions from a new moving bed boiler burning biomass can vary significantly because of factors such as the type of biomass being burned, the moisture content of the biomass, temperature on the grate, and quantity of primary air. Due to the less uniform combustion with a grate than with a fluid bed, there will be more unintended by-products of combustion i.e. CO, NO_x, VOC etc. as well as the need for greater total excess air which will also result in an absolute reduction in in boiler efficiency for the stoker in the order of 1-4% compared to BFB.

As with all biomass combustion plants, the availability is primarily dependent on the quality of feedstock meeting the OEM’s specification, rather than the feedstock itself. For moving bed systems, low ash content feedstock means the bed is exposed to furnace temperature that can reduce availability. The plant’s lifetime is dependent on the operator carrying out operation and maintenance in accordance with the OEM’s instructions. There will generally be no difference in lifetime between the feedstocks listed.

Sensitivity Analysis - Power

The cost curve in Figure 16 corresponds to the total investment cost per unit of output capacity for a moving bed power plant. It has been derived from quotes provided by suppliers (2010 – 2014) for all-in overnight EPC costs.

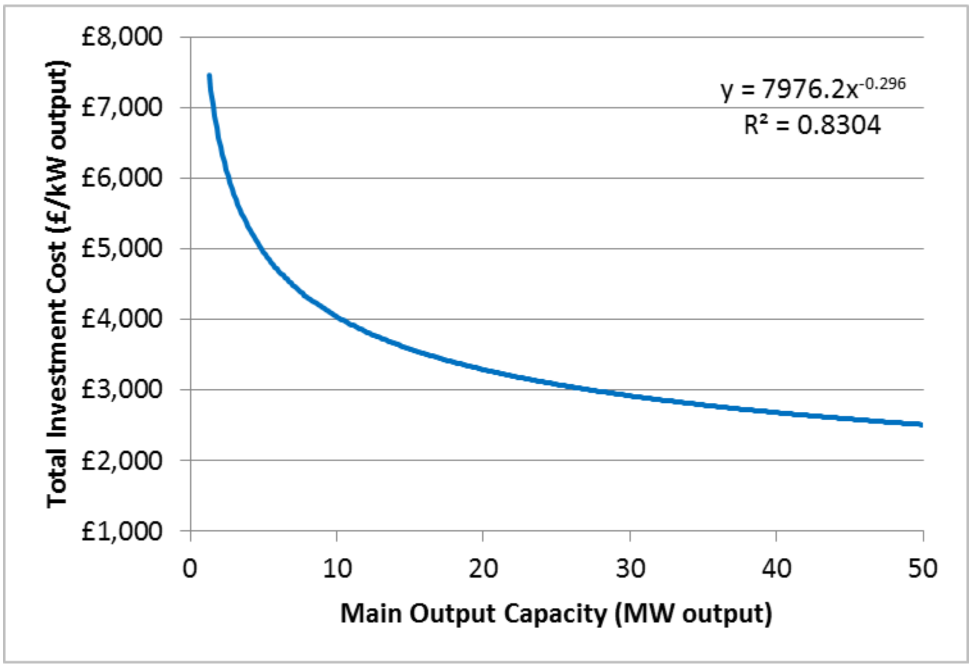

Figure 16: Moving bed (power) total investment cost vs. output power capacity (derived from four supplier quotes (2010 - 2014) and B&V data for all-in overnight EPC costs)

Total operational costs for moving bed power technologies include fixed costs (insurance, maintenance parts and labour) and variable costs (operations labour, additional fuel, reagents, water). Figure 17 shows the relationship between annual total operational cost for a moving bed per unit of energy output and the main output capacity rating.

Except where the parameter of interest is shown on the x axis, the curves in Figure 17 to Figure 24 are all created using the following base values: Main output capacity = 20 MW_e; station capacity factor of 85%; feedstock moisture content (WB) = 50%; feedstock ash content (DB) = 1.5%; feedstock nitrogen content (DB) = 0.35%; feedstock sulphur content (DB) = 0.05%; and a feedstock chlorine content (DB) = 0.04%.

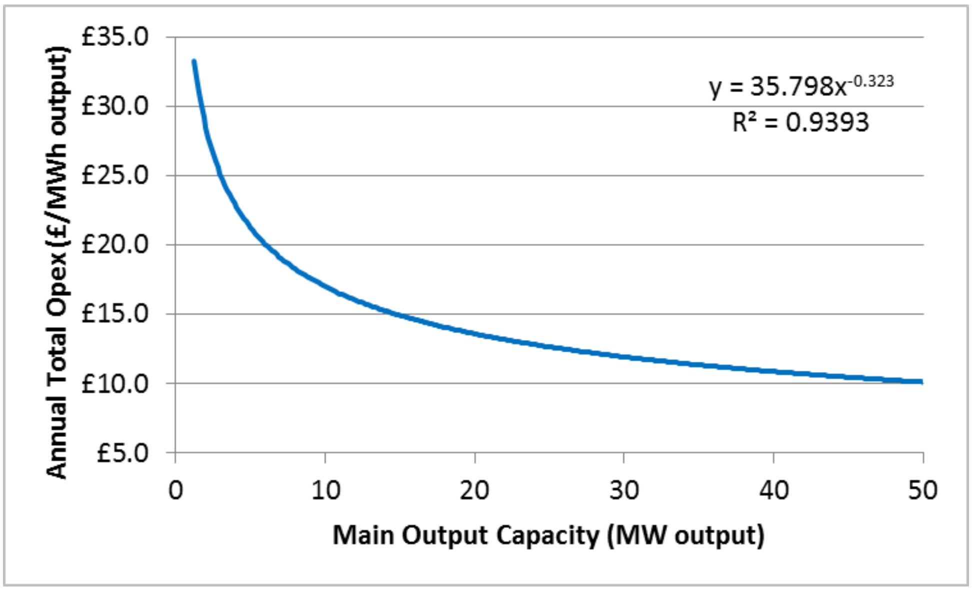

Figure 17: Moving bed (power) annual operating cost per unit of electricity output vs. output power capacity (derived from B&V industry data)

Figure 18 below highlights the change in overall moving bed station efficiency (LHV) over varying feedstock moisture content. The efficiency is calculated as total electrical energy output over fuel energy input, as such parasitic loads post-generation are not included. The curve is considered to be a representative average across the moving bed potential rated output range.

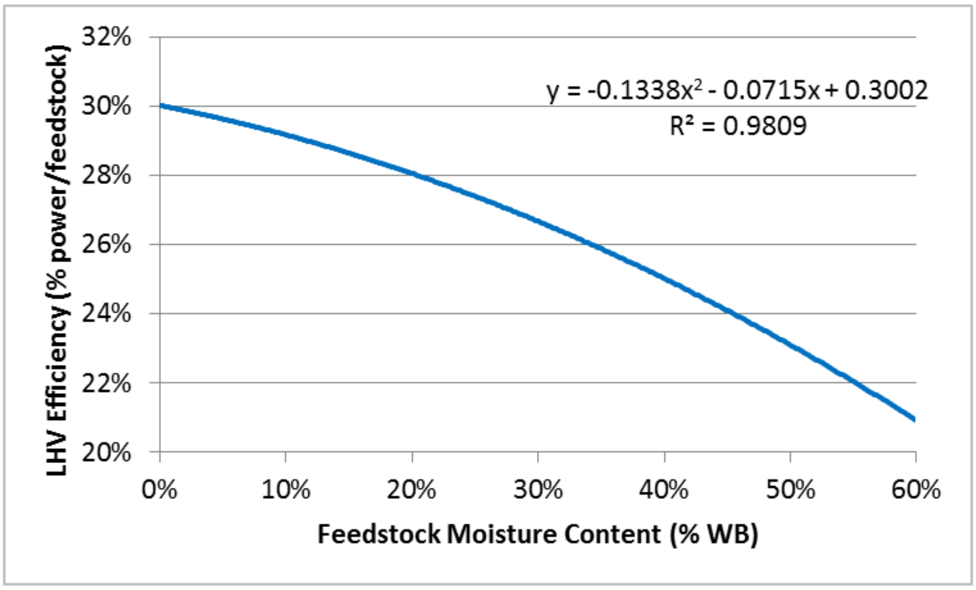

Figure 18: Moving bed (power) efficiency vs. feedstock moisture content (derived from B&V industry data, considering a representative average across the technology potential rated output range)

The total operational costs for moving bed power technologies (as previously described for Figure 17) are related to feedstock moisture content, as shown in Figure 19 below.

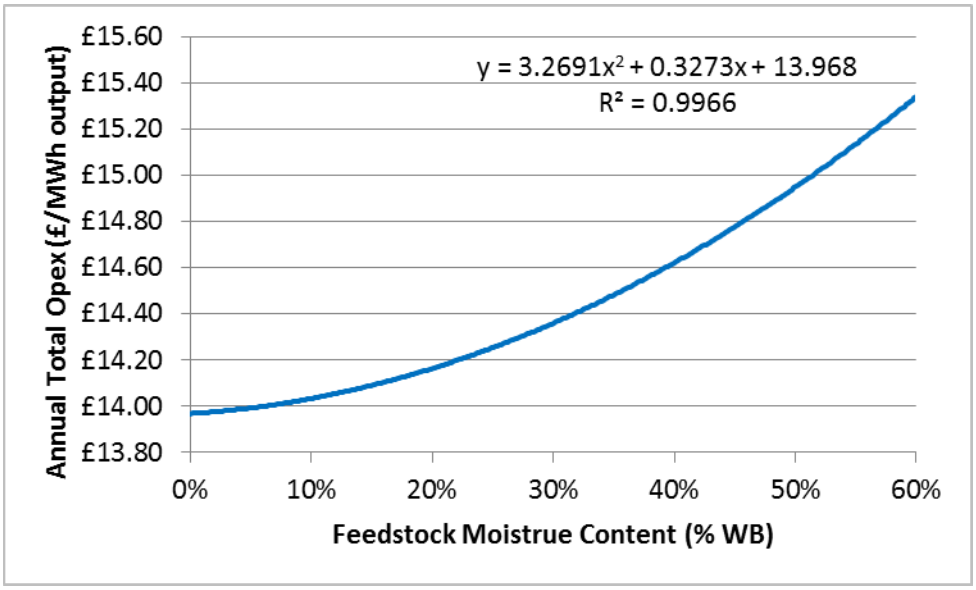

Figure 19: Moving bed (power) annual operating cost per unit of electricity output vs. feedstock moisture content (derived from B&V industry data)

Annual operational costs for moving bed power stations vary with the percentage of ash present within the feedstock, as shown in Figure 20 below.

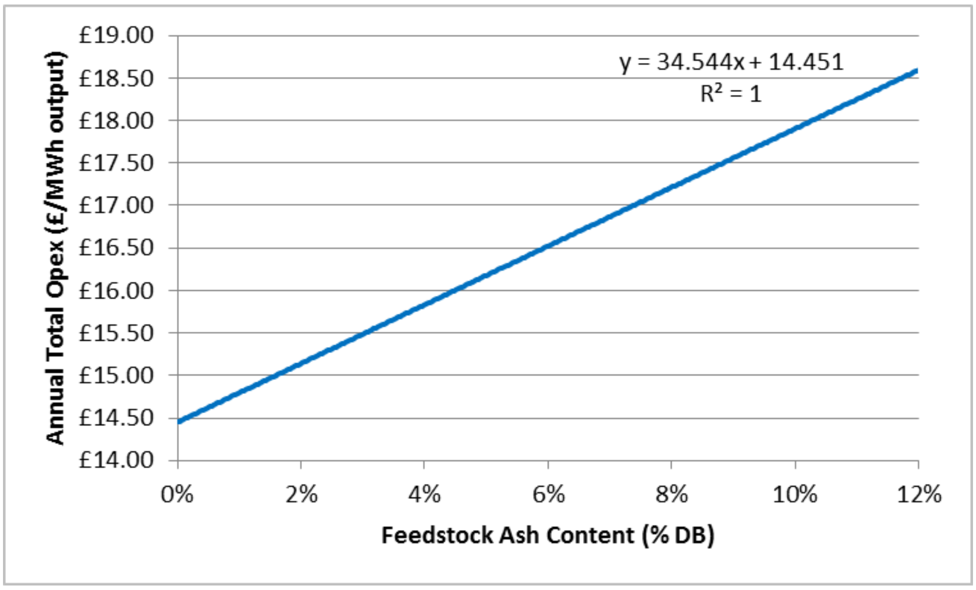

Figure 20: Moving bed (power) annual operating cost per unit of electricity output vs. feedstock ash content (derived from B&V industry data, and calculation based on section 2.4.2.3)

Figure 21 below displays the relationship between nitrogen content within the biomass feedstock and the annual cost of NO_x abatement via urea injection.

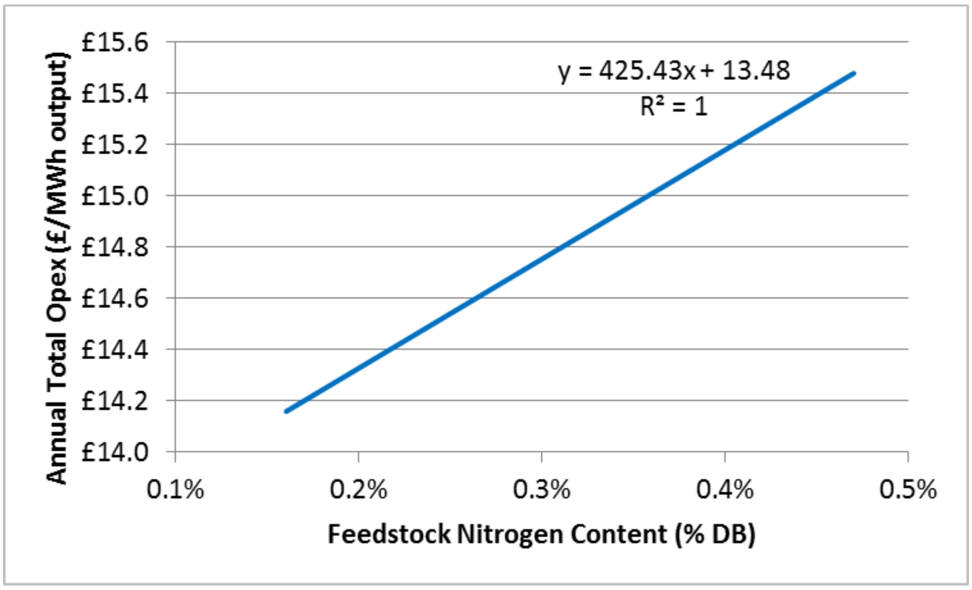

Figure 21: Moving bed (power) annual operating cost per unit of electricity output vs. feedstock nitrogen content (calculation based on section 2.4.2.4)

Figure 22 below displays the relationship between sulphur content within the biomass feedstock and the annual cost of abatement via lime addition.

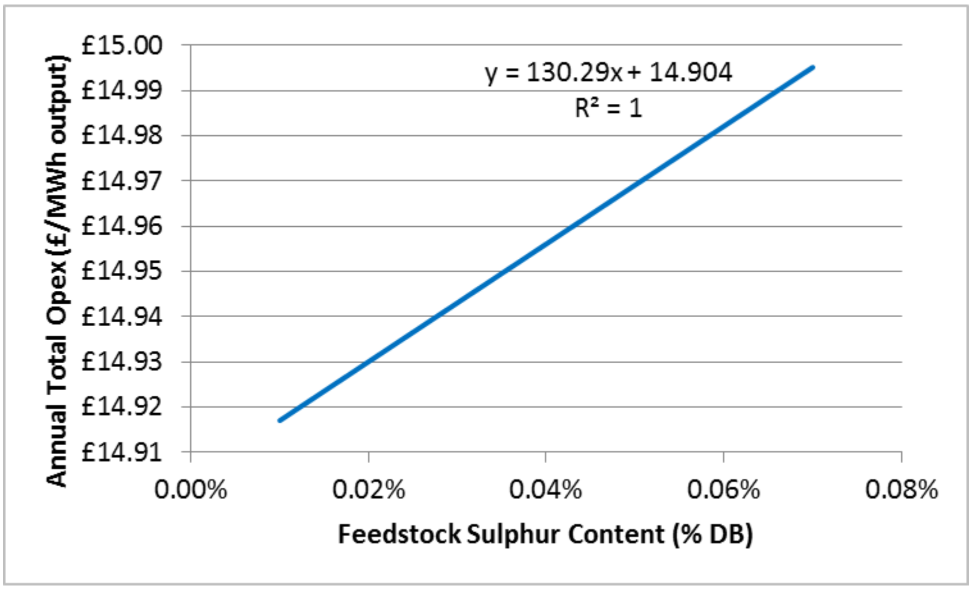

Figure 22: Moving bed (power) annual operating cost per unit of energy output vs. feedstock sulphur content (calculation based on section 2.4.2.4)

Figure 23 below displays the relationship between chlorine content within the biomass feedstock and the annual cost of abatement via lime addition.

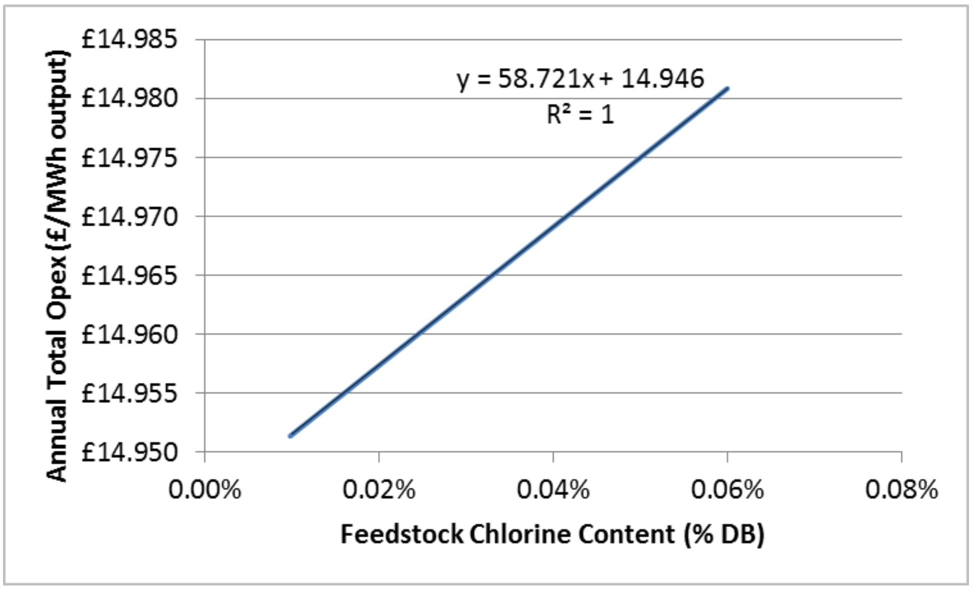

Figure 23: Moving bed (power) annual operating cost per unit of energy output vs. feedstock chlorine content (calculation based on section 2.4.2.4)

Figure 24 below displays the relationship between the alkali index (kg K_2O and Na_2O per GJ energy) and the annual total operational cost for a moving bed per unit of energy output.

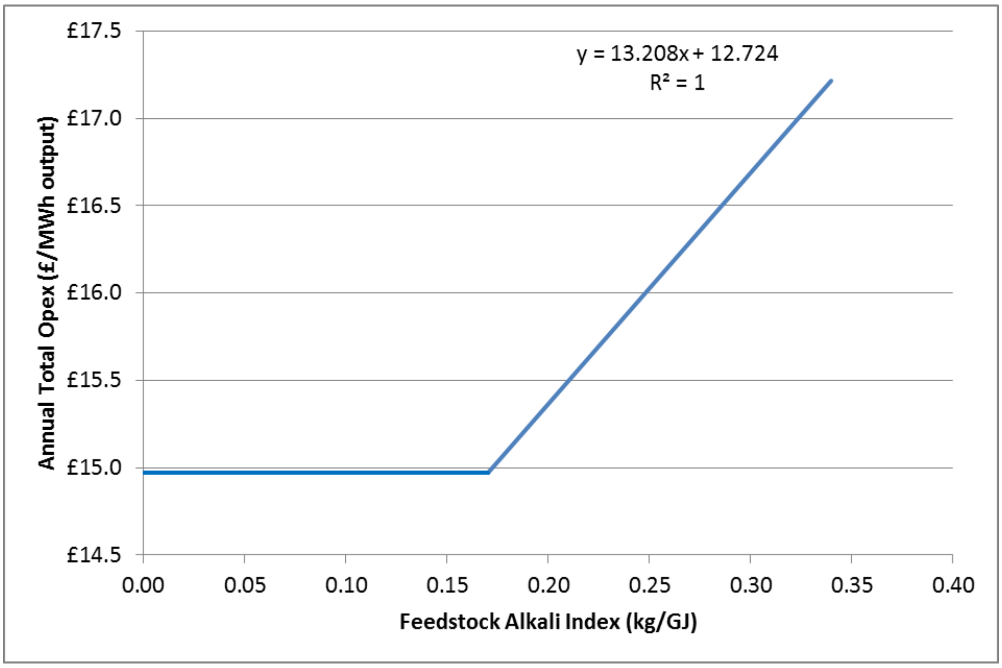

Figure 24: Moving bed (power) annual operating cost per unit of energy output vs. feedstock alkali index (derived from formula based on [28] and section 2.4.2.5)

Sensitivity Analysis - CHP

Due to the inability to export heat over long (inter-city) distances, the thermal output capacity of a plant is typically dependent on the site local demand profile. To optimise financial returns, plant is often operated to maximize returns from electrical generation with the heat utilized wherever possible. Heat-to-power ratios can range from 0.6:1 to 10:1[29] and therefore the amount of useful heat available for capture from the power generation process is considered too variable across sites to offer meaningful comparison and therefore the following sensitivities are on an electrical power output basis.

The cost curve in Figure 25 corresponds to the total investment cost per unit of electrical output capacity for a moving bed CHP plant. It has been derived from quotes provided by suppliers (2010 – 2014) for all-in overnight EPC costs.

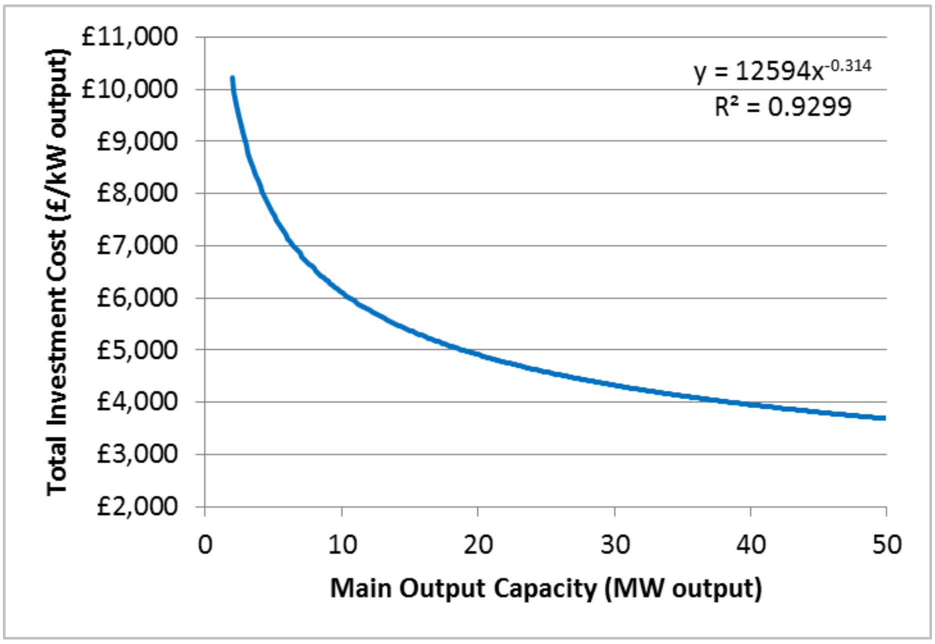

Figure 25: Moving bed (CHP) total investment cost vs. main output electrical capacity (derived from four supplier quotes (2010 - 2014) and B&V data for all-in overnight EPC costs)

Total operational costs for moving bed CHP technologies include fixed costs (insurance, maintenance parts and labour) and variable costs (operations labour, additional fuel, reagents, water). Figure 26 shows the relationship between annual total operational cost for a moving bed per unit of electrical energy output and the main electrical output capacity rating.

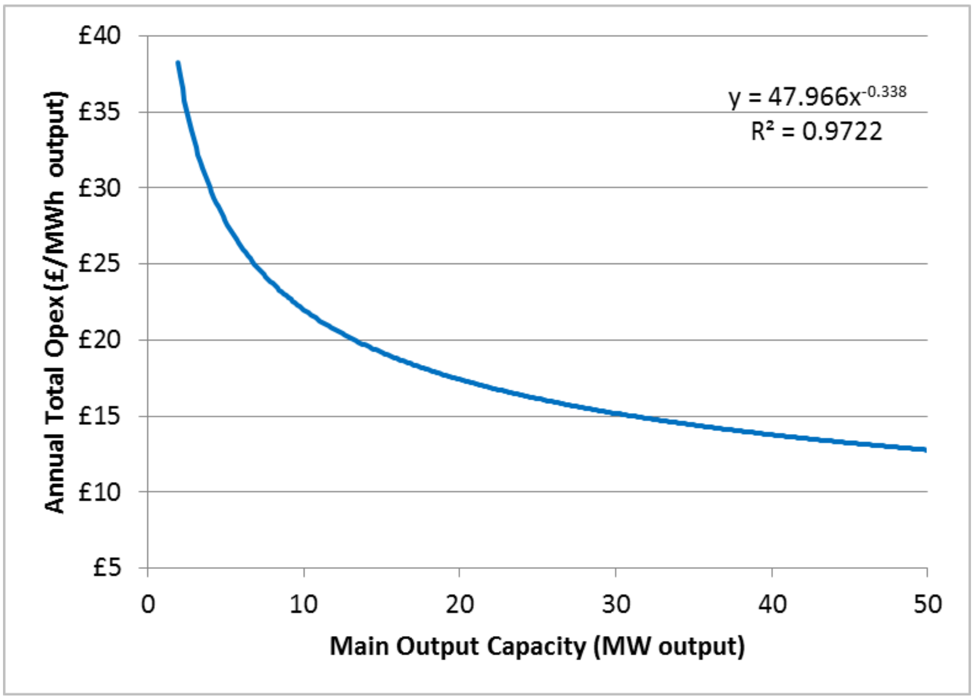

Figure 26: Moving bed (CHP) annual operating cost per unit of electricity output vs. power output capacity (derived from B&V industry data)

Except where the parameter of interest is shown on the x axis, the curves in Figure 26 to Figure 33 are all created using the following base values: Main output capacity = 16 MW_e; station capacity factor of 85%; feedstock moisture content (WB) = 50%; feedstock ash content (DB) = 1.5%; feedstock nitrogen content (DB) = 0.35%; feedstock sulphur content (DB) = 0.05%; and a feedstock chlorine content (DB) = 0.04%.

Figure 27 below highlights the change in electrical efficiency (LHV) of a moving bed CHP over varying feedstock moisture content. The efficiency is calculated as total electrical energy output over fuel energy input, as such parasitic loads post-generation are not included. As stated previously, the thermal efficiency is dependent on the amount of useful heat captured and is considered site specific and therefore too variable for further comparison.

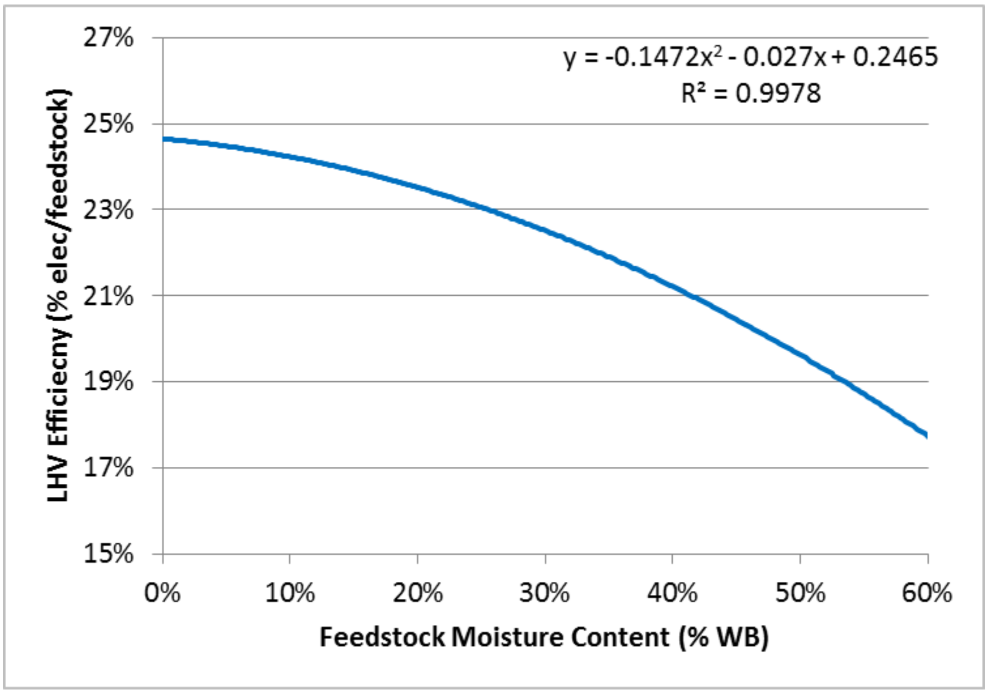

Figure 27: Moving bed (CHP) electrical efficiency vs. feedstock moisture content (derived from B&V data, considering a representative average across the technology potential rated output range).

The total operational costs for moving bed CHP technologies (as previously described for Figure 26) are related to feedstock moisture content as shown in Figure 28 below.

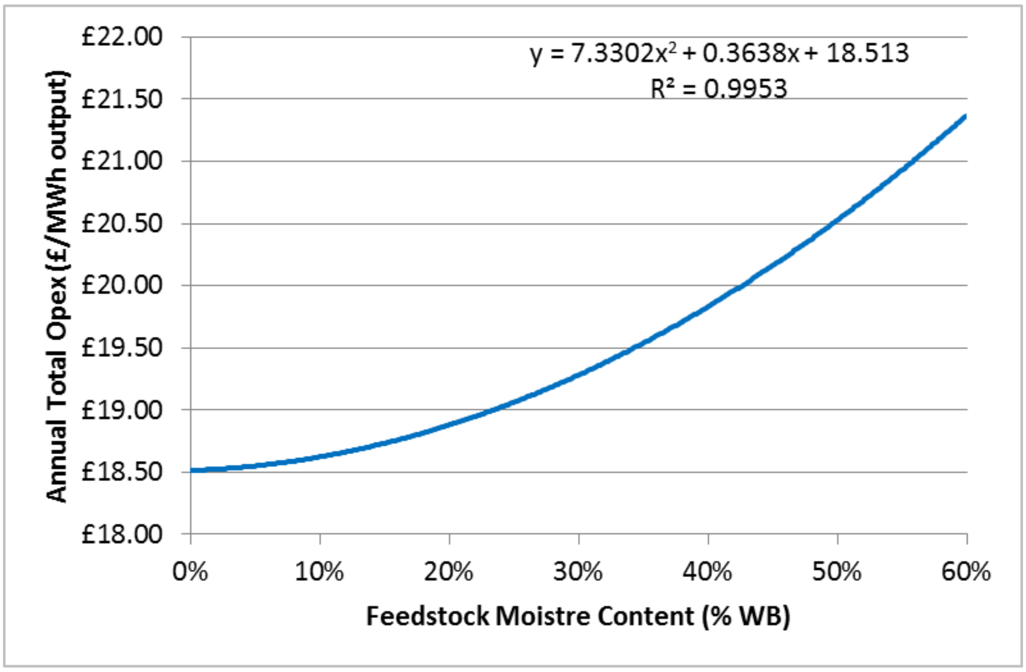

Figure 28: Moving bed (CHP) annual operating cost per unit of electricity output vs. feedstock moisture content (derived from B&V industry data)

As per Figure 29 below, annual operational costs for moving bed CHP stations vary with the percentage of ash present within the feedstock.

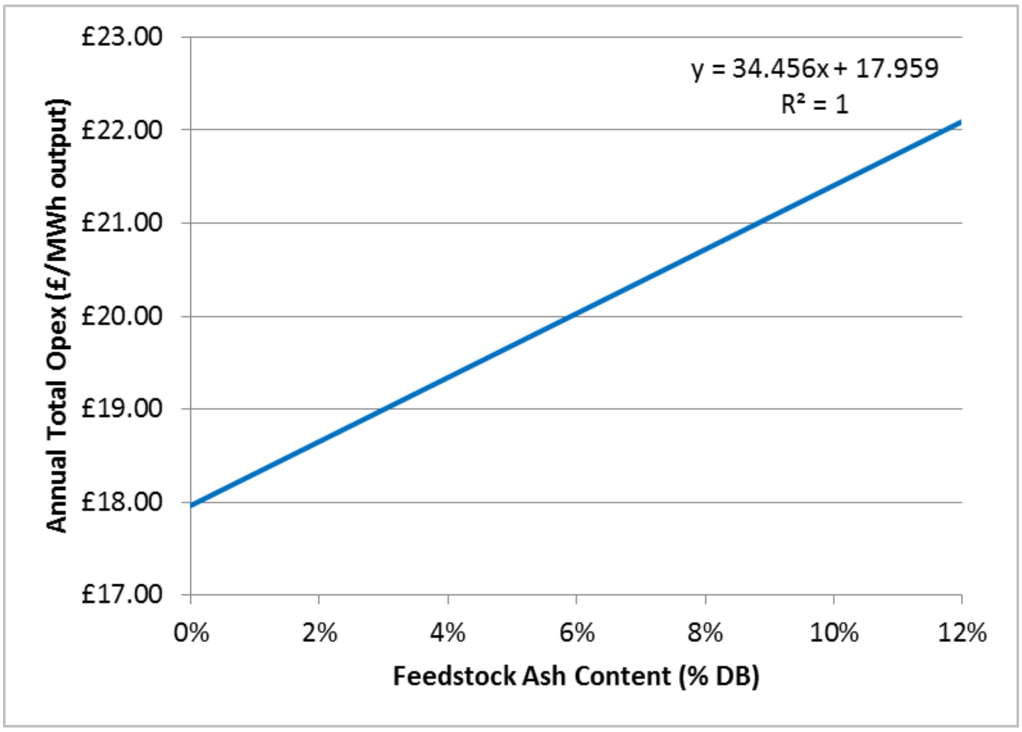

Figure 29: Moving bed (CHP) annual operating cost per unit of electricity output vs. feedstock ash content (derived from B&V industry data)

Figure 30 below displays the relationship between nitrogen content within the biomass feedstock and the annual cost of NO_x abatement via urea injection.

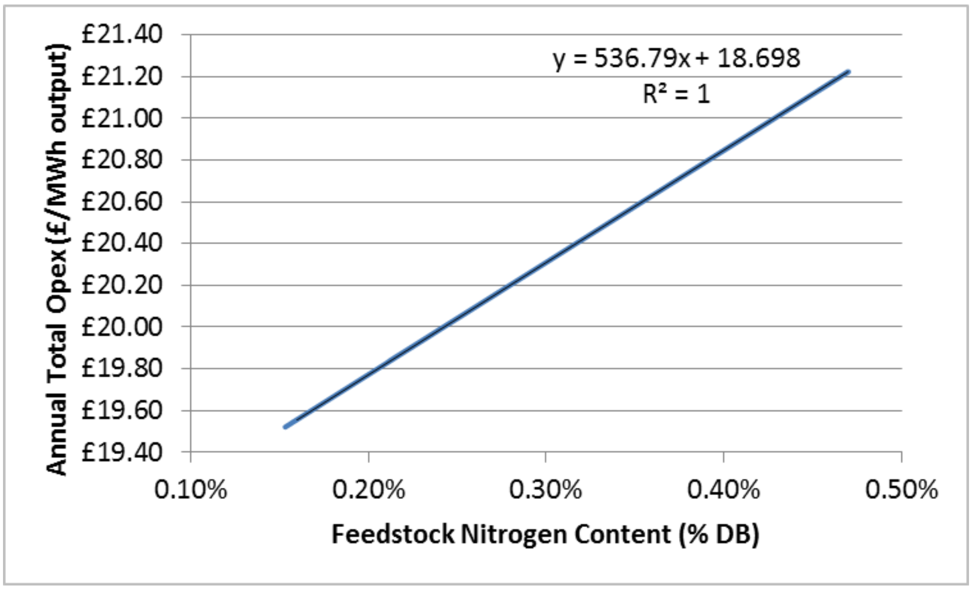

Figure 30: Moving bed (CHP) annual operating cost per unit of electricity output vs. feedstock nitrogen content (calculation based on section 2.4.2.4)

Figure 31 below displays the relationship between sulphur content within the biomass feedstock and the annual cost of abatement via lime addition.

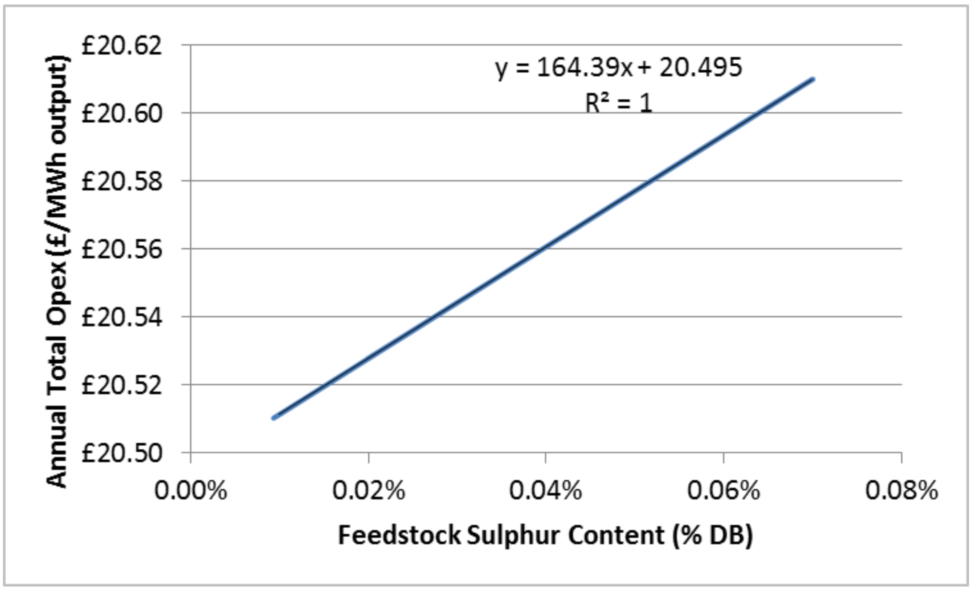

Figure 31: Moving bed (CHP) annual operating cost per unit of electricity output vs. feedstock sulphur content (calculation based on section 2.4.2.4)

Figure 32 below displays the relationship between chlorine content within the biomass feedstock and the annual cost of abatement via lime addition.

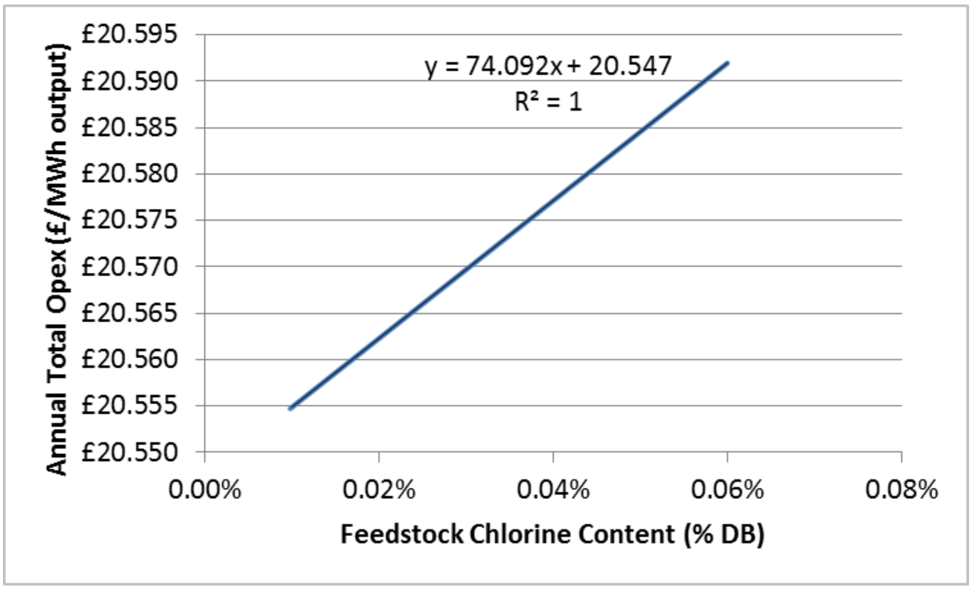

Figure 32: Moving bed (CHP) annual operating cost per unit of electricity output vs. feedstock chlorine content (calculation based on section 2.4.2.4)

Figure 33 below displays the relationship between the alkali index (kg K_2O and Na_2O per GJ energy) and the annual total operational cost for a moving bed per unit of energy output.

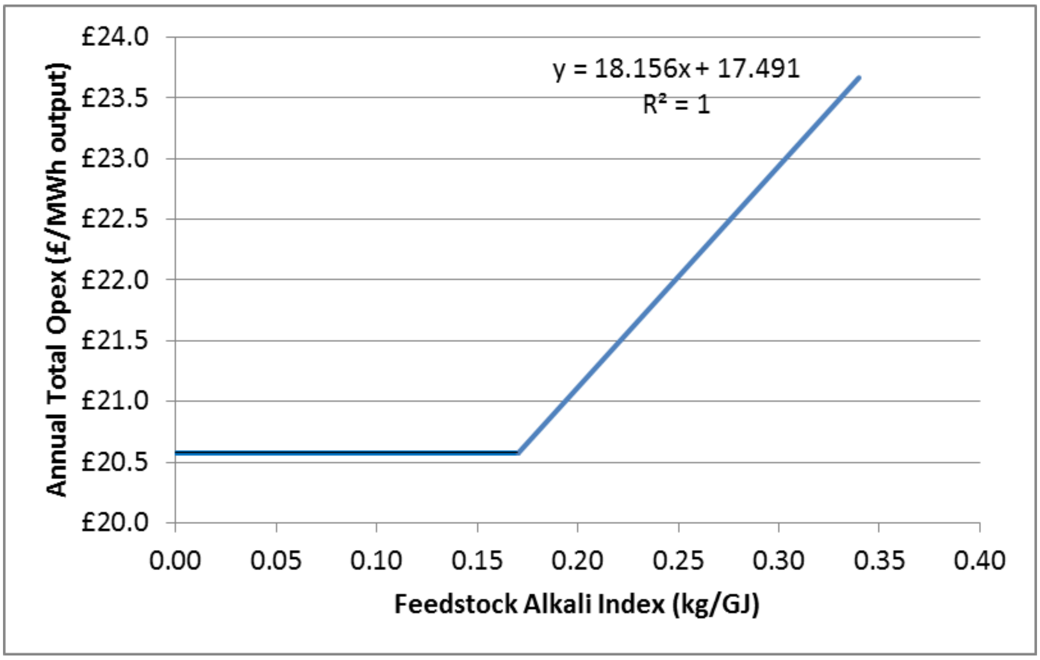

Figure 33: Moving bed (CHP) annual operating cost per unit of electricity output vs. feedstock alkali index (derived from formula based on [28] and section 2.4.2.5)

Sensitivity Analysis - Heat

The cost curve in Figure 34 corresponds to the total investment cost per unit of output capacity for a moving bed heat boiler. It has been derived from quotes provided by suppliers (2009 – 2014) for all-in overnight EPC costs.

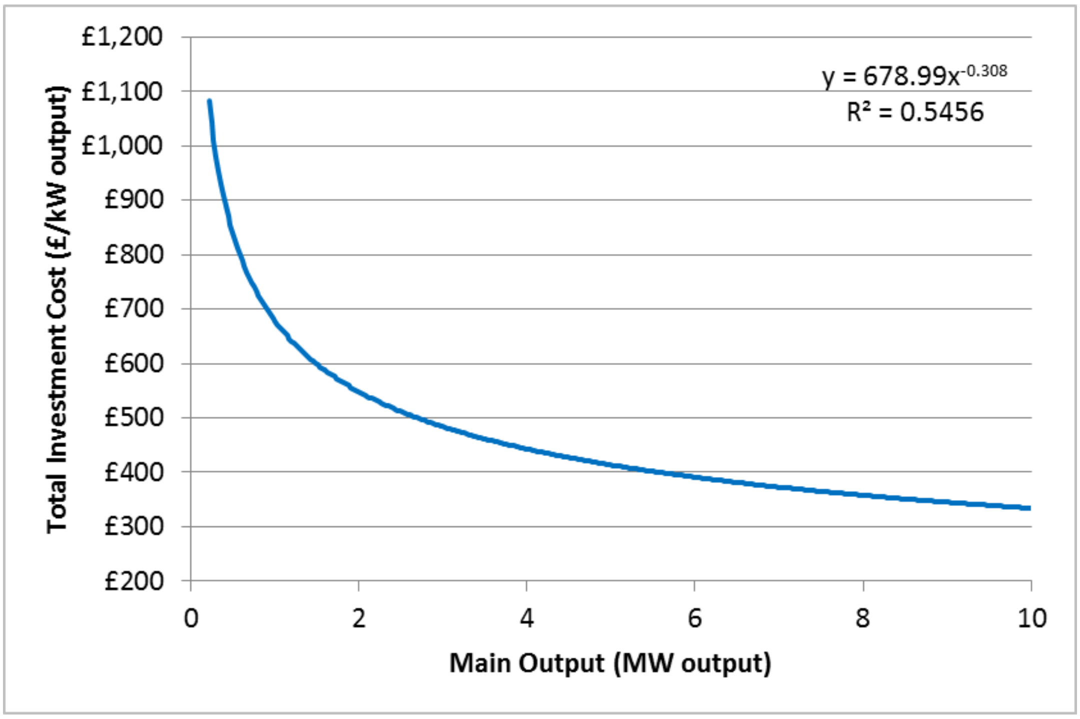

Figure 34: Moving bed (heat) total investment cost vs. output heat capacity (derived from 25 supplier quotes (2009 - 2014) and B&V data for all-in overnight EPC costs).

Total operational costs for moving bed heat technologies includes fixed costs (insurance, maintenance parts and labour) and variable costs (operations labour, additional fuel, chemical dosing). For LTHW/MTHW boilers it is considered that water use is negligible and that the operating thermal output range does not require the addition of reagents. Figure 35 shows the relationship between annual total operational cost per unit of heat output and the main output capacity rating for a moving bed (heat) system.

Except where the parameter of interest is shown on the x axis, the curves in Figure 35 to Figure 39 are all created using the following base values: Main output capacity = 1 MW_{th}; station capacity factor of 60%; feedstock moisture content (WB) = 50%; feedstock ash content (DB) = 1.5%; feedstock nitrogen content (DB) = 0.35%; feedstock sulphur content (DB) = 0.05%; and a feedstock chlorine content (DB) = 0.04%.

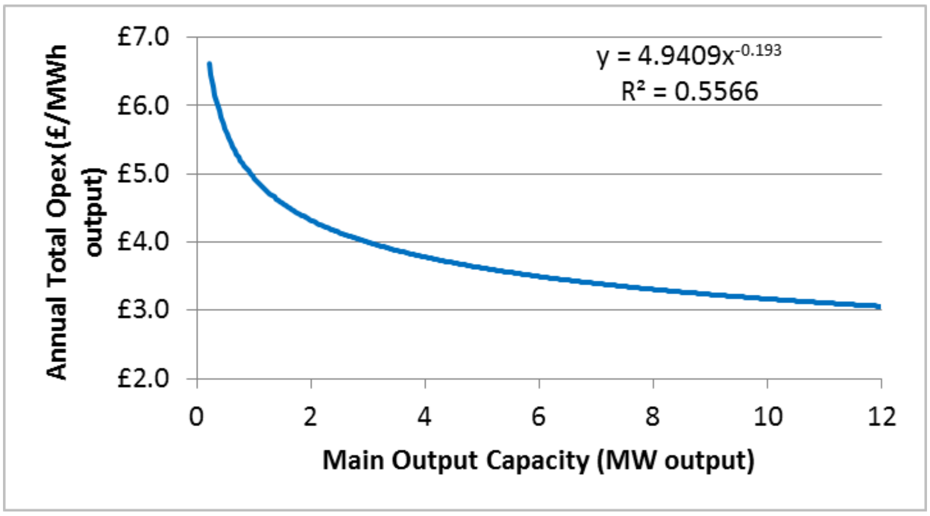

Figure 35: Moving bed (heat) annual operating cost per unit of heat output vs. output heat capacity (derived from B&V industry data)

Figure 36 below highlights the change in overall moving bed boiler efficiency (LHV) over varying feedstock moisture content. The efficiency is calculated as total heat energy output over fuel energy input. The curve is considered to be a representative average across the moving bed potential rated output range.

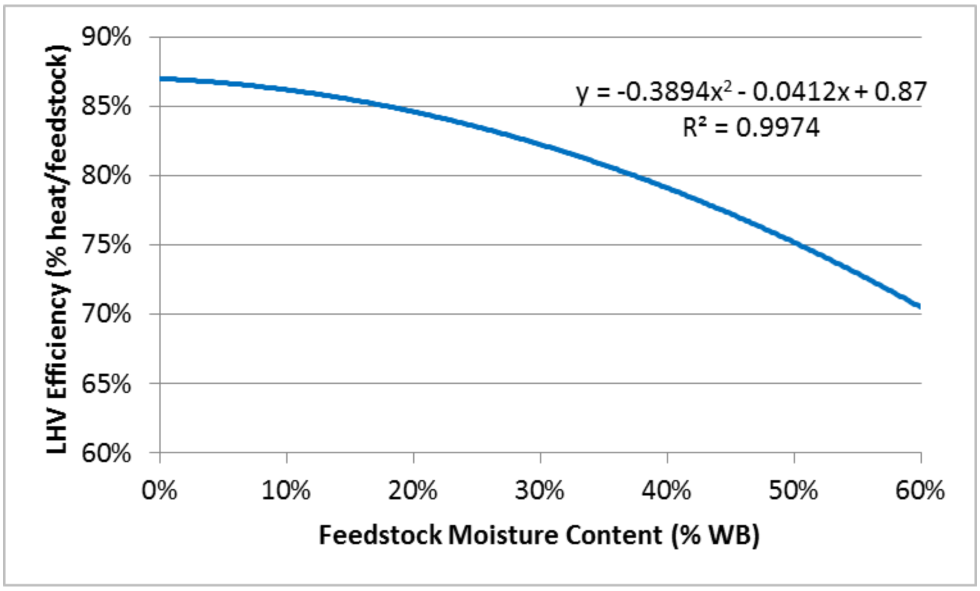

Figure 36: Moving bed (heat) efficiency vs. feedstock moisture content (derived from B&V industry data considering a representative average across the technology potential rated output range).

The total operational costs for moving bed heat technologies (as previously described for Figure 35) are related to feedstock moisture content as shown in Figure 37 below.

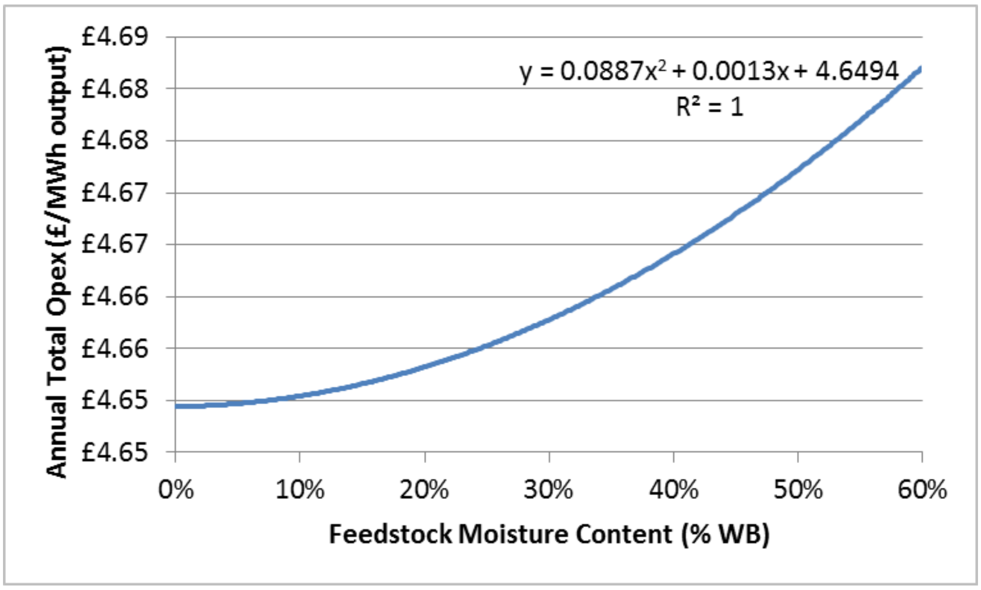

Figure 37: Moving bed (heat) annual operating cost per unit of heat output vs. feedstock moisture content (derived from B&V industry data)

As per Figure 38 below, annual operational costs for moving bed power stations vary with the percentage of ash present within the feedstock.

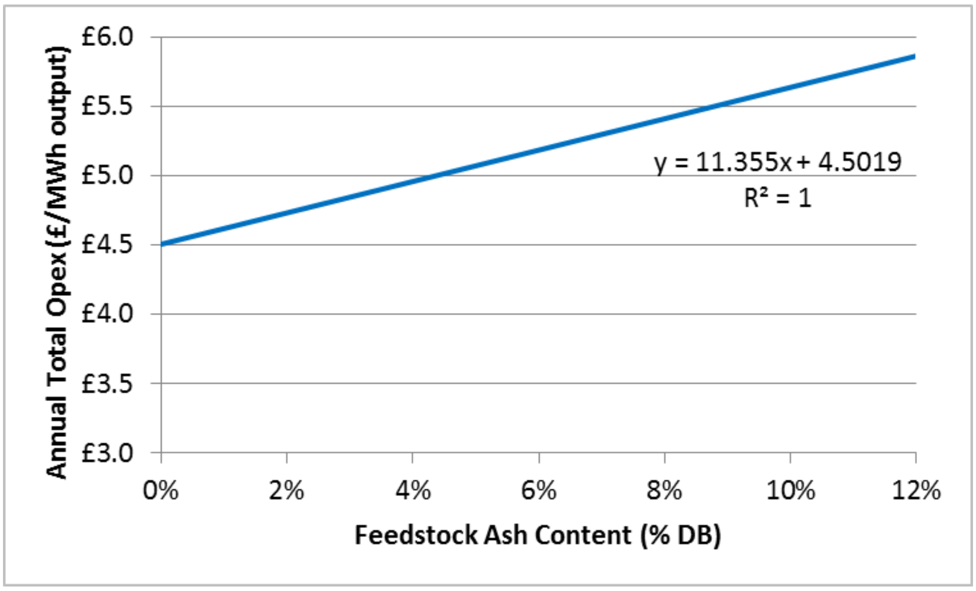

Figure 38: Moving bed (heat) annual operating cost per unit of heat output vs. feedstock ash content (derived from B&V industry data)

As described above, it is considered that reagent use is negligible across the output range available for heat only boilers. Figure 39 below displays the relationship between the alkali index (kg K_2O and Na_2O per GJ energy) and the annual total operational cost for a moving bed per unit of energy output.

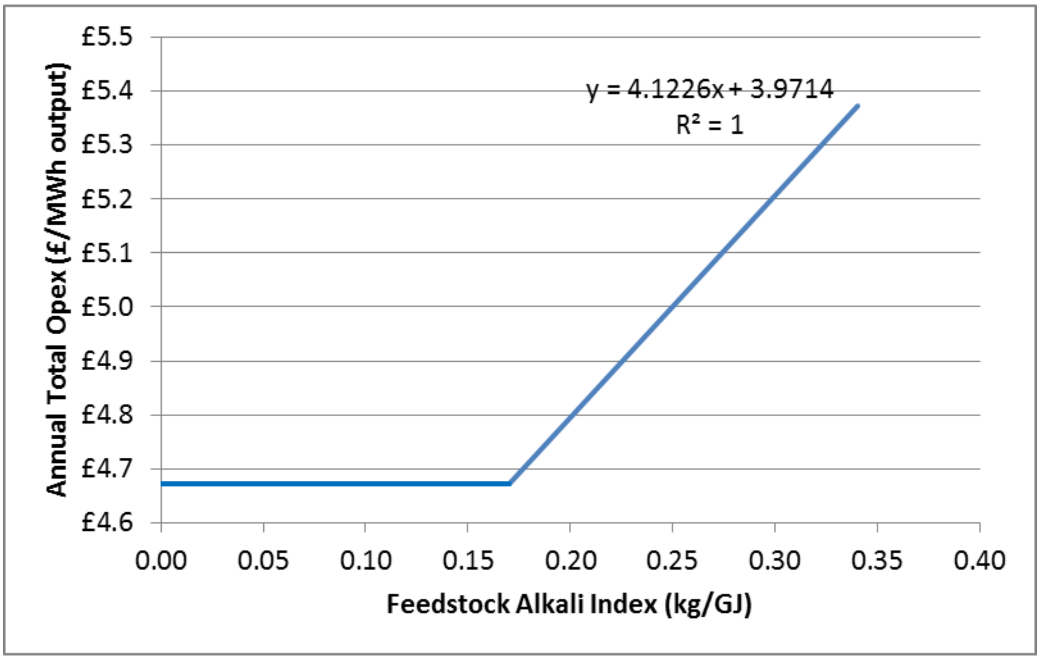

Figure 39: Moving bed (heat) annual operating cost per unit of heat output vs. feedstock alkali index (derived from formula based on [28] and section 2.4.2.5)

3.2.4 Available options for improvement

Grate fired boilers are a well-developed technology and have evolved over 150 years and their design and operation is well understood, robust and reliable. However, they are generally less efficient than more modern technologies such as BFB and CFB, as fuel burn out is dependent on the residence time on the grate and bed depth. Some of the more significant improvements in moving bed boiler design are listed in Table 10 and summarised below.

As the design is well understood, the Capex for such plant is well defined and has been optimised, in the design operating range (below 50 MW_e) the cost will be more competitive that for CFB or BFB boilers. Suppliers of fluidised bed technology would be reluctant to quote for boilers in this size range due the expected non-competiveness unless emission requirements are a major consideration or fuel supplies are unsuitable for grate boilers.

Due to the lower effectiveness of mixing fuel and sorbent prior to or on a grate (compared to that in a fluidised bed), emissions abatement typically comprises flue gas treatment. In order to compete with BFB on final emissions to atmosphere, moving bed boilers may need to incorporate higher cost emission control equipment that could offset the lower grate costs. Improvements to emission control equipment to reduce cost would therefore be advantageous to moving bed cost effectiveness.

Table 10: Improvements Analysis of Moving Bed Boilers

Issue arising due to biomass characteristic	Options to ameliorate	Evaluation of effect	Long Term Improvements
Poor combustion conditions from high moisture content feedstock	Longer grate area; Controllable bed speed; Variable fuel feed rates; Pre-heat primary combustion air.	Control over residence time will increase drying zone and improve burnout of wet fuels; Consistent bed depth across grate; Heated air supply to increase drying time.	Increased tolerance of fuel specs; Improved efficiency; Reduced emissions and environmental impact.
Variable fuel specifications - High fines content; High oversize content.	Secondary emissions control – multicyclone; Avoidance of spreader stoker feed with high fines; Vibrating grate for oversize.	Key to avoid entrainment of fines as danger of fires in baghouse / downstream Recommended to improve fuel screening beforehand to avoid this situation occurring; Reduction of moving parts aids movement of oversize – increasing residence time of grate to ensure full burnout.	Increased tolerance of fuel specs; Reduced emissions and environmental impact.
Slagging in furnace from biomass ash and chemical composition.	Separate zones for primary and secondary air supply. Controlled modulation of air flows; Control of airflow to regulate furnace temperature and ensure flow consistency.	Optimum oxygen supply to handle changing combustion conditions; Improved mixing and residence time to burn off volatiles.	Increased tolerance of fuel specs; Reduced emissions and environmental impact; Improved efficiency; Improved availability; Reduced maintenance Opex.
Grate damage from dry fuels.	Zoning of primary air across grate; Recirculation of inert flue gases (if primary air is preheated); Water cooled grate.	Addition of temperature probes across bed to highlight hotspots and reduce O_2 supply accordingly; Water cooled grate most expensive but provides best availability.	Reduced maintenance Opex; Improved availability.
Variations in emissions due to inconsistent fuel feed.	Complete combustion (as above); Control of furnace temperature conditions; Inclusion of secondary abatement (SNCR or similar) as required	Key to optimise the combustion conditions to minimise polluting emissions in first instance; Secondary abatement options in higher sulphur contents.	Increased tolerance of fuel specs; Reduced emissions and environmental impact;
Fouling of heat exchanger tubes from biomass ash and chemical composition	Automated soot blowing; Plate cleaning; Turbulator coils within the tubes.	Sootblowing cost prohibitive for smaller boilers; Movable turbulator coils improve residence time (heat transfer) and physically agitate deposits to prevent deposits (vertical HX).	Reduced maintenance Opex; Improved availability; Improved efficiency.
Low ash content in fuels reducing protection of grate against higher temperatures	Zoning of primary air across grate; Recirculation of inert flue gases (if primary air is preheated); Water cooled grate.	Addition of temperature probes across bed to highlight hotspots and reduce O_2 supply accordingly; Water cooled grate most expensive but provides best availability	Reduced maintenance Opex; Improved availability.

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[27] Marutzky, R. and Seeger, K. (1999), Energie aus Holz und anderer Biomasse, DRW-Verlag Weinbrenner, Leinfelden-Echtlingen, Germany

[28] Fahmi R., Bridgwater A. V., Darvell L. I., Jones J. M., Yates N., Thain S., et al. (2007). The effect of alkali metals on combustion and pyrolysis of Lolium and Festuca grasses, switchgrass and willow. Fuel 86, 1560–1569 10.1016/j.fuel.2006.11.030.

[29] Carbon Trust (2010), Introducing combined heat and power (CTV044).

3.3 Bubbling Fluidised Bed

Fluidised Bed Combustion Overview

Fluidised bed combustion involves the combustion of a fuel in a bed of an inert material (often sand). The sand and fuel is kept in suspension by passing a current of air upwards through the bed, this current of air causes the bed of sand to be fluidised. Fuel is added to the bed in a consistent manner and is immersed in the high temperature sand bed. Oversize material including ash is withdrawn from the bed on a continuous basis, cooled and sized, before recycling back suitable material into the bed where possible. Oversize material is discarded.

There are two main types of fluidised bed combustor:

- Bubbling Fluidised Bed (BFB): Bubbling beds tend to use low fluidising velocity and the bed behaves similarly to a pan of boiling water.
- Circulating Fluidised Bed (CFB): Circulating fluidised beds use much higher fluidising velocities in which the sand and fuel are carried upwards through the furnace. At the top of the furnace, the solids are removed from the gases in a cyclone and the solids are returned to the bed by gravity.

Ash is removed from a BFB via two routes: extracted as oversize from the bed remediation system; and fly ash that is carried over. The ratio of ash carried over to that which is removed from the bed is a function of the particle size distribution of the bed ash. If there is high fines content, there tends to be more fly ash, whereas if there is little fine material, there may be more ash removed from the bed system.

The combustion temperatures that apply in fluidised beds are somewhat lower than those that apply in moving bed technologies. Fluidised bed and freeboard temperatures when burning biomass materials tend to be in the range of 800 to 900°C. At these temperatures, the levels of release of alkali metals in the combustion gases and the degree of fusion of the ash tends to be significantly lower than those that apply at the higher temperatures in moving bed combustors. Where the ash is exposed to higher temperatures in a grate furnace, there may be increased fouling of heat transfer surfaces.

In fluidised bed boilers the control of bed temperature is important. If fluidisation is not uniform, there is a tendency for hot and cold spots to form. This can lead to uneven combustion and variations in gaseous emissions and, in the worst case, agglomeration of the bed (caused by the low melting temperature ash components). If left unchecked, this can lead to bed 'slumping' in which case it may be necessary for the operators to bring the boiler off-line to remove and replace the bed material.

3.3.1 Technology description

In BFB boilers, fuel feeders discharge either to chutes that drop the fuel into the bed or to fuel conveyors that distribute the fuel to feed points around the boiler. The speed of the feeders is modulated to maintain output when the fuel conditions or loads change. The fluidised bed consists of fuel, ash from the fuel, inert material (e.g., often sand), and possibly a sorbent (e.g., limestone) to reduce sulphur emissions.

The fluidised state of the bed is maintained by hot primary air flowing upward through the bed. The bottom of the bed is supported by water-cooled membrane walls with specially designed air nozzles that uniformly distribute the air. The amount of air is just sufficient to cause the bed material to lift and separate. In this state, circulation patterns occur, causing fuel discharged on top of the bed to mix throughout the bed. Because of the turbulent mixing, heat transfer rates are very high and combustion efficiency is good. Consequently, combustion temperatures can be kept low compared to other conventional fossil fuel fired boilers. The bed may also be operated in a sub stoichiometric mode, with flue gas recirculation to the primary air nozzles and additional air added in the freeboard to complete combustion. Low bed temperatures and air staging reduce NO_x formation. Low bed temperature is also an advantage with biomass fuels because they may have relatively low ash fusion temperatures. Low ash fusion temperatures can lead to excessive boiler slagging or bed material agglomeration.

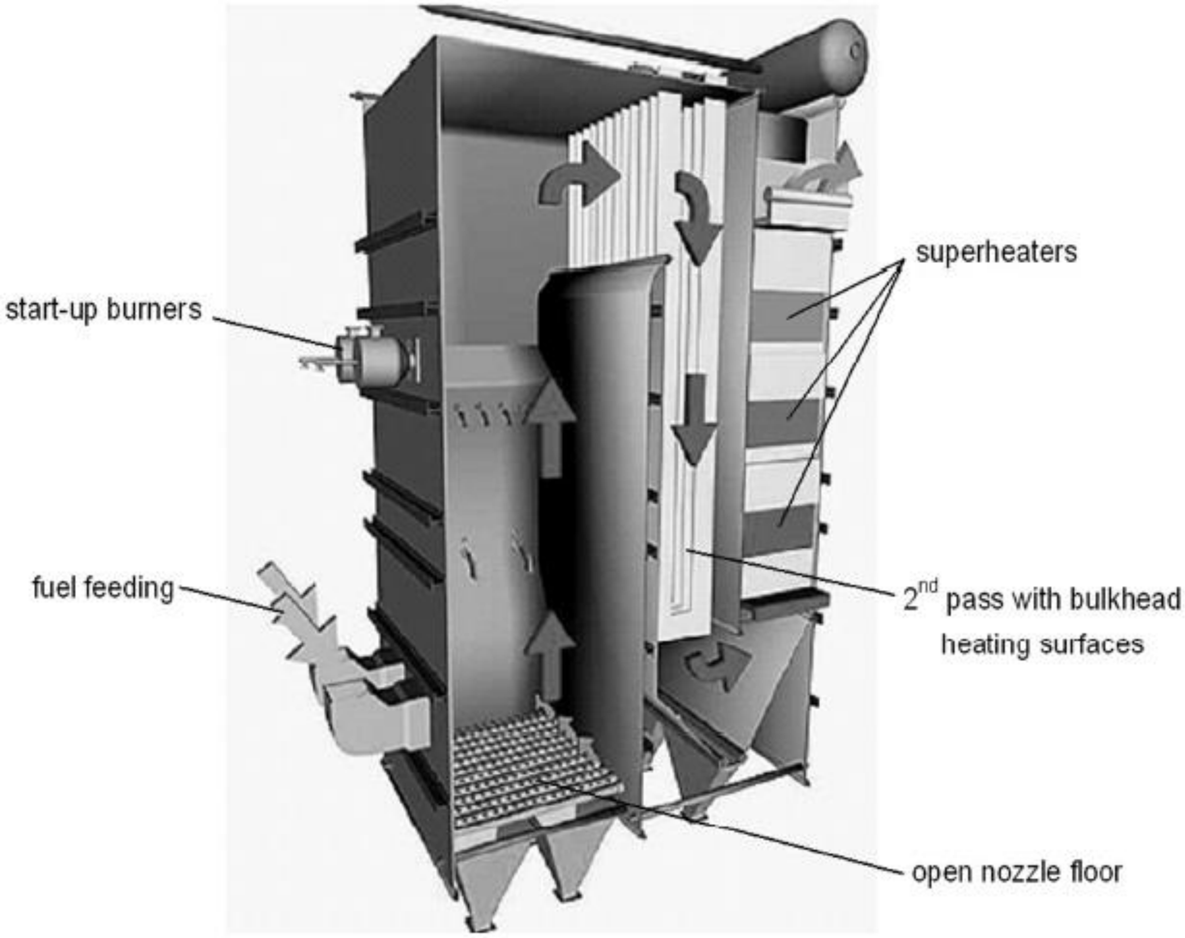

Figure 40: Typical Bubbling Fluidised Bed combustion boiler[30]

In a BFB boiler, the unit is generally designed to have flue gas velocities through the bed of less than 3m/s. This low velocity minimizes the amount of large solid material entrained in the flue gas stream. The management of tramp material (dirt and other non-combustibles) and agglomerates in the bed is very important for reliable long-term operation. The removed bed material is screened to separate the tramp materials from the inert bed material, and the reclaimed inert material is recycled back to the bed. Additionally, a flue gas recirculation system recirculates inert flue gas to the bed to ensure proper fluidisation, even during periods of low load operation when less oxygen is needed for fuel combustion. A disadvantage of BFB boilers is the large auxiliary power requirement for the fluidising air fans. This typically increases operations and maintenance (O&M) costs compared to equivalent-sized moving bed systems.

Because of the lower fluidising velocities compared to CFB boilers, the BFB boiler requires significantly more fuel and sorbent feed points that can be operated independently to ensure uniform mixing of bed materials.

Hot sand in the bed effectively dries out and volatilizes the introduced fuel. Smaller, less dense fuel and ash particles and volatile gases then travel to a zone of secondary combustion. This results in c. 60% of the combustion occurring in the bed and c. 40% occurring above the bed. Overfire air is required to ensure complete fuel combustion. The over-bed combustion results in higher flue gas temperatures above the bed, which increases the radiant heat transfer to water walls and suspended superheater surfaces compared to a CFB boiler. This area above the dense bed resembles the open burning of volatile fuel components in a moving bed boiler.

Low fluidising velocities typical with BFB combustors mean the bed materials maintain a relatively high solid density. These results in a well-defined bed surface, with only a small fraction of the solids entrained in the flue gas stream leaving the bed. The bed, therefore, retains most of the heat of combustion and this high thermal inertia that compensates for variations in nonhomogeneous fuels, including variations in heating value and moisture content, is thus well suited for lower heating value, moist fuels such as biomass. This results in a consistent heat output and flue gas quality. The high heat transfer of the fluid bed medium also provides high carbon burnout. With proper design, BFB boilers should be capable of processing a diverse mix of fuels simultaneously (e.g., a mixture of wood waste, agricultural residues, and biosolids).

Boiler efficiency is affected by the composition of the fuel being burned as well as the boiler design. The typical boiler efficiency (output to feedstock ratio) for BFB combustion units firing biomass is c. 75% when using feedstock at c. 45% moisture content.

Emission control is required regardless of the fuel; with control of PM_{10} typically accomplished with a fabric filter. However, because of the lower combustion temperatures (compared to a moving bed), thermal NO_x emissions from a BFB boiler that burns biomass will generally be less than 0.31kg/MWh – further detail on NO_x is given in Section 3.2.3. In addition, the operating temperature of a BFB is usually within the temperature range that allows a selective non-catalytic reduction (SNCR) system to be effective. In certain cases, for plant burning clean woody fuels NO_x control other than air staging is not always required to meet current EU emission limits. In time, with the introduction of ever more stringent gaseous emissions limits, consideration may be given to the more expensive SCR process.

3.3.2 Development status and timescales

The use of BFBs for biomass combustion has mainly occurred within the past 30 years, and in recent years the technology has become more robust to meet the more demanding design criteria of biomass. Much of the early development of BFB was carried out in the UK but this lead was soon overtaken particularly in Scandinavia and the USA. Whereas the initial development was in the combustion of poor quality coals, the Scandinavians were quick to realise the potential for biomass combustion, especially in the pulp and paper industry where BFBs have been developed for black liquor and bark.

BFBs are a mature commercial technology (TRL 9) with proven service records developed in the 1970s and 1980s as a means of burning a range of solid fuels including materials containing higher than average ash or sulphur, such as coals, and high moisture fuels such as sewage sludge and other biomass fuels.

Electricity only

The largest single BFB boiler firing 100% biomass has been in operation since 2002; it is located in Finland. This unit has c. 80 MW_e electric generating capacity and was provided by Metso. At present in the UK, the largest BFB biomass electricity only plant is the 31 MW_e “Wilton 10” plant in Middlesbrough, owned by Sembcorp[31,32.].

Vendors offer BFB systems as large as 100 MW_e, but have yet to install any units of that size. Above 100 MW_e, a CFB would generally be recommended as the technology is better suited for large fuel volume addition or bed make up. The fuel may be fed into the circulating media and this is much more easily achieved than feeding across a large moving bed or bubbling bed.

The minimum economic scale of a BFB is generally c. 5-10 MW_e; however, below c. 50 MW_e the more conventional moving bed boilers are typically more cost competitive.

BFB CHP

Depending on the specification of the local heat requirement, heat from Biomass CHP plant is typically extracted either from back pressure steam turbines or condensing steam turbines. Although technically a minor amount of heat can be extracted and used post-electricity generation, for a site to be classed as “good quality” CHP, a nominal amount of heat needs to be utilised in addition to the electricity.

The largest BFB system providing CHP in the UK is the Iggesund Paperboard mill in Workington, with a capacity of 49 MW_e. In 2011, the 44 MW_e Steven's Croft plant owned by E.ON near Lockerbie in Scotland was converted to CHP by enabling a supply of heat to a sawmill situated approximately 1km from the plant.

BFB Heat

As stated above, BFB boilers tend to be more cost effective than moving bed above c. 50 MW_e (i.e., c. 200 MW_{th} input). There is a very limited market requirement for heat generation at this scale, so heat deployment is unlikely. The TEABPP team did not gather techno-economic data, as none of the BFB boiler manufacturers appear to offer the technology for hot water applications.

3.3.3 Impact of different feedstock parameters on operation and cost

Sizing

BFB boilers are technically capable of burning a wide variety of biomass fuels provided the fuel is sized appropriately. The high thermal inertia of the bed allows for good flexibility. However, a limiting factor may be the material handling system, which would generally require the material to be small sized prior to introduction to the boiler feed hopper. A BFB can generally only tolerate fuel sized in a narrower band than moving bed technologies; stringy materials do not feed out consistently onto the bed unless smaller sized and to a tighter range.

In a review of vendor offerings of BFB combustors, the three-dimensional limits are generally shown as c. 250mm x 50mm x 25mm. This may require more screening and sizing operations to ensure that no dimension of the fuel exceeds the recommended upper limits. A limitation on fines is also generally proposed such that less than 30% mass fraction should pass through a 5mm screen.

Availability Characteristics

Because biomass-fired plants are similar in design to modern coal-fired power generation systems, it is reasonable to expect that biomass-fired plants would perform in a similar manner to coal-fired plants of the same scale and provide similar reliability, provided that biomass-specific design considerations are implemented and proper operation and maintenance (O&M) practices are applied. Based on the experience of previous biomass-fired power plants, it is likely that initial operations may be hampered by issues associated with fuel handling and boiler feeding systems. However, in general, these issues have been able to be resolved. Following the resolution of these technical issues, these facilities have demonstrated reliable operation.

It is expected that new fluidised bed plants should be expected to operate at reliabilities similar to new coal units, with expected availabilities around 90%.

Compared to coal, biomass fuels are generally less dense, have lower energy content, and are more difficult to handle. With some exceptions, these qualities generally mean that biomass fuel is naturally disadvantaged compared to fossil fuels. The fuel constituents will also impact the range of bed temperatures needed for combustion. Lower volatile matter fuels will require higher bed temperatures for initial and complete combustion. Bed temperature also affects the effectiveness of sorbent materials present; there is an optimum temperature for sulphur capture so the bed temperature has to be accurately maintained in order to maintain sulphur capture efficiency.

Unlike coal, biomass fuels typically contain only trace amounts of toxic metals, such as mercury, cadmium, and lead. On the other hand, facilities that fire biomass must still cope with some of the same pollution issues as larger coal-fired plants with various air quality control technologies used to manage these pollutants.

The majority of biomass materials of industrial interest have similar key chemical characteristics, which have a key influence on the high temperature corrosion processes that any biomass combustion system must be designed to withstand.

As with all biomass combustion plants, the availability is primarily dependent on the quality of feedstock meeting the OEM’s specification, rather than the feedstock itself. In particular, the feedstock should be properly sized and contain minimum contraries (stones). The plant’s lifetime is dependent on the operator carrying out operation and maintenance in accordance with the OEM's instructions. There will generally be no difference in lifetime between the feedstocks listed.

Proximate Analysis - Moisture Content and Volatile Content

An important biomass fuel parameter that is dependent on the selected fuel and the time of year is moisture content. When higher moisture fuels are burned in the combustor, higher fuel feed rates are required to maintain the steam flow. With the increased fuel-burn rate, additional flue gas is produced, requiring a larger boiler vessel. Primary airflow is increased, raising the lower bed material temperature to evaporate the additional water in the bed. As a result, the balance of air to the secondary airflow (overfire air) is reduced, lowering the upper furnace temperature and maintaining the overall furnace temperature.

Conversely, when the moisture in the combustor decreases, primary airflow is decreased, lowering the bed temperature. Simultaneously, overfire air is increased, raising the upper furnace temperature and therefore maintaining the overall furnace temperature. Temperature control for decreased moisture in the bed can also be controlled by flue gas recirculation (FGR). Flue gas is recirculated and mixed with fluidising air, thereby reducing the amount of available oxygen. As a result, FGR acts as a heat sink and lowers the bed temperature.

The volatile matter content also affects the split between the primary airflow at the bottom of the combustion bed and secondary (or over-bed) airflow that is introduced above the dense bed section. Higher volatile matter will increase the need for secondary airflow to reduce carbon monoxide (CO) emissions from combustion. Lower volatile matter content increases the need for primary air in the bed material to promote better carbon burnout.

All of these variations in fuel constituents will result in the need for greater flexibility in equipment design. Wide variations in fuel characteristics will result in the need to optimize equipment design and performance around the fuel that is expected to be predominantly used.

High moisture content in biomass can reduce efficiency of combustion processes and may necessitate the need for supplemental fuel. In addition, boiler efficiency is negatively impacted by high moisture fuels. Fuel that is too wet may not burn. Biomass with a moisture content of up to c. 60-65% by weight can be burned in some BFB combustors while maintaining stable combustion without the use of a supplemental fuel.

In a review of vendor offerings of BFB combustors, the typical fuel moisture content upon which the boiler efficiency is to be guaranteed is 45% moisture. Higher moisture content can be tolerated but may result in poorer control of combustion and therefore efficiency.

Proximate Analysis - Ash Content

Though the fluid bed is expected to be in excess of 95% inert bed material (with the balance being fuel and sorbent), the fuel ash mineral analysis will also impact the boiler and auxiliary equipment design.

For both CFB and BFB combustors, solids entrained in the gas stream are captured by mechanical dust collectors, cyclones, or U-beam separators, and recycled back into the combustor to maintain bed inventories and improve combustion efficiencies. The recycle ratio is defined as the ratio of the recycled mass flow to the fuel feed mass flow. In a BFB combustor, the mass flow of the recycled solids can be up to c. five times the fuel feed rate, compared to a CFB combustor where the mass flow can be 10 to 100 times the fuel feed rate.

Burning biomass with very low ash content may require the injection of sand to supplement the bed material, because the ash content of the fuel will not be sufficient to maintain bed level and size consistency. In addition, the use of very low ash fuels may result in issues with bed erosion and agglomeration if sorbent and sulphur reaction products concentrate.

Sulphur

SO_2 formation in the combustion process occurs as a result of the combustion of the sulphur contained in the fuel. Compared to coal-fired applications, most biomass fuels typically have very low sulphur and, in certain cases, sorbent-injection is not required to control emissions.

The addition of a sorbent (such as limestone) in the bed material is a method of controlling SO_2 emissions through a series of temperature-sensitive reactions with conversion rate peaking at typical temperatures between 840 - 870°C. Maintaining the bed temperature is therefore critical for effective sulphur removal in a BFB boiler, albeit not so important for biomass fuels.

A typical sorbent size of 3mm is required for BFB combustors which can achieve up to 90% removal of SO_2. However, it should be noted that sorbent addition to the bed is more effective in CFBs due to the increased mixing of material within the furnace.

An alternative to application directly into the bed is injection of sorbent within the bag filter or an SO_2 polishing scrubber upstream of the particulate removal system. Lime slurry is atomized into fine droplets that come in contact with the remaining SO_2 in the flue gas stream from the boiler. The SO_2 is absorbed into the atomized slurry droplets from the flue gas and reacts with calcium in the slurry. Effective operation yields a dry, free-flowing by-product that may be recycled to improve reagent utilization.

Although polishing scrubbers are not typically required to control SO_2 emissions from biomass fired plants, inclusion may be considered best practice to control HCl and the small amounts of SO_2 that are produced by wood combustion.

A dedicated control strategy is important as SO_2 removal improves as fluidising airflow increases. High airflows increase both fuel-to-limestone contact and oxygen content. When the fluidising air is reduced (e.g., to control the bed temperature), the limestone feed rate may be increased to compensate for the reduced airflow, to maintain the required ratio for the reduced airflow condition. However, if the fluidising airflow is too low, the sulphur capture reaction has neither adequate oxygen nor fuel-to-limestone contact to complete the reaction. At this point, the SO_2 controls may detect that not enough SO_2 is being removed and may further increase the limestone feed, which further decreases the reaction rate.

The fuel sulphur content and associated sorbent requirement would directly impact sorbent use. In cases of low fuel sulphur content, pre-sized (pulverized) limestone delivered to the plant may be the most economical method of sorbent supply. As sulphur content increases and sorbent requirements increase, limestone gravel deliveries sized for pulverizing at the plant may be the most economical. If pulverized limestone is delivered to the plant, pneumatic transport to storage silos would be required. If bulk delivery of limestone gravel for sizing at the plant is the sorbent supply, outdoor storage of limestone and conveyance with either mobile equipment or belt conveyors would be expected.

Nitrogen

During combustion, NO_x formation occurs in two ways: thermal or fuel-bound. The rate of thermal NO_x formation is strongly governed by gas temperature and, to a lesser extent, by residence times and excess air levels (affecting oxygen availability). The rate of thermal NO_x reaction increases exponentially at temperatures above 1200°C. Since temperatures are typically maintained in the range of 840 - 870°C, thermal NO_x is significantly less when compared to other technologies such as moving grate boilers.

Fuel-bound NO_x depends on the amount of nitrogen in the fuel and the availability of oxygen for the conversion of fuel nitrogen to NO_x. Managing oxygen levels by increasing or decreasing the amount of primary and secondary air (staging) can control the amount of fuel NO_x formed to a certain extent.

When NO_x controls are needed, SNCR may be incorporated. This process involves spraying an ammonia (or urea) based chemical reagent into the flue gas stream in the upper portion of the furnace at optimum reaction temperatures ranging between 815 – 1200°C. The NO_x reacts with the ammonia to form stable N_2 molecules. SNCR provides NO_x removal efficiencies of 50-70%. Because of the longer flue gas residence time at optimum temperature for SNCR performance, a CFB boiler with SNCR would provide better NO_x control than a BFB boiler for the equivalent fuel.

As with SO_2 removal, NO_x control varies with the fuel moisture and heating value. However, SO_2 removal generally improves as fluidising air increases, but NO_x control becomes less effective as fluidising air increases. Experience shows that increases in the amount of limestone tend to increase NO_x emissions. Therefore, the metering and control of limestone and fluidising air must consider the effects of these items on the NO_x control reactions.

Alkali compounds

The ash from certain biomass fuels can have high levels of alkali components. The alkali components of ash, particularly potassium and sodium compounds such as potassium oxide (K_2O) and sodium oxide (Na_2O), cause the ash to remain sticky at a much lower temperature than coal ash. This increased stickiness creates the potential for serious slagging and fouling problems. In fluidised bed technologies, high alkali content can also lead to bed agglomeration.

To remove the sticky material from the boiler reactor surfaces, it is required to perform soot blowing, implement operational procedures such as slag shedding, or have regularly scheduled outages to manually clean the unit. While none of these factors are critical flaws with regards to technical feasibility, they do present significant maintenance and availability burdens that need to be accounted for. These concerns can be substantially reduced if the potential for alkali deposition is properly considered during boiler design.

The problems associated with alkali materials in biomass vary widely between different biomass fuels. To a certain extent, slagging potential can be determined by analysis of fuel properties. However, the slagging tendency of a particular fuel cannot be predicted from fuel properties alone. Boiler design and operating conditions (especially temperature) have a large impact on the nature of deposits. Temperatures of 760°C and below have been shown to significantly reduce deposition.

Potassium is the most common alkali metal in herbaceous biomass fuels and is responsible for much of the fouling and corrosion found in biomass boilers. One alternative that can be considered, particularly for fluidised bed conversion technologies, is the addition of limestone or other additives (such as magnesium oxide) to the fuel feed which react during combustion to form a eutectic with higher melting temperature. The limestone works to alter the composition of compounds formed by the alkalis and hence their ability to form sticky deposits (note effects may be negative).

Chlorine

The chlorine content of biomass materials can vary significantly with wood pellets regularly <0.01% and values >0.5% not uncommon in straw. High chlorine content in some biomass fuels can lead to high temperature corrosion. Large biomass combustion plants routinely operate with steam conditions of 85bar and 480 - 520°C. However, if the fuel contains significant amounts of chlorine (>0.1%), temperatures within the boiler may need to be limited to minimize corrosion of reactor surfaces resulting in a reduction in steam-grade conditions produced to less than 425°C.

Sensitivity Analysis – Power

The cost curve in Figure 41 corresponds to the total investment cost per unit of output capacity for a bubbling fluidized bed power plant. It has been derived from quotes provided by suppliers (2007) for all-in overnight EPC costs.

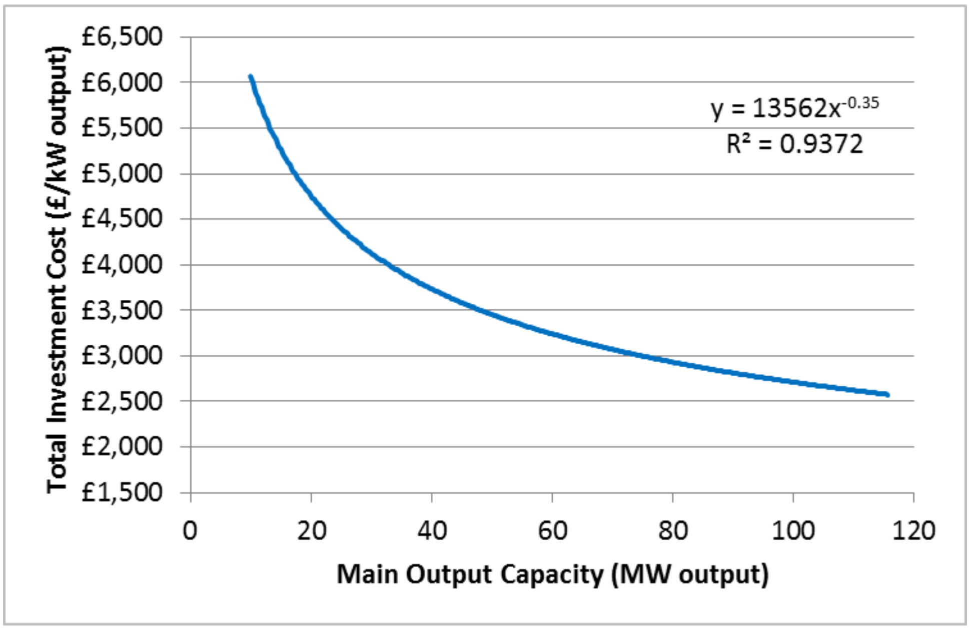

Figure 41: BFB (power) total investment cost vs. main output capacity curve (derived from five supplier quotes (2007) and B&V data for all-in overnight EPC costs)

Total operational costs for BFB power technologies include fixed costs (insurance, maintenance parts and labour) and variable costs (operations labour, additional fuel, reagents, water). Figure 42 shows the relationship between annual total operational cost for a BFB per unit of energy output and the main output capacity rating.

Except where the parameter of interest is shown on the x axis, the curves in Figure 42 to Figure 49 are all created using the following base values: Main output capacity = 50 MW_e; station capacity factor of 85%; feedstock moisture content (WB) = 50%; feedstock ash content (DB) = 1.5%; feedstock nitrogen content (DB) = 0.35%; feedstock sulphur content (DB) = 0.05%; and a feedstock chlorine content (DB) = 0.04%.

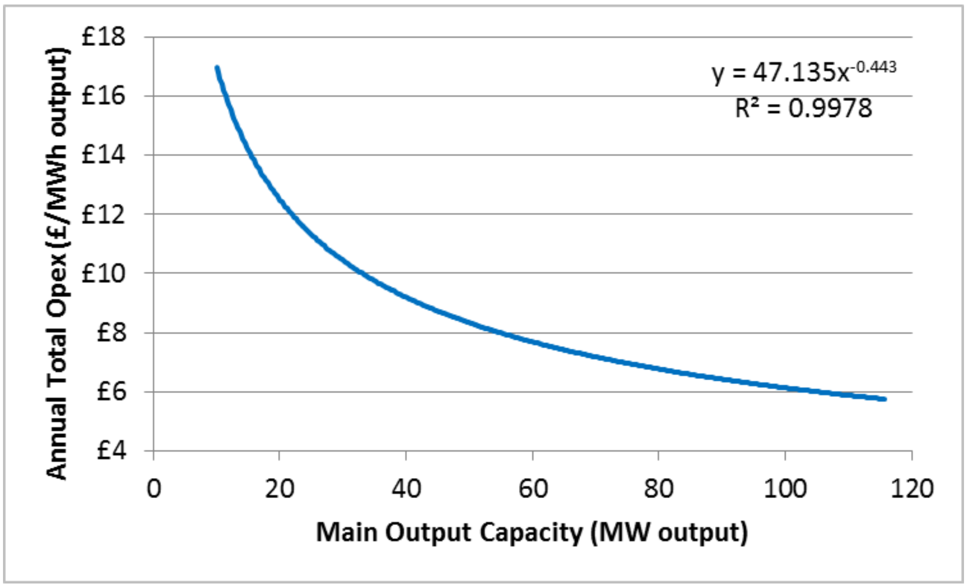

Figure 42: BFB (power) annual operating cost per unit of electricity output vs. output power capacity (derived from B&V industry data)

Figure 43 below highlights the change in overall BFB station efficiency (LHV) over varying feedstock moisture content. The efficiency is calculated as electrical energy output over fuel energy input; as such parasitic loads post-generation are not included. The curve is considered to be a representative average across the BFB potential rated output range for a condensing turbine with air-cooled condenser.

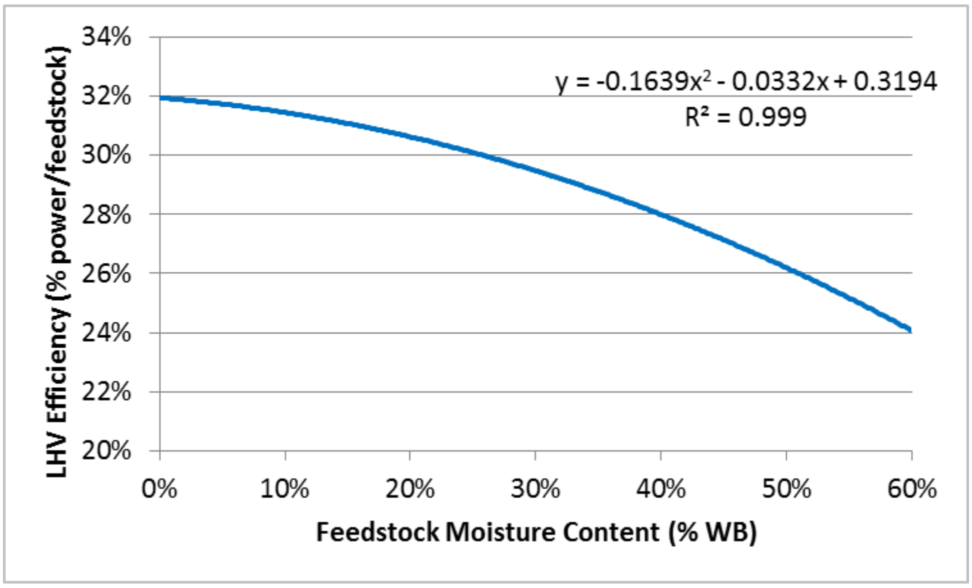

Figure 43: BFB (power) efficiency vs. feedstock moisture content (derived from B&V industry data considering a representative average across the technology potential rated output range)

The total operational costs for BFB Power technologies (as previously described for Figure 42) are related to feedstock moisture content as shown in Figure 44 below.

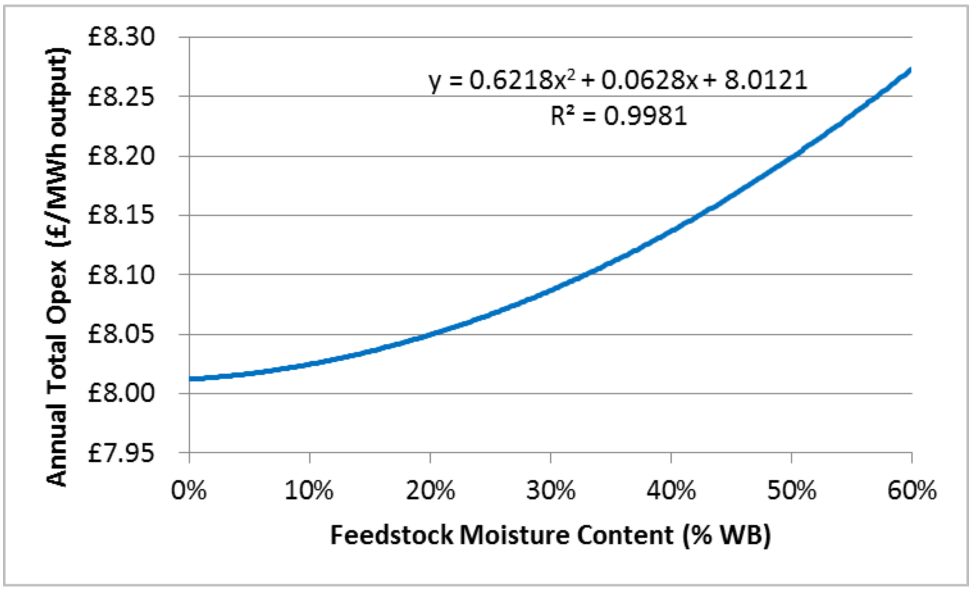

Figure 44: BFB (power) annual operating cost per unit of electricity output vs. feedstock moisture content (derived from B&V industry data)

As per Figure 45 below, annual operational costs for BFB Power stations vary with the percentage of ash present within the feedstock.

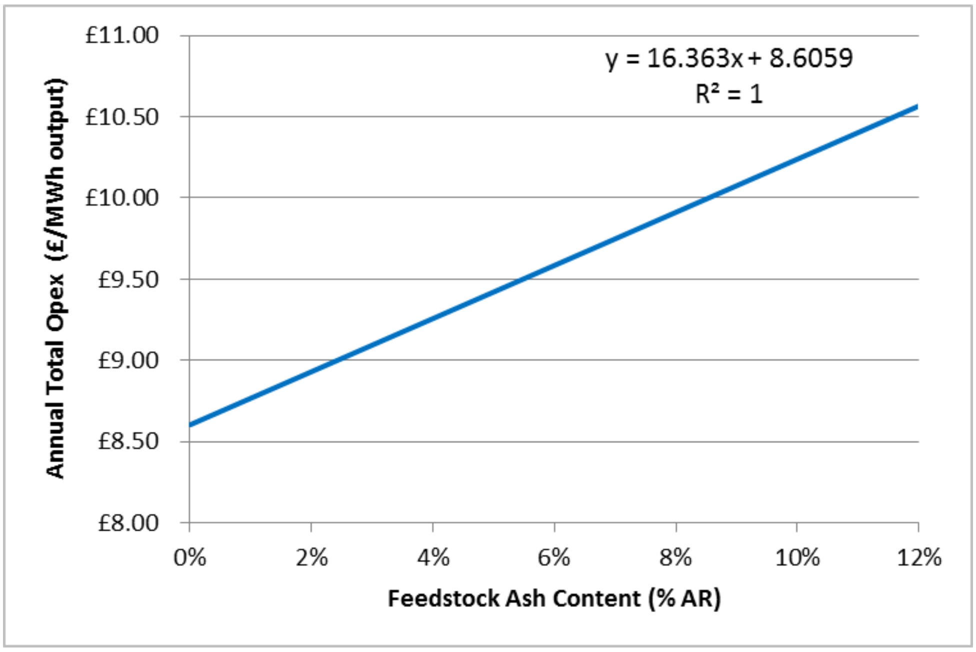

Figure 45: BFB (power) annual operating cost per unit of electricity output vs. feedstock ash content (derived from B&V industry data)

Figure 46 below displays the relationship between nitrogen content within the biomass feedstock and the annual cost of NO_x abatement via urea injection.

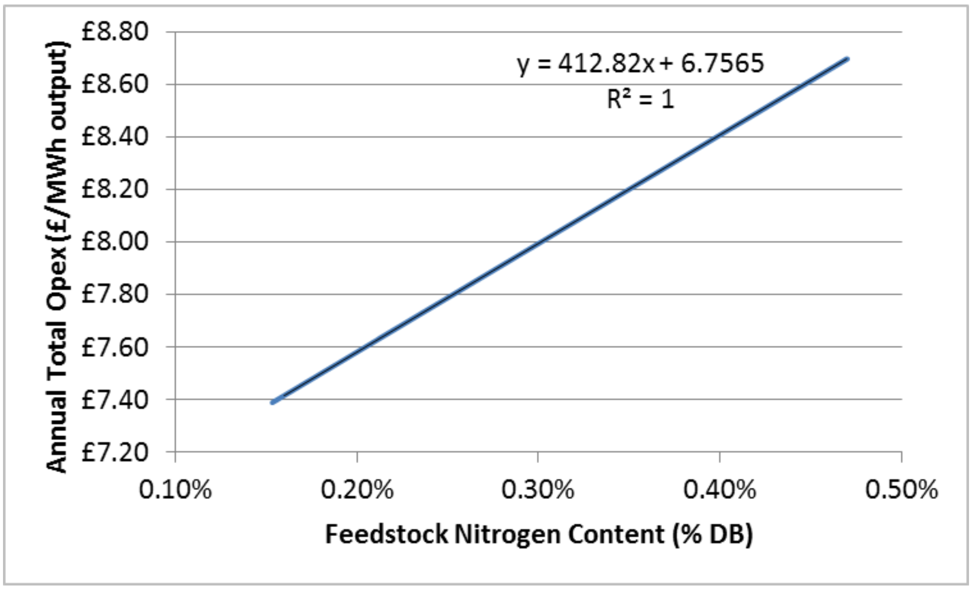

Figure 46: BFB (power) annual operating cost per unit of electricity output vs. feedstock nitrogen content (calculation based on section 2.4.2.4)

Figure 47 below displays the relationship between sulphur content within the biomass feedstock and the annual cost of abatement via lime addition.

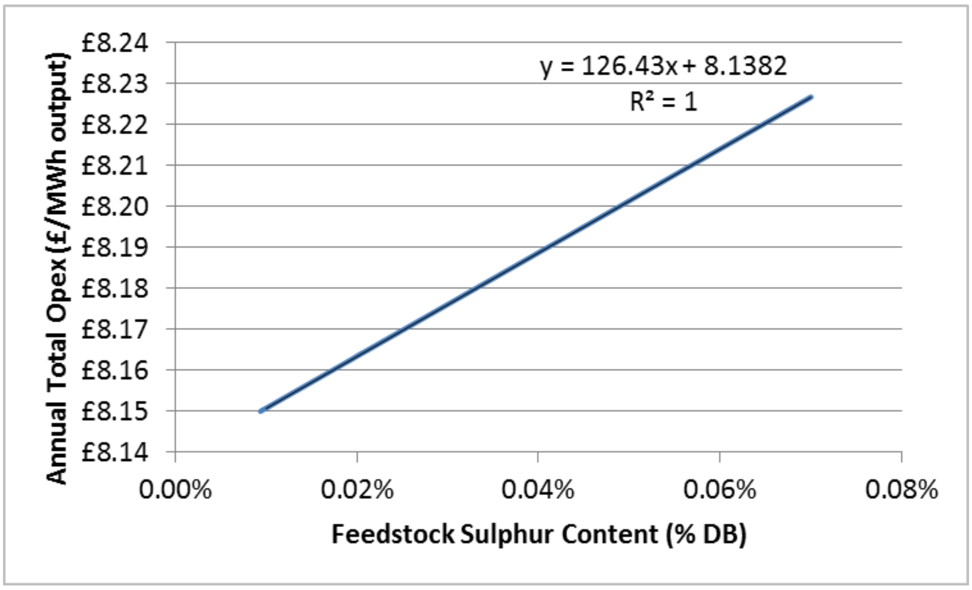

Figure 47: BFB (power) annual operating cost per unit of electricity output vs. feedstock nitrogen content (calculation based on section 2.4.2.4)

Figure 48 below displays the relationship between chlorine content within the biomass feedstock and the annual cost of abatement via lime addition.

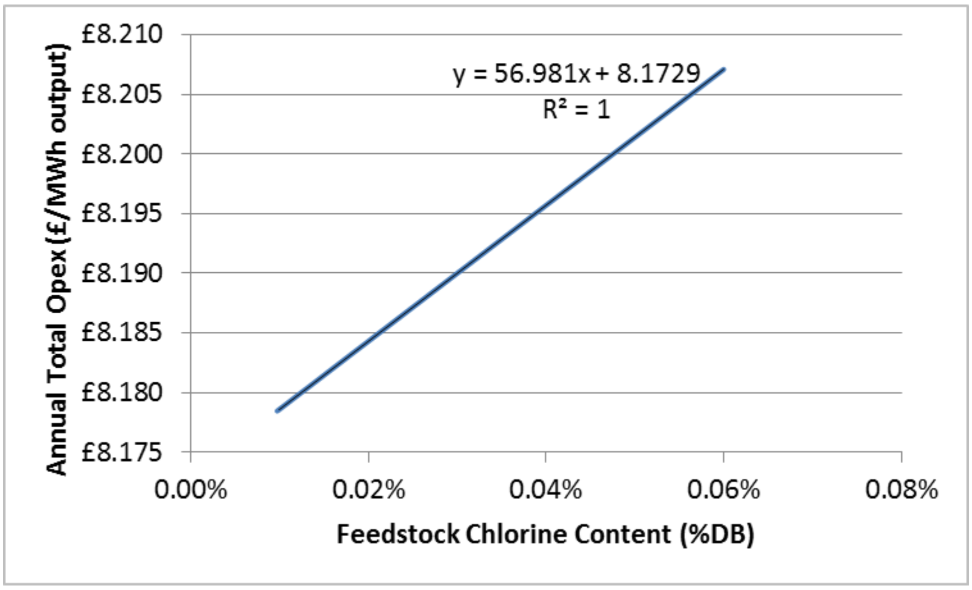

Figure 48: BFB (power) annual operating cost per unit of electrical output vs. feedstock chlorine content (calculation based on section 2.4.2.4)

Figure 49 below displays the relationship between the alkali index (kg K_2O and Na_2O per GJ energy) and the annual total operational cost for a BFB per unit of energy output.

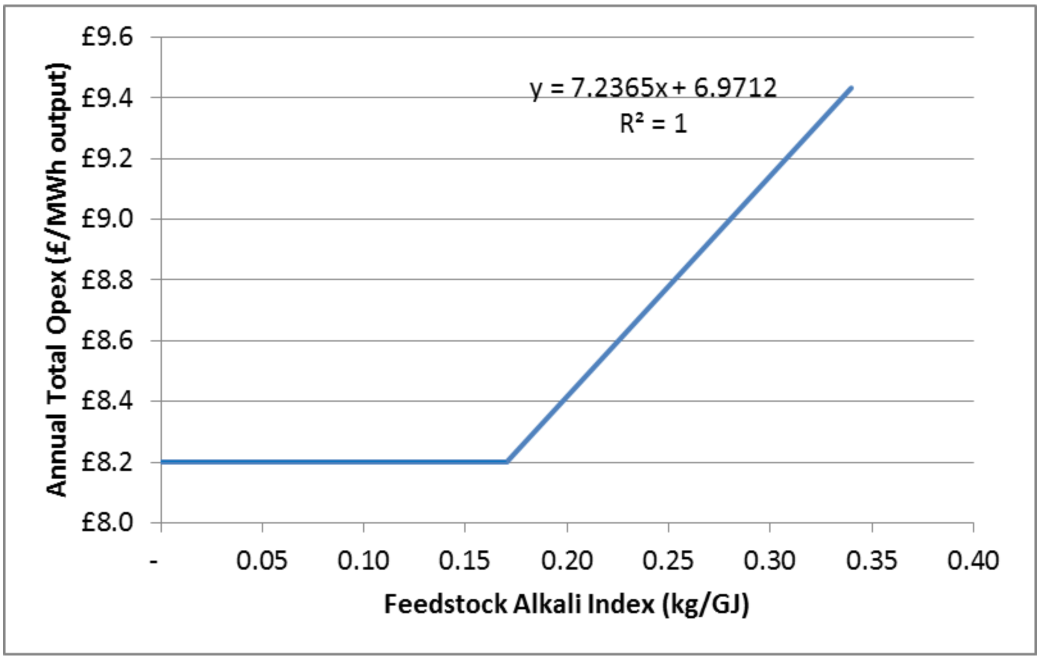

Figure 49: BFB (power) annual operating cost per unit of electricity output vs. feedstock alkali index (derived from formula based on [33] and section 2.4.2.5)

Sensitivity Analysis – CHP

Due to the inability to export heat over long distances, the thermal output capacity of a plant is typically dependent on the site local demand profile. To optimise financial returns, plant is often operated to maximize returns from electrical generation with the heat utilized wherever possible.

The amount of useful heat available for capture from the power generation process is considered too variable across sites to offer meaningful comparison and therefore the following sensitivities are on an electrical power output basis.

The cost curve in Figure 50 corresponds to the total investment cost per unit of electrical output capacity for a BFB CHP plant. It has been derived from quotes provided by suppliers (2007 – 2011) for all-in overnight EPC costs.

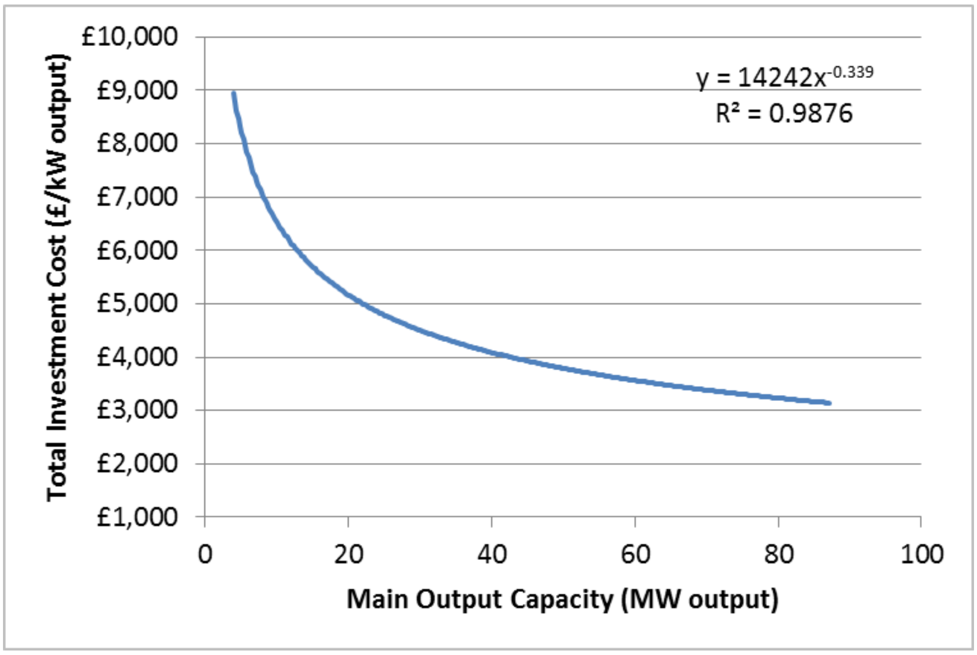

Figure 50: BFB (CHP) total investment cost vs. output power capacity (derived from five supplier quotes (2007 - 2011) and B&V data for all-in overnight EPC costs)

Total operational costs for BFB CHP technologies include fixed costs (insurance, maintenance parts and labour) and variable costs (operations labour, additional fuel, reagents, water). Figure 51 shows the relationship between annual total operational cost for a BFB per unit of electrical energy output and the main electrical output capacity rating.

Except where the parameter of interest is shown on the x axis, the curves in Figure 51 to Figure 58 are all created using the following base values: Main output capacity = 36 MW_e; station capacity factor of 85%; feedstock moisture content (WB) = 50%; feedstock ash content (DB) = 1.5%; feedstock nitrogen content (DB) = 0.35%; feedstock sulphur content (DB) = 0.05%; and a feedstock chlorine content (DB) = 0.04%.

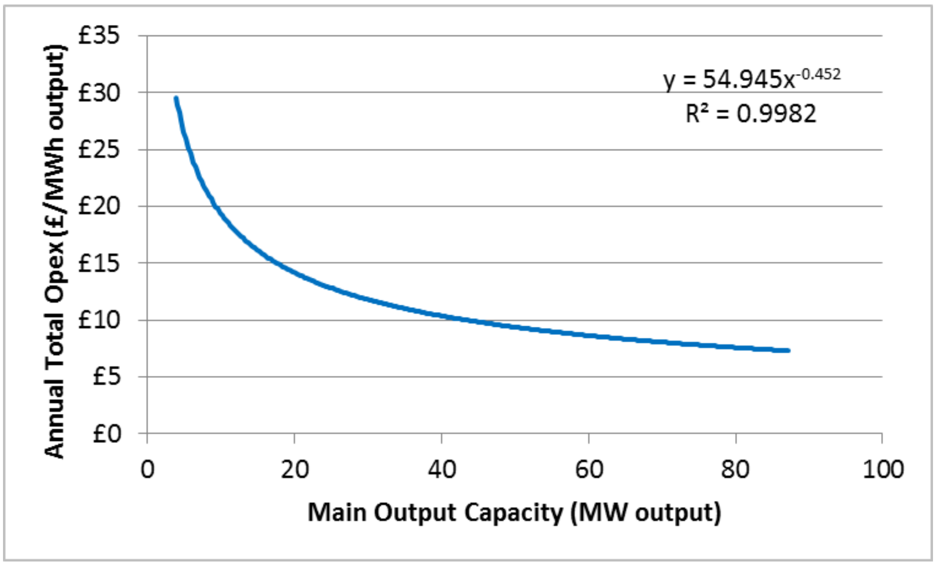

Figure 51: BFB (CHP) annual operating cost per unit of electricity output vs. output power capacity (derived from B&V industry data)

Figure 52 below highlights the change in electrical efficiency (LHV) of a BFB CHP over varying feedstock moisture content. The efficiency is calculated as electrical energy output over fuel energy input; as such parasitic loads post-generation are not included. The curve is considered to be a representative average across the BFB potential rated output range and, as stated previously, the thermal efficiency is dependent on the amount of useful heat captured and is considered site specific and therefore too variable for further comparison.

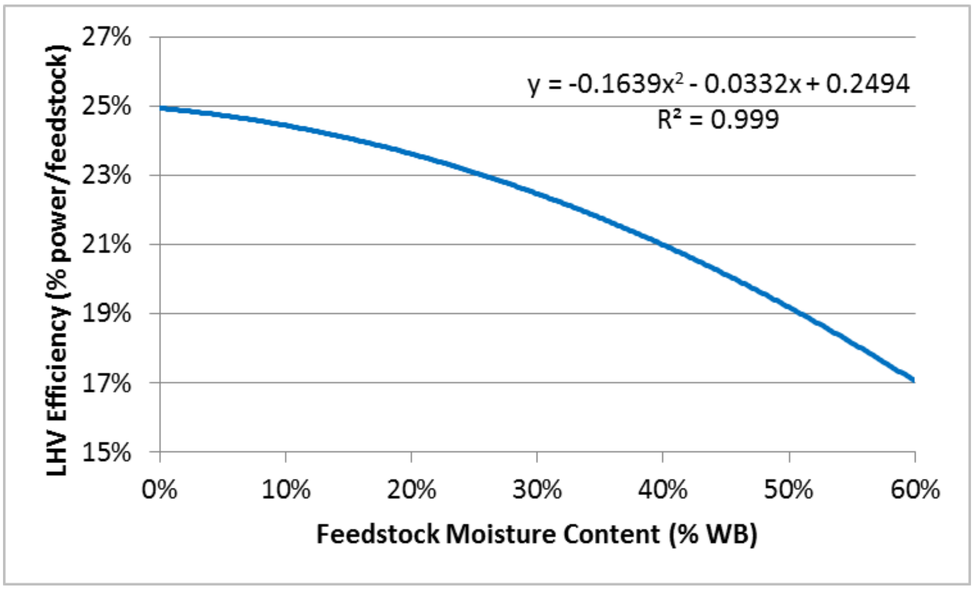

Figure 52: BFB (CHP) electrical efficiency vs. feedstock moisture content (derived from B&V industry data considering a representative average across the technology potential rated output range)

The total operational costs for BFB CHP technologies (as previously described for Figure 51) are related to feedstock moisture content as shown in Figure 53 below.

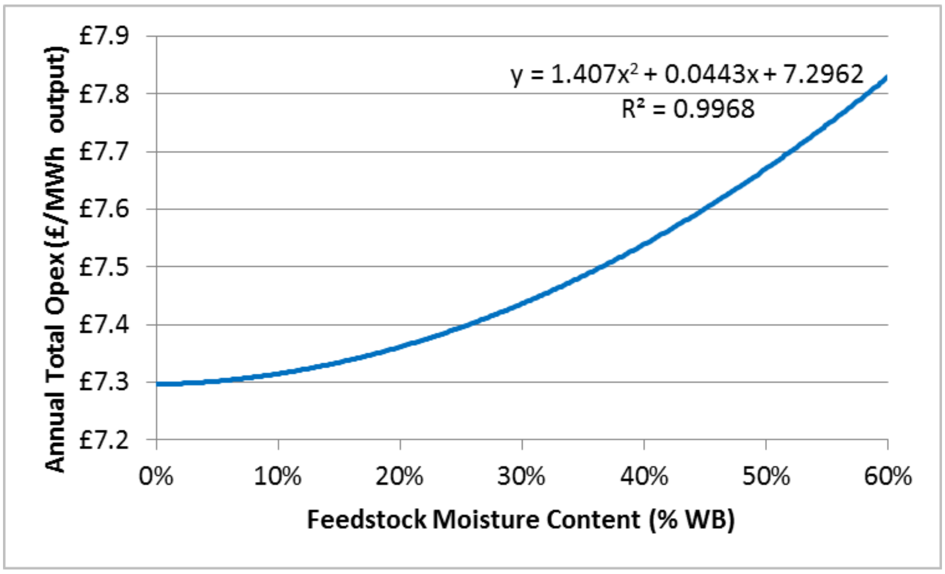

Figure 53: BFB (CHP) annual operating cost per unit of electricity output vs. feedstock moisture content (derived from B&V industry data)

Annual operational costs for BFB CHP stations vary with the percentage of ash present within the feedstock, as shown in in Figure 54.

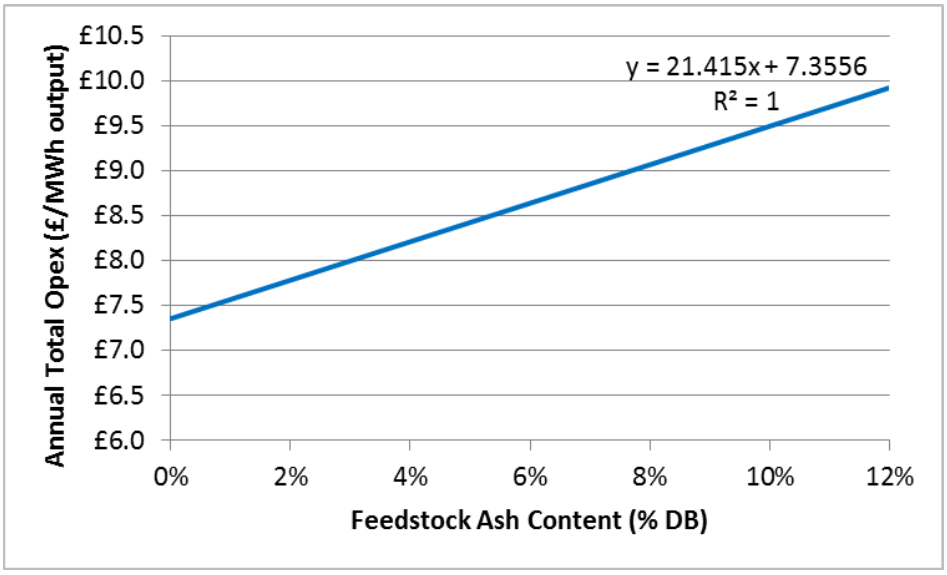

Figure 54: BFB (CHP) annual operating cost per unit of energy output vs. feedstock ash content (%DB) (derived from B&V data)

Figure 55 below displays the relationship between nitrogen content within the biomass feedstock and the annual cost of NO_x abatement via urea injection.

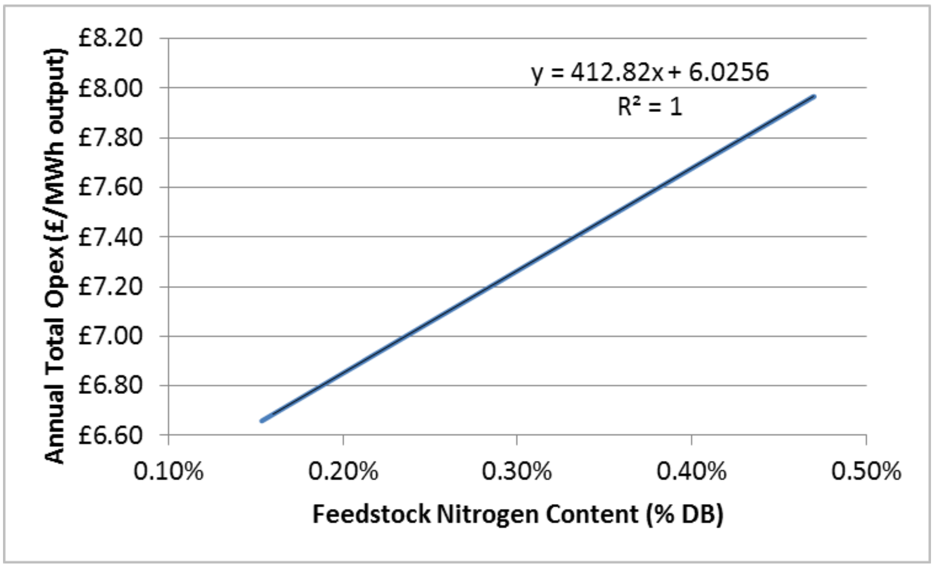

Figure 55: BFB (CHP) annual operating cost per unit of energy output vs. nitrogen content within the biomass feedstock (calculation based on section 2.4.2.4)

Figure 56 below displays the relationship between sulphur content within the biomass feedstock and the annual cost of abatement via lime addition.

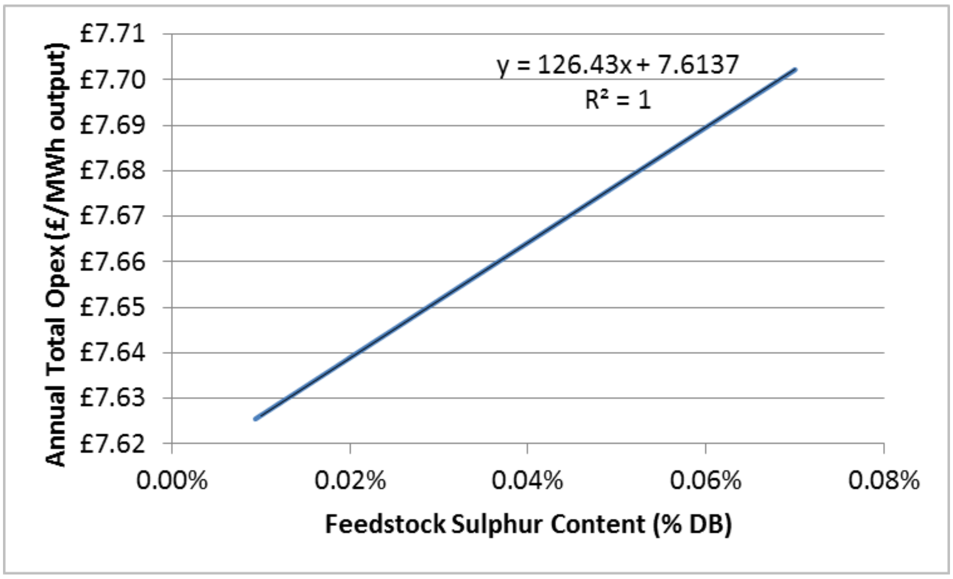

Figure 56: BFB (CHP) annual operating cost per unit of energy output vs. sulphur content within the biomass feedstock (calculation based on section 2.4.2.4)

Figure 57 below displays the relationship between chlorine content within the biomass feedstock and the annual cost of abatement via lime addition.

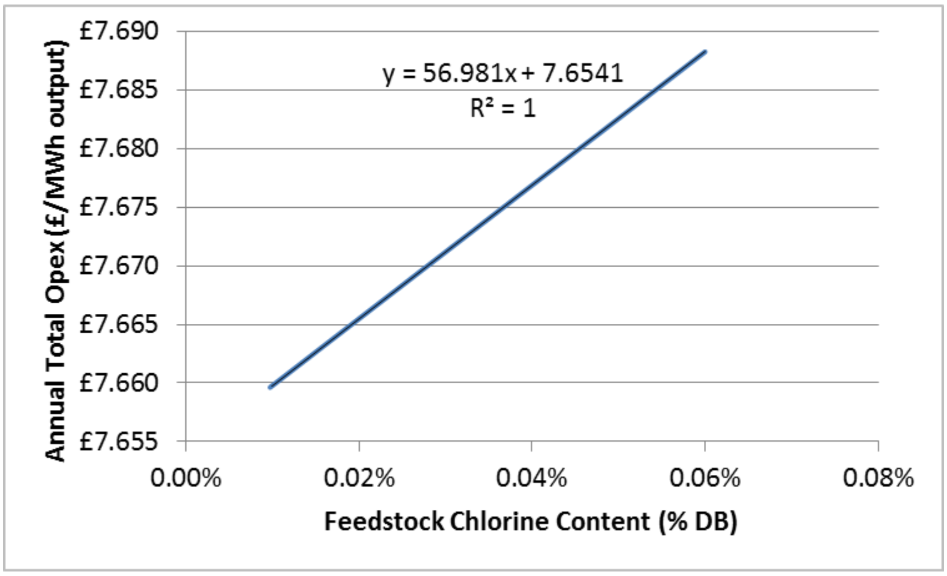

Figure 57: BFB (CHP) annual operating cost per unit of electricity output vs. feedstock chlorine content (calculation based on section 2.4.2.4)

Figure 58 below displays the relationship between the alkali index (kg K_2O and Na_2O per GJ energy) and the annual total operational cost for a BFB per unit of energy output.

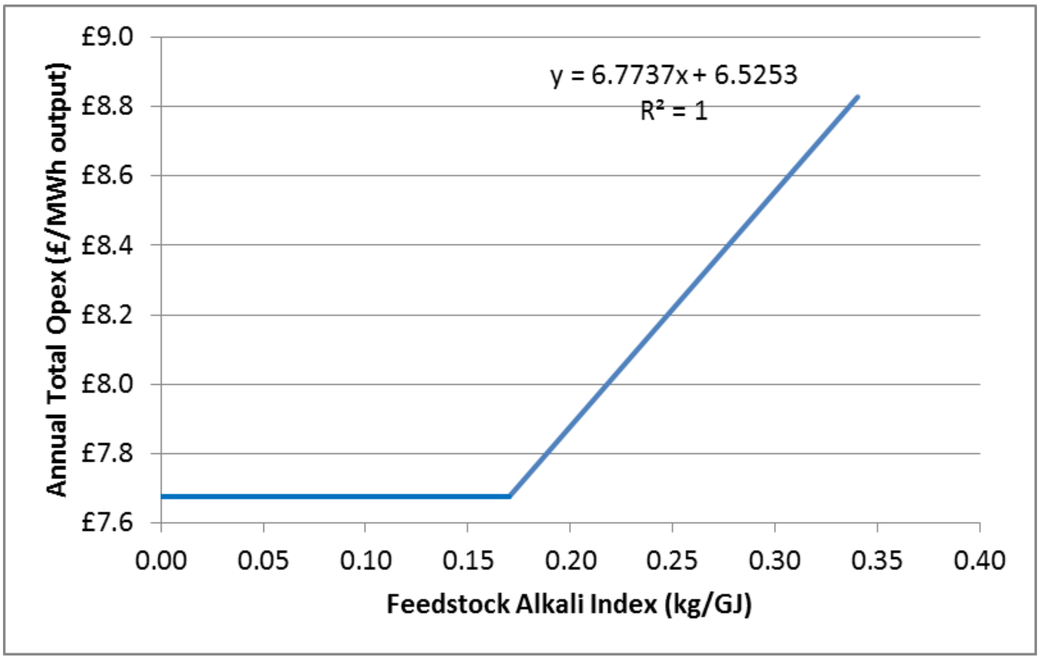

Figure 58: BFB (CHP) annual operating cost per unit of energy output vs. feedstock alkali index (derived from formula based on [33] and section 2.4.2.5)

3.3.4 Available options for improvement

Some of the more significant improvements in biomass fired BFB boiler design are shown in Table 11 and summarised below. The early generations of BFB boilers were typically fitted with significant heat transfer surfaces in the bubbling bed as well as numerous under-bed injection points for introducing fuel and sorbent. Bed ash extraction provisions were not numerous or flexible regarding the ability to remove only selects portions of the bed material. The original installations found that agglomerated bed material was difficult to remove and that erosion of the in-bed heat transfer material was a significant maintenance issue. The less aggressive fluidising action of the BFB compared to a higher-flow CFB was the basis for the need for multiple fuel and sorbent input locations to ensure uniform bed inventory. For coal or other sulphur bearing fuels, this resulted in the need for large numbers of under-bed feed points for fuel and sorbent. These multiple feed points were found to be prone to blockage, which resulted in less than ideal bed material distribution.

For low ash and low sulphur biomass fuels, the need for in-bed sorbent feed has been replaced with the need for increased bed makeup material (sand) feed. Since the bed composition is more than 95% inert sand/ash material, numerous feed points are less critical. This simplifies the material feed system and provides more reliable performance since sulphur capture in the bed is not required. The complexity of the fuel feed system is also reduced. This is due both to the absence of the sulphur combustion by-product reaction with sorbent as well as the removal of the in-bed heat transfer surface, which has increased reliability because tube leaks no longer occur from the in-bed heat transfer surfaces. Because there are no in-bed heat transfer surfaces, the uniformity of combustion in the bed (to provide an even heat transfer to the in-bed surfaces) has become less critical. This has enabled a substantial reduction in the number of fuel feed points and the relocation of these feed points above the bed (rather than below the bed), which was standard for the first generation units. Rather than utilising the in-bed heat transfer surface to help maintain bed temperature, the approach is now to provide a flue gas recirculation system to help maintain bed fluidisation with lower oxygen content fluidising air, which limits fuel combustion and heat release. This amount of flue gas recirculation is controlled in order to maintain the proper bed temperature across the unit load range.

Another major improvement in BFB boiler technology is in the design of the bed fluidising nozzle support system. The initial units were provided with solid floor systems that provided the pressure boundary between the fluidising air nozzles and the combustor. The floor system did not allow for flexibility in draining select portions of the bed material. The “Hybex” bottom design (as proposed by Metso) and the “Open Bottom BFB” design (offered by B&W) offer significant improvements and flexibility to the original solid floor systems of older units. This enables more reliable and flexible provisions for controlling and draining bed material from BFB boilers.

Table 11 : Improvements Analysis of BFB Boilers

Issue arising due to biomass characteristic	Options to ameliorate	Evaluation of effect	Long Term Improvements
Poor bed conditions from non-homogeneous feedstock	Multiple fuel input points; Blown in over bed instead of screwed in under bed; Variable fuel and bed feed rates.	Improved distribution reduces cases of unburnt fuel in larger size / higher MC; Better mixing with sorbents (if added) in bed; Consistent bed depth will reduce hot / cool spots; Increases chances of entrained unburnt fines in product gas.	Reduced maintenance Opex; Reduced bed medium turnover; Reduced sorbent requirements; Improved combustion efficiency; Increased tolerance of fuel specs (not fines); Reduced emissions and environmental impact (not particulates).
Biomass primarily low sulphur	Reduces requirements for sorbent added directly to bed; If sorbent is required many designs incorporate addition at post combustion stage (bag filter or final scrubbing).	Inert content of bed is significantly increased requiring fuel to be well distributed to avoid hot/cold spots.	Reduced sorbent requirements; Reduced emissions and environmental impact.
Biomass oversize and ash affecting bed conditions	Numerous bed thermocouples to monitor bed conditions; Multiple independent locations for bed extraction; “Open” combustor floor bottom for multiple and reliable bed material drains.	Reliable diagnosis of hot and cold spots; Reduced slagging; Better refractory life; Reduced bed agglomeration / bed slump.	Reduced maintenance Opex; Improved boiler availability.
Dry fuels increasing bed temperatures;	Removal of heat exchanger surfaces from reducing zones within the bed; Varying combustion air make up – addition of recirculated flue gas.	Better corrosion and erosion resistance of furnace heat exchanger surfaces; Increased control of O_2 content and reaction rates.	Reduced maintenance Opex; Improved combustion efficiency. Improved boiler availability / longer plant life; Reduced emissions and environmental impact.
Increased volatile release with biomass (compared to coal).	Secondary combustion air control; Sorbent addition outside of bed.	Increased control of heat release in freeboard; Increase rate of complete combustion.	Reduced emissions and environmental impact; Improved combustion efficiency.

Heat Exchanger Materials

The standard materials of construction for the finishing and secondary superheater materials are adequate for the temperatures and pressures of steam generation for power production, but without special consideration for potential high corrosiveness of biomass combustion. Depending on the level of fire-side corrosion when combusting woody biomass, service lives using standard superheater materials have been demonstrated of 5 to 10 years, improving to up to 20 years of life when using more corrosion-resistant materials. The higher corrosion resistant materials may also enable higher steam temperatures and more efficient steam cycles.

Bed Temperature Control and Ash Extraction

With low ash and low sulphur content fuels such as biomass, a supplemental inert bed material feed system is required. With low ash biomass fuels, there is reduced need for draining the bed material unless it becomes oversized due to agglomeration or the injection of excessive non-combustible oversized material. The vast majority of the fuel ash is removed as fly ash in the fabric filter.

With the numerous bed thermocouples that are provided with a typical modern BFB boiler, the bed performance can be continuously monitored and areas of poor fluidisation can be detected and selectively drained to eliminate the oversized materials. Multiple independent extraction points are a benefit for this.

The bed temperature thermocouple system is expected to enable better early detection of bed agglomeration and enable selective material removal. The large openings between the fluidisation nozzles and large hoppers below the nozzles are expected to provide more flexibility in removing oversized material without interfering with the operation of the fluidisation nozzles.

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[30] The handbook of biomass combustion and co-firing (2008), edited by Sjaak van Loo and Jaap Koppejan, Earthscan London.

[31] Wilton 10 Biomass-fueled Boiler factsheet http://media.corporate-ir.net/media_files/otcbb/FWHLF.OB/releases/release032305b.pdf

[32] Wood Burning Power Station at Wilton 10, United Kingdom http://www.power-technology.com/projects/Wood-Burning/

[33] Fahmi R., Bridgwater A. V., Darvell L. I., Jones J. M., Yates N., Thain S., et al. (2007). The effect of alkali metals on combustion and pyrolysis of Lolium and Festuca grasses, switchgrass and willow. Fuel 86, 1560–1569 10.1016/j.fuel.2006.11.030.

3.4 Circulating Fluidised Bed

3.4.1 Technology description

In CFB boilers, fuel feeders discharge biomass either to chutes that drop the fuel into the bed or to fuel conveyors that distribute the fuel to feed points around the boiler and improve mixing. The speed of the feeders is modulated to maintain output when the fuel conditions or loads change. As with BFB boilers, the fluidised bed consists of fuel, ash from the fuel, inert material (e.g., often sand), and possibly a sorbent (e.g., limestone) to reduce sulphur emissions if required.

In CFB systems, the fluidising velocity is maintained at 4-6m/s to prevent a dense bed from forming and to encourage carry-over of solids from the bed. This velocity allows entrainment of bed material and turbulent mixing throughout the combustor. The entrained bed material is separated from the flue gas stream leaving the combustor either by impact or cyclone separation. Once removed from the flue gas stream, the material is returned to the combustor via a loop seal. Fuel is typically fed pneumatically into the combustor near the bottom of the unit and/or in the solids return leg.

The CFB technology is better suited for higher fuel input capacities than moving bed and BFB combustion because the injection of fuel and sorbent or bed makeup material into the circulating media is much easier than evenly spreading the feed across a large grate or bubbling bed. In addition, the greater depth of combustion zone, potentially covering the entire boiler height, increases the unit capacity.

A typical CFB boiler with features included for biomass fuels is illustrated in Figure 59.

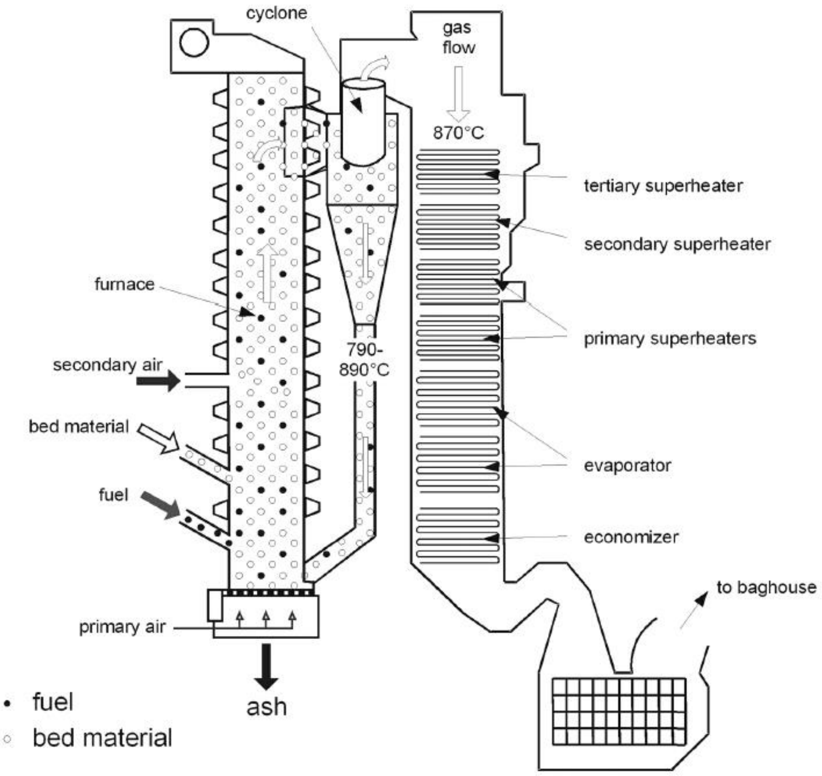

Figure 59: Typical CFB for Biomass Combustion[34]

As with moving bed and BFB boilers, the biomass fuel rapidly devolatilises as it is injected into the boiler. This results in combustion occurring throughout the combustor. Overfire air is required to ensure complete fuel combustion. The higher concentration of bed material throughout the combustor (compared to the BFB technology) absorbs the heat of combustion as it occurs above the over bed air injection point. The major heat transfer mechanism to the combustor water walls is conduction from the bed material to the wall surfaces and in-bed heat transfer surfaces as opposed to radiant heat with the moving bed and BFB technologies.

3.4.2 Development status and timescales

Fluidised Bed Boiler Design Evolution

Fluidised bed combustion was developed in the 1970s and 1980s as a means of burning a range of solid fuels including materials containing higher than average ash or sulphur such as coals, and high moisture fuels such as sewage sludge and other biomass fuels. Application of the CFB technology to biomass combustion was developed in Scandinavia and the USA during the 1980s. By the 1990s CFB technology was fully developed allowing projects above 100 MW_e to be developed for biomass fuels, particularly for the pulp and paper industry in Scandinavia. The current status is therefore TRL 9.

The evolution of circulating bed design has been primarily driven by the combustion of lower moisture content fuels such as coal and petroleum coke (petcoke). Foster Wheeler has stated that of 350 CFB boiler projects, 50 were designed for biomass applications, and 50 for wastes applications[35]. The majority of these applications are in Scandinavia and the USA. As with BFB boilers, CFB units also offer a high degree of fuel flexibility and offer a suitable and economical technology for burning biomass, particularly at larger scales (i.e. 100 - 300 MW_e).

Early CFB units[36] were in the size range appropriate for most biomass plants (i.e. 10-50 MW_e), whereas the present use of CFB technology is more focused primarily on large fossil fuel fired units to enable single boiler steaming capacities in excess of 250 MW_e in the United States and 400 MW_e in Europe. The largest (fossil fired) unit currently in operation globally has a supercritical steam flow rate in excess of 1,900tonnes/h at 254bar and 571°C with an electric output of 600 MW_e at the Baima power plant, China.

CFB Electricity only

The 125 MW_e Kaukaan Voima power plant at the UPM-Kymmene Paper Mill site in Lappeenranta, Finland, which started up in 2009, was for a time the largest biomass-fired single CFB boiler in the world. However, in 2012, a larger 205 MW_e plant (also supplied by Foster Wheeler) started up in Polaniec, Poland[35]. The CFB boiler has a steam flow of 570tph, steam pressure of 127.2bar and a temperature of 565°C. The steam generated by reheating has a pressure of 20bar and a temperature of 565°C.

For 100% biomass, CFB technology is typically recommended above 100 MW_e as the technology is better suited for feeding large volumes of biomass and for bed inventory management. The fuel may be fed into the circulating media and this is much more easily achieved than feeding across a large grate or bubbling bed.

CFB CHP

CFB boilers tend to be cost effective compared to BFB above c. 75 MW_e. However, there is generally a limited market requirement for heat generation at this scale so CHP deployment is likely to be limited.

A plant at the Tullis Russell paper mill in Scotland has completed commissioning (2015), with an output of 50 MW_e and a thermal demand of 120 MW_{th}. Unfortunately, the paper mill has recently closed and the plant is now operating in power only mode. This is currently the only biomass-fired CFB plant currently operating in the UK, although several large biomass power projects using CFB technology have been proposed in the past. Many of these projects have been abandoned, due to reduction in subsidies, the introduction of RO capacity cap, and the requirements of the CfD regime for biomass power stations to include CHP. MGT Power are planning a 299 MW_e CFB combustion plant on Teesside using mainly imported forestry pellets, but with an onsite drying facility for locally sourced wood chip. The Anglesey Biomass Power Station project is also looking to build a 299 MW_e biomass-fired CFB plant, also providing heat to a local eco-park.

CFB Heat

As stated above, CFB boilers tend to be cost effective above c. 75 MW_e (i.e., c. 300 MW_{th}). There is generally a very limited market requirement for heat generation at this scale so heat deployment is unlikely. The TEABPP team did not gather techno-economic data, as none of the CFB boiler manufacturers appear to offer the technology for hot water applications.

3.4.3 Impact of different feedstock parameters on operation and cost

There are many similarities between BFB and CFB technologies and as such many of the effects on boiler operation and cost due to varying feedstock parameters are applicable for both systems. Detail on the techno economic impacts relevant to general fluidised bed systems are in the BFB chapter previous. Impacts specific to CFB systems are outlined below:

Sizing

CFBs can tolerate fuel sized in a narrower band than moving beds, but the action of the fuel in the boiler is quite different. Table 12 illustrates fuel requirements. Fuel size limiting factors may also exist with the feed system to the boiler feed hopper. The material handling system would generally require the material be sized smaller prior to introduction to the boiler feed hopper. This would likely be located after reclaim from storage.

The large CFB boilers are technically capable of burning a wide variety of biomass fuels as well as coal, provided the fuel is sized appropriately.

As with BFB, plant availability is dependent on the feedstock being properly sized and containing minimum contraries (stones).

Table 12: Fuel Flexibility Features

Fuel Parameters	CFB	BFB
Moisture Content	Capacity de-rate above c. 50%	No de-rate with design fuel range up to c. 55%
Fuel Size Limits, Maximum Size	Max sum of length + width + thickness = 200mm	No data
Fuel Size Limits, Forest Residue	No data	200 x 50 x 25 mm
Fuel Size Limits, Bark	No data	250 x 50 x 25 mm
Fuel Size Limits, Minimum Size	80% greater than 3mm	70% greater than 6mm

Nitrogen

Compared to a BFB, the lower operating temperature of a CFB as well as the increased residence time provided by the cyclone or impact separator allows a SNCR system to be very effective. The freeboard area above the dense bed of the BFB results in an area of combustion of the volatile fuel components that are released as the fuel is devolatilized in the bed. This area above the dense bed of a BFB resembles the open burning of volatile fuel components in a moving bed boiler. However, with CFB technology, there is sufficient bed density in this location that the bed material absorbs this combustion heat so that the combustion temperatures of CFB boilers are not as high as those of a BFB boiler. This enables the CFB boiler to operate with lower peak temperatures and lower uncontrolled NO_x emissions. The flue gas residence time in the particulate separation portion of the CFB as well as the lower temperature is expected to result in better SNCR performance to reduce the NO_x formed as an unintended combustion by-product in the CFB boiler. Reduced NO_x and improved SNCR performance means reduced urea consumption.

NO_x control is required regardless of the fuel, and the prevailing technology for NO_x control is SNCR. Because of the lower combustion temperatures (compared to a moving bed and BFB), NO_x emissions from a CFB boiler that burns biomass will generally be less than 0.23kg/MWh.

Sulphur

The improved mixing of bed material, fuel and sorbent feed allows for significant advantages of sulphur capture and decreased sorbent consumption.

Sensitivity Analysis - Power

The cost curve in Figure 60 corresponds to the total investment cost per unit of output capacity for a circulating fluidized bed power plant. It has been derived from quotes provided by suppliers (2007 – 2011) for all-in overnight EPC costs.

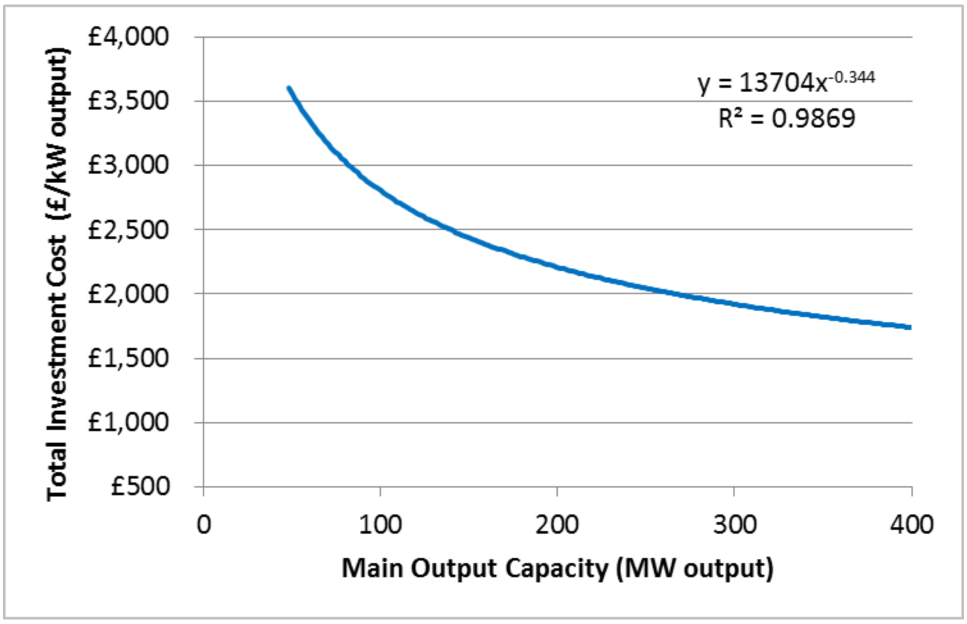

Figure 60: CFB (power) total investment cost vs. output power capacity (derived from seven supplier quotes (2007 - 2011) and B&V data for all-in overnight EPC costs)

Total operational costs for CFB Power technologies include fixed costs (insurance, maintenance parts and labour) and variable costs (operations labour, additional fuel, reagents, water). Figure 61 shows the relationship between annual total operational cost for a CFB per unit of energy output and the main output capacity rating.

Except where the parameter of interest is shown on the x axis, the curves in Figure 61 to Figure 68 are all created using the following base values: Main output capacity = 100 MW_e; station capacity factor of 85%; feedstock moisture content (WB) = 50%; feedstock ash content (DB) = 1.5%; feedstock nitrogen content (DB) = 0.35%; feedstock sulphur content (DB) = 0.05%; and a feedstock chlorine content (DB) = 0.04%.

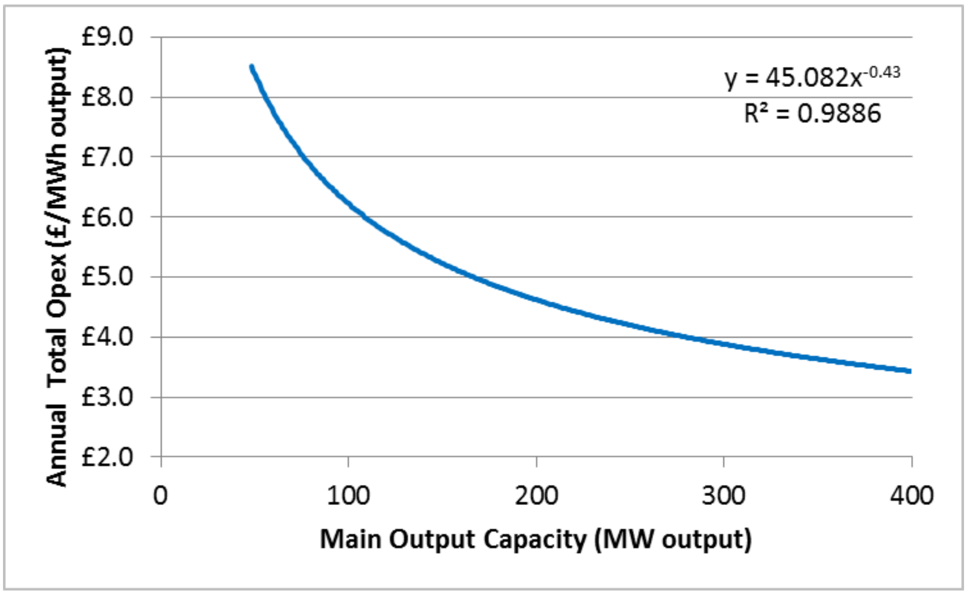

Figure 61: CFB (power) annual operating cost per unit of electrical output vs. output power capacity (derived from B&V industry data)

Figure 62 below highlights the change in overall CFB station efficiency (LHV) over varying feedstock moisture content. The efficiency is calculated as electrical energy output over fuel energy input, as such parasitic loads post-generation are not included. The curve is considered to be a representative average across the CFB potential rated output range.

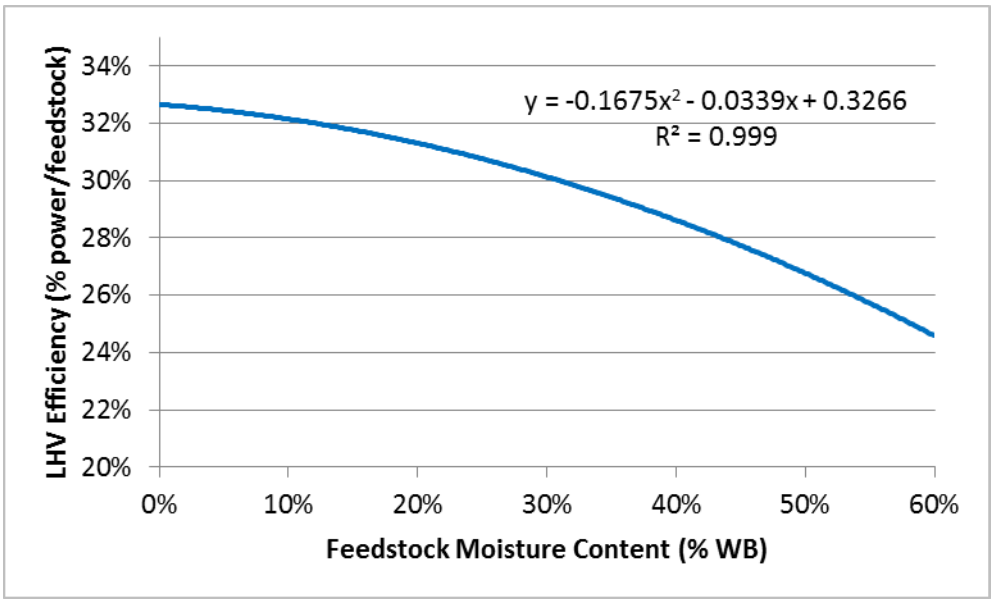

Figure 62: CFB (power) efficiency vs. feedstock moisture content (derived from B&V industry data considering a representative average across the technology potential rated output range)

The total operational costs for CFB Power technologies (as previously described for Figure 60) are related to feedstock moisture content as shown in Figure 63 below.

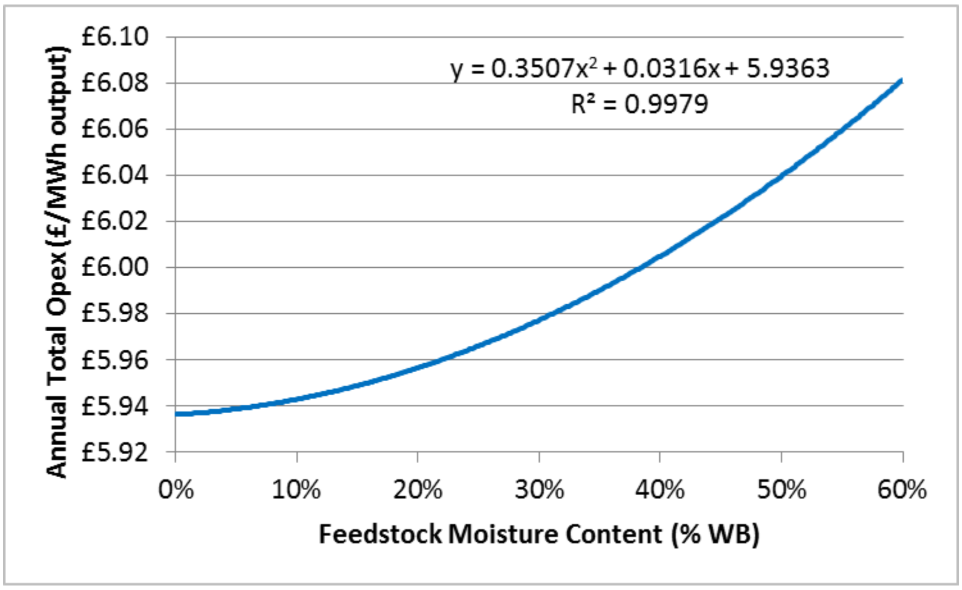

Figure 63: CFB (power) annual operating cost per unit of electricity output vs. feedstock moisture content (derived from B&V industry data)

Annual operational costs for CFB Power stations vary with the percentage of ash present within the feedstock, as shown in Figure 64 below.

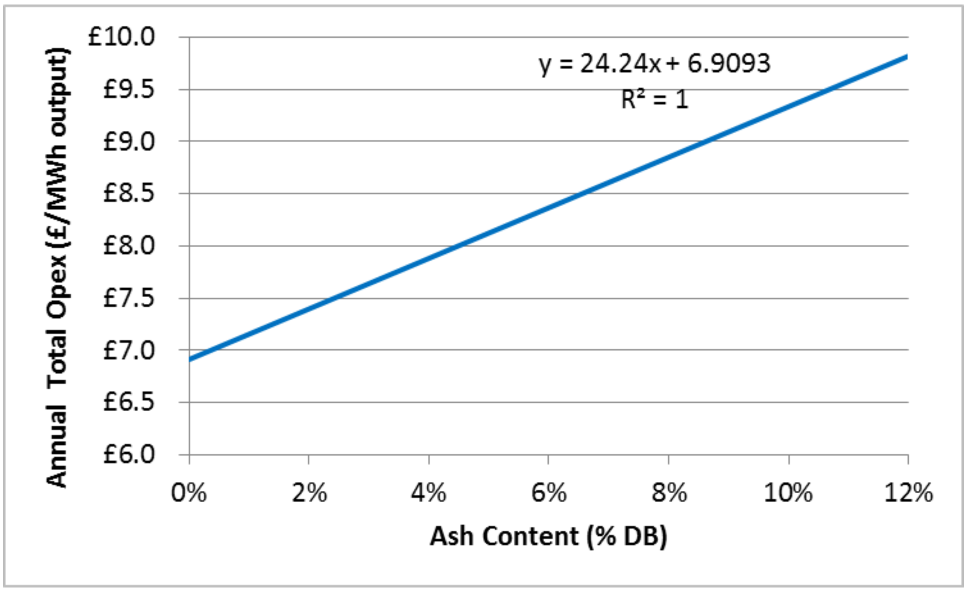

Figure 64: CFB (power) annual operating cost per unit of electricity output vs. feedstock ash content (derived from B&V industry data)

Figure 65 below displays the relationship between nitrogen content within the biomass feedstock and the annual cost of NO_x abatement via urea injection.

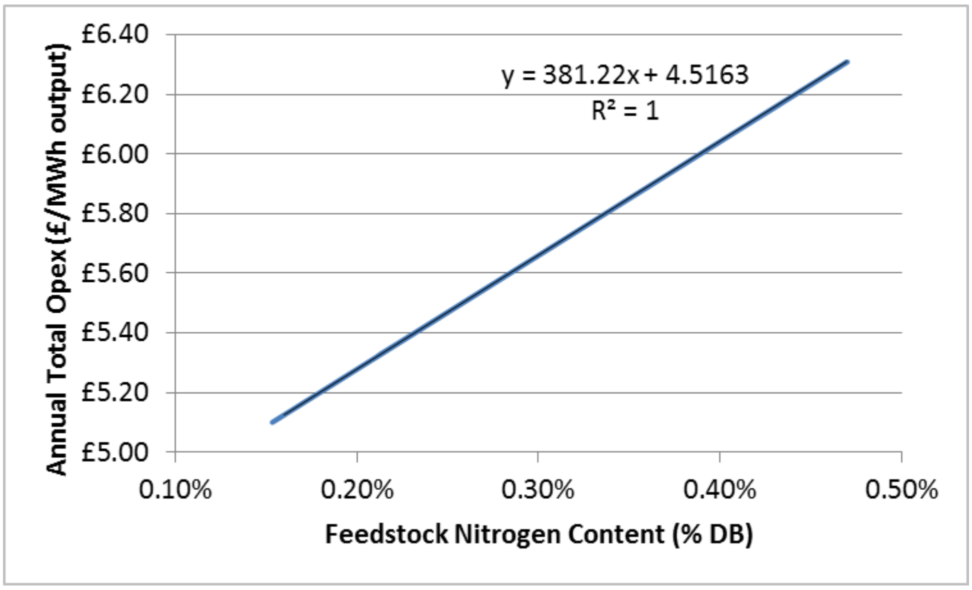

Figure 65: CFB (power) annual operating cost per unit of electricity output vs. feedstock nitrogen content (calculation based on section 2.4.2.4)

Figure 66 below displays the relationship between sulphur content within the biomass feedstock and the annual cost of abatement via lime addition.

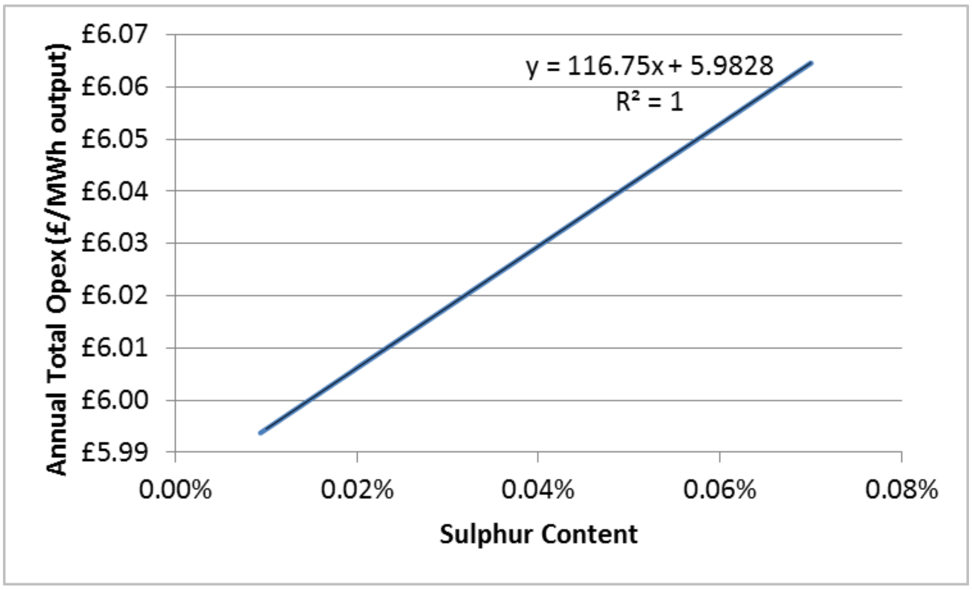

Figure 66: CFB (power) annual operating cost per unit of electricity output vs. feedstock sulphur content (calculation based on section 2.4.2.4)

Figure 67 below displays the relationship between chlorine content within the biomass feedstock and the annual cost of abatement via lime addition.

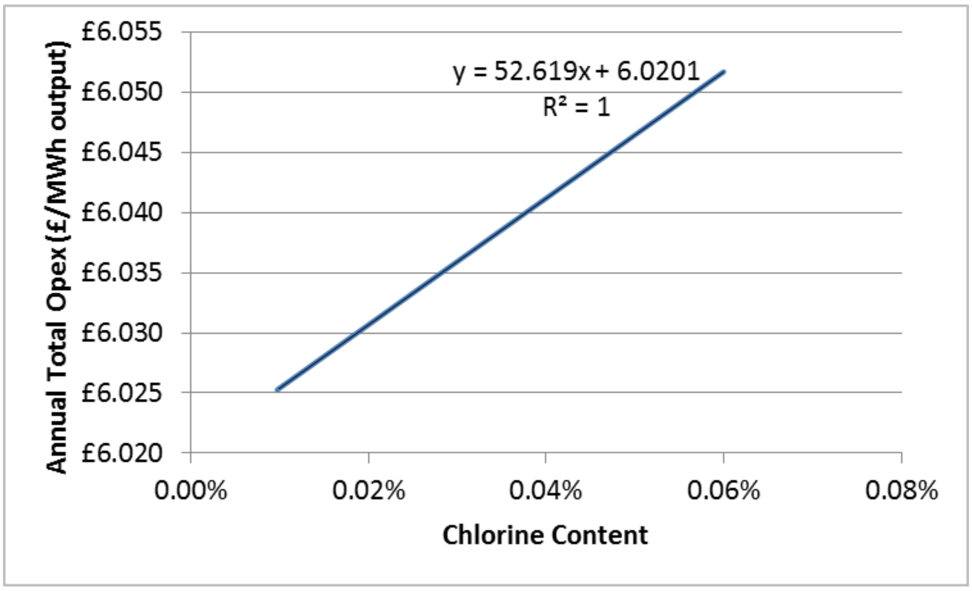

Figure 67: CFB (power) annual operating cost per unit of electricity output vs. feedstock chlorine content (calculation based on section 2.4.2.4)

Figure 68 below displays the relationship between the alkali index (kg K_2O and Na_2O per GJ energy) and the annual total operational cost for a CFB per unit of energy output.

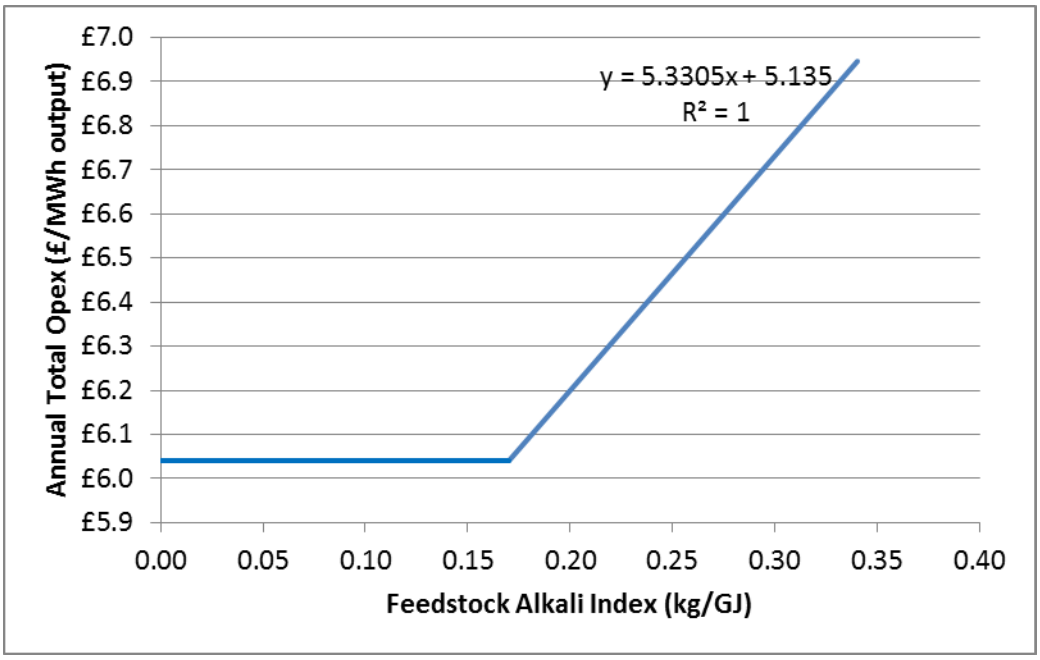

Figure 68: CFB (power) annual operating cost per unit of electricity output vs. feedstock alkali index (derived from formula based on [37] and section 2.4.2.5)

3.4.4 Available options for improvement

Major improvements have occurred with CFB boilers since the initial units were installed in the 1970s and 1980s.

CFB boilers share many of the same advantages as BFB boilers with regard to fuel flexibility, combustion efficiency, and emissions. Some of the more significant improvements in CFB boiler design are listed in Table 13 and summarised below.

Table 13 : Improvements Analysis of CFB Boilers

Issue arising due to biomass characteristic	Options to ameliorate	Evaluation of effect	Long Term Improvements
Poor bed conditions from non-homogeneous feedstock	Multiple fuel input points; Variable fuel and bed feed rates.	Improved bed / fuel ratio reduces cases of unburnt fuel in larger size / higher MC; Better mixing with sorbents (if added) in bed;	Reduced bed medium turnover; Reduced sorbent requirements; Improved combustion efficiency.
Biomass primarily low sulphur	Reduces requirements for sorbent; If required, sorbent can be added post-combustion (see BFB) but improved mixing in CFB means typically added in furnace.	Inert content of bed is increased.	Reduced sorbent requirements; Reduced costs of post-combustion plant; Reduced emissions and environmental impact.
Biomass oversize and ash affecting bed conditions	Panels installed to smooth transition from refractory to tube wall; Multiple independent locations for bed extraction; “Open” combustor floor bottom for multiple and reliable bed material drains; Particle cyclones are top supported steam cooled.	Reliable diagnosis of hot and cold spots; Reduced slagging; Better refractory life; Reduced bed agglomeration / bed slump.	Reduced maintenance Opex; Improved boiler availability.
Dry fuels increasing bed temperatures	Removal of heat exchanger surfaces from reducing zones within the bed; Varying combustion air make up – addition of recirculated flue gas	Better corrosion and erosion resistance of furnace heat exchanger surfaces; Increased control of O_2 content and reaction rates.	Reduced maintenance Opex; Improved combustion efficiency. Improved boiler availability / longer plant life; Reduced emissions and environmental impact.
Increased volatile release with biomass (compared to coal)	Secondary combustion air control.	Increased control of heat release in freeboard; Increase rate of complete combustion.	Reduced emissions and environmental impact; Improved combustion efficiency.

Capex

Although manufacturers might quote small-sized CFB boilers, these units generally cost more than other combustion technologies, making them difficult to justify for smaller biomass plants. In general, biomass CFB boilers are not expected to be economically competitive at scales less than 75-100 MW_e.

Parasitic load

The higher fluidising velocity of the CFB boiler combustor does require higher auxiliary power requirements than a comparatively sized BFB boiler. A disadvantage of CFB boilers compared to moving bed and BFBs is the larger auxiliary power requirement for the primary fluidising air fans as well as fluidised air blowers for loop seal operation. There is opportunity here for reducing the parasitic load and station efficiency.

Boiler efficiency

As a general rule, the higher fluidising velocity and greater mixing of bed material with a CFB boiler enable lower excess air operation and lower unburned carbon losses than a BFB boiler burning similar fuels, which results in a higher boiler efficiency for CFB boilers.

This results in a consistent heat output and flue gas quality. The high heat transfer of the fluid bed medium also provides high carbon burnout leading to increased combustion efficiency and therefore boiler efficiency.

The typical boiler efficiency for circulating bed combustion units firing biomass is approximately 70-80% (heat output to input biomass). Efficiency improvements have been made with the introduction of steam- or water-cooled cyclones to reduce the heat loss of the original air-cooled units.

Fluidised bed ash extractors have been developed to recover a large portion of the heat previously lost with the water-cooled screw type ash extractor systems.

Availability

New fluidised bed plants should be expected to operate at reliabilities similar to new pulverised coal units, with expected target equivalent availabilities around 90%.

Availability improvement features have included panels in the combustor sidewalls to smooth the interface between the refractory covered and bare wall tube panels. Availability improvements were also made with the conversion to steam- or water-cooled cyclones to reduce refractory maintenance and expansion joint issues, which were prevalent with the original bottom-supported cyclone design. Design provisions to enable inert bed material feed on CFB boilers burning low ash fuels (such as clean biomass) have been made to reduce the agglomerating nature of bed material that becomes too highly concentrated in reactive compounds, such as calcium carbonate, sulphates, and other reactive alkali components.

The prime CFB design evolution has been for increasingly larger capacities as well as increased steam temperatures, reheat for steam cycle heat rate improvement, and higher pressures for once-through supercritical steam generating capability.

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[34] The handbook of biomass combustion and co-firing (2008), edited by Sjaak van Loo and Jaap Koppejan, Earthscan London.
[35] Timo Jäntti and Kalle Nuortimo, 2012, “Foster Wheeler Advanced Bio CFB Technology for Large Scale Biomass & Peat Firing Power Plants” http://www.fwc.com/getmedia/5ab3ec92-6d93-44e2-a57b-05f8e0b066f1/TP_CFB_12_04.pdf.aspx?ext=.pdf
[36] The early generation of CFB boilers were typically fitted with significant in-bed heat transfer surfaces in the combustor in the lower density portions of the bed as well as a relatively small number of above bed feed points for fuel and sorbent. Bed ash extraction provisions were provided at a small number of points in the combustor sidewalls. The orientation of the bed fluidisation nozzles helps to direct bed material to the bed drain points. The original installations utilised bottom-supported, heavy refractory covered, air-cooled particle separators (cyclones) that required excessive maintenance and introduced additional heat loss that reduced the operating boiler efficiency. The combustor sidewalls were subject to rapid erosion at the interface between the refractory covered and bare water wall membrane tubes. In units that burn low ash fuels, bed material agglomeration and surface deposition occurred as a result of the chemical reaction-based agglomeration of the bed materials rather than the solid-to-liquid phase change reactions of ash particles of conventional pulverised coal or moving bed boilers. Bed ash extraction systems (typically of the water-cooled type) were provided, which were a source of heat loss as well as maintenance requirements due to the aggressive environment of the high temperature ash stream.
[37] Fahmi R., Bridgwater A. V., Darvell L. I., Jones J. M., Yates N., Thain S., et al. (2007). The effect of alkali metals on combustion and pyrolysis of Lolium and Festuca grasses, switchgrass and willow. Fuel 86, 1560–1569 10.1016/j.fuel.2006.11.030.

3.5 Dust Suspension

3.5.1 Technology description

Dust suspension combustion is analogous to pulverized coal combustion. The biomass feedstock is dried (if necessary) and ground in grinding mills to a small size, of the order of 1 mm and smaller. This biomass powder is then entrained in an air stream through burners into the combustor. Combustion then occurs at a high rate in the volume of the chamber producing heat, alongside the gaseous combustion products and ash. The residence time is short, of the order of a few seconds, and the particles are not recirculated. This requires that the design of the burners and combustor need to achieve the complete burnout of the particles within this residence time whilst maintaining a stable flame.

The heat is transferred to water tubes in the combustor walls and convective heating surfaces. This leads to the production of high temperature steam that is used in a turbine generator to produce electricity. The concept is illustrated in Figure 69, where the coal is replaced by biomass, or co-firing of coal and biomass. A variation of this is a combined heat and power unit, where in addition to the electricity generation, some of the steam produced is used to heat water in order to provide space heating.

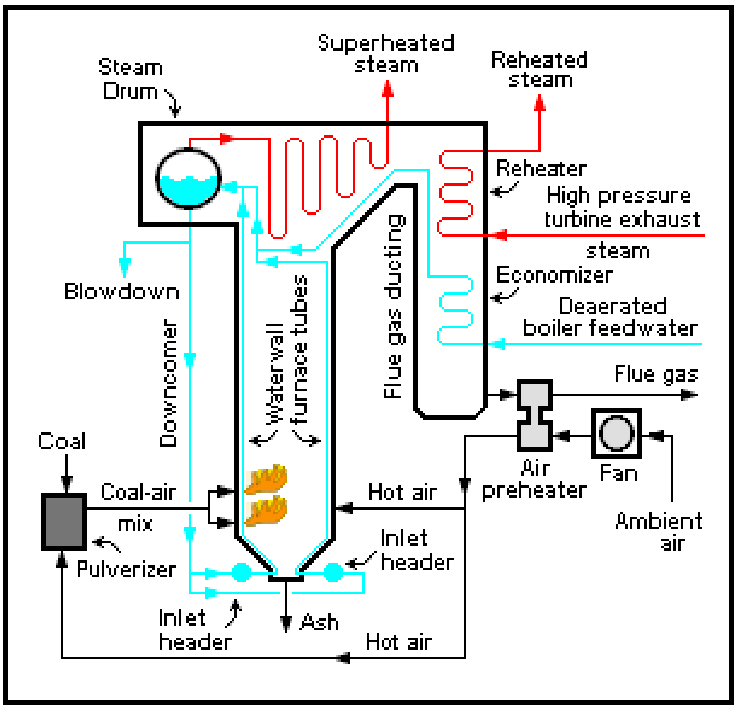


Figure 69: Simplified diagram of a conventional coal-fired steam generator[38]

Dust combustion plants are in general capable of delivering an additional 5% of their rated load capacity as primary load support within 30 seconds when operating at loads between 50% and 90%[39], by utilising certain water/steam buffers within the unit. Secondary load support takes over after about 5 minutes and is able to sustain this load rise and increase it if necessary.

The dust combustion costs include pollutant removal technologies applied to meet the Large Combustion Plant Directive: Selective Catalytic Reduction and inside furnace control for NO_x removal, SO2 removal via a lime spray dryer, flue gas desulphurization and Cold-side Electrostatic Precipitation (CEP) for particulate removal. In addition, a mercury removal system via carbon injection is used and cooling water is recirculated through a wet cooling tower to maintain plant efficiency and reduce water consumption.

3.5.2 Development status and timescales

Power

This is a mature technology and can be considered at TRL 9 in the context of co-firing, but TRL 8 at the level of 100% biomass firing. TRL 9 is likely to be achieved in less than 5 years, as further units are converted. Pulverised coal technology has been around since the 1910s. Commercial power plant capacities are in the range from ~400 to 1000 MW_e.

In the USA, four direct coal and biomass co-firing facilities were operational during the 2000's, along with at least 38 in which the concept had been tested[40]. In the UK, all the major pulverised coal power plants have co-fired a small percentage of biomass in the past, but many have either stopped with reduction in government support, or have moved towards 100% biomass firing/full conversion.

Drax has already converted two of its 6 units to pure biomass firing in 2013 and 2014, and has advanced plans to convert the 3rd unit in 2015/16 (and is also evaluating options for converting a 4th unit). These two units generate 645 MW_e each and are the largest dedicated biomass dust combustion boilers in the world. Lynemouth Power's 420 MW_e plant is also due to be converted, under the CfD regime – Eggborough and Uskmouth also have plans for conversion, but have not secured support.

In 2014, 7.9 TWh of power representing 29% of the total Drax station output was generated from biomass[41]. Drax sourced 78% of its biomass requirement from outside of the UK, consuming 60% of all US wood pellet exports. E.ON's Ironbridge plant was also fully converted to biomass in October 2013 (two 450 MW_e units), but a fire in one turbine hall has halved capacity until the plant shuts at the end of 2015 under the Large Combustion Plant Directive (LCPD). RWE's Tilbury B plant suffered a major fire[42], after full biomass conversion in 2011 of its 3 units (to 750 MW_e). Although it was repaired and came back online, it was shut in 2013. The closure was due to the plant being opted-out of the LCPD and RWE deciding not to re-permit the plant after it used its available running hours.

Dust formation, off-gassing and self-heating are well-recognised and important safety issues regarding the use of wood pellets in dust suspension combustion. There are risks of explosions and fires due to the generation of dusts during the handling, transport and milling of pellets. In addition, since biomass is prone to self-heating, the large-scale storage of biomass in piles or silos can result in spontaneous ignition. The prevention of explosions and fires require the implementation of safety measures, which include the design of explosion containment equipment (e.g. hammer mills) and control of the oxygen concentration in pulverisers, cyclones, conveyors, silos, etc. by using inert gases such as nitrogen, steam or CO_2[43].

Feasibility studies into conversion to 100% biomass firing have been performed at many sites worldwide, with Doosan Babcock being involved on 7 projects of this type since 2009, including several in Ontario, Canada[44]. Denmark, Belgium and the Netherlands have also converted or co-fired several stations. Conversion of existing coal assets is significantly cheaper and faster than building a new dedicated biomass plant using dust combustion technology and hence the focus globally is on the conversion of aging coal assets. Based on the experience of Drax, the conversion process for individual boilers can be completed in about one year, as opposed to the 3+ year construction time for new build.

Despite these conversion activities, no new build 100% biomass dust suspension plants are imminent; Drax having abandoned in 2012 its plans for new biomass plant development due to changes in government subsidies[45] to favour conversion.

The performance and economic data for new build dust combustion technology therefore rely on theoretical estimates from coal conversion projects, rather than real-world operational experience of new build engineering studies.

CHP

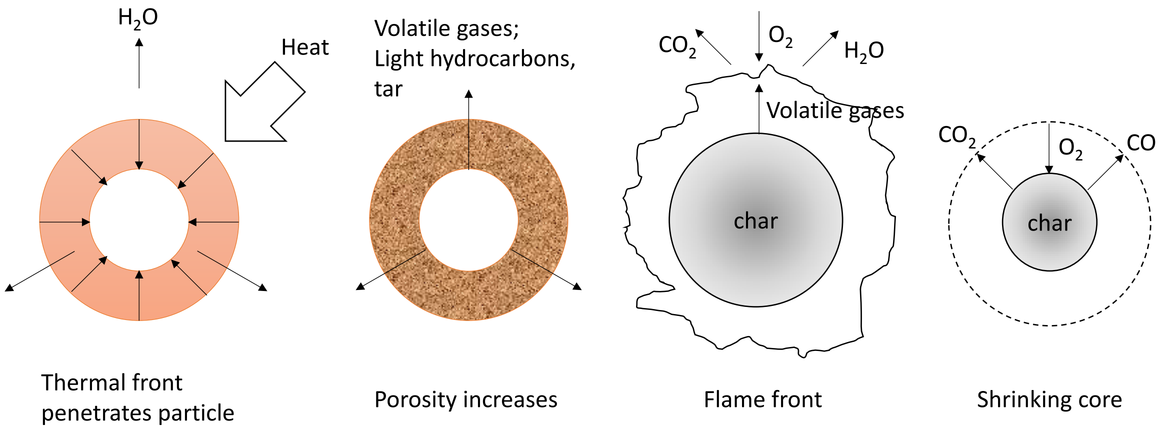
Combined heat and power represents a much smaller segment of biomass usage, typical sizes being from 1 to 100 MW_e[46]. However this market is typically supplied by other technologies, such as fixed and vibrating grate stoker boilers and BFB combustors, which dominate this sector[47]. Dust combustion systems are typically too large to have CHP applications.

3.5.3 Impact of different feedstock parameters on operation and cost

Biomass dust suspension combustion is in principle equivalent to pulverized coal combustion. However, properties of the biomass dust can be very different from coal, leading to different handling issues such as a significantly higher volume to be handled and stored as well as a susceptibility to degradation and spontaneous combustion, and a different combustion behaviour in the system. The main factors influencing operation are the moisture content, particle size, volatiles fraction, LHV, density and ash content, as discussed subsequently.

Moisture Content

Fuel moisture content influences the time taken for the particle to ignite and then to combust completely. Figure 70 illustrates the processes involved in particle combustion, where the particle undergoes a combination of drying, devolatilisation and combustion processes leading to a stratification of the particle composition.



Thermal front penetrates particle

Porosity increases

Flame front

Shrinking core

Figure 70: Particle combustion stages, adapted from Brown[48]

For a given particle size, the higher the moisture content, the more energy is required to dry it to the point that ignition occurs. This impacts the timescale of the process, leading to combustion to begin further away from the burner inlet. This has consequences in terms of ignition and flame stability, and the residence time required in the burner to achieve complete combustion. For 100% biomass firing, the moisture content should be reduced to 10% or less.

Particle Size and shape

One of the major differences between pulverized coal combustion and biomass dust suspension concerns the particle size and shape. Whereas coal can be ground to a very fine size of the order of 75 µm or less, the fibrous nature of biomass makes it difficult to reduce its size. Paulrud et al.[49] found from an investigation of wood powder produced from a variety of industrial impact and knife mills, 50% volume fractions ranging from about 250 to 750 µm. This impacts on the burnout properties in a similar way to moisture content, in that it takes longer to completely combust the particle. Studies have been performed into the grinding performance using a pilot scale hammer mill of various biomass samples as a function of moisture content[50]. They investigated various biomass samples at 8 or 12 % by weight of moisture, with hammer mill screen sizes ranging from 0.8 to 3.2 mm and showed that specific energy requirements can vary considerably between ~10 to 70 kWh/t, and it is not always the case that drier material is easier to grind.

Particle shape manifests itself in various ways, firstly fuel bridging may occur during the fuel feed process which causes instabilities in the fuel flow rate, secondly, finer particles may build up on surfaces inside the combustion chamber leading to instabilities in emissions when they become dislodged and combust inside the combustion chamber as effectively a larger particle. Finally, particle shape impacts upon drying and devolatilisation. Flake like particles have a larger surface to volume ratio leading to faster drying, devolatilisation and combustion than equivalent spherical particles.

Volatile Content

Wood dust has a much higher volatile matter content than coal. This has the effect of enhancing combustion close to the burner as these volatiles are released, leading to the stretching of the flame zone for a biomass dust flame due to this initial intense volatiles combustion followed by the effect of the larger particle sizes having a slower burnout time downstream[51]. The high volatile content and hence reactivity also has a safety implication in the use of biomass, from the risk of smouldering fires in biomass storage, and a requirement of sufficiently low air temperatures in milling to minimise volatiles release[52].

LHV and density

Biomass through its lower LHV and density compared to coal has about two thirds of the heating value of coal on a mass basis. This has implications for plant capacity and output power in systems originally designed for pulverized coal and can also decrease combustion efficiency. Combustion efficiency can be mitigated by modifications to both fuel and air feeding. These modifications consist of adjusting the air preheating to encourage faster particle drying and to increase the swirl number of the burner in order to improve the flame stability and increase the residence time of the particles in the flame zone to encourage complete combustion.

Alkali Metals

The characteristics of dust suspension firing involving high temperature of the order of 1200 °C [53,54], short residence time and small particle size lead to equilibrium being approached selectively, with the ash matter being strongly fractionated. Ash transformations in dust suspension firing of biomass have been studied by Nordgren et al.[55], both experimentally and using an equilibrium model at two global temperatures of 1000 and 1200 °C. They concluded that reactions involving condensed phases are kinetically limited compared to reactions between gaseous ash compounds. Thus gas phase reactions leading to chlorides, sulphates and carbonates are favoured at the expense of purely condensed phase reactions. However, reactions between condensed and gaseous phase components such as KCl or KOH with silica will occur leading to molten potassium silicates, if these condensed components then settle on the bottom of the boiler/furnace/furnace walls, then conditions for obtaining equilibrium are more favourable.

For the white wood pellets considered in the base case, with a moisture content of 10%, an ash content of 0.6 % and an alkali index of 0.15, fouling is not an insurmountable problem. In the simple analysis discussed below, effects on Opex are not assumed to occur until a threshold of 0.17 is reached. However, this will limit what are acceptable biomass feedstocks for this process. Any biomass with an alkali metal content above approximately 3 g/kg will probably lead to slagging, and certainly above approximately 7 g/kg[56], thus ruling out any herbaceous feedstock. Feedstock boundary conditions, including moisture and size are summarised in Table 14.

Table 14: Dust Suspension Combustion feedstock boundary conditions

Parameter	Value
Size range (mm)	< 1
Moisture content (%)	< 10
Alkali metal content (g/kg)	< 3

Sensitivities

The key sensitivity is the Capex as a function of scale, as illustrated in Figure 71. In this case, the base case costs are derived from a 400 MW pulverized coal plant cost[39], with a 0.68 power law applied.

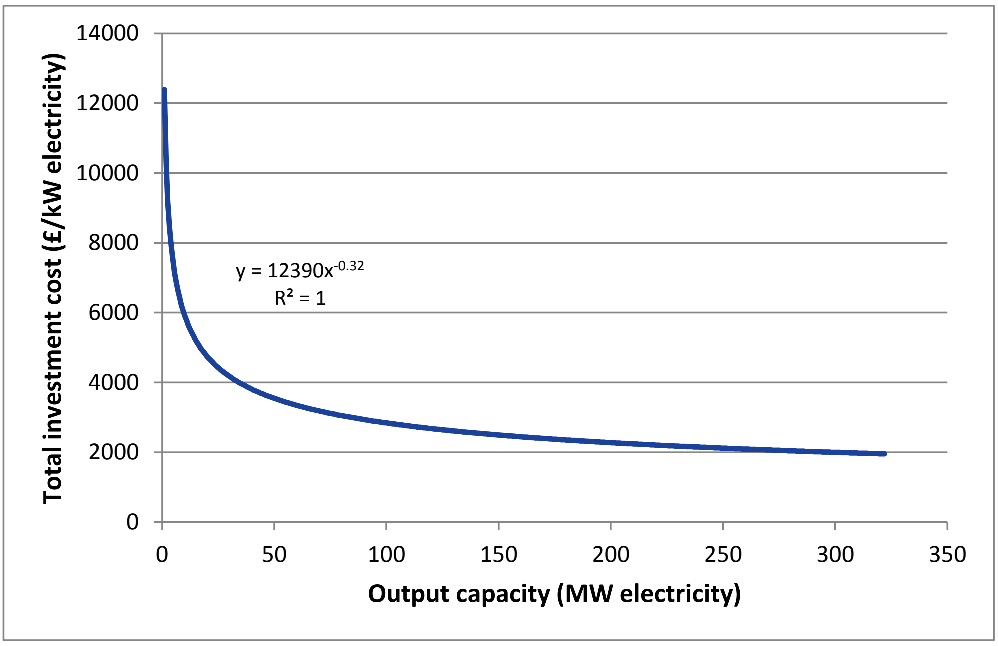


Figure 71: Dust combustion total investment cost vs. output power capacity (data based on [39])

Total Opex as a function of biomass ash content (base case of 0.6 %) is illustrated in Figure 72, based on the increased maintenance costs that higher ash contents impose on the plant. the base case Opex is composed of the following: Insurance (1.5% of TIC), maintenance: parts (2% of TIC), maintenance: labour (60% of operations labour), operations labour: UK labour rates up to 2015[57] & Wessel correlation[59], reagents/materials[61] and water[61].

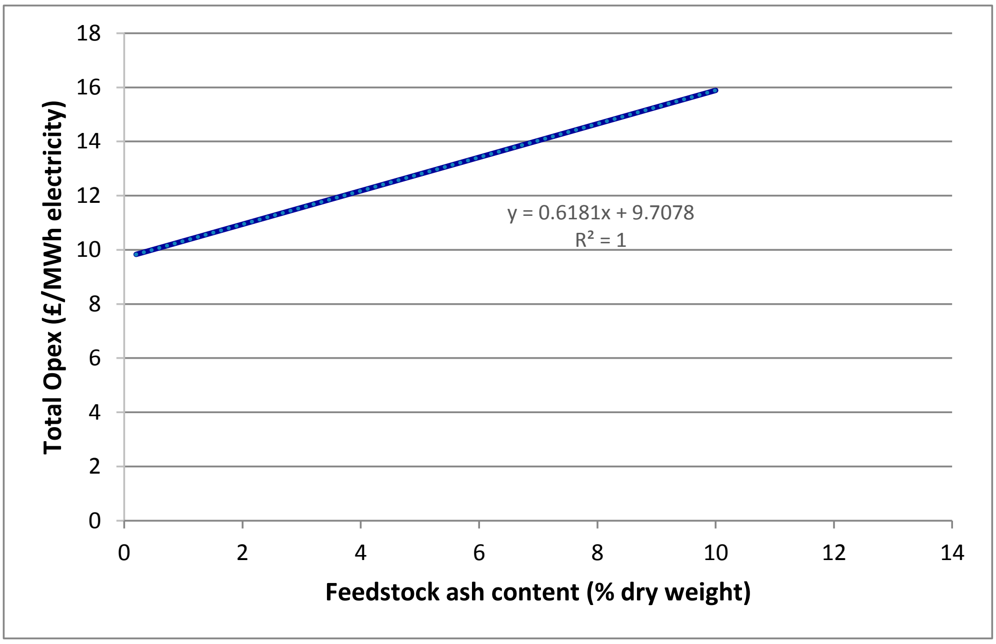


Figure 72: Dust combustion total opex vs. feedstock ash content (based on [58, 59, 61] and Sheffield calculations from section 2.4.2.3)

Figure 73 illustrates the variation of total Opex with Alkali index, making the assumption that once the threshold of an alkali index of 0.17 is crossed at which fouling is probable, there is then a linear increase in total Opex until a factor of 1.15 increase is reached at the threshold of 0.34 at which fouling is almost certain. In this example, the base case conditions equate to an alkali index of 0.154.

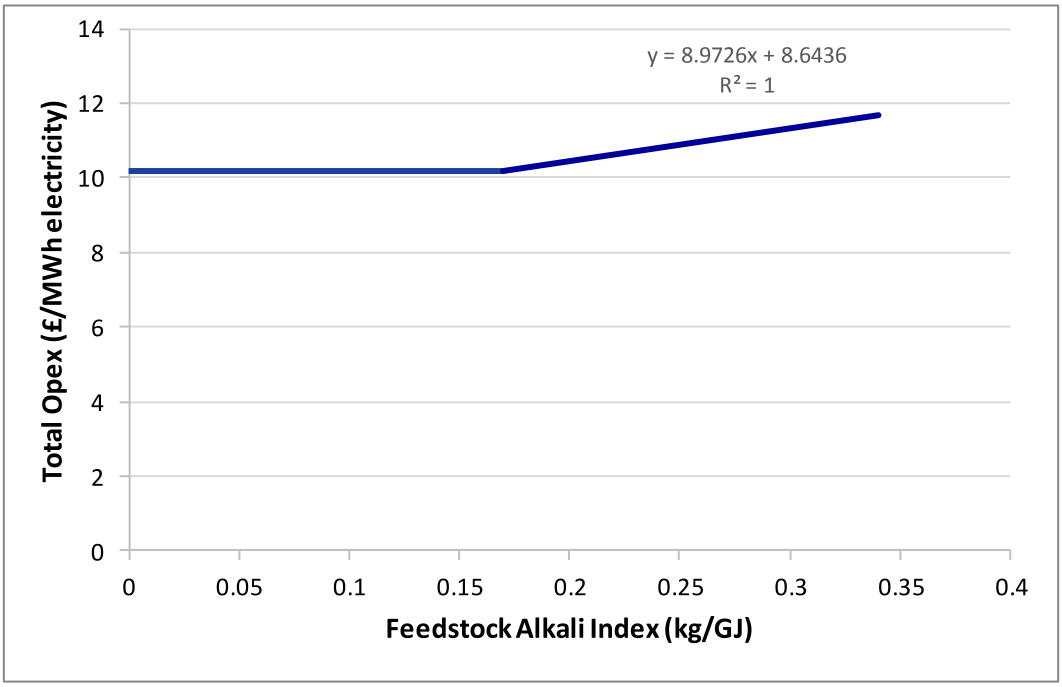


Figure 73: Dust combustion total opex vs. feedstock alkali index (formula based on [60] and section 2.4.2.5)

Figure 74 illustrates the variation of efficiency vs moisture content (base case of 10%, given the white wood pellets used). Potential emissions of NO_x, sulphur and chlorine into the atmosphere are accounted for by a procedure of calculating the Opex for the mitigation of each component, using a urea based NO_x mitigation cost and a lime based mitigation cost for both sulphur and chlorine.

Figure 75, Figure 76 and Figure 77 show the total Opex as a function of biomass nitrogen (base case of 0.45 %), sulphur (base case of 0.01 %) and chlorine (base case of 0.01 %) content respectively, with the importance decreasing in the order of N>S>Cl.

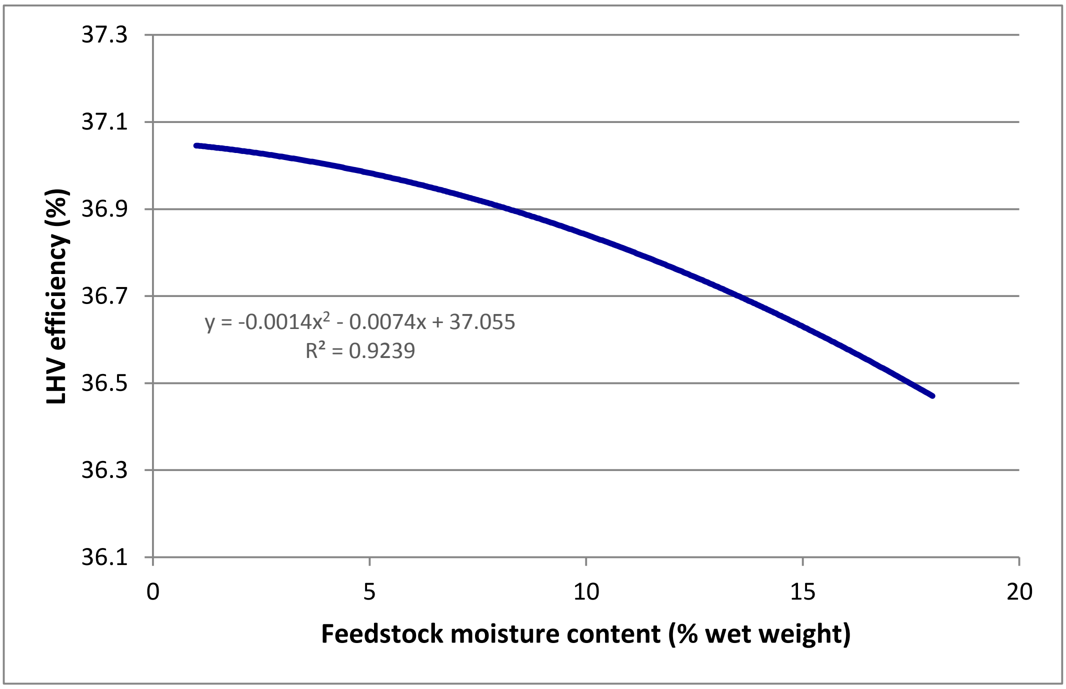


Figure 74: Dust combustion electrical efficiency vs. feedstock moisture content (based on an IECM[61] calculation by Sheffield)

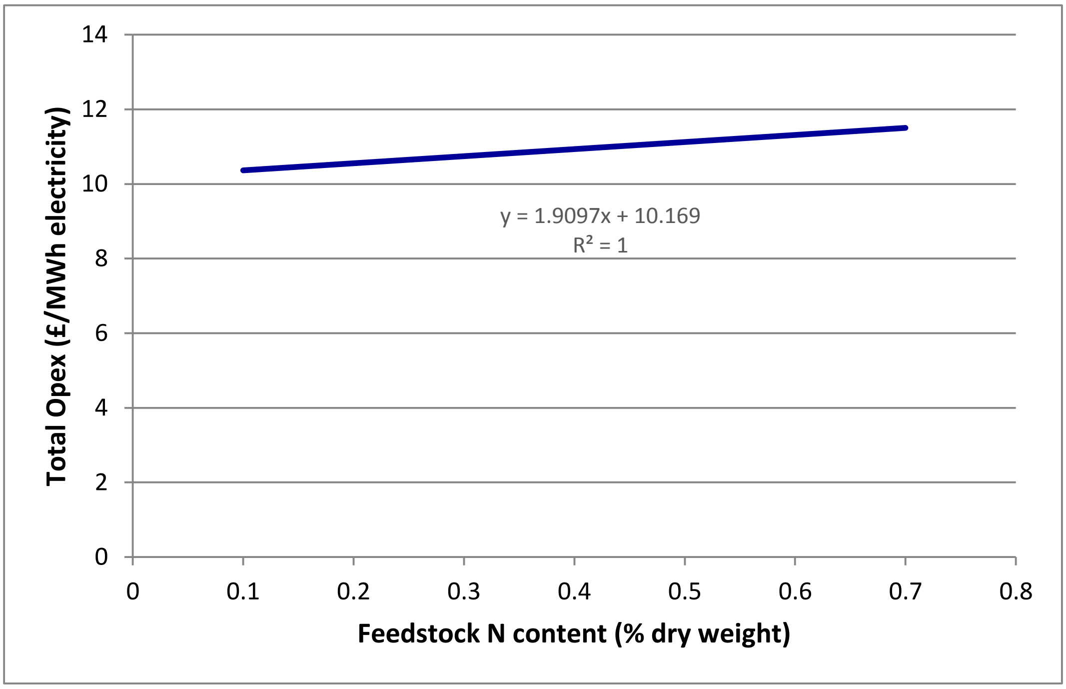


Figure 75: Dust combustion total opex vs. feedstock N content (calculation based on section 2.4.2.4)

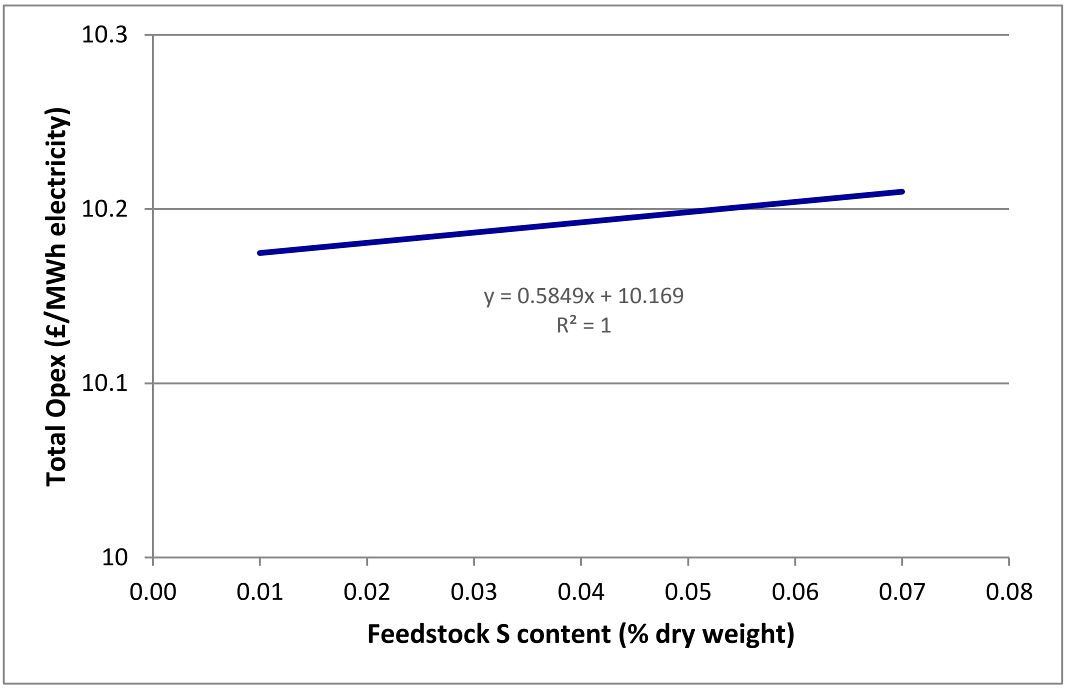


Figure 76: Dust combustion total opex vs. feedstock S content (calculation based on section 2.4.2.4)

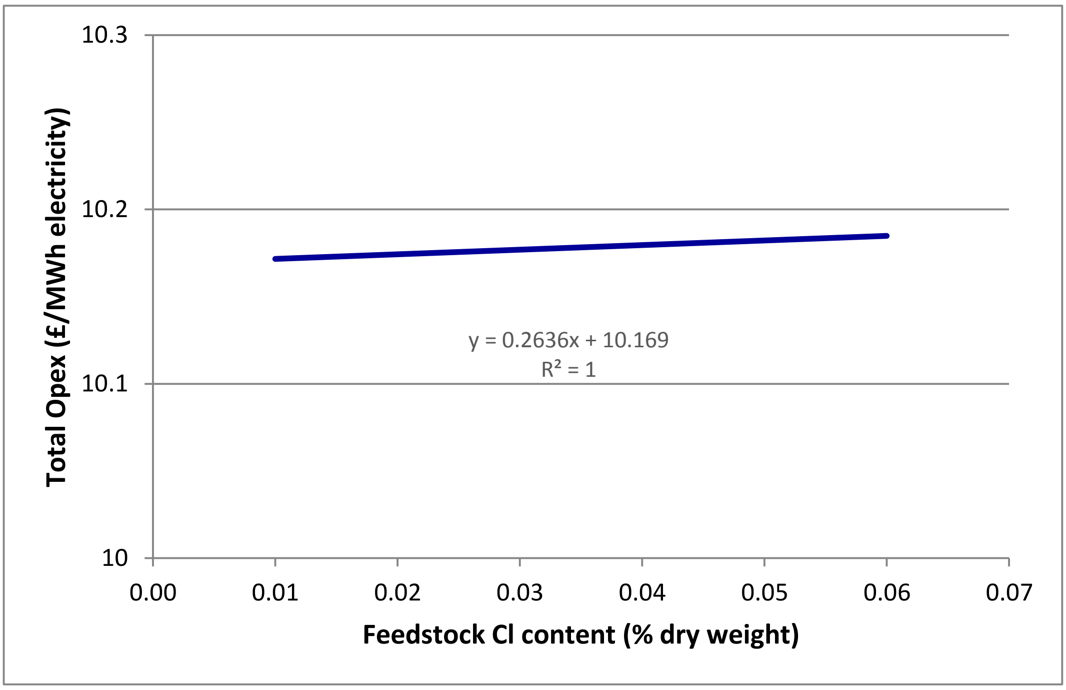


Figure 77: Dust combustion total opex vs. feedstock Cl content (calculation based on section 2.4.2.4)

3.5.4 Available options for Improvement

Given the fact that the current implementations of this technology are restricted to conversion of existing pulverised coal fired power stations, the main options for improvement in this scenario are to modify the properties of the biomass feedstock to have a closer resemblance to coal. The use of torrefied pellets over standard pellets for example is one option, the increased energy density compared to wood pellets leads to lower volumes needing to be handled at the power plant for the equivalent energy input. The increased cost of such an option may be offset against the reduced energy requirement of grinding torrefied wood[62]. Additionally, the decrease in particle size distribution of the ground torrefied pellets will improve the properties of the dust with respect to its entrainment into the combustor and its subsequent combustion. It will have a reduced timescale for drying, devolatilisation and burnout in the combustor, therefore becoming closer in behaviour to pulverized coal, for which the burner and combustor were originally designed.

For a new build implementation of a pulverized dust combustion plant, the main option for improvement would be optimisation of the burner and possibly the furnace design to better accommodate the different combustion characteristics of pulverized biomass compared to pulverized coal. As discussed in section 3.5.3, a combination of the increased moisture content, volatile content and size range of ground biomass stretches out the observed combustion zone such that designs optimised for pulverized coal are no longer appropriate. Customised burners have already been developed by Doosan Power Systems to address this[44]. Their assessment also concluded that there were no significant changes to the actual furnace and boiler performance, there is still a requirement for a finely ground low moisture content biomass feedstock that cannot be mitigated and the only reasonable method of handling ash deposition and corrosion is by control of the fuel specification. Reduction of the biomass feedstock ash and alkali content by a washing process is a theoretical possibility, but will introduce its own problems at this scale in terms of the water quantities required and the subsequent waste water treatment, and also the stringent moisture content requirement either for efficient grinding of the biomass and also its subsequent combustion.

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4 Conversion technologies – gasification

4.1 Updraft Gasifiers

Fixed bed/moving bed gasifiers

Updraft gasifiers (this Section) and downdraft gasifiers (Section 4.2) are both designed as moving-bed gasifiers (also confusingly known as fixed bed gasifiers). These gasifiers are called *fixed-bed* or *moving-bed* because the biomass is fixed onto a grate that then moves through the reactor. The biomass on the grate is typically placed in a refractory lined shaft reactor.

4.1.1 Technology description

In an updraft gasifier, also known as counter-flow gasification, biomass enters from the top of the reactor and a grate at the bottom of the reactor supports the reacting bed. The oxidising agent - air, oxygen or steam - enters from the bottom of the gasifier below the grate and diffuses up through the bed of biomass and char (carbon, short hydrocarbons and water). Some of the char at the bottom of the reactor undergoes complete combustion producing CO_2 and H_2O at approximately 1000°C (oxidation zone in Figure 78). As these hot gases travel upwards through the bed they react endothermically with the unreacted char, are reduced to H_2 and CO and thereby cooled to approximately 750°C (gasification zone). These gases (reducing gases) pyrolyse the descending dry biomass (pyrolysis zone) and moving upwards dry the incoming wet biomass (drying zone). Most of the remaining heat is used in this drying step and thereby little energy is lost as heat in the gas. The pyrolysis zone involves the thermal breakdown of larger hydrocarbon molecules of biomass into smaller condensable and non-condensable gases. One of the products that form as a result of pyrolysis is tar, which is a sticky liquid produced by the condensation of some of the vapour. The presence of tar in the product stream is difficult to handle in industry and requires additional syngas clean-up before use in a gas engine or gas turbine[63].

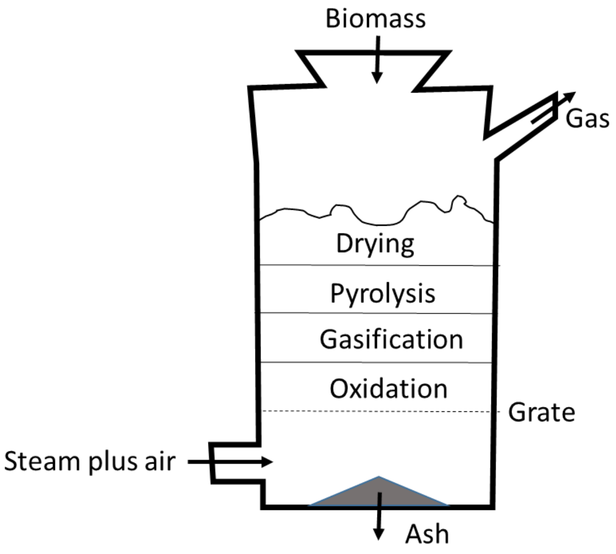

Figure 78: Schematic of an updraft gasifier, derived from[63]

Updraft gasification typically produces syngas with 10-20% tar by weight, which is one of the principal disadvantages of this type of gasifier[64]. However, the Harboøre biomass plant[65] has implemented gas clean-up technology and has now fitted two 648 kW_e Jenbacher gas engines (quoted in a different document as 650 kW_e and 740kW_e[66], indicating a potential recent upgrade of one engine), demonstrating significant progress with biomass tar clean-up.

There are two main types of updraft gasifier: dry-ash and slagging.

Dry-Ash Gasifier

This type of gasifier is a pressurized dry-ash updraft gasifier that was first developed by Lurgi, a process development company in 1931. The name dry-ash comes from the fact that the ash produced is not molten (liquid). The Lurgi gasifier has a peak temperature of 1200°C in the combustion zone, a maximum gasification temperature of 700 to 900°C and a reactor pressure of 3 MPa. The residence time of coal in this gasifier is 30 to 60 minutes. There are three options for gasification mediums in the Lurgi gasifier: steam and oxygen, steam and air, or steam and oxygen-enriched air. The typical steam/fuel carbon ratio is high at 1.5.

The dry-ash gasifier has been used to produce 55 million Nm^3/day of syngas from coal at SASOL in South Africa. With coal, the Dry-ash gasifier screens the fuel to sizes between 3 and 40mm. Information from (Babcock & Wilcox Vølund)[65] indicates that they have operated the Harboøre plant successfully on a commercial basis since 2001. The plant originally started simply for district heating (producing 4 MW_{th} of syngas), was converted to CHP in 2001, and produces (max) (650 + 740 =) 1390 kW_e of electricity, with an overall electrical efficiency based on gas output of ~ 35 %. However, typical operating data are given for the plant[66] (Table 15) which are at odds with the efficiencies and outputs stated above (most likely, the values above are the maximum full-load values). The stated electrical efficiency was 28 %, with 53 % thermal efficiency and 94 % “total efficiency” (presumably, the gasification efficiency, though this seems high). An order has been received for a 4 MW_e plant, to be situated in Southern Italy.

Slagging Gasifier

Developed by British Gas/Lurgi, the slagging gasifier works on the same principle as the dry-ash gasifier, but at much higher combustion temperatures (1500-1800°C). This high temperature melts the ashes rendering it molten and hence requires a much lower steam/fuel ratio of 0.58. Typically oxidation agents steam and/or oxygen are used and are also introduced at 3 MPa. When using coal, the fuel is first crushed to 5 to 80 mm pieces before being fed into the reactor. As biomass is more reactive, large pieces would be allowable.

The hearth load of an updraft gasifier is typically limited to 2.8 MW_{th}/m^2 or 150 $kg/m^2/hr$ for biomass. This is lower than that of an updraft gasifier for coal. The height of the updraft gasifier is usually greater than its diameter and the height-to-diameter is typically above 3:1. The diameter is typically limited to 3 – 4 metres.

4.1.2 Development status and timescales

The technology readiness level of the updraft gasifier is set at 9 as it is the oldest, simplest gasifier technology and is the principal gasification technology which has been used for coal over 150 years[67] and was first used with biomass feedstock around the time of World War I. These moving-bed gasifiers were built to operate vehicles, boats, trains and electric generators during World War I, as they were small and easy to construct.

Twelve Lurgi dry-ash gasifiers have been in continuous operation for 24 years. In operation in the Great Plains Gasification facility in North Dakota, over 54 billion standard cubic feet of syngas are produced annually from coal. A number of these gasifiers are also in operation by Sasol in South Africa (some of which have been operational since the 1960s) using bituminous coal to produce syngas, which is then converted to liquid fuel through the Fischer Tropsch operation[68].

Finnish companies SME and VTT developed nine Bioneer updraft gasifiers, constructed in 1982-1986, of which 8 are in operation in Finland and one in Sweden. These gasifiers were designed to produce 4-5 MW_{th} using wood/peat as the feedstock. Danish company, Babcock & Wilcox Vølund constructed an updraft gasification plant called the Harboøre plant in 1996 fuelled with non-pre-treated wood chips producing 3.7 MW_{th}. The plant has been in continuous operation since 1996, operating on average for 8,000 hours per year. The average electrical efficiency of this plant is 23%, is capable of running at 25% load and 11-13% of the energy from the wood chips is converted to heavy tar. The Babcock & Wilcox Vølund gasification technology is used to produce energy for three CHP plants in Japan by JFE Corporation each of 8 MW_{th}, 9 MW_{th} and 11 MW_{th} outputs[69].

Nexterra, a Canadian/UK based company has developed an updraft gasifier with over 150,000 hours of commercial operation. These gasifiers are designed for a range of outputs from 2 to 40 MW_{th} and 2-15 MW_e as well as a range of feedstocks, although which feedstocks is not explicitly stated[70]. In April 2015, Nexterra made a financial agreement to develop an updraft gasification plant for the Welland Waste Wood Power Plant in Northamptonshire, UK. This plant is scheduled to be operational by 2017 converting 60,000 tonnes of dry locally sourced waste wood (diverted from landfill) into 9 MW_e of net electricity[70].

4.1.3 Impact of different feedstock parameters on operation and cost

The effects of the different parameters associated with the feedstock are provided in a quantitative and qualitative fashion. Quantitative estimates have been made below for a number of the parameters. For those where it was not possible to obtain a reliable estimate, a qualitative discussion is presented below. Table 15 below compares some of the key characteristics and limits of updraft and downdraft gasifiers when using biomass.

Note that for all the gasifiers, our cost and other performance parameters are those for the production of clean syngas. Availability and Opex are for the base feedstocks; high ash and/or high alkali metal content feedstocks would increase downtime and Opex. The latter is modelled using the threshold concept.

Table 15: Characteristics of fixed-bed gasifiers for biomass[63]

Fuel (wood)	Updraft	Downdraft
Moisture wet basis (%)	60 max	25 max
Ash content (dry basis %)[71]	25 max	6 max
Ash melting temperature (Celsius)	>1000	>1250
Particle Size of biomass (mm)	5-500	20-100
Application range (thermal input)	1 MW – 15 MW	50 kW – 2MW
Gas exit temp (Celsius)	200-400	700
Tar (g/m^3)	30-150	0.015-3.0
Gas LHV (MJ/m^3)	5.0-6.0	4.5-5.0
Hot-gas efficiency (%)	90-95	85-90
Turn down ratio	5-10	3-4
Hearth load (MW_{th}/m^3) specific gasification rate	2.8	
Oils and tar (kg/kg dry feed)	**0.05-0.15 –** *considered too high for engine use*[72]	0.001-0.01
Char loss (kg/kg dry feed)	0.01-0.02	0.02

Plant size

The plant size primarily affects the Capex. A 0.79 power law has been estimated given the mixed nature of the equipment. This gives rise to a cost curve as in Figure 79.

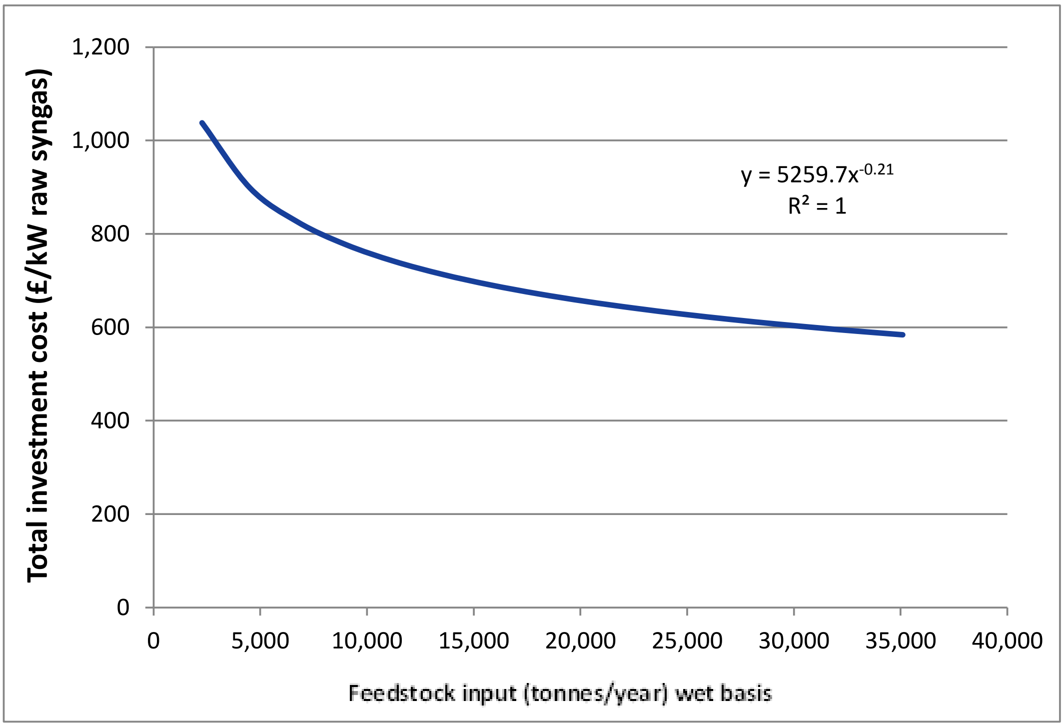

Figure 79: Updraft gasifier total investment cost vs. plant scale (based on data from [73] and [74], using three data points from different sources, so harmonisation was required)

Feedstock conditions

Moisture content

The moisture content affects the efficiency due to the latent heat of vaporisation of the water in the biomass. The relationship is illustrated in Figure 80. Note that this efficiency is the energy output of cleaned, polished syngas divided by the LHV feedstock input – it is not just the cold gas efficiency of the gasifier alone (which is a higher value). Syngas cleaning imposes a large efficiency loss for updraft gasifiers in particular, given their very high tar content syngas (up to 20%) and the large energy consumption needed for tar cracking. Syngas clean-up is necessary in order to meet specifications for use in engines, which should ideally be less than 100 ppm (100 mg/Nm^3)[67].

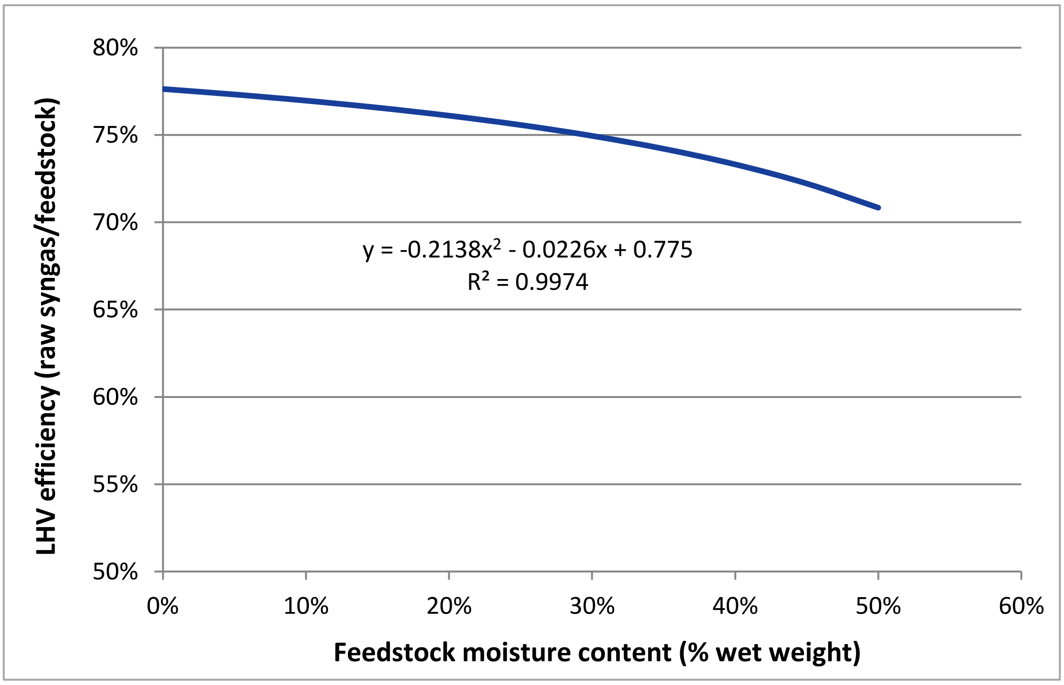

Figure 80: Updraft gasifier efficiency vs. feedstock moisture content (based on[73,74])

Particle Size

Owing to the long residence time of the particles within updraft and downdraft gasifiers, these are more tolerant of a range of particle sizes. However, some important issues are highlighted by Reed[67]. Essentially, the particles need to be of sufficient size that they will pass down the conical neck of the gasifier, but not so large that they pass through unreacted (which would clearly reduce the gasifier hot or cold gas efficiency by decreasing the burnout). Hence particle size is essentially a feasibility rather than cost issue. Uniform particle size is beneficial. Some agitation or distribution can be beneficial to keep the fuel moving, but too much will cause the bed to move through the gasifier too rapidly and cause unburned fuel[67]. It is notable that the Harboøre plant includes such a distribution system[73].

Feedstock composition

The effect of ash, sulphur, nitrogen and chlorine in the feedstock is felt in Opex via disposal or mitigation costs; here it is assumed a level of clean-up suitable for pipelines/turbines. To obtain the base Opex, the following assumptions are used:

- Base values of the impurity values as per the assumed feedstock composition base cases
- Insurance was assumed at 1% of TIC
- Other Opex, e.g. labour, fuels, reagents are process specific and determined on a process by process basis

For the base Opex, the O&M, labour, fuels and reagents costs are from[75]. These incremental effects on Opex are illustrated below, showing how different amounts of biomass constituents in the feedstock alter the total plant Opex. Ash handling stands out as the largest cost to mitigate, followed in order by sulphur, nitrogen and chlorine contents.

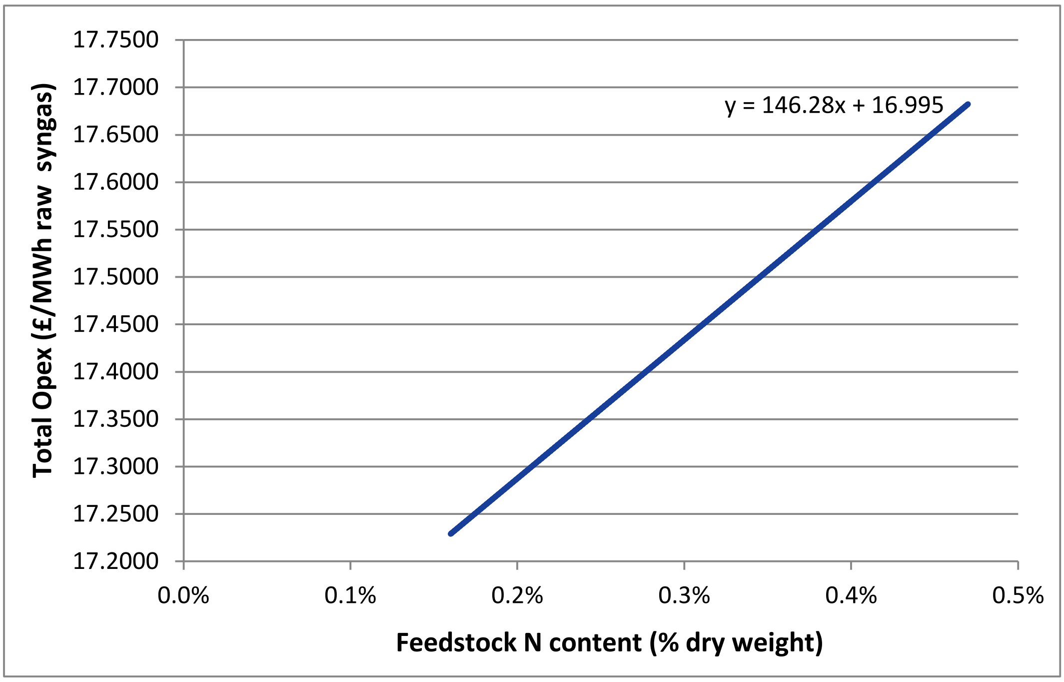

Figure 81: Updraft gasifier annual total opex vs. feedstock N content (labour, O&M, fuels and reagents from[75], calculation based on section 2.4.2.4)

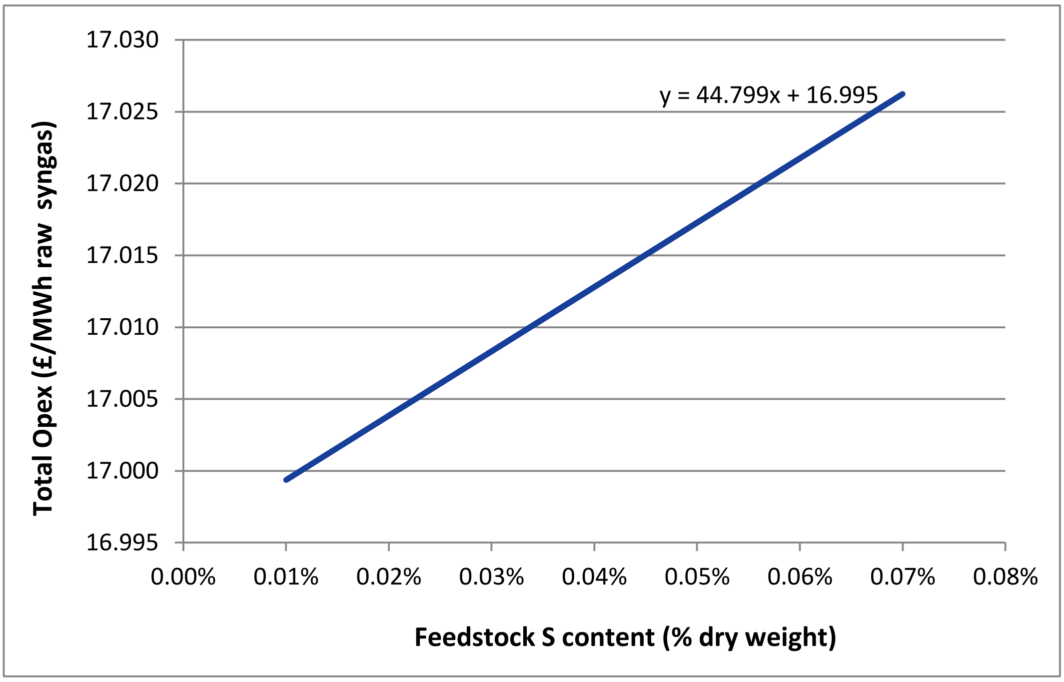

Figure 82: Updraft gasifier annual total opex vs. feedstock S content (labour, O&M, fuels and reagents from[75], calculation based on section 2.4.2.4)

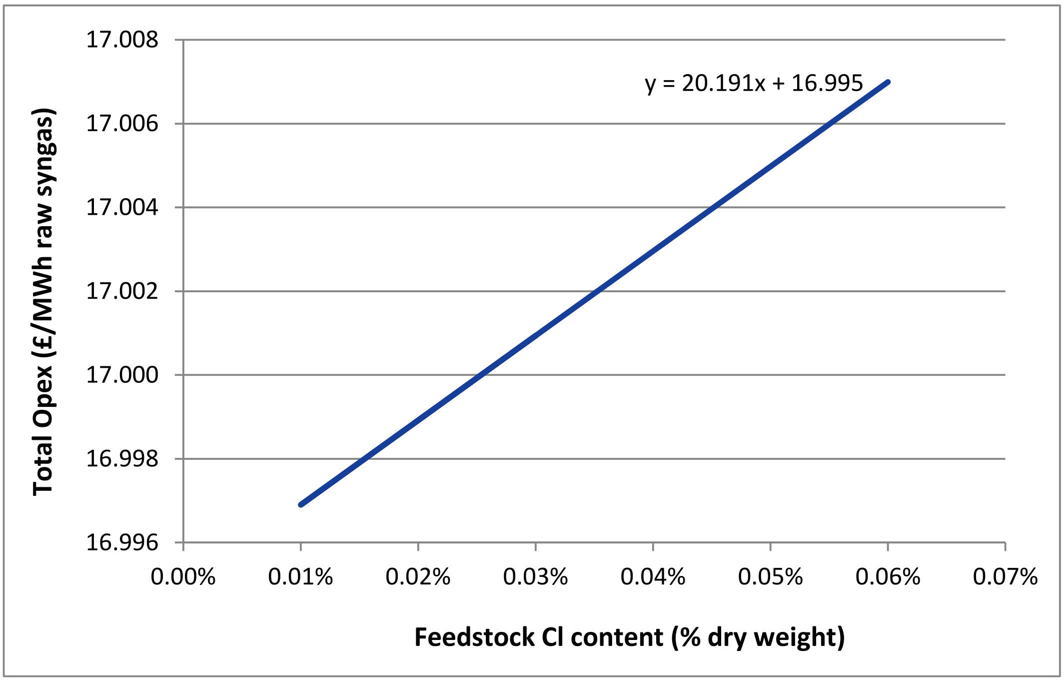

Figure 83: Updraft gasifier annual total opex vs. feedstock Cl content (labour, O&M, fuels and reagents from[75], calculation based on section 2.4.2.4)

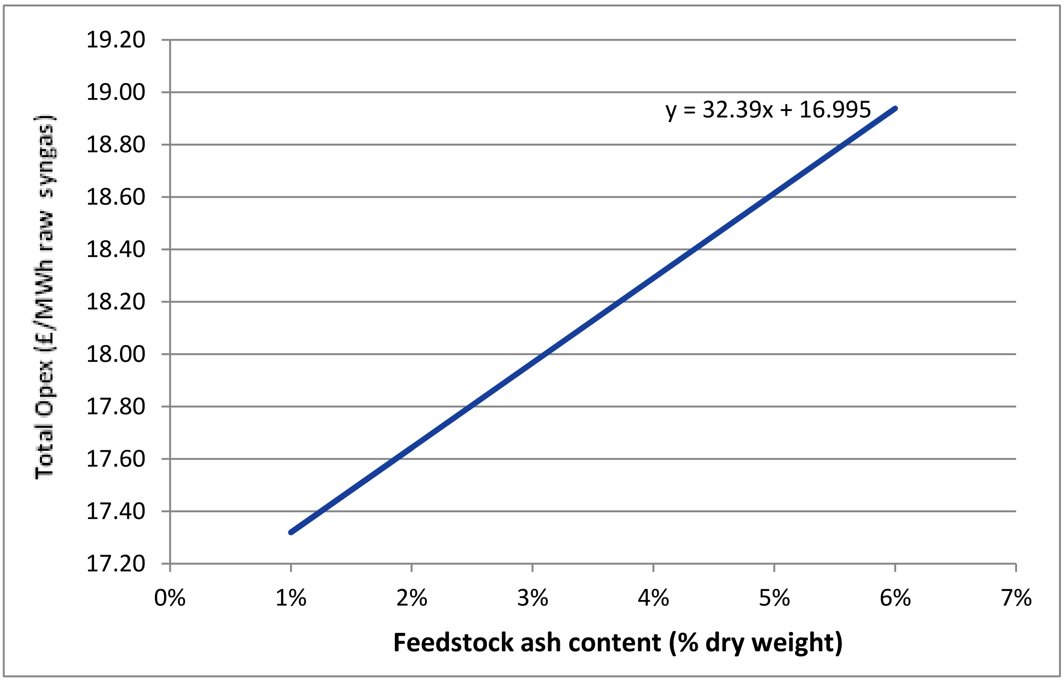

Figure 84: Updraft gasifier annual total opex vs. feedstock ash content (labour, O&M, fuels and reagents from[75], calculation based on section 2.4.2.3)

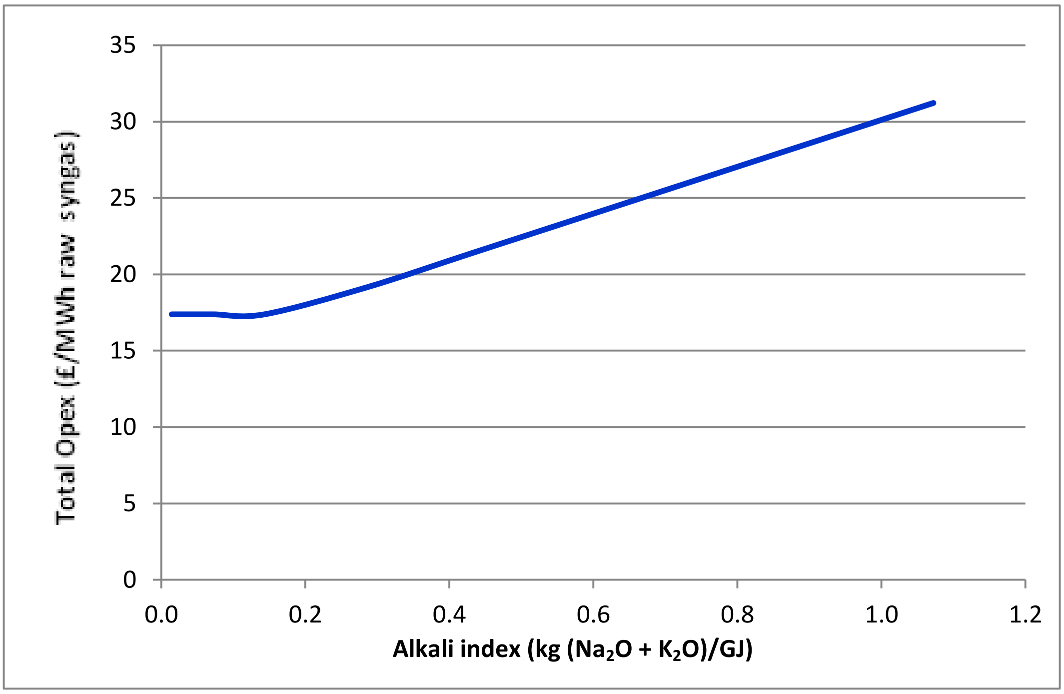

Figure 85: Updraft gasifier annual total opex vs. feedstock alkali index (labour, O&M, fuels and reagents from[75], formula based on [76] and 2.4.2.5)

Ash Melting Temperature

The ash melting temperature has been described in the introduction.

Within the context of updraft and downdraft (also entrained-flow) gasifiers, the slagging characteristics of the ash are very important, but the parameterisation is highly complex, depending on the fuel characteristics (frequently, the ratio of acidic and basic species within the ash) and the achieved temperature in the bed. It is not possible to simply give a simple formula – testing is frequently required. The development of slagging indices for gasifiers (of whichever type) is an area of continued ongoing research – the slag viscosity needs to be below a certain value for updraft and downdraft slagging gasifiers (else blockages can occur), needs to be within a range for an entrained flow slagging gasifier (essentially, too thin or too thick and the slag will not form a protective coating on the gasifier wall) and for a dry ash updraft gasifier the ash has to be solid.

4.1.4 Available options for improvement

Basic updraft gasifiers are well-established and widely used internationally. There are some potential areas of improvement such as thermal management and residence time management which can either enable tuning of operation to the feedstock and/or result in a product with a tighter specification. However, the main issue with biomass gasification in updraft gasifiers is the high level of tar in the syngas product due to the high volatility of typical biomass fuels and the lack of a substantial homogeneous tar cracking/heterogeneous char reforming zone. The level of tar in the syngas resulting from updraft gasification of biomass, as presented in Table 7, can be between 30,000 ppm and 150,000 ppm by weight, which prevents these gases from being suitable for direct use in gas engines or turbines. However, the high cold gas efficiency of biomass updraft gasification products, make it more suitable for direct firing.

In Harboøre, Denmark, Babcock & Wilcox Vølund developed a biomass updraft gasifier in combination with gas engines for use in a CHP demonstration plant that proved to obtain a very low tar content syngas. They did this by burning the product gas from the reactor (containing 150 mg/Nm^3) in a Low-NO$_x$ burner into a hot-water district heating boiler. The gas then at 75 C is cooled serially through two district heating cooled shell and tube heat exchangers, and as it is cooled a lot of the tar, water and particles are separated from the product gas. This is finally cleaned for remaining water/tar aerosols in a wet electrostatic precipitator (ESP).[77] This combined process achieves a syngas product with only 25 mg/Nm^3 of tar and dust, which is largely suitable for fuelling gas engines.

One approach to dealing with the tar cracking problem is that of Nexterra, who in collaboration with Alaska Power & Telephone (AP&T) are conducting a project that combines an updraft biomass gasifier with a tar cracking system (essentially a two-stage gasifier). GE Jenbacher has then approved the resultant syngas exiting the conditioning system for use in their gas engines.[78]

A study by Rao et al.[79] has also shown that replacing wood chips – conventional biomass fuel – with low density refuse derived fuel and using an inclined grate (as opposed to a horizontal one), reduces the tar content by 45 % and increases the cold gas efficiency from 65% to 73%[80]. A study by Na et al.[81] showed that increasing the height of the reactor bed increased the syngas content (H$_2$ and CO), but was limited to 700 mm.

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4.2 Downdraft Gasifiers

4.2.1 Technology description

The downdraft gasifier, also known as a concurrent gasifier has the same mechanical design as the updraft gasifier but instead of introducing the oxidising agent (air, steam, oxygen) at the base, it enters just above the combustion zone (or with the biomass, from the top) and flows downwards.

The main advantage of downdraft gasifiers is that up to 99% [82] of the tar formed is consumed as it finds favourable conditions for cracking therefore, compared with updraft gasifiers, less syngas clean-up is required. In addition, the downdraft design requires only 20-30 minutes to ignite and for the plant to reach operating temperature (shorter than the updraft gasifier).

However, downdraft gasifiers are typically limited to diameters of up to 1.5 metres, in order to ensure adequate penetration of nozzle gas into the hearth (throat or reactor)[83]. Hence, they are inherently limited in size, typically up to about 2 MW_{th} biomass input. This rules out CCGT applications, but still allows use with smaller gas engines, either for power or CHP, along with syngas boilers for heating.

There are two main types of downdraft gasifiers: the throated (Imbert) gasifier and the throatless gasifier.

Throated Gasifier

The first downdraft gasifier model is the “Imbert” gasifier or the throated gasifier named after its inventor. A typical throated gasifier is represented in Figure 86. These Imbert gasifiers were first invented in the 1920s. Similarly to the updraft gasifier, the biomass enters at the top and with the heat coming from the reactions taking place below is dried. The Imbert gasifier has a closed top that opens when more biomass is to be inserted. Typically two radially directed air nozzles are located at the throat (the constricted area) where it burns and pyrolyses some of the biomass, most of the tars and oils and some of the charcoal remaining below the nozzles at 1200-1400 °C. The constricted zone (or hearth zone) at which the air nozzles enter allows for improved insulation in the combustion section resulting in lower tar production and a higher efficiency over a wide range of operating conditions. The hot combustion gases (CO_2 and H_2O) then pass through the hot char at 800-1200°C where they are reduced to fuel gases (CO and H_2). The reactions involved include Boudouard reactions and steam gasification. Due to the endothermic nature of the gasification reactions, these gases are cooled to below 800 °C. At the bottom of the reactor the ash and unconverted char pass through the bottom of the grate and are sent for disposal. The main purpose of the throat (or hearth constriction) is that it ensures that all the gases pass through the hot zone at the constriction resulting in maximum mixing and minimum heat loss. This in turn results in a high amount of tar cracking, and if tarry gas is still produced the constriction is simply made smaller, either by design or retrofit.

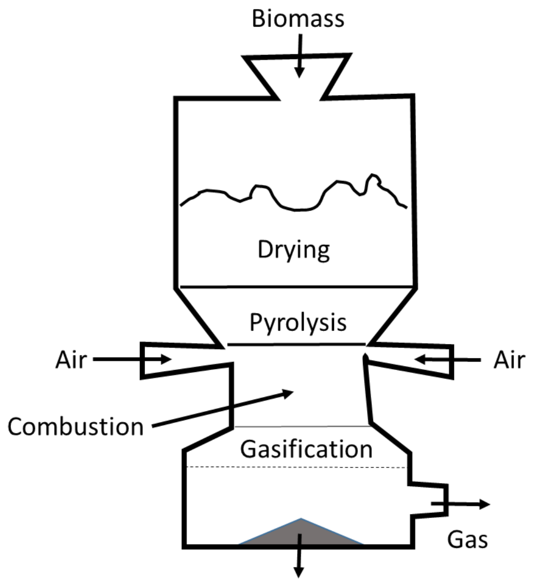

Figure 86: Schematic of a throated (Imbert) gasifier, derived from[83]

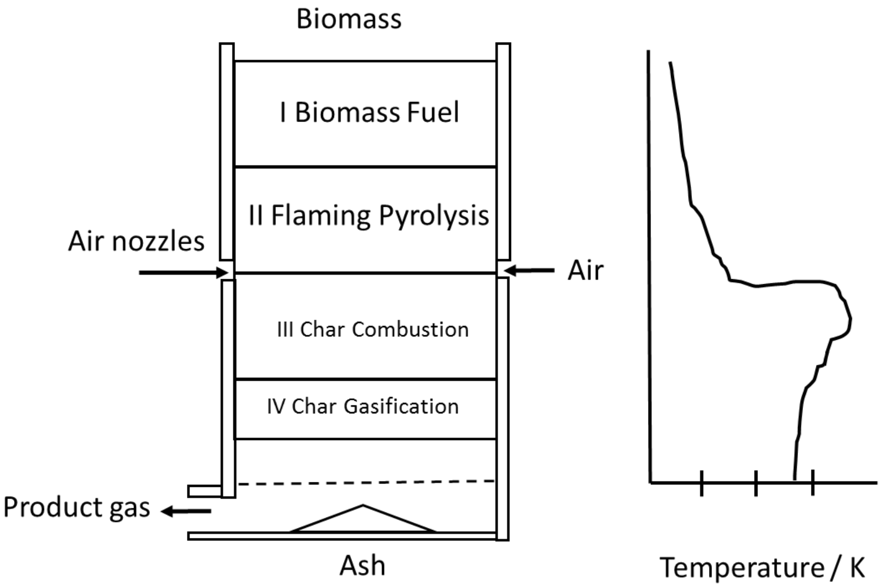

Figure 87**: Schematic of a throatless downdraft gasifier and temperature gradient along the height shown on the right, derived from**[83]

However, the disadvantages of a throated downdraft gasifier include[83]:

- Hearth constriction limits the range of biomass fuel shapes that can be successfully gasified
- Low moisture content acceptable (maximum limit of 25% moisture)
- Large amount of ash and dust are present in the product gas
- Although tar content is low, it can still reach 5 g/Nm^3 which requires additional gas clean up. Attempts to increase the scale of the Imbert gasifiers have also led to a drastic increase in tar production.

Throatless (also known as stratified or open top)

The open top throatless or stratified downdraft gasifier was developed in the 1980s[67], long after the first Imbert gasifier, to overcome many of the difficulties of the throated downdraft gasifier. A typical throatless gasifier is represented in Figure 87 above, consisting of a cylindrical vessel. This type of gasifier was developed as a result of a collaboration between researchers at SERI, UC Davis, the Open University in London, the Buck Rogers Co. in Kansas and in Florida. This gasifier is designed such that, contrary to the older Imbert gasifier, the top of the cylindrical vessel is exposed to the atmosphere and the walls are totally vertical. Its open top allows for uniform access of air or oxygen to the flaming pyrolysis zone. The stratified gasifier goes through the same reaction steps as the Imbert gasifier as the biomass flows down the reactor (biomass drying, pyrolysis of the dried biomass, combustion at the oxidation zone of the volatiles and part of the solid biomass and finally combustion gases from the oxidation zone are reduced when passing over the rest of the solid biomass at high temperature). Some advantages this throatless design has over the Imbert design include[85]:

- Easier to feed fuel from the open top and otherwise troublesome fuels are easy to get through the cylindrical shape without causing bridging or channelling in the reactor
- Easy access for instruments to measure conditions within the reactor bed
- No extreme (high/low) temperatures seen thanks to the uniform passage of air while maintaining a high average temperature (800-1200°C)
- Easy to manufacture due to cylindrical form
- In principle, can be scaled to larger diameters since it operates as a plug flow reactor (examples include a 0.6m internal diameter stratified gasifier by the Buck Rogers Co. in Kansas and a 0.77m internal diameter one producing 750 kW of power by Syngas Systems).

However, without a constricting throat zone, there is less insulation in the combustion section/more radiative losses, and hence throatless designs typically have slightly higher tar production and a lower efficiency than Imbert designs of the same scale.

4.2.2 Development status and timescales

The downdraft gasifier has a technology readiness level of 9, as it has been in used since the 1920s when the Imbert (throated gasifier) was first deployed. The throated downdraft gasifier with biomass was typically used for vehicle operation during World War II at which point 1 million of them were mass-manufactured at a cost of $1000 each at the time. Throughout history, these small-scale biomass gasifiers have been used in times of crisis when there was a need to provide an alternative method of energy generation: for example before World War II when Germany blocked the transportation of oil to Western Europe and in 1973 as a result of the OPEC oil crisis 1973[84].

Downdraft gasifiers have small capacities ranging from 10kW to 2 MW_{th} input and are typically inexpensive to manufacture compared to other types of gasifiers[85]. Currently, there are few large-scale commercial downdraft gasifiers. The largest downdraft gasifier of 1 MW_{th} output is situated in the USA, developed by CLEW (Camp Lejeune Energy from Wood)[86] with hogged wood as feedstock and has operated continuously for over 500 hours at $600-$1200/kWh[87]. Several facilities exist at the small and medium scale in Europe and the numbers have increased significantly over the past few years[88] as shown in Table 16 (although noting that some of this deployment is also updraft and fluidised bed gasifiers).

The stratified downdraft gasifier (also called "open-top" or "topless" gasifier) was first developed in the 1980s as a result of a collaboration between researchers at SERI, UC Davis, the Open University in London, the Buck Rogers Co. in Kansas and in Florida. The technology is now well established and a number of these are in operation. Buck Rogers Co. of Kansas has successfully operated a 0.6m internal diameter stratified downdraft gasifier. In addition, Syngas Systems Inc. operate a 0.77m internal diameter biomass stratified downdraft gasifier to produce 750 kW of power using air and oxygen suction (Reed, 1988).

Table 16: The number of biomass gasification facilities in Europe in 2013[89]

Country	Biomass gasification facilities in operation (total capacities)	Planned/under construction biomass gasification facilities
Germany	160 (>70 MW_{th} + 24 MW_e)	150
Austria	6 (19 MW_{th} + 6 MW_e)	2
Finland	3 (137 MW_{th} + 1.8 MW_e)	2
Denmark	8 (12 MW_{th} + 1.4 MW_e)	2
Other EU countries	31	15

4.2.3 Impact of different feedstock parameters on operation and cost

The parameters influencing downdraft gasifier operation and performance are:

- Low moisture content of fuel (maximum 25 wt.%)
- Dry-ash basis maximum 6%
- Higher ER reduces the amount of “useful” components in the product stream (H_2, CO)
- Typical biomass particle size should be less than 50 mm according to Martinez et al.[90], or between 20-100 mm according to Basu[63].

For a stratified (or throatless) downdraft gasifier, high versatility and high efficiency is demonstrated with solid fuels of poly-dispersed nature such as rice husk. For a rice husk downdraft gasification, Martinez et al.[90] found an optimal cold-gas efficiencies of 58% and 60%, depending on the throat size, as shown in Table 18. For gasification of cashew nut shells in a downdraft gasifier the optimum specific rate of gasification was found to be 167 kg/(h.m^2) for 70% gasification efficiency. Availability and Opex are for the base feedstocks; high ash and/or high alkali metal content feedstocks would increase downtime and Opex. The latter is modelled using the threshold concept.

Oxidising agent

Using air as an oxidising agent in a downdraft gasifier inherently produces gases with a high concentration of nitrogen and hence a product gas with a low calorific value of approximately 4-6 MJ/Nm^3.[84] A typical gas composition resulting from the gasification of biomass in a downdraft reactor comprises of: 15-20% H_2, 15-20% CO, 0.5-2% CH_4, 10-15% CO_2 with the remainder composed of nitrogen, oxygen and hydrocarbons. However, a downdraft gasifier with steam or oxygen as the oxidising agent produces a gas with higher calorific value of 10-15MJ/Nm^3 [91]containing a much higher fraction of combustible gases[64].

The producer gas equilibrium composition for a biomass $C_1H_{1.57}O_{0.78}N_{0.0056}S_{0.0001}$ in a downdraft gasifier with air as the oxidising agent and a gasification equivalence ratio of 3.5 (different from total equivalence ratio ER) is presented in Table 17[91]. The methane content is under predicted compared with experimental results and typically has a volumetric fraction of 2% to 4.5%.

Table 17: Producer gas equilibrium composition for a biomass $C_1H_{1.57}O_{0.78}N_{0.0056}S_{0.0001}$ and a gasification equivalence ratio of 3.5

Producer gas composition	Volumetric fraction
N_2	0.41
CO_2	0.09
H_2O	0.06
CO	0.24
H_2	0.2

Table 18: Design characteristics of downdraft gasifiers and experimental results published[103]

Biomass	Diameter (mm)		Height of reactor (m)	ER	Combustion zone temperature (°C)	Gas composition (%)			Heating value (MJ/Nm3)	Yield (Nm3/kg)	Power (kW)	Cold efficiency (%)
	Reactor	Throat				CO	H_2	CH_4				
Wood chips	1000	500	2.5	1.66[c]	n.a.	26.5	7.0	2.0	5.06[b]	1.44	448.04	48.77[h]
						22.1	13.4	2.9	5.59[b,d]	1.86	765.15	69.42[h]
Rice husk	152		n.a.	0.40	n.a.	n.a.	n.a.	n.a.	3.91[f]	2.13[g]	8.20	58.11
	203		n.a.	0.39	n.a.	n.a.	n.a.	n.a.	4.02[f]	2.10[g]	14.83	58.78
	244		n.a.	0.40	n.a.	n.a.	n.a.	n.a.	4.00[f]	2.17[g]	21.40	60.44
	343		n.a.	0.41	n.a.	n.a.	n.a.	n.a.	3.98[f]	2.22[g]	42.89	61.49
Wood chips	600	200	2.5	0.287	1000	n.a	n.a	n.a	5.19[a]	n.a	44.93	76.68
Hazelnut shells	450	135	0.81	1.51[e]	1025	16.8	14.12	1.7	4.55[a]	1.97	9.17	51.53
Rubber wood	920	100	1.15	1.9[c]	1000	20.2	18.3	1.1	n.a	n.a	n.a	n.a
Sawdust	270		1.1	0.26	900	19.48	18.89	3.96	6.32[a]	1.99[h]	n.a	62.5[h]
Pine wood blocks	350	n.a	1.3	0.28	1108	25.53	28.93	6.82	4.76	n.a	n.a	n.a
Wood chips	440	350	2	1.3[c]	1460	9.4	14.8	1.2	3.8[b]	n.a	n.a	n.a
Rice husk	30		n.a	1.5[c]	1000	n.a	n.a	n.a	4.2[a]	n.a	n.a	60
Wood chips	250	70	1.05	0.32	900	19.48	18.89	3.96	6.32[a]	n.a	n.a	62.5
Wood waste	310	150	1.1	0.205	1050	22	14	0.1	6.34[a]	1.62	7.28	55

n.a.: not available; [a] Higher heating value; [b] Lower heating value; [c] Air/fuel ratio in Nm^3/kg; [d] Dry, inert free; [e] Air/fuel ratio (Nm^3/kg), fuel is dry, ash free; [f] Lower heating value at 25°C; [g] At 25°C; [h] Dry basis.

Plant size

The plant size primarily affects the Capex. A 0.79 power law is estimated given the mixed nature of the equipment. This gives rise to a cost curve as in Figure 88. The Capex is higher than for updraft gasifiers, mainly due the smaller scale.

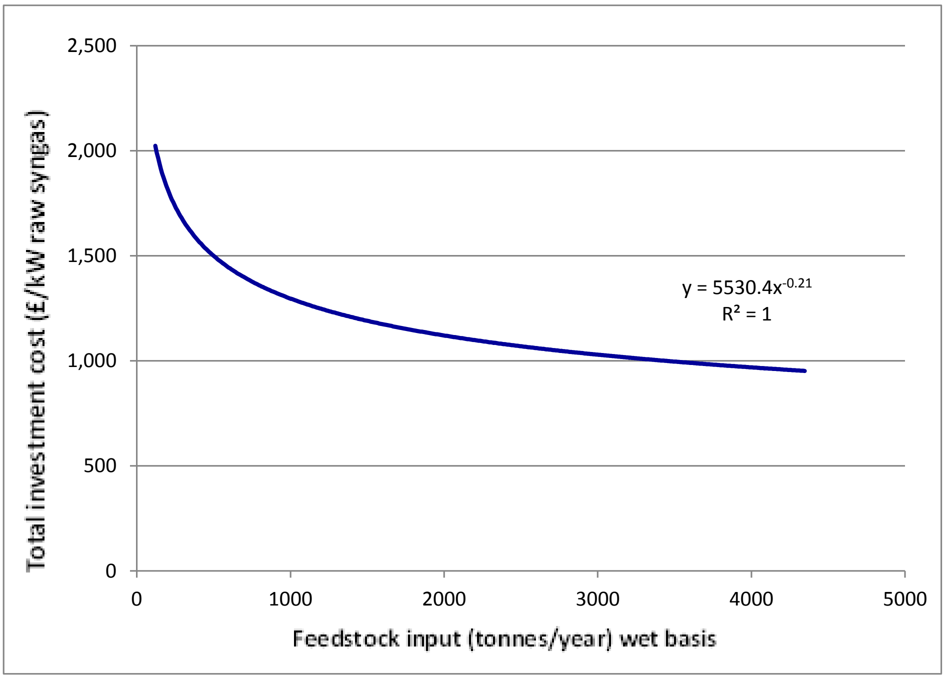

Figure 88: Downdraft gasifier total investment cost vs. plant scale (based on data from [92,93])

Feedstock conditions

Moisture content

The moisture content affects the efficiency due to the latent heat of vaporisation of the water in the biomass. The relationship is illustrated in Figure 89. This efficiency to clean, polished syngas is higher than for an updraft gasifier, since the low presence of tars in the syngas leads to a much more efficient syngas cleaning process.

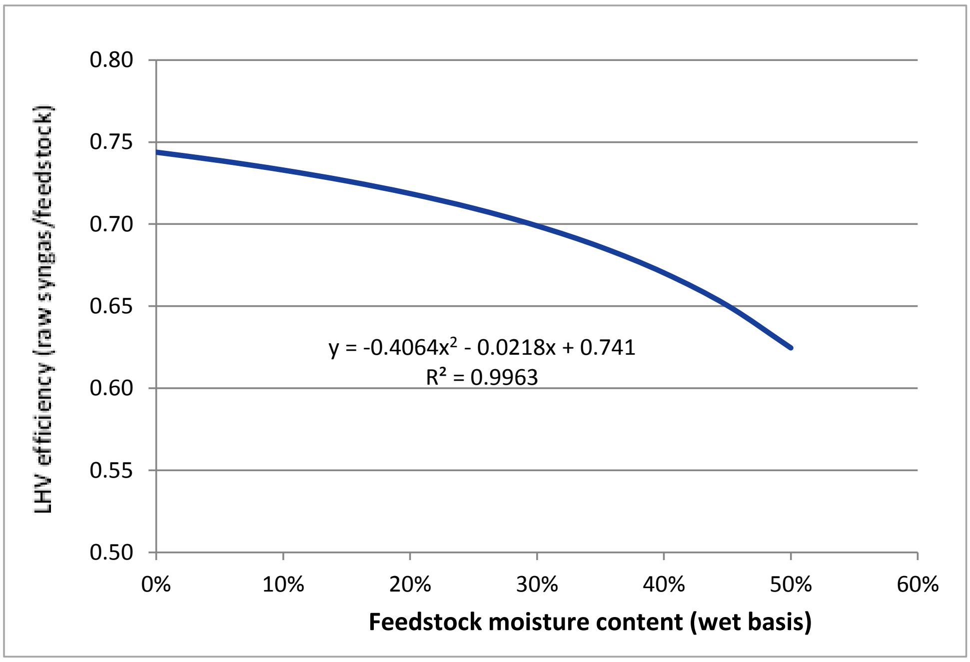

Figure 89: total investment cost efficiency vs moisture content (based on data from [74,73])

Feedstock composition

The effect of ash, sulphur, nitrogen and chlorine in the feed is felt in Opex via disposal or mitigation costs; here it is assumed a level of clean-up suitable for gas turbines. To obtain the base Opex, the following assumptions are used:

- Base values of the impurity values as per the assumed feedstock composition base cases
- Insurance was assumed at 1% of TIC
- Other Opex, e.g. labour, fuels, reagents are process specific and determined on a process by process basis

For the base Opex, the O&M, labour, fuels and reagents costs are from[94]. These effects on total plant Opex are illustrated below.

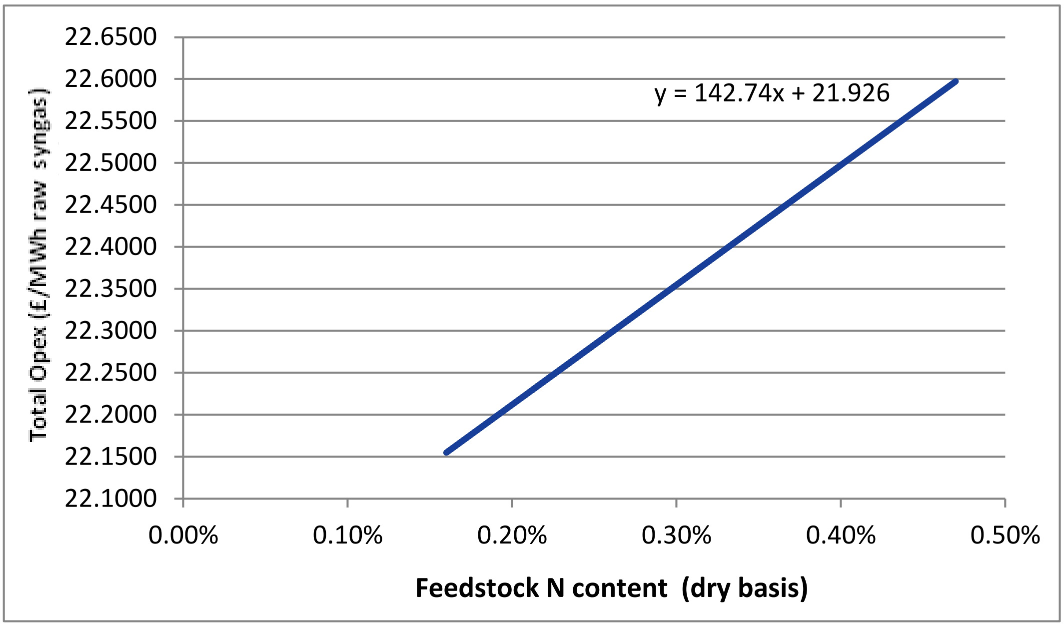

Figure 90: Downdraft gasifier annual total opex vs. feedstock N content (labour, O&M, fuels and reagents from[94], calculation based on section 2.4.2.4)

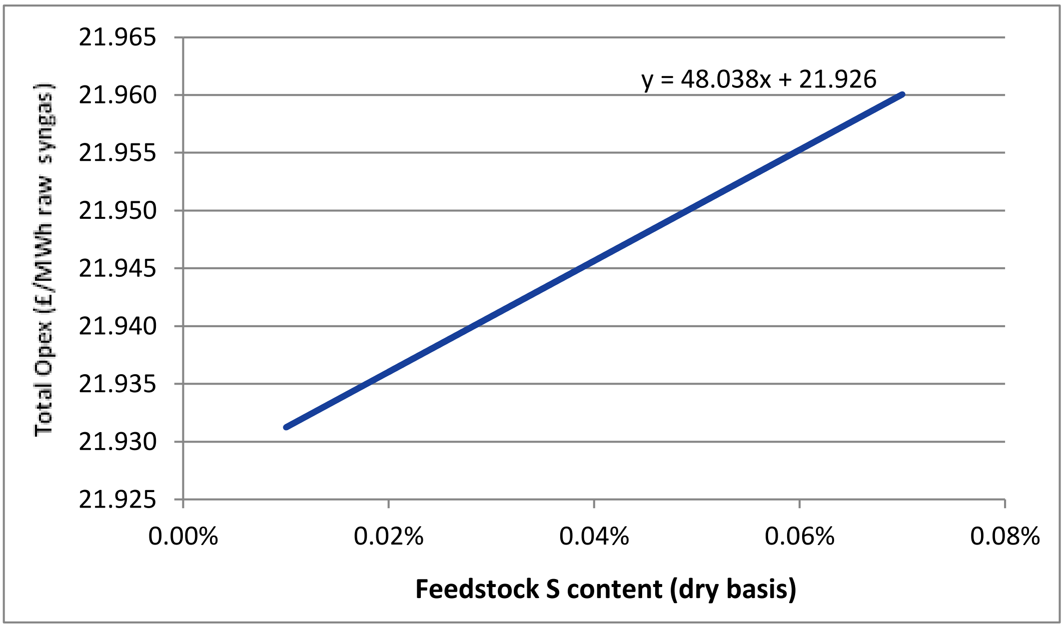

Figure 91: Downdraft gasifier annual total opex vs. feedstock S content (labour, O&M, fuels and reagents from[94], calculation based on section 2.4.2.4)

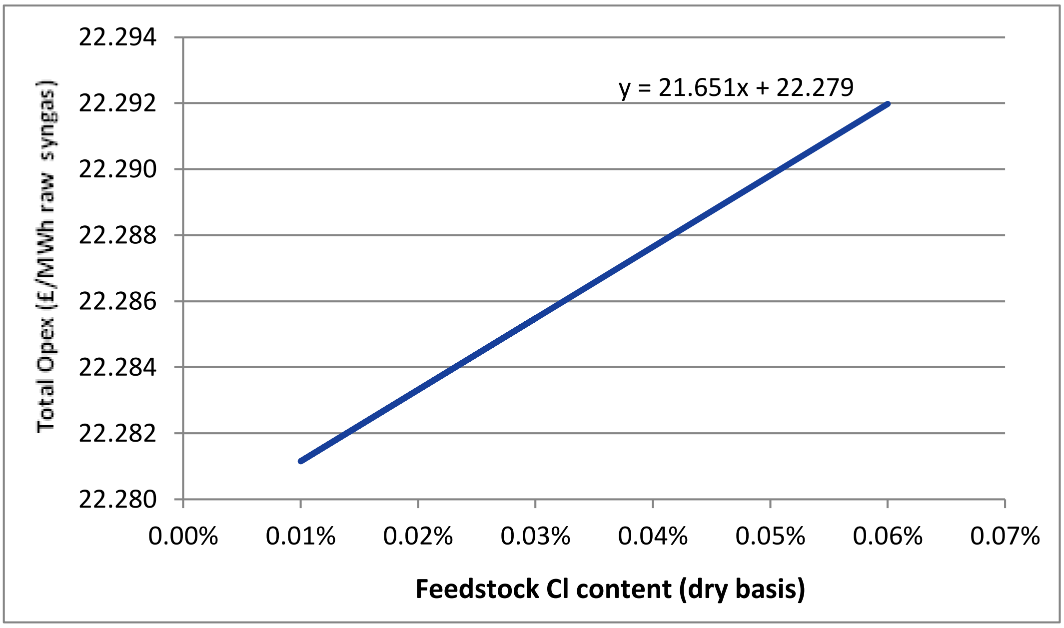

Figure 92: Downdraft gasifier annual total opex vs. feedstock Cl content (labour, O&M, fuels and reagents from[94], calculation based on section 2.4.2.4)

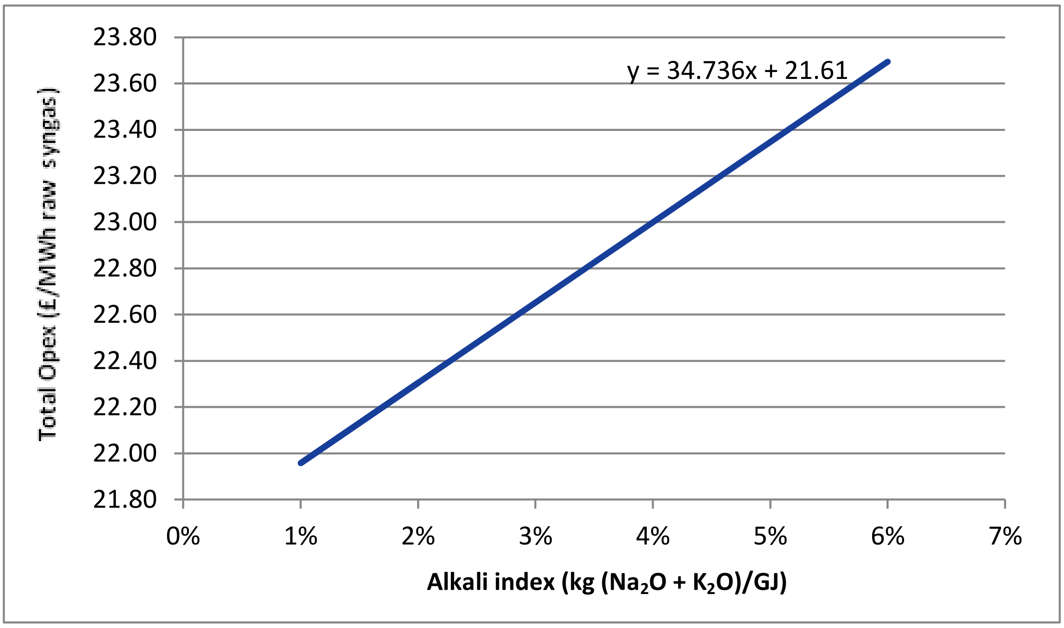

Figure 93: Downdraft gasifier annual total opex vs. feedstock ash content (labour, O&M, fuels and reagents from[94], calculation based on section 2.4.2.3)

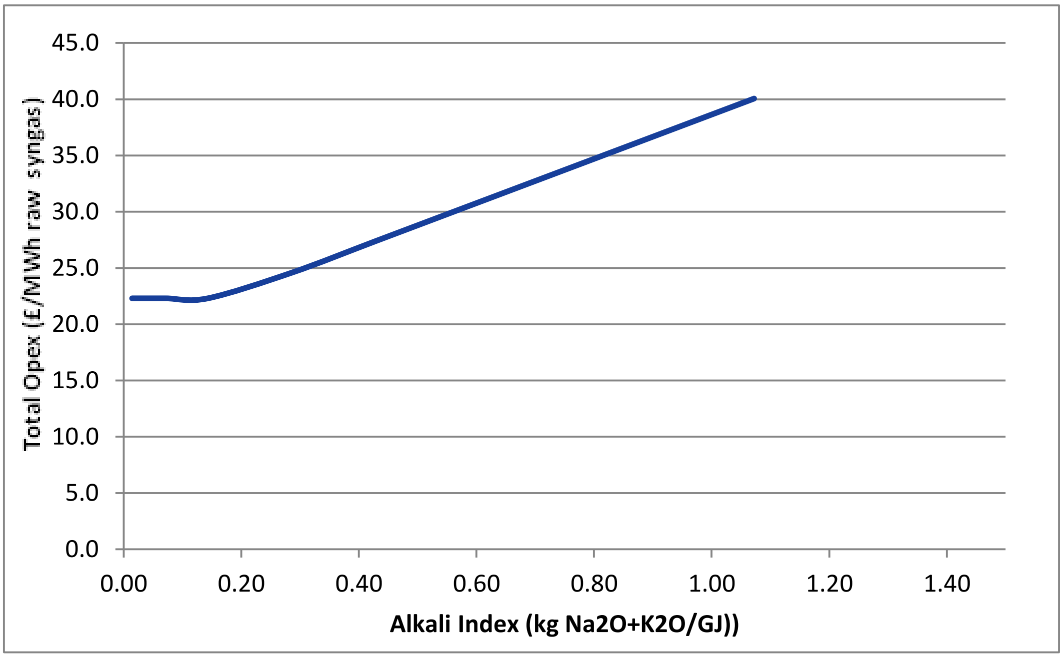

Figure 94: Downdraft gasifier annual total opex vs. feedstock alkali index (labour, O&M, fuels and reagents from[94], formula based on [95] and section 2.4.2.5)

The composition of the producer gas from downdraft gasification will be heavily dependent on the gasification temperature and the equivalence ratio. The figures below show how these parameters affect the quality of the syngas. F_{rg} in the figures below refers to the gasification equivalence ratio, i.e. the ratio of air present in the gasification step within the gasifier to the amount of biomass fuel (as opposed to the total equivalence ratio (ER) which refers to the amount of total air supplied to the system per kg of biomass fuel).

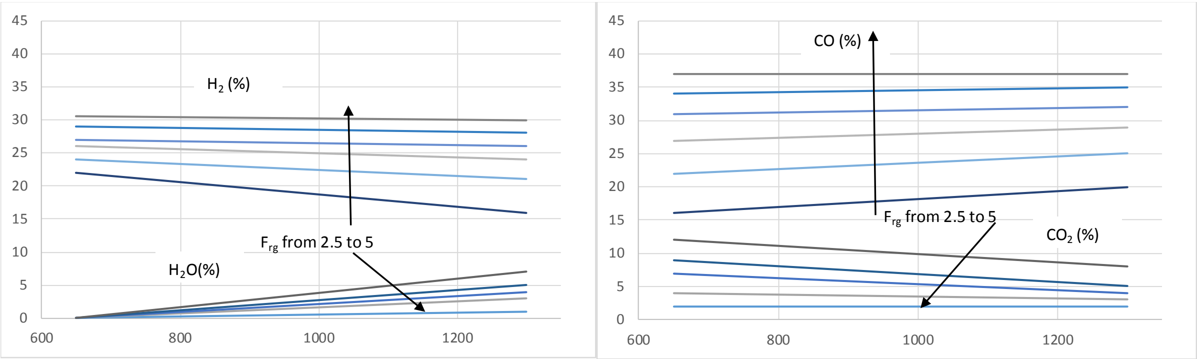

Figure 95: Effects of temperature and gasification equivalence ratio on the producer gas equilibrium composition. Y-axis is % composition, X-axis is temperature in K (redrawn from[91])

4.2.4 Available options for improvement

Typical downdraft gasifiers, although they produce significantly lower tar content syngas than updraft reactors, still require further tar cracking in order to have a syngas that is suitable for engine use. In addition, downdraft gasifier design remains at a small scale (maximum 1 MW_e, so not suitable for gas turbine applications), with relatively low cold gas efficiency and also has little flexibility with the type of biomass feedstock it can process (being particularly strict on the moisture content of the biomass).

A number of downdraft gasification reactor designs have been investigated and developed to overcome some of these limitations.

Many design improvements have focused on minimising tar production or increasing tar cracking within the process. DeLaCotte developed a tar-recycling gasifier that incorporates a gas-combustion chamber on the side of the principal reactor (fuel chamber). As the fuel travels down the gasifier, the pyrolysis products are aspirated out the top to a side gas-combustion chamber in which air is inserted. The pyrolysis products thereby burn completely at high temperature in the absence of solids. The hot combustion products coming out of the gas combustion chamber are then re-injected at the centre of the gasifier. A quarter of the gas moves upwards to enable the pyrolysis of the biomass fuel and the remaining three quarters travel down. These gases (CO_2 and H_2O) react with the char and produce syngas. It is the particularly high temperatures achieved in the gas combustion chamber that result in a significant reduction in tar content.

The Viking gasifier of the Technical University of Denmark (DTU) has developed a two-stage gasification process that allows for the production a very low tar and dust content (<5mg/Nm^3) and has a high energy efficiency (25% from biomass to net electricity)[96]. The Viking gasification system consists of two reactors: the first dries and pyrolyses the biomass and the hot pyrolysis products then enter a second reactor in which the products are partially oxidised in pre-heated air then reduced and cooled down to 800 °C. The main part of the tar decomposes at 1100 C. The final produced gas is cooled down to 90 °C by passing through several heat exchangers (heating up the air) and district heating. That fuel gas is then clean enough to fuel a gas engine and the hot exhaust gas produced from it is used to dry the biomass entering the first reactor. The pilot plant was first successful at a small scale with a 17.5 kW_e output but has now been successful in upscaling to 150 kW_e.

Xylowatt Gasification Technology in Belgium has also developed the novel downdraft biomass gasifier design "Notar". This downdraft reactor introduces two separate nozzles with air blowing through which separates the reactor into the three gasification stages: pyrolysis, combustion and reduction. This design has achieved a tar content of the exiting syngas at less than 100 mg/Nm^3 and a wood to gas conversion of over 98%.[97,96]

In spite of these improvements in the design and product of downdraft biomass gasifiers, they remain limited in their thermal and electrical outputs.

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[85] These gasifiers are designed on the basis of characteristic design parameters that include: specific grate gasification rate, hearth load and space velocity. These are defined below.

- The **equivalence ratio (ER)** is one of the most important variables in the gasification process for fixed-bed gasifiers. ER is the ratio of actual air to fuel ratio divided by the stoichiometric air to fuel ratio and affects the quality of the syngas produced[85].
- The **hearth load** (specific grate gasification rate or space velocity) is the amount of producer gas to be obtained per unit cross sectional area of the throat (or the smallest area of cross section in the reactor). The hearth load of a moving-bed gasifier can be expressed in terms of the fuel gasified, the volume of product gas that is produced, or the amount of energy produced.
- The performance of a gasifier is defined by the quality and quantity of gas produced. The performance is usually measured by the **gasification efficiency**. This can be calculated in three ways, by the Cold-gas efficiency, Hot-gas efficiency or Net gas efficiency
- The **yield** of a gasifier is used to measure the volume of producer gas specific production per kg of feedstock supplied to the system.
- The **turndown ratio** is the ratio of the highest practical gas generation rate to the lowest practical rate.

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4.3 Bubbling Fluidised Bed Gasifiers

Fluidised Bed gasifiers

There are three main types of fluidised bed gasifiers considered in this study; bubbling fluidised beds (BFB – this Section), circulating fluidised beds (CFB – Section 4.4) and dual fluidised beds (Dual FB – Section 4.5). These operate on the same principle, namely that the granular inert bed material is kept in motion or suspended by a gasifying agent, which can be air, oxygen or steam, which flows through the bed material. The bed material must be both inert and capable of promoting heat transfer and not agglomerate too much. As such, materials such as sand, alumina or dolomite are commonly used.

When compared to fixed-bed gasifiers, fluidised bed gasifiers generally have smaller volumes for the same output of product gas due to having better rates of heat exchange across the gasifier and a more uniform temperature profile across the bed than that within fixed-bed gasifiers due to the mixing within the bed material. In combination with the simpler scale up than fixed bed technologies, fluidised bed technology can be used simply for larger scale projects that would be infeasible or would not be cost effective to achieve with a fixed bed gasifier.

4.3.1 Technology description

The bubbling fluidised bed was developed in 1921 and has since been used commercially primarily for the gasification of coal to produce syngas. Within a bubbling fluidised bed, the gasifying agent's velocity is controlled so that it the gas velocity is just greater than the minimum fluidisation velocity of the bed material (usually 0.5-1m/s), leading to a fairly smooth bed surface with entrained small particles.

The feedstock biomass is fed into the bed at the side, below the surface of the bed material, while the gasifying agent is added from the bottom of the bed material. As one of the by-products, ash produced from the gasifier is removed, usually from either the bottom of the reactor or an adjacent cyclone through which the product gas flows. The ash content within the reactor is dependent on the nature of the feedstock, with different feeds having different ash fractions, which can inhibit reaction and it is therefore important to consider ash content when selecting feedstock. A high ash content can lead to poor thermal performance and can also act to inhibit the action of the bed material if concentrations increase too far. Bed temperature is kept deliberately lower in gasifiers operating with biomass than coal in order to prevent ash fusion and agglomeration, which would lead to a reduction in heat transfer across the bed and operational downtime. Note that there are differences in temperatures seen between coal and biomass gasification, i.e. coal gasifiers are hotter and need more steam than biomass (due to the higher intrinsic oxygen content in biomass).

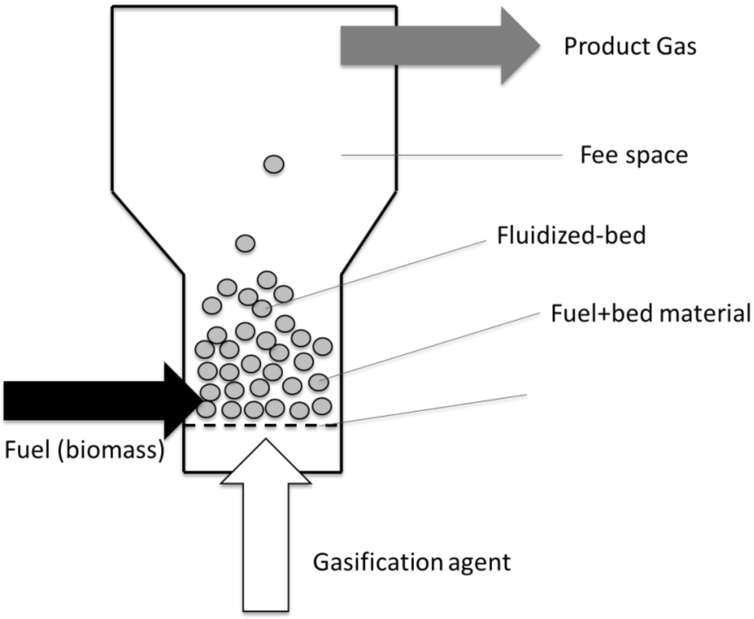

Figure 96: Schematic of Bubbling Fluidised Bed Gasifier, derived from[98]

Bubbling fluidised bed gasifiers can operate within a range of temperatures, with a maximum temperature of around 900°C in order to prevent ash fusion and agglomeration which could lead to severe reductions in the conversion of the biomass to syngas. These gasifiers can operate at either atmospheric pressure, or at high pressure, usually in the region of 20-30 bar. The choice of gasifying agent is important as it has a large effect on the composition of the product syngas, with air producing products containing large proportions of nitrogen, reducing their energy intensity compared to the use of oxygen or steam as the gasifying agent. Gasifier operating pressure affects not only gasifier equipment size and cost, and the syngas composition, but also the interfaces to the rest of the power plant including the necessary clean-up systems[99]. Gas turbines operating downstream from the gasifiers require elevated pressures, therefore high pressure gasifiers lead to more compact compressors. High pressure gasification also favours hot, pressurised syngas clean-up, which can be more efficient for an overall plant perspective than lower-pressure cleaning[100].

Bubbling fluidised bed gasifiers act similarly to the beds in continuously stirred reactors, allowing some of the biomass and tar to slip from the bed, though this effect can be reduced by having a uniform introduction of feedstock and utilising the bed material properly.

Overall, BFBs have been found to be cheaper to operate and construct than CFBs due to the smaller fans and compressors required for the gasifying agent, which can be either oxygen or air (in some cases in combination with steam). Air tends to be most commonly used unless larger scale applications are under consideration. This is especially true of BFBs which operate at atmospheric pressure or with only a small degree of pressurisation. Because there is no recirculation of the product, there are fewer energy losses compared to circulating designs, giving better heat exchange as the bed is more uniform. BFBs can also potentially produce a more uniform syngas when using variable feedstocks, though careful design is required in order to be able to achieve this. The more difficult operation of fluidised bed reactors is also lessened slightly when using bubbling beds instead of circulating ones, due to the lack of recirculation and the more uniform temperature profile.

4.3.2 Development status and timescales

BFB gasifiers have been in development and operation since the 1920s and are used commercially on a very wide scale. As well as being used for the gasification of biomass, these units have been used in processes to gasify both coal and municipal solid waste (MSW), although entrained flow or fixed-bed gasifiers are typically preferred for coal gasification. Biomass has been used as a feedstock in a number of plants operating using bubbling fluidised bed technology since the 1980s, however it has not been used on a commercial scale save a few small plants, though a large number of pilot plants and demo plants have been constructed, giving this technology a TRL of 8. It is estimated that it will take around ten years to reach TRL 9, with fine tuning of the systems, and experience from further deployment needed.

It is important to note that bubbling fluidised bed gasifiers have been used more extensively and on a commercial scale for gasification of municipal solid waste (often focusing on waste destruction instead of high-efficiency energy generation), but there have been issues with reliability and costs. Many pilot scale plants were closed prematurely due to high operating costs and the low availability of the gasifier, making cost data unreliable and lacking key information in many cases. Research into the use of bubbling fluidised beds for gasification is ongoing and there are a small number of small-scale commercial facilities, although these are often used for niche applications and may not be suitable for larger scale heat & power operations.

Bubbling fluidised bed gasifiers are particularly suitable for mid-scale operation up to a thermal output of 25 MW_{th} if at atmospheric pressure, although pressurised systems can obtain higher throughputs, as shown below.

The largest BFB gasifier plant currently known to be in operation is Enerkem's first commercial plant in Edmonton, Canada, which uses up to 48 MW_{th} of sorted MSW feedstock. The pressurised O_2-blown plant started up in 2014, producing syngas for methanol and ethanol production. Close behind in scale is Carbona/Andritz's Skive CHP plant, which started in mid-2008, using wood pellets in a O_2/steam-blown pressurised BFB gasifier to produce syngas for CHP gas engines. The plant is producing 6 MW_e of power and 11.5 MW_{th} of district hot water from 19.5 MW_{th} of wood pellets (although the maximum fuel input of the plant is 28 MW_{th} input)[101]. Energy Products of Idaho also built 4 pressurised BFB plants in the 1980's from 2-28 MW_{th} input, but these have shut down[102]. ThermoChem Recovery International also have an atmospheric pressure, indirectly heated 14 MW_{th} input black liquor gasifier at the Trenton Normapac mill.

As with the majority of biomass gasification, the technology is mainly used in Northern Europe, Canada and the USA, particularly Scandinavia.

No information was found in the literature about the timeframe required in order to develop and construct a commercial plant, since most of the facilities where bubbling fluidised bed gasifiers are used with biomass are pilot or demonstration scale and not commercial, and the few commercial scale facilities have not disclosed much data regarding the construction and commissioning phase. Plant construction is however expected to be on a slightly longer timescale than an equivalent sized BFB combustion plant, given the additional gas handling and system complexity.

4.3.3 Impact of different feedstock and other parameters on operation and cost

Some of the key feedstock requirements and characteristics of operating fluidised bed gasifiers are shown in Table 19 – for example, for BFB gasifiers, the biomass must be crushed to a standardised size depending on the reactor configuration, with a maximum size generally between 10-50mm.

Table 19: Characteristics of Fluidised Bed Gasifiers

Fuel (Wood chips)	Bubbling FB	Circulating FB	Dual FB
Max moisture content (%)	30	70	~38
Max ash content (%)	5	5	5
Max feed size (mm)	63	50	30
Gas exit temperature (°C)	750	800	850
Syngas LHV (MJ/kg)	4-6 (air blown)	5-6 (air blown)	12 (steam-based)
Cold gas efficiency (%)	80	75	75

Due to the high velocities of the gasifying agent and the subsequent fluidisation of the solid bed material, fluidised bed reactors are able to attain an extremely high degree of solid-gas mixing when compared to other gasifiers, leading to a uniform temperature profile. This high degree of mixing combined with a high thermal inertia of the bed material make fluidised bed gasifiers much less sensitive to variations in the feed than other gasifiers, such as particle size, density or the moisture and ash contents, so can be used for lower quality fuel sources such as waste-wood, RDF or MSW. This property of fluidised bed gasifiers is particularly useful for gasification of biomass, as the feedstock is much more variable than coal, for example, and the high tolerance of feed states allows these units to be used flexibly with a variety of feedstocks. The higher rates of heat transfer in the bed material enables a higher conversion of the feedstock to the product gas compared to updraft gasifiers, resulting in lower outputs of unconverted carbon and heavy hydrocarbon molecules, which are highly polluting and can lead to fouling of the units. Availability and Opex are for the base feedstocks; high ash and/or high alkali metal content feedstocks would increase downtime and Opex. The latter is modelled using the threshold concept; the former may have implications for bed material fouling.

However, this process flexibility and performance increase over fixed bed gasifiers comes at a cost; the specific investment cost and power consumption due to the compression requirements of the gasifying agent. This has led to a number of projects being cancelled or pilot schemes scrapped due to the high costs of gasification making the process highly uneconomical. In addition, fluidised beds are more complex to design and operate than fixed bed reactors, and many of the benefits outlined above may be negated if the gasifier is run sub-optimally. Compared to downdraft gasifiers, fluidised bed gasifiers will generally still have a higher output of tar and dust particles, though tar levels are lower than those found if using updraft gasifiers. The higher average gas temperatures can also cause problems with other impurities such as alkali metal compounds, which will be vaporised and can greatly increase corrosion rates, necessitating a sorbent in order to attempt to reduce this effect. Bed and equipment erosion also presents challenges given the high velocities, particularly in CFB gasifiers.

Plant size

The plant size primarily affects the Capex. A 0.79 power law is estimated given the mixed nature of the equipment. This gives rise to a cost curve as in Figure 97.

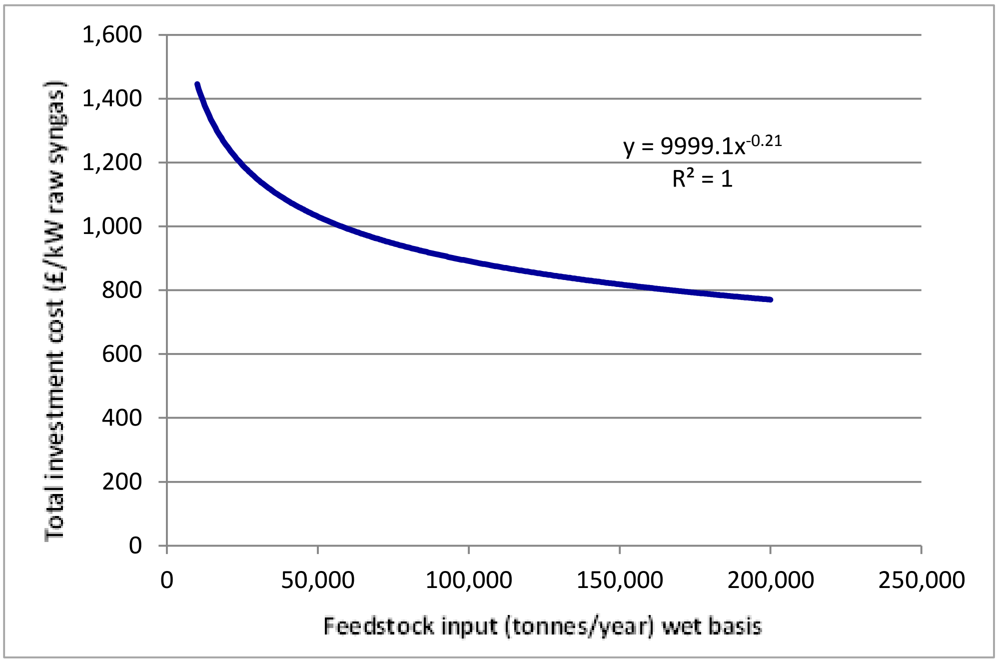

Figure 97: Total investment cost vs. plant scale (based on data from[74,103])

Feedstock conditions:

Moisture content

The moisture content affects the efficiency due to the latent heat of vaporisation of the water in the biomass; the effect is felt either through direct loss of efficiency or through the need for pre-heating. The relationship is illustrated in Figure 98.

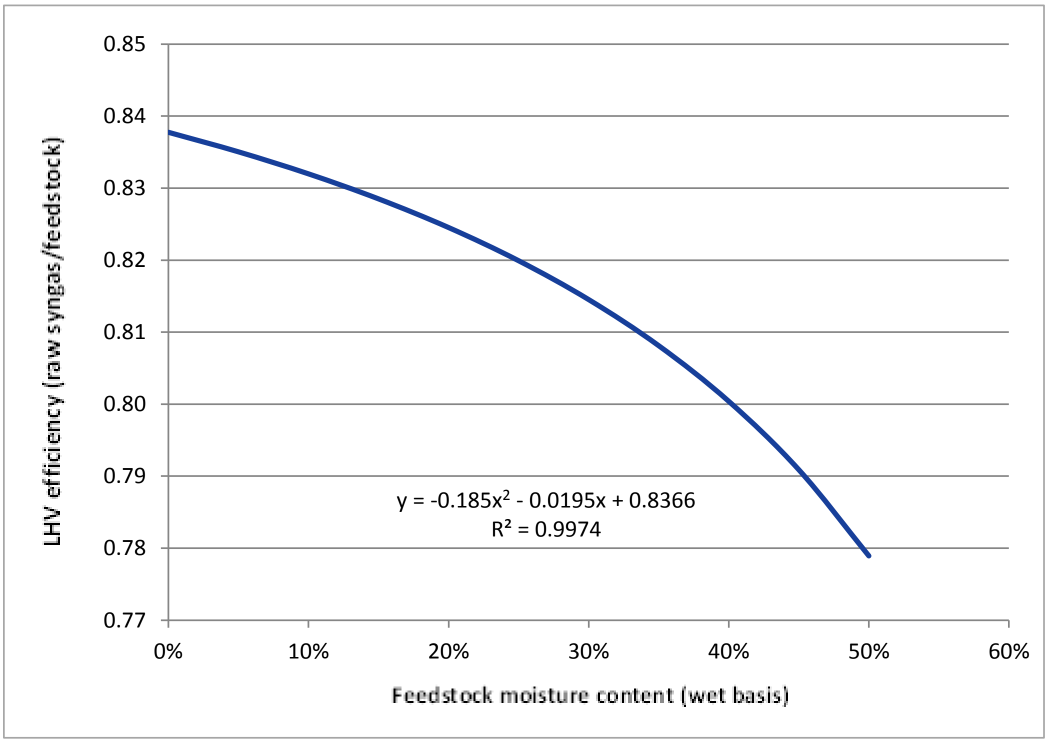

Figure 98: BFB gasifier efficiency vs. feedstock moisture content. Note that ~30% is the recommended limit (data from [104])

Feedstock composition

The effect of ash, sulphur, nitrogen and chlorine in the feed is felt in Opex via disposal or mitigation costs; here it is assumed a level of clean-up suitable for pipelines/turbines. To obtain the base Opex, the following assumptions are used:

- Base values of the impurity values as per the assumed feedstock composition base cases
- Insurance was assumed at 1% of TIC
- Other Opex, e.g. labour, fuels, reagents are process specific and determined on a process by process basis

The labour, O&M, fuels and reagents are from [105]. These incremental effects on Opex are illustrated below, showing how different amounts of biomass constituents in the feedstock alter the total plant Opex.

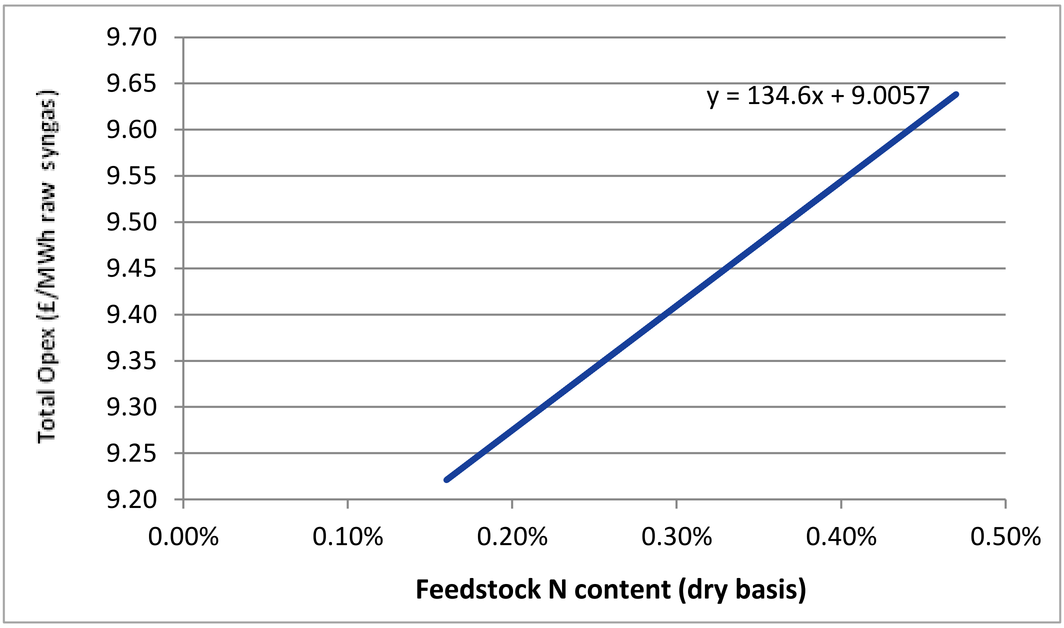

Figure 99: BFB gasifier annual total opex vs. feedstock N content (labour, O&M, fuels and reagents from[105], calculation based on section 2.4.2.4)

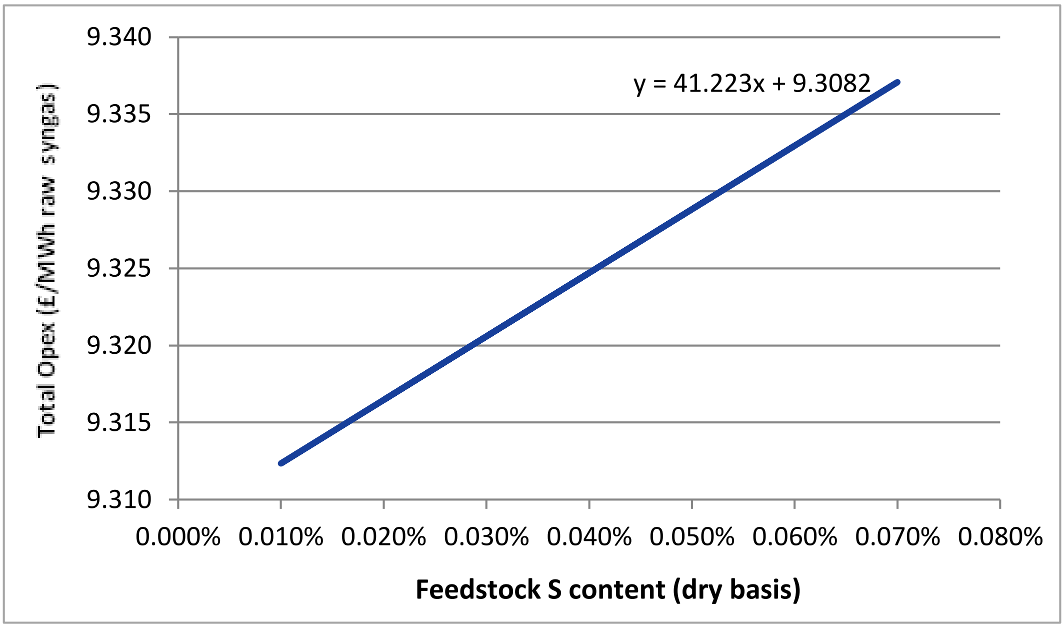

Figure 100: BFB gasifier annual total opex vs. feedstock S content (calculation based on section 2.4.2.4)

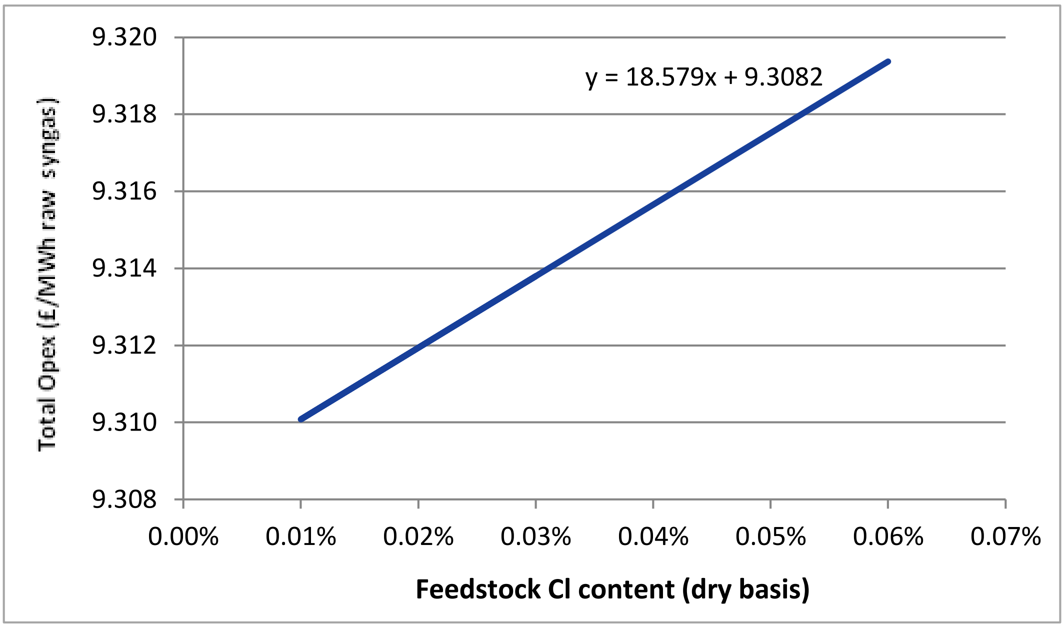

Figure 101: BFB gasifier annual total opex vs. feedstock Cl content (labour, O&M, fuels and reagents from[105], calculation based on section 2.4.2.4)

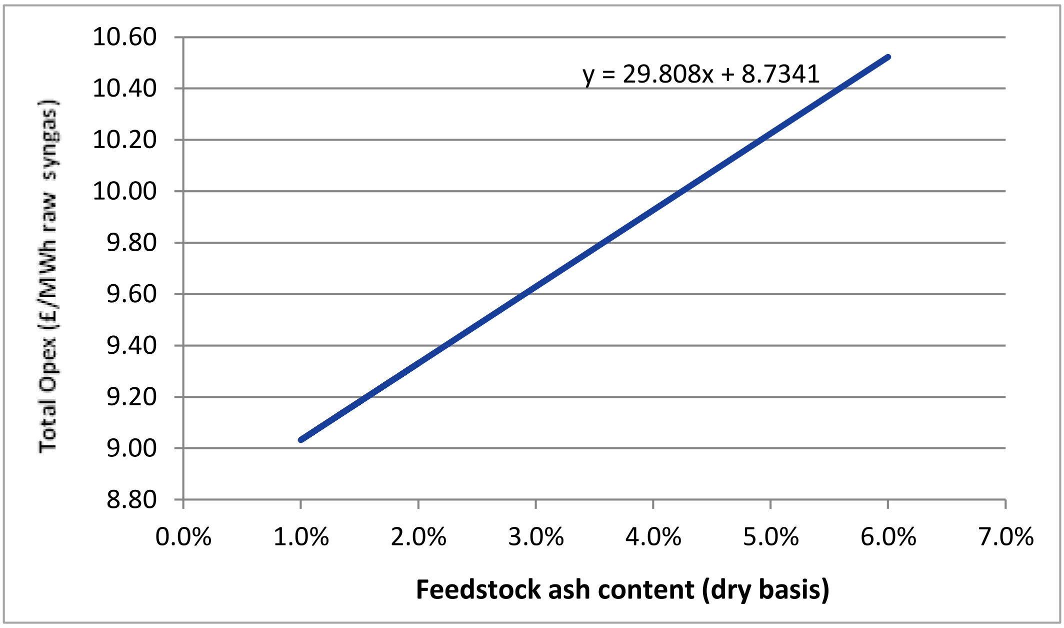

Figure 102: BFB gasifier annual total opex vs. feedstock ash content (labour, O&M, fuels and reagents from[105], calculation based on section 2.4.2.3)

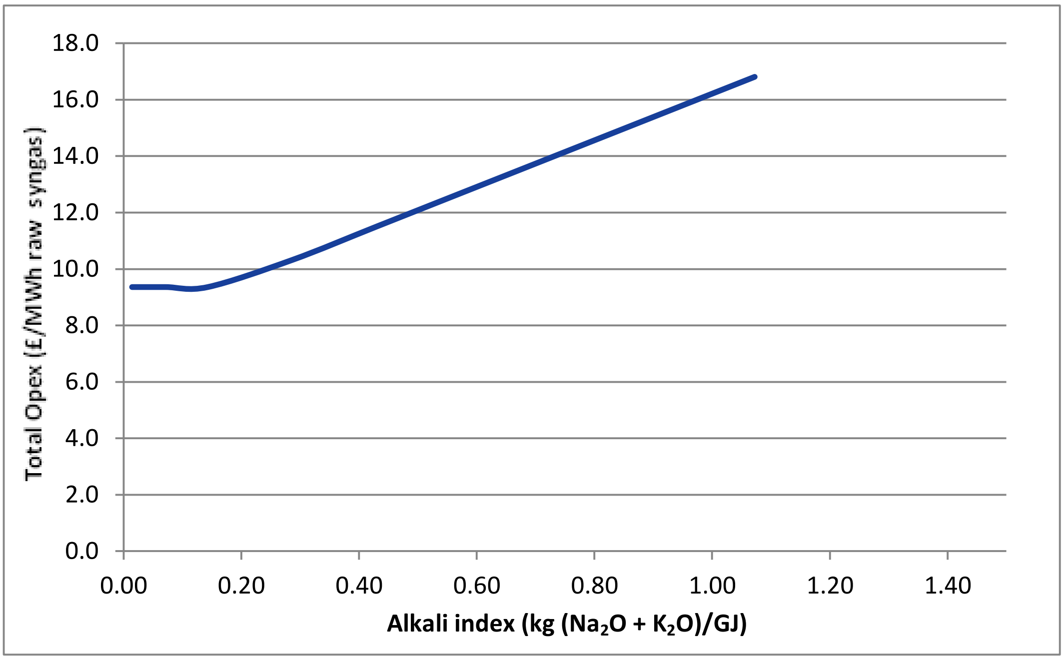

Figure 103: BFB gasifier annual total opex vs. feedstock alkali index (labour, O&M, fuels and reagents from[105], formula based on [106] section 2.4.2.5)

4.3.4 Available options for improvement

Being a more mature technology than the other fluidised bed gasifier types, there is less research into new and innovative ways to improve performance of bubbling fluidised beds compared to the other gasifiers.

The key challenges with respect to this project include the ability to deal with a broader range of feedstocks while producing a better product e.g. through good thermal management to maintain temperatures at the desired level and avoid operability problems and to reduce the tar content of the product and therefore reduce the need for downstream processing.

For example, Shen et al. showed that by introducing catalysts into a fluidised bed, rice husk char or rice husk ash can be converted into syngas[107]. Biomass tar can be converted easily by using co-pyrolysis with supported nickel-iron catalysts, with a potential maximum tar conversion efficiency of up to 93% at 800°C. These catalysts are further reduced by reducing gases such as CO, enhancing the catalytic performance of the tar conversion process. Another advantage of adding these metal oxides to the system is that by adding to the bed material, the catalyst can also improve the overall fluidisation behaviour of the bed material compared to a bed of just silica.

Modelling of fluidised beds at the particle level is in its infancy, but within a decade it is likely that direct simulation of full beds will be possible. This will allow hotspot formation to be examined, which is a frequent precursor to bed agglomeration. Recent work in the UK has examined the potential for agglomeration of sand beds when different feedstocks are burned, since the interactions of the different ashes and their propensity to slag are still not perfectly known.

As previously mentioned, since bubbling fluidised beds have a more fixed design and have been in operation for much longer than circulating or dual fluidised beds, there is considerably less innovation into new designs, with the majority of innovation coming via use of novel catalysts in order to reduce waste production, such as mentioned above. These changes are minor and can be implemented relatively easily in to existing processes or new designs, and as such have a TRL of 6.

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[107] In situ catalytic conversion of tar using rice husk char/ash supported nickel–iron catalysts for biomass pyrolytic gasification combined with the mixing-simulation in fluidized-bed gasifier. Yafei Shen, Peitao Zhao, Qinfu Shao, Fumitake Takahashi, Kunio Yoshikawa

4.4 Circulating Fluidised Bed Gasifiers

4.4.1 Technology description

The second major type of fluidised bed, the circulating fluidised bed gasifier, has a longer residence time than the bubbling fluidised bed, so is more suitable for high-volatility fuels. The principle of a circulating fluidised bed is to have two integrated units in which the process takes place, with the bed material able to circulate between the two. In the first unit, called the riser, the bed material is kept fluidised by the gasifying agent, with a higher velocity than that found in a bubbling fluidised bed, between 3.5-5.5 m/s instead of 0.5-1 m/s, which allows the bed material to be fluidised to a much greater degree. This entrains smaller particles from the bed material, which leave the riser and pass into a cyclone with the syngas. The cyclone separates the syngas from the solid bed material, with the syngas removed from the gasification equipment and the bed material recycled back into the riser.

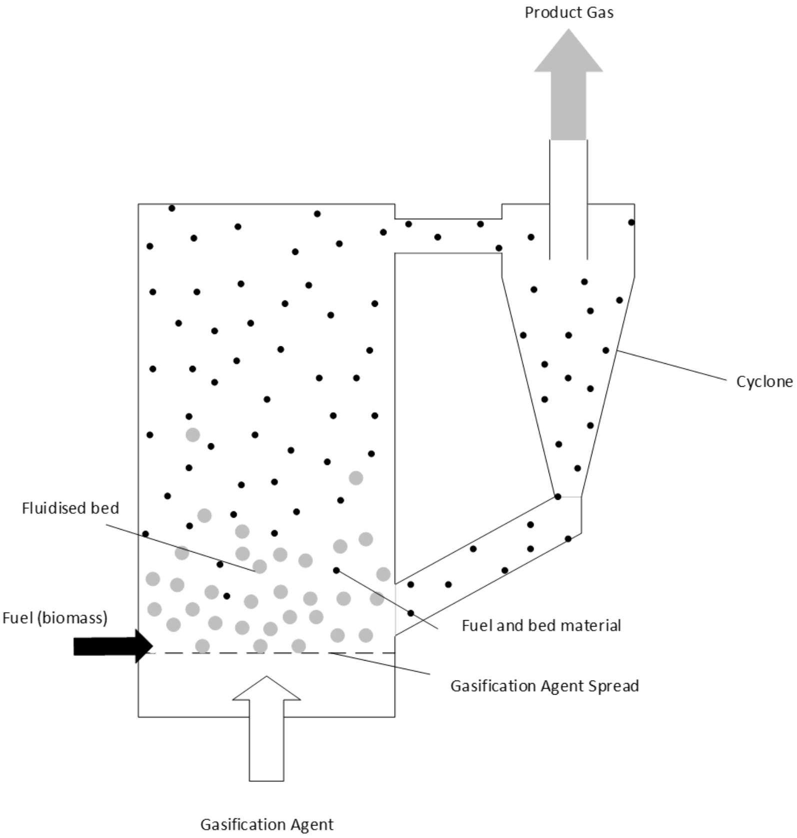

Figure 104: Schematic of Circulating Fluidised Bed Gasifier, derived from[98]

The riser operates at between 800-900°C, depending on the feedstock and flowrates of gasifying agent, with the hot product gas from the cyclone leaving at a lower temperature of 650-750°C. The unit will usually not operate at temperatures above 900°C due to the potential of particulate matter such as ash fusing to form large agglomerated particles, which would hamper heat exchange and reduce the conversion of biomass to the product.

CFBs have, in principle, a number of advantages in their operation compared to BFBs. Circulating beds are more able to process larger volumes of lightweight, fine feed than a bubbling bed, and are as such argued to be suitable for finer particles. CFBs can be scaled up to a larger capacity than atmospheric or low pressure BFB gasifiers, due to a much larger range of acceptable gas velocities and the use of the entirety of the reactor volume. The char conversion and carbon burn-outs of circulating bed gasifiers are reported to be higher than those of bubbling bed gasifiers, with carbon conversions of up to 96% demonstrated from wood based feedstocks. However, this advantage of using circulating beds over bubbling beds has been disputed, as a well-designed BFB gasifier is able to achieve similar conversions, as the recirculation is mostly lost as the friable biomass is discarded from the cyclone in a CFB, thus reducing the potential improvement possible over the BFB gasifier.

4.4.2 Development status and timescales

Most of the information regarding circulating fluidised bed gasification comes from pilot scale facilities, with only a small number of commercial facilities appearing to be currently operating, of which most are only viable as they operate as niche facilities due to local conditions. Because of this, this technology has been assigned a TRL of 7, as there are no significant large scale commercial applications which suggest that a more widespread rollout of the technology is imminent, except for a few niche uses. There are however several pilot scale plants working globally. The expected commercial scale could reach 100 MW_{th}, with around 10-20 years required to reach TRL 9. This is because there is still work to be done on thermal management and solids handling, and also because build rates have been slow.

The largest CFB gasifiers currently thought to be in operation globally are the pair of Foster Wheeler plants built in Lahti, Finland in 1997 (up to 70 MW_{th}) and in Ruien, Belgium (up to 90 MW_{th}) in 2002. These co-fire the syngas produced in large pulverised coal boilers – however, it is unclear whether these atmospheric pressure gasifiers are still in operation (the feed-in tariff for Ruien stopped in 2013). Several older plants are now known to have now shut, including Uhde pressurised HTW's 120 MW_{th} peat to ammonia plant in Oulu, Finland and their 135 MW_{th} MSW/lignite plant in Berrenrath, along with a host of other smaller lime kiln plants that Foster Wheeler built in the 1980s.

The 18 MW_{th} CHRISGAS pressurised BIGCC demonstration in Värnamo operated from 1993 to 1999, but was mothballed after funding delays meant an O_2-fed upgrade to biofuel production did not go ahead. NSE Biofuels' 12 MW_{th} pilot for BTL production only operated for 2 years, before closing and the decision to not pursue the technology.

CFB technology has mainly been developed and used in northern European countries, predominantly Scandinavia, with little deployment further afield. The TEABPP team are not aware of any biomass-fired CFB gasifier developers in North America or Asia.

No information was found in the literature about the timeframe required in order to develop and construct a commercial plant, since most of the facilities where circulating fluidised bed gasifiers are used with biomass are pilot scale and the few demonstration or commercial scale facilities have not disclosed much data regarding the construction and commissioning phase. It could be expected that that plant construction timescales would be slightly longer than an equivalent sized CFB combustion plant, given the additional gas handling and system complexity.

4.4.3 Impact of different feedstock parameters on operation and cost

Availability and Opex are for the base feedstocks; high ash and/or high alkali metal content feedstocks would increase downtime and Opex. The latter is modelled using the threshold concept; the former may have implications for bed material fouling.

Plant size

The plant size primarily affects the Capex. A 0.79 power law is estimated given the mixed nature of the equipment. This gives rise to a cost curve as in Figure 105.

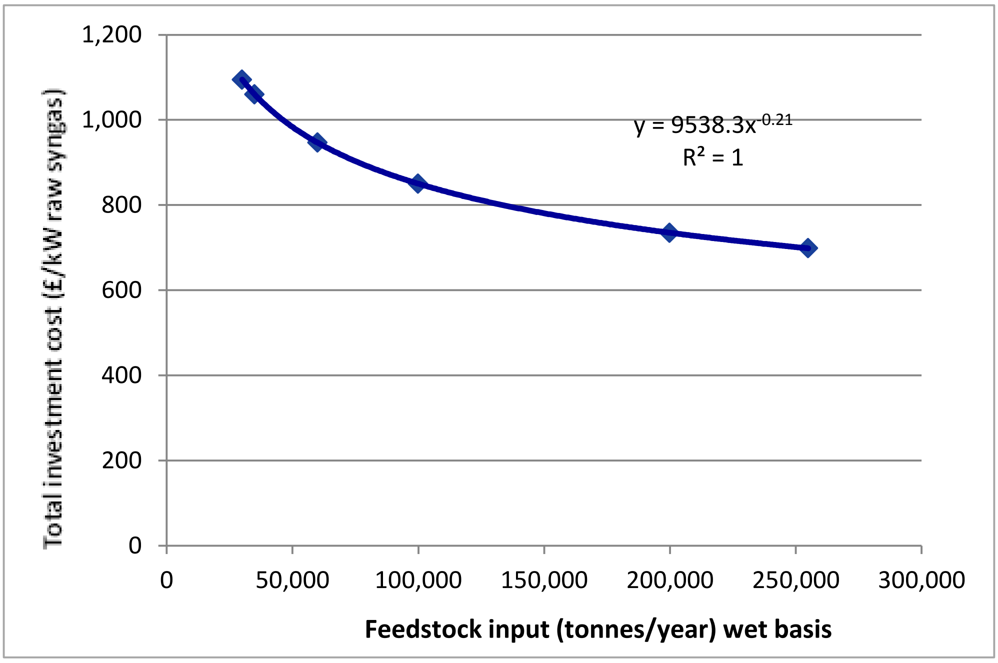

Figure 105: CFB total investment cost vs. plant scale (based on data from[74,103])

Feedstock conditions

Moisture content

The moisture content affects the efficiency due to the latent heat of vaporisation of the water in the biomass. The relationship is illustrated in Figure 106.

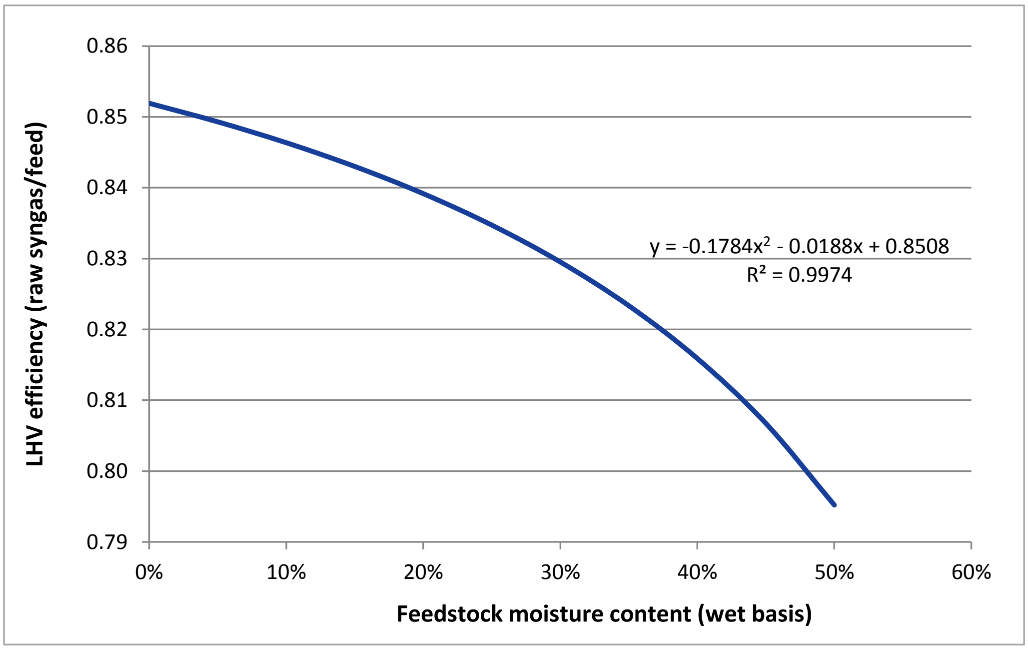

Figure 106: CFB gasifier efficiency vs. feedstock moisture content. High moisture content biomass is potentially usable (data from [104] and ICON calculations)

Feedstock composition

The effect of ash, sulphur, nitrogen and chlorine in the feed is felt in Opex via disposal or mitigation costs; here it is assumed a level of clean-up suitable for pipelines/turbines. To obtain the base Opex, the following assumptions are used:

- Base values of the impurity values as per the assumed feedstock composition base cases
- Insurance was assumed at 1% of TIC
- Other Opex, e.g. labour, fuels, reagents are process specific and determined on a process by process basis

For the base Opex, the O&M, labour, fuels and reagents costs are from[108]. These incremental effects on Opex are illustrated below, showing how different amounts of biomass constituents in the feedstock alter the total plant Opex.

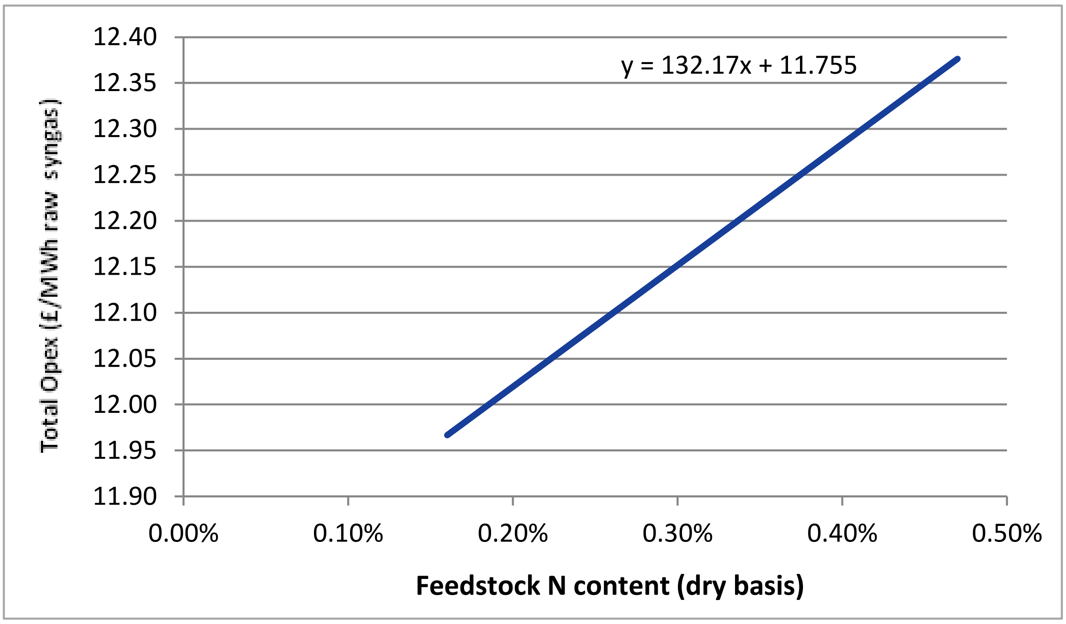

Figure 107: CFB gasifier annual total opex vs. feedstock N content (labour, O&M, fuels and reagents from[108], calculation based on section 2.4.2.4)

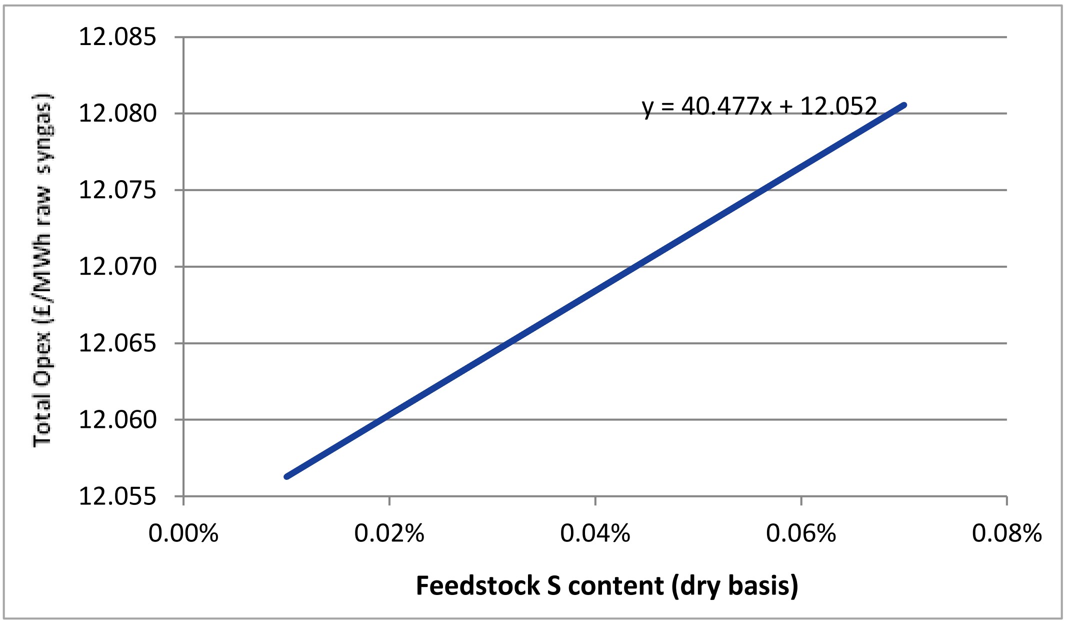

Figure 108: CFB gasifier annual total opex vs. feedstock S content (labour, O&M, fuels and reagents from[108], calculation based on section 2.4.2.4)

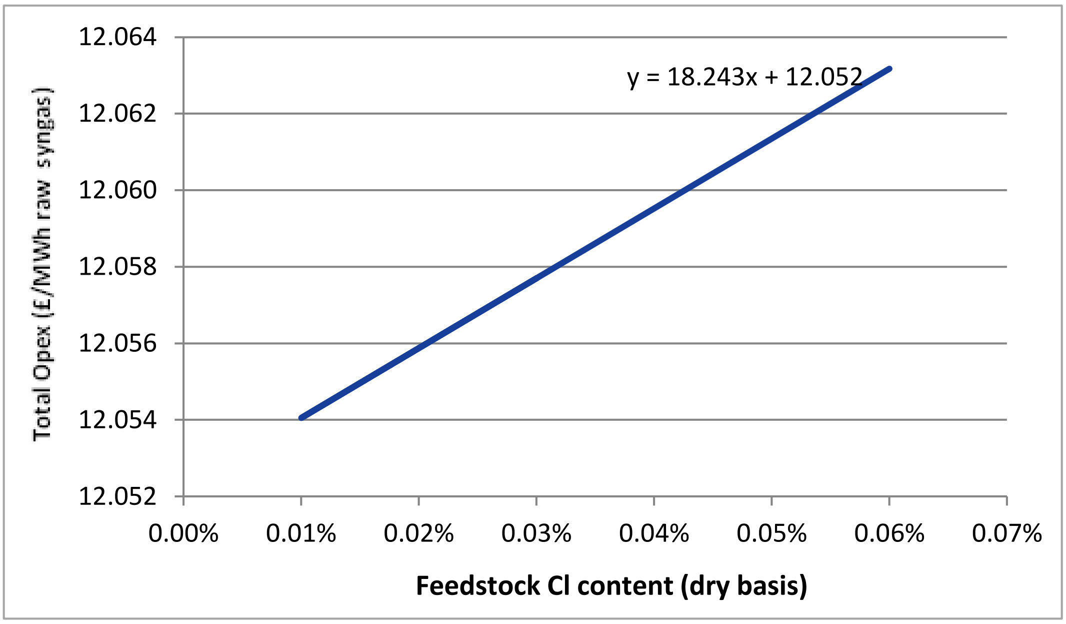

Figure 109: CFB gasifier annual total opex vs. feedstock Cl content (labour, O&M, fuels and reagents from[108], calculation based on section 2.4.2.4)

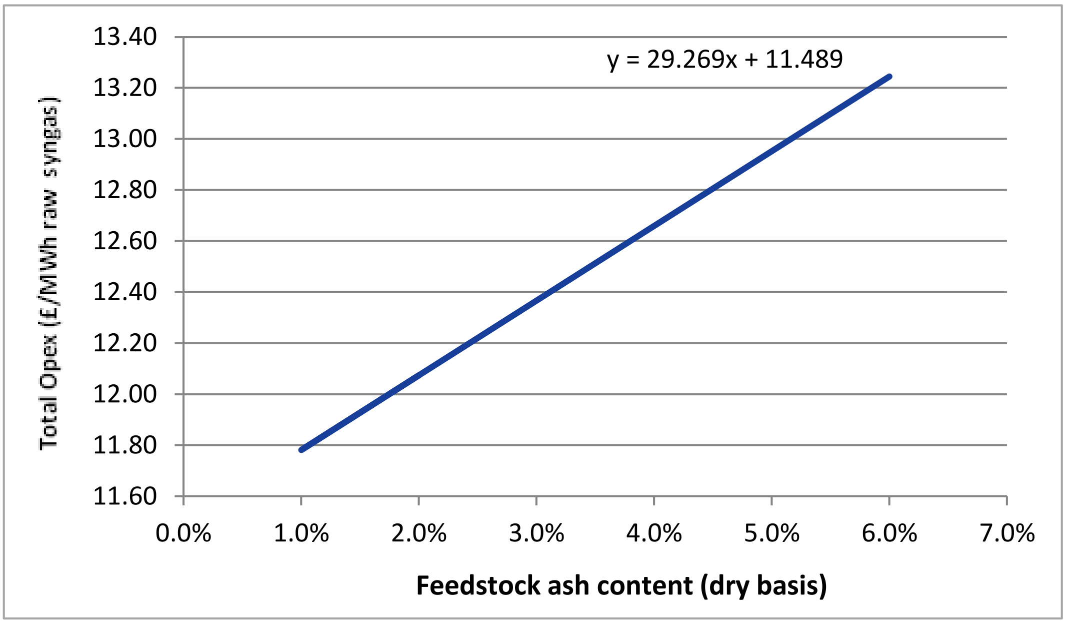

Figure 110: CFB gasifier annual total opex vs. feedstock ash content (labour, O&M, fuels and reagents from[108], calculation based on section 2.4.2.3)

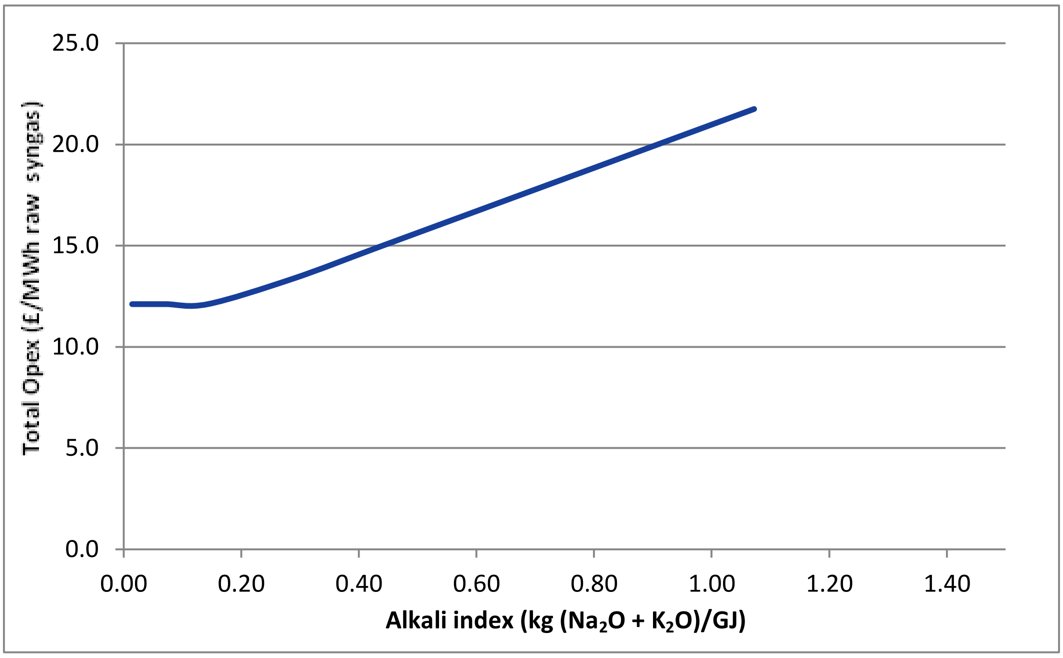

Figure 111: CFB gasifier annual total opex vs. feedstock alkali index (labour, O&M, fuels and reagents from[108], formula based on [109] and 2.4.2.5)

4.4.4 Available options for improvement

The main areas for improvement in CFB are relatively common with those for BFB. There are two aspects of operation that are key to gasification efficiency and gas usability. Firstly, either avoiding the production of tar, or destroying any tar that does get formed. This allows the gas from the gasifier to be used in subsequent electricity generation technologies – and in general the more complex the technology the higher the efficiency and the lower the tolerance to tar. Thus, a modified internal combustion engine can be used, a turbine is better and a fuel cell best of all but most likely to have significant issues with tars. Most uses of fuel gas require a tar content of magnitude 0.05g/Nm^3 or less[110]. Research focuses on either adding some form of catalytic material into the CFB which acts to enhance cracking (and potentially to reduce emissions of unburned CO / NO_x), or potentially modification of the gasifier itself (for example, addition of a secondary high temperature stage to enhance cracking). Secondly, to improve the gasification efficiency, it is necessary to convert as much of the material as possible to gases rather than char. Research here focusses on pressure, temperature, particle size, steam concentration, oxygen concentration – essentially varying parameters to ensure that the optimal gasification conversion is achieved without starting to combust the char. Depending upon the use proposed for the gas, the ratio of H_2 to CO might be of importance (chemicals, liquid fuel production and gaseous fuel production all have different requirements). Secondary issues (though extremely important to operation) include ensuring that fouling and slagging is minimised; this may well be a clear area where pre-processing could reduce the propensity of the char / ash to foul.

Examples of current / previous research efforts in this area include: Matsuoka et al. investigated a system where a pyrolyser and gasifier were combined in a circulating fluidised bed, with the pyrolyser and gasifier isolated from each other[111]. The feed of coal char is kept separated from the main gasification chamber so as not to interact with volatiles such as H_2 in the product gas, with this isolation giving a lower partial pressure of hydrogen in the gasification chamber and a *decrease in tar vapour in the gasifier*. These factors combined increase the overall char gasification rate while not increasing the temperature.

Xiao et al. carried out an experimental assessment of the gasification of animal waste in internally circulating fluidised beds as a multistage gasification process[112]. In this process, the bed material acts as the medium for both char combustion and internal heat exchange. In this manner, low temperature operation is able to produce a hydrogen rich, low tar product gas. When assessing the performance under a range of temperatures and S/B ratios, it was found that tar production could be reduced by operating at low temperatures and a lower S/B ratio. By increasing the S/B ratio, the LHV of the product gas is decreased, *but there is an increase in the H_2/CO ratio* of the product gas, meaning it could be used as a feedstock in chemical synthesis processes instead of just as a fuel.

Simanjuntak and Zainal investigated the effects of air blown gasification of sawdust using a two-compartment cylindrical circulating fluidised bed[113]. In this system, the gasification zone is in the annulus of the gasifier and combustion occurs in the draft tube. *Fluidisation ratio and equivalence ratio can be manipulated to control the internal temperature of the gasifier*. Char, gasification residue and the hot bed material is able to provide the necessary heat for gasification to the gasification area of the vessel. Compared to previous studies, the gasifier is shown to be able to produce a product gas with comparably higher concentrations of CH_4 and CO.

On a larger scale, Thunman et al. have investigated using ilmenite as an oxygen carrier in an experimental 12 MW_{th} circulating fluidised bed[114]. This was with the aim of reducing emissions of hydrocarbons and CO as much as possible for use with biomass. *By adding up to 40% by weight of ilmenite to the bed material, product gas concentrations of CO and NO were reduced by 80% and 30% respectively*. By using ilmenite mixed in with the bed material, measurements have shown that there is a reduction in concentrations of both CO and also all hydrocarbons across a furnace when adding ilmenite to the bed material compared to a bed material of just silica due to the oxidising effect of the ilmenite.

These technologies show some potential for improving the process efficiency or to reduce waste products generated within the gasifier, but all bar the last example have only been operated on a lab scale at best, with a TRL of 3-5, while the last example with the ilmenite has been tested on a larger scale and has a TRL of 5-6.

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[108] Knoef, 2012, Handbook of Biomass Gasification, biomass technology group.

[109] Fahmi R., Bridgwater A. V., Darvell L. I., Jones J. M., Yates N., Thain S., et al. (2007). The effect of alkali metals on combustion and pyrolysis of Lolium and Festuca grasses, switchgrass and willow. Fuel 86, 1560–1569 10.1016/j.fuel.2006.11.030.

[110] Fjellerup J., Ahrenfeldt J., Henriksen U. and Gøbel B., —Formation, Decomposition and Cracking of Biomass Tars in Gasification, Technical University of Denmark, 2005

[111] Enhancement of coal char gasification using a pyrolyzer–gasifier isolated circulating fluidized bed gasification system. Koichi Matsuoka, Sou Hosokai, Koji Kuramoto, Yoshizo Suzuki

[112] Multi-stage biomass gasification in Internally Circulating Fluidized-bed Gasifier (ICFG): Test operation of animal-waste-derived biomass and parametric investigation at low temperature

[113] Experimental study and characterization of a two-compartment cylindrical internally circulating fluidized bed gasifier. J.P Simanjuntak, Z. A. Zainal

[114] Using an oxygen-carrier as bed material for combustion of biomass in a 12-MWth circulating fluidized-bed boiler. Henrik Thunman, Fredrik Lind, Claes Breitholtz, Nicolas Berguerand, Martin Seemann

4.5 Dual Fluidised Bed Gasifiers

4.5.1 Technology description

The third major type of gasifier is the dual fluidised bed gasifier, which unlike the previous two classes utilises two separate fluidised beds. The biomass feed is fed into base of the gasification unit into the bed material which is typically fluidised by steam (although note that ECN's plant described in the next section operates in pyrolysis mode).

In "classic" gasifiers (e.g. downdraft), it is combustion of mainly pyrolysis gases that provides the energy for the system, while the syngas is made by reduction of char. However, in this type of gasifier, the syngas is made from mostly pyrolysis gases (whether reacted or not) plus some char reduction, while the heat is provided by char combustion in the separate combustor.

The fluidised bed material promotes gasification to produce the syngas product, with some of the bed material and char circulated into the combustor. Air is added to the combustor, which is designed as a fast fluidised bed and the char is combusted to heat the bed material. The bed material is then removed from the combustor flue gas stream in a cyclone and is returned to the gasifier. This process allows the gasifier to be heated by the combustor, which can also burn an additional fuel in order to supply the necessary heat to the process. One of the main attractions of the dual fluidised bed is that the product gas produced has no nitrogen in it, so the proportions of hydrogen and carbon monoxide in the syngas are higher, corresponding to a greater heating value of the gas when compared to other fluidised bed operations (unless they were to go to the expense of using oxygen as a gasifying agent). Temperature is controlled via the combustor, and hence by the input of additional fuel. The flue gas contains the majority of the particulate matter associated with biomass systems, producing a cleaner syngas than other fluidised bed systems.

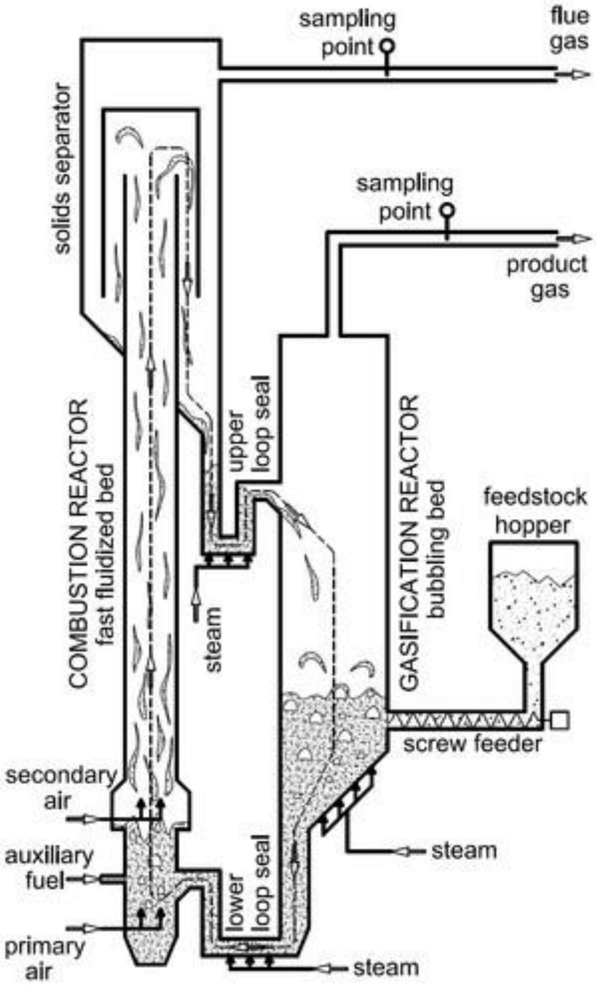

Figure 112: Schematic of Dual Fluidised Bed Gasifier[115]

In terms of their advantages when compared to other types of fluidised bed reactor, dual fluidised beds perform more similarly to circulating fluidised beds, but with the added advantage of a cleaner syngas product with a higher heating value due to the lack of nitrogen and more hydrogen present (due to the use of steam). However, since dual fluidised beds are a newer technology than the other two types, there is less information available about their operation and challenges associated with long term use. They have the potential to be more complex to control than circulating beds and bubbling beds by virtue of having the fuel addition rate as another input variable that must be controlled.

4.5.2 Development status and timescales

Dual fluidised bed gasifiers have been developed more recently than BFB and CFB gasifiers, starting in the mid 1990's. While not quite at the same level of development as either circulating or bubbling fluidised bed reactors, there are a small number of pilot plants, including ECN's MILENA 1 MW_{th} pilot plant and the facility at Güssing (operated for over 80,000 hours at 4 MW_{th} input). Regarding the former, the product gas is entirely made up of gaseous products of pyrolysis with some steam reactions but not much. The downstream separation (solvent extraction) system strips out the C6+

which is recycled back to the combustion side of the gasifier. Hence the syngas is a mix of C6-gases unlike the typical gasification syngas.

The developer and owner of the Güssing technology, Güssing Renewable Energy GmbH, are developing similar gasifiers based on the same principles in India and Thailand. The only demonstrations of the technology currently known to be operating are a 14 MW_{th} demonstration with ORC CHP in Senden, Germany, and the GoBiGas 32 MW_{th} wood pellet gasifier in Gothenburg, that entered operation in 2014, to produce syngas for methanation and gas grid injection. This GoBiGas project is working with Metso, Repotec and Haldor Topsoe, and has a 128 MW_{th} Phase 2 in planning.

In the US, a 73 MW_{th} dual fluidised wood gasifier was only tested for about 1,000 hours at the McNeil facility in Vermont between 1999 and 2001[116]. This SilvaGas technology was bought out by Rentech, who recently abandoned the technology. A similarly sized project in Montgomery, New York never progressed past breaking ground.

Although dual fluidised beds do not have as long a history of operation as BFBs or CFBs, they now have a number of operating pilot and demonstration facilities, so the technology has been assigned a TRL of 6-7. Dual FB gasification benefits from the experience gained with BFB and CFB, but is yet to be demonstrated at high pressure as is the case with several BFB and CFB developers – however, if pressurised Dual FB systems were developed, they have the potential to produce a high value, nitrogen free, pressurised syngas.

There is still little information in the literature regarding the duration of development and construction which can be attributed to any commercial scale plant. Development to full commercialisation is dependent on the successful demonstration of the technology in the recently built facilities (and other projects worldwide). 30 MW_{th} input is expected to be the minimum commercially viable scale for biomethane production and there will be power and CHP applications in this scale range, hence the technology is not far off proving itself at commercial scale. However, reaching mass-deployment at TRL 9 will take considerably longer, most likely around 15-20 years.

4.5.3 Impact of different feedstock and other parameters on operation and cost

Availability and Opex are for the base feedstocks; high ash and/or high alkali metal content feedstocks would increase downtime and Opex. The latter is modelled using the threshold concept; the former may have implications for bed material fouling.

Plant size

The plant size primarily affects the Capex. A 0.79 power law is estimated given the mixed nature of the equipment. This gives rise to a cost curve as in Figure 113.

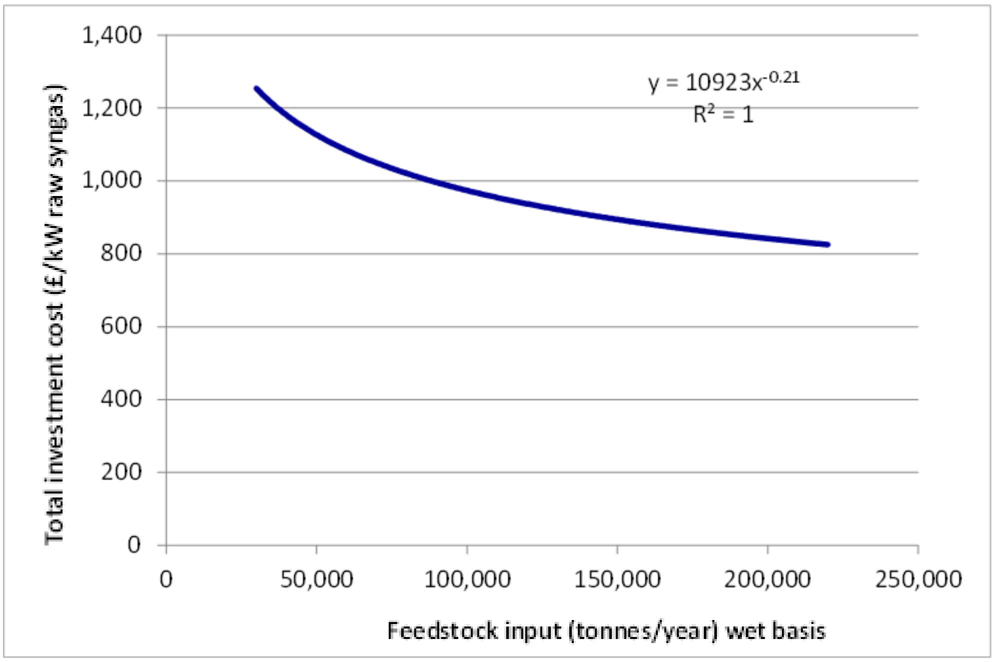

Figure 113: Dual FB gasifier total investment cost vs. plant scale (data from[74],[103])

Feedstock conditions

Moisture content

The moisture content affects the efficiency due to the latent heat of vaporisation of the water in the biomass. The relationship is illustrated in Figure 114.

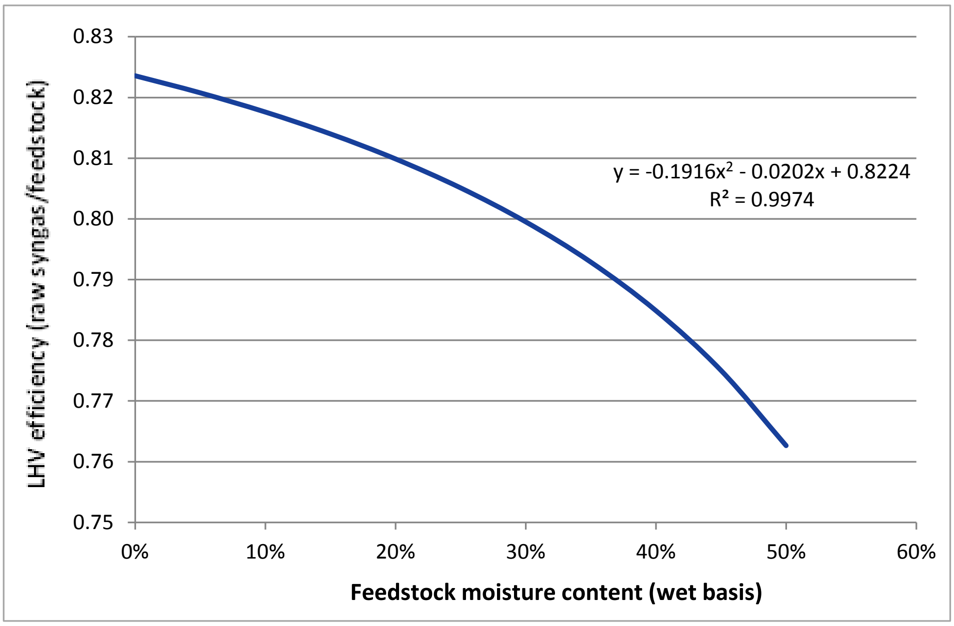

Figure 114: Dual FB gasifier efficiency vs. feedstock moisture content. Maximum moisture content of up to 38% is recommended (data from [104] and ICON calculations)

Feedstock composition

The effect of ash, sulphur, nitrogen and chlorine in the feed is felt in Opex via disposal or mitigation costs; here it is assumed a level of clean-up suitable for pipelines/turbines. To obtain the base Opex, the following assumptions are used:

- Base values of the impurity values as per the assumed feedstock composition base cases
- Insurance was assumed at 1% of TIC
- Other Opex, e.g. labour, fuels, reagents are process specific and determined on a process by process basis

For the base Opex, the O&M, labour, fuels and reagents costs are from[117]. These incremental effects on Opex are illustrated below, showing how different amounts of biomass constituents in the feedstock alter the total plant Opex.

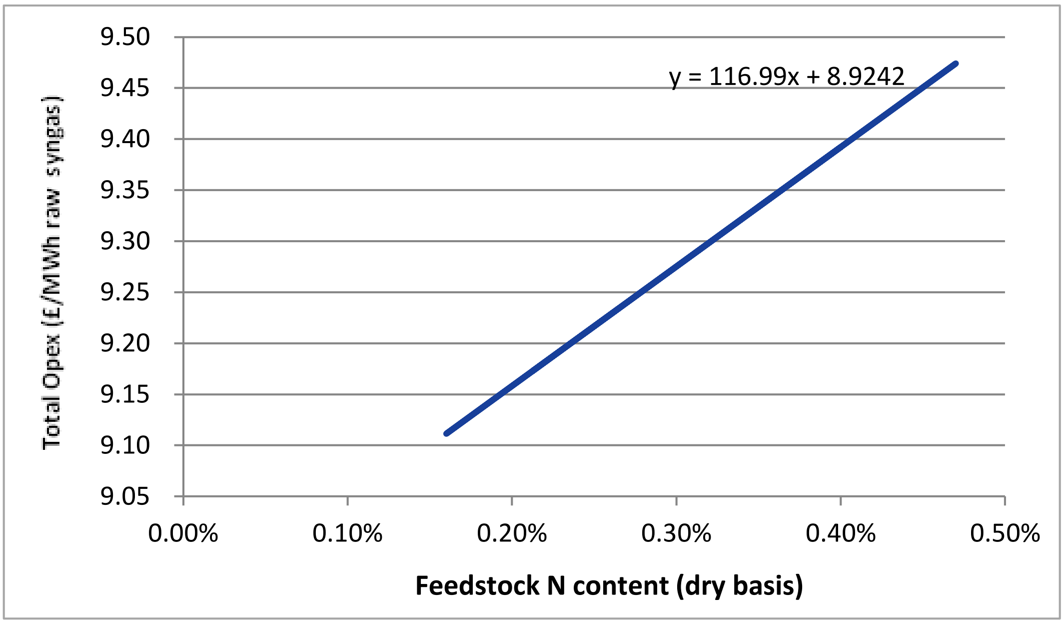

Figure 115: Dual FB gasifier annual total opex vs. feedstock N content (labour, O&M, fuels and reagents from[117], calculation based on section 2.4.2.4)

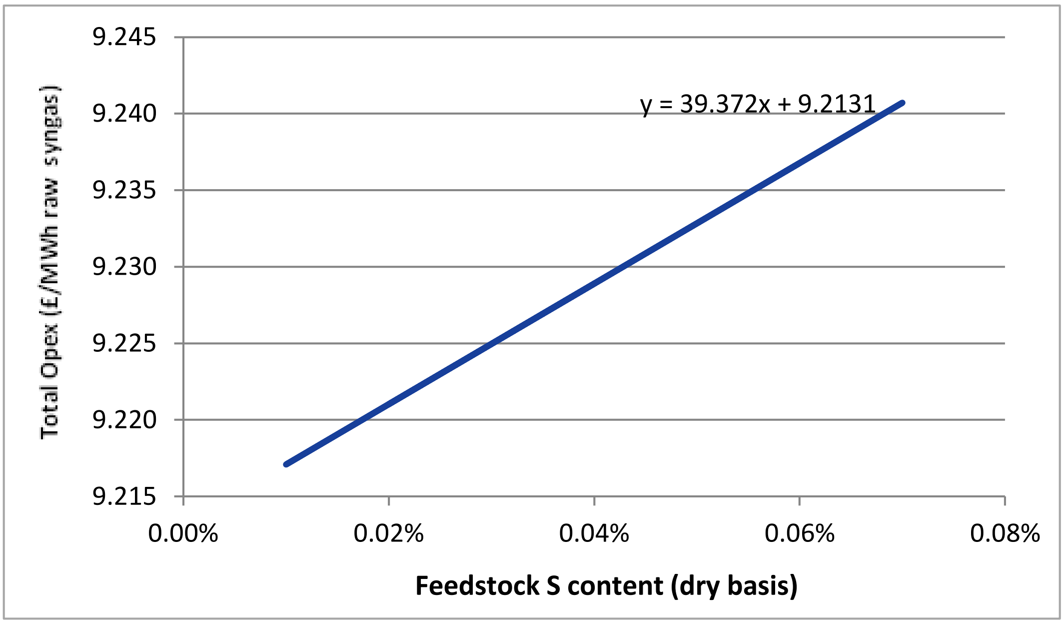

Figure 116: Dual FB gasifier annual total opex vs. feedstock S content (labour, O&M, fuels and reagents from[117], calculation based on section 2.4.2.4)

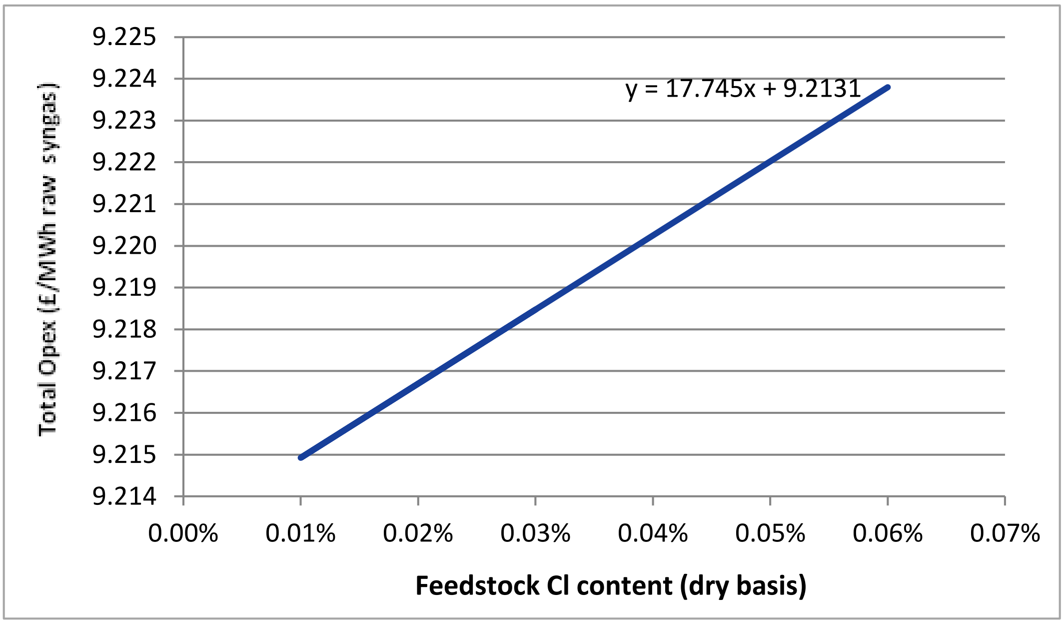

Figure 117: Dual FB gasifier annual total opex vs. feedstock Cl content (labour, O&M, fuels and reagents from[117], calculation based on section 2.4.2.4)

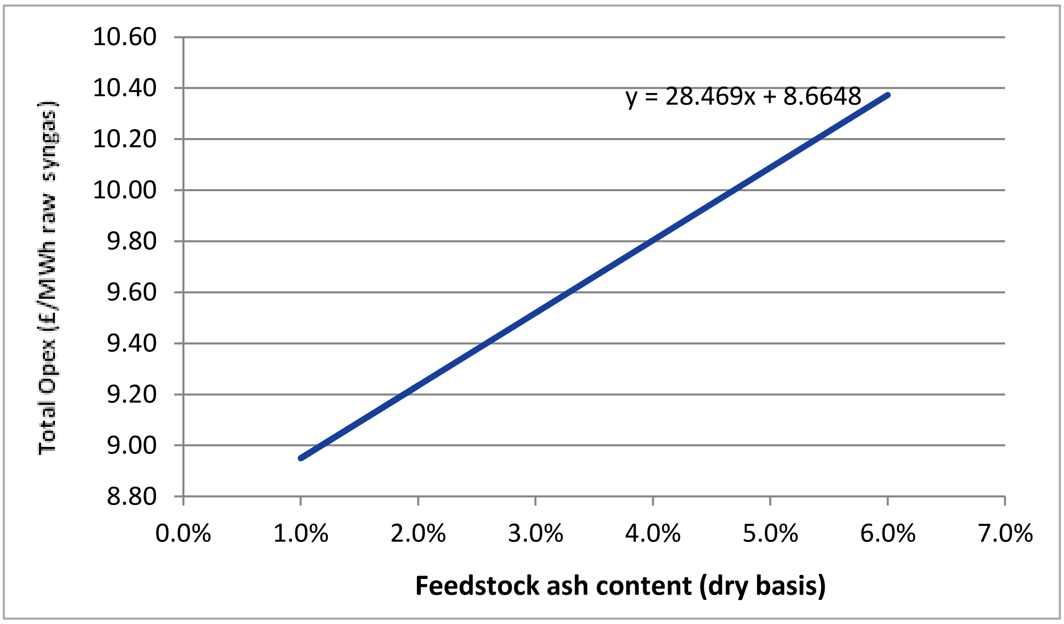

Figure 118: Dual FB gasifier annual total opex vs. feedstock ash content (labour, O&M, fuels and reagents from[117], calculation based on section 2.4.2.3)

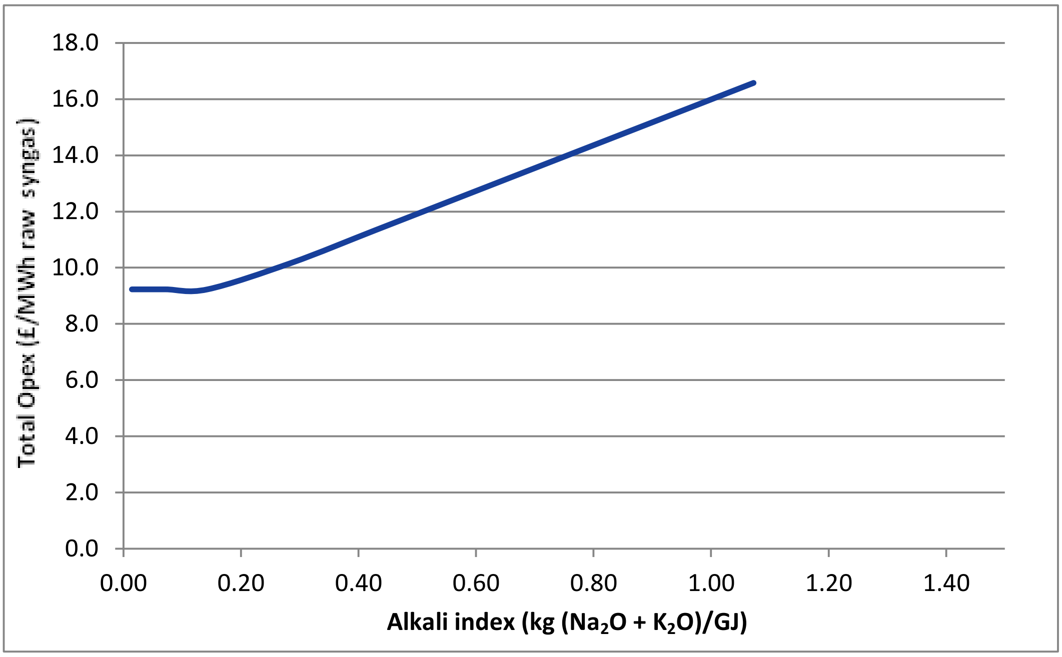

Figure 119: Dual FB gasifier annual total opex vs. feedstock alkali index (labour, O&M, fuels and reagents from[117], formula based on [118] and section 2.4.2.5)

4.5.4 Available options for improvement

The issues discussed for CFB are similar to those for DFB, tar should be minimised, char gasification maximised, unburned hydrocarbons reduced and fouling minimised. The DFB is a modification of a bubbling fluidised bed, essentially designed as a two-stage reactor, and is designed to answer the issues above, most particularly tar emissions.

Xu et al. proposed using a two-stage gasifier in the dual fluidised bed in order to increase efficiency and reduce tar output[119]. The fuel is fed into the bottom chamber which acts like the gasifier in the standard dual fluidised bed and the top chamber has the hot gas and bed material from the combustion chamber fed into it. This facilitates the gas upgrading reactions in this chamber, before the bed material is recirculated back into the chamber below to complete the cycle. This has the dual benefit of a higher gasification efficiency and the production of a gas with a lower tar content than the standard design. *Tests on small experimental setups indicated that the cold gas efficiency was increased by 7% and the tar in product gas reduced by 25%.*

Shaw and Pang studied the effect of *calcite loading* on dual fluidised bed gasification using a woodchip feed, with the gasifier temperature set at 710-750°C and the combustor at 800°C. By increasing the calcite loading to 100% and increasing the steam/biomass ratio, it is possible to increase the H_2/CO ratio from 0.9 to 4.2, while *decreasing tar output in the product gas from 5 to 0.7 g/Nm*^{3}, allowing target specifications for downstream processes to be met. However, these benefits come at the expense of having a larger attrition rate, which increases from 0.5-4.1 kg/h when calcite loading is increased from 0-100%, which may cause operability problems in the gasification system when operated on larger scales[120]. A simple pre-treatment step may simply be to mix in a small proportion of calcite[121]

Göransson and Söderlind carried out preliminary experimental tests of 150kW$_{th}$ gasifier[122]. This was set up as a dual fluidised bed, consisting of a circulating fluidised bed gasifier and a bubbling fluidised bed combustor. This research investigates the effect on the product gas of increasing the ratio at which steam is added compared to biomass, the S/B ratio. The experimental results show that an increase of the S/B ratio leads to an increase in the H_2/CO ratio, although this will reduce the production of CH_4. This produces a syngas rich in hydrogen, making it suitable for use as a feedstock for chemical synthesis, and not simply as a fuel, potentially increasing the product value.

Since normal methods to increase tar conversion may be insufficient to reach the level at which no more tar treatment is required downstream before burning, Gómez-Barea et al. have proposed using a three stage gasifier instead of the normal one or two chambers[123]. In this gasifier, the separation between the different zones will allow a greater overall char conversion by isolating the gasifier from the combustor. As a consequence of this, this gasifier could exhibit greater process efficiency than other designs, and since the majority of tars would be converted internally, it would be able to produce a product gas which could be burned directly. Because of the advantages of this design, it would be especially suitable for use with fuels with high ash content or where bed material discharge requirements due to fouling give large reductions to process efficiency, as both of these problems are mitigated against with this design.

The new technologies discussed here are either conceptual or lab/pilot-scale, so would not be suitable for deployment in the near future without a concerted effort in investment and testing in order to see if there would be any discernible process benefit compared to other gasifier types, with these technologies being equivalent to having TRLs between 3 and 5.

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[115] TU Wien EU-Project "Fecundus" http://www.vt.tuwien.ac.at/chemische_verfahrenstechnik_und_wirbelschichttechnik/gasification_and_gas_cleaning/projects/DE/

[116] Booth M.S., 2013, Burning Money: Biomass Gasification and the DOE Loan Guarantee Programhttp://www.pfpi.net/wp-content/uploads/2013/05/PFPI-Gasification-and-DOE-loan-guarantees.pdf

[117] Catolica et al, (2009) “Economic analysis of a 3MW biomass gasification power plant”. Proceedings of the ASME 2009 3rd International Conference of Energy Sustainability.

[118] Fahmi R., Bridgwater A. V., Darvell L. I., Jones J. M., Yates N., Thain S., et al. (2007). The effect of alkali metals on combustion and pyrolysis of Lolium and Festuca grasses, switchgrass and willow. Fuel 86, 1560–1569 10.1016/j.fuel.2006.11.030.

[119] Two-stage dual fluidized bed gasification: Its conception and application to biomass, Guangwen Xu, Takahiro Murakami, Toshiyuki Suda, Yoshiaki Matsuzaw, Hidehisa Tani

[120] The influence of calcite loading on producer gas composition and tar concentration of radiata pine pellets in a dual fluidised bed steam gasifier, Woei L. Shaw, Shusheng Pang

[121] Ongoing research at Imperial College is investigating such additions

[122] Experimental test on a novel dual fluidised bed biomass gasifier for synthetic fuel production. K. Göransson, U. Söderlind, W. Zhang

[123] - Improving the performance of fluidized bed biomass/waste gasifiers for distributed electricity: A new three-stage gasification system. Alberto Gómez-Barea, Bo Leckner, Angel Villanueva Perales, Susanna Nilsson, Diego Fuentes Cano

4.6 Entrained Flow Gasification

4.6.1 Technology Description

In entrained-flow gasifiers[124], fine biomass[125] feed and the oxidant (oxygen[126]) and/or steam are fed co-currently to the gasifier. This results in the gases surrounding or entraining the biomass particles as they flow through the gasifier in a dense cloud. Entrained-flow gasifiers operate at high temperature and pressure—and extremely turbulent flow—which causes rapid feed conversion and allows high throughput. The gasification reactions occur at a very high rate (typical residence time is on the order of few seconds), with high carbon conversion efficiencies (98-99.5%). These gasifiers have been developed for coal, and only limited testing with biomass has been performed. There are a number of reasons for the lack of application of these coal reactors to biomass, but the high cost of feed preparation to reduce moisture content to low levels and reduce the particle size, along with low heat content, is the primary concern. Refractory life is also a concern for biomass feeds with high potassium content.

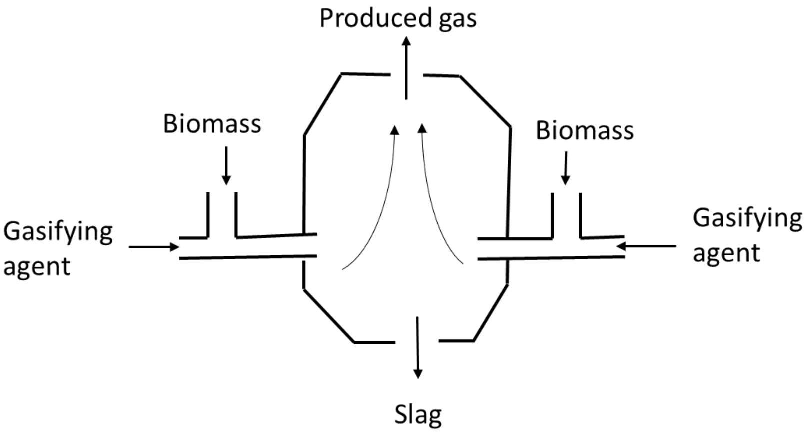

Figure 120: Entrained flow gasifier, derived from[127]

The tar, oil, phenols and other liquids produced from devolatisation of biomass inside the gasifier are decomposed into hydrogen (H_2), carbon monoxide (CO) and small amounts of light hydrocarbon gases; the remaining char is oxidised. Entrained-flow gasifiers have the ability to handle practically any biomass feedstock and produce a clean, tar-free syngas. Given the high operating temperatures, gasifiers of this type melt the biomass ash into vitreous inert slag[128]. The high temperatures involved in this type of gasification tend to shorten the life of system components, including gasifier vessel refractory[129]. Also, it may be necessary to add fluxes or blend feedstock parameters to achieve good slagging characteristics.

The fine biomass feed can be fed to the gasifier in either a dry or slurry form102[125]. The former uses a lock hopper system, while the latter relies on the use of high-pressure slurry pumps. The slurry feed is a simpler operation, and requires less feedstock drying, but it introduces water into the reactor which needs to be evaporated. The result of this additional water is a product syngas with a higher H_2 to CO ratio, but with a lower gasifier thermal efficiency.

The on-site feed preparation system (e.g. moisture and size adjustment) will need to take into consideration the gasifier requirements and the ranges of characteristics of the potential incoming feedstocks and therefore be designed accordingly. The main application for entrained flow gasifiers in this TEABPP study will be for the production of electricity, via combined cycle turbines – given the very large scale of the gasifier, applications for the heat (or CHP) are unlikely.

4.6.2 Development status and timescales

Entrained flow slagging gasifiers have been developed since 1950 and have been selected for the majority of commercial-sized coal IGCC applications. Slagging entrained flow gasifier manufacturers include Shell, Texaco, Krupp-Uhde, Siemens (formerly Future-Energy, formerly Babcock Borsig Power and Noell), E-gas (formerly Destec and Dow), MHI (Mitsubishi Heavy Industries), Hitachi and Linde (formerly Choren, formerly UET).

One example of a large-scale entrained flow gasifier was the 600 MW_{th} coal-fired Shell gasifier in Buggenum, the Netherlands[130]. This IGCC plant was owned by the utility company NUON, producing electricity with a net efficiency of 43%, until its closure in 2013[131]. Regular tests were performed with wood dust, sewage sludge and chicken manure up to c. 10% co-firing on an energy basis from 2006. Tests more recently carried out co-firing up to 70% with torrefied and steam exploded pellets (although using the existing lock hopper system presented several feeding difficulties). Another example is a 130 MW_{th} Noell gasifier (25 bar) which operated on wastes, oils and sewage sludges on the premises of the Schwarze Pumpe power plant in Germany from 1995 until 2007[132,133].

Several biofuel developers attempted to commercialise biomass-fired EF technology in the late 2000’s. CHOREN built a 1 MW_{th} pilot in 1997, then commissioned a 45 MW_{th} input “Beta” plant in Freiburg, Germany, that operated sporadically during 2008-2011, before the company went bankrupt. Linde bought the technology, and although offered to other projects (e.g. Forest BTL), has yet to be taken up. Range Fuels failed during trying to commission their 25 MW_{th} gasifier in 2011. Their Soperton site was bought by Lanzatech, who are now instead installing a staged pyrolysis technology (from Concord Blue). Pearson Technology also built a small 1 MW_{th} pilot in Livingston, with the technology taken over by Rentech (under the ClearFuels brand), before being abandoned.

However, KIT has successful operated a 5 MW_{th} pyrolysis-oil fired EF gasifier from Siemens/Lurgi at their Karlsruhe site since 2013[134], as part of their “bioliq” pilot process to produce methanol and bioDME[135]. Siemens also has another 5 MW_{th} operational testing facility in Freiburg. MHI has also carried out straw tests at 0.4 MW_{th} pilot scale in Japan[136].

Despite a few successful pilot plants using biomass, there have not been any successful demonstrations at scale. Most of the BTL projects planned in Europe using EF gasifiers have been abandoned or placed on hold due to policy uncertainty. Given the high quality syngas (meaning low tar content), the focus to date has been on biofuels applications. The TEABPP team are not aware of any power plants using a wholly biomass-fired EF gasifier in operation globally, with only some sporadic co-firing known to have occurred recently. Given the recent shutdown of the Buggenum plant (which would have qualified as TRL 7), the current technology status is judged to be TRL 6.

The lack of real world, large-scale examples presents a challenge in evaluating the performance and economics of the technology. The range of commercial scales is expected to be c. 100 MW_{th} to 2,000 MW_{th} biomass input. As with major power & heat plants, project timelines of over 2 years can be expected to build a commercial plant.

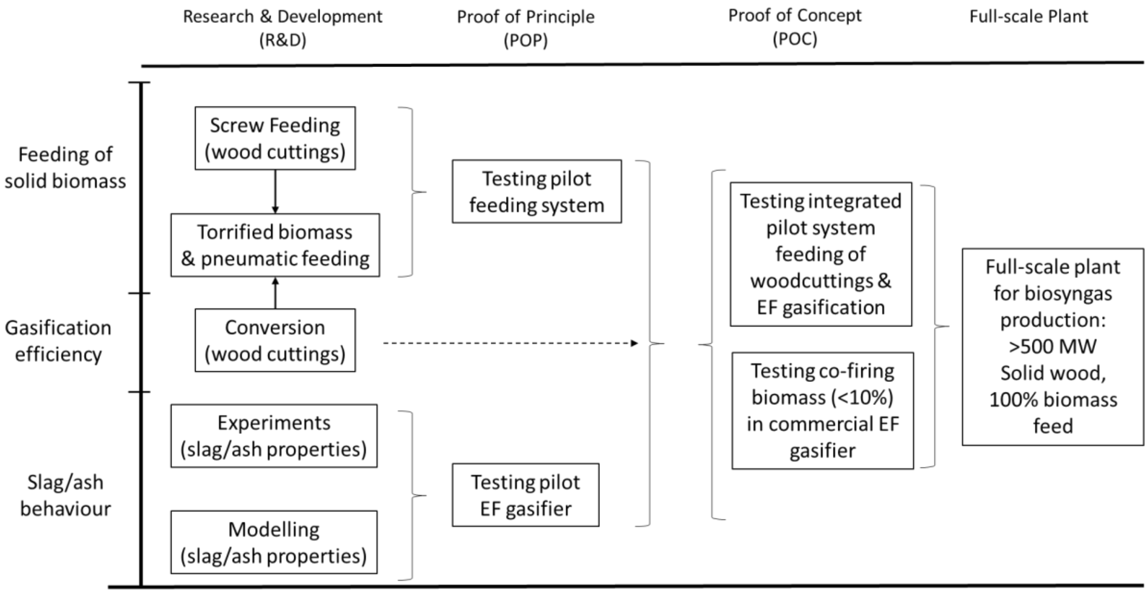

Figure 121: Approach to developing large-scale EF plants using solid biomass[137]

4.6.3 Impact of different feedstock and other parameters on operation and cost

Availability and Opex are for the base feedstocks; high ash and/or high alkali metal content feedstocks would increase downtime and Opex. The former is modelled using the threshold concept and high temperature corrosion is more of a concern with this technology.

Plant size

The plant size primarily affects the Capex. A 0.79 power law is estimated given the mixed nature of the equipment. This gives rise to a capital cost curve as shown in Figure 122. These costs are much lower than other gasifier types, mainly due to the very large scales.

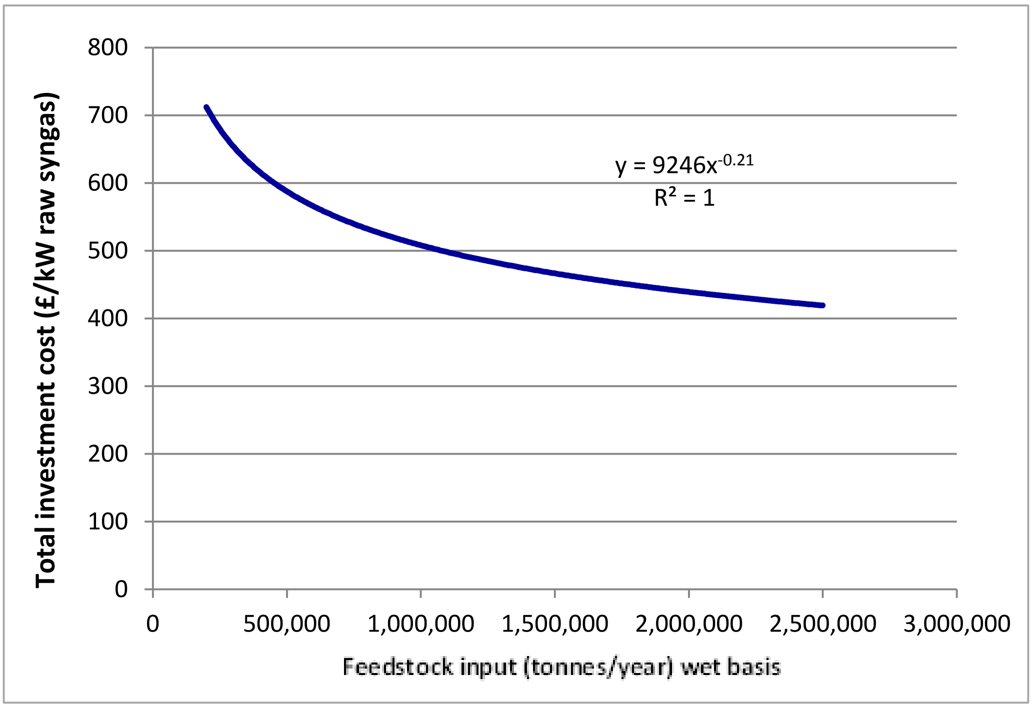

Figure 122: Entrained flow gasifier total investment cost vs plant scale (based data from[74,104])

Feedstock conditions

Moisture content

The moisture content affects the plant biomass to syngas efficiency due to the latent heat of vaporisation of the water in the biomass. The relationship is illustrated in Figure 123. Note that feedstocks greater than 15% moisture content can be fed into the plant, but these will require drying down to 15% before use. This efficiency is the yield of cleaned, polished syngas output divided by the LHV feedstock input – it is not just the cold gas efficiency of the gasifier alone (which is a higher value). Entrained flow gasifiers produce a very low tar syngas; hence syngas clean-up is efficient.

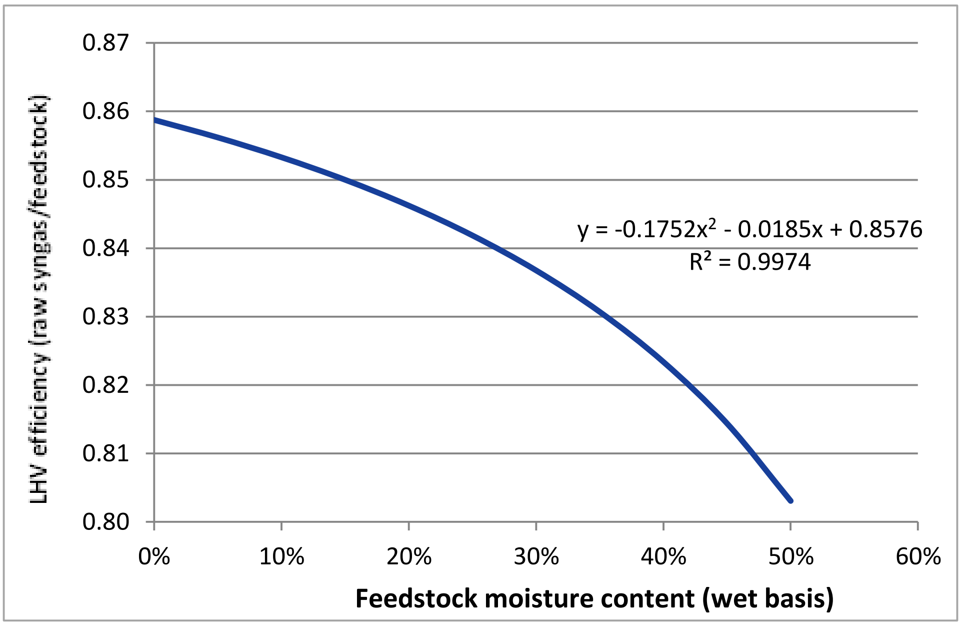

Figure 123: Entrained flow gasifier efficiency vs. feedstock moisture content. Note 15% is an accepted maximum (based on data from[74,104])

Feedstock composition

Biomass EF gasification pilots and demonstrations have focused on using wood (wood chips, forestry residues, sawdust, waste wood, etc.) as the preferred feedstock, although other materials tested include RDF pellets, sorted MSW, sewage sludge, straws and grasses. In general, EF gasifiers can accept a mixture of feedstocks, but under the designed operating conditions, this mixture should not change significantly over time, hence feedstock storage is usually necessary to ensure the supply of quality controlled biomass is achieved.

The biomass received usually undergoes a process of drying, storage, blending and sizing. Entrained flow gasifiers have the ability to gasify practically any fuel, but fuels with lower moisture and ash content are favoured to reduce oxygen consumption. Due to a short EF residence time, large feedstock particles would lead to unconverted biomass, and a high feedstock moisture content would lower gasification efficiency. EF gasifiers therefore have the most stringent feedstock requirements of the gasifier types considered. A typical EF biomass gasifier needs a fuel with less than c. 15% moisture content. EF coal gasifiers need a particle size of 50-100µm, however because biomass is much more reactive than coal, biomass particles can be sized as large as 1mm.

Referring to any type of gasifier, it is quite difficult to meet the stringent syngas quality requirements of a gas turbine (or potential pipeline) without syngas cleaning and conditioning. Due to the high temperatures present within an EF gasifier, hydrogen and carbon monoxide are strongly favoured over methane within the gasification reactions. CO_2 yields are reduced at higher temperatures and tars and hydrocarbons are cracked into smaller components. The result is a high cleanliness syngas that needs very little cleaning for tars.

The effects of temperature on the gas composition including H_2, CO, CO_2 and C_xH_y (light hydrocarbons including C_1 and C_2 species) are presented in Table 20.

Table 20: Effects of temperature on the syngas yield for different types of biomass, derived from[137]

Fuel	T (°C)	H_2	CO	CO_2	C_xH_y
		kg/kg used biomass			
weech wood	1000	0.02	0.415	0.118	0.103
Dp =0.31-0.44 mm	1200	0.043	0.477	0.13	0.037
	1400	0.052	0.614	0.062	0.009
beech wood	1000	0.021	0.415	0.11	0.101
Dp =0.73-0.90 mm	1200	0.041	0.458	0.14	0.038
	1400	0.05	0.596	0.062	0.012
cypress sawdust	600	0.003	0.21	0.046	0.044
Dp <0.50 mm	800	0.01	0.462	0.068	0.113
	900	0.016	0.478	0.071	0.118
	1000	0.021	0.478	0.073	0.078
	1100	0.032	0.486	0.083	0.046
	1200	0.04	0.546	0.066	0.038
	1400	0.047	0.701	0.015	0.002
rice husk	700	0.005	0.209	0.095	0.028
Dp <0.40 mm	800	0.006	0.235	0.105	0.034
	900	0.011	0.303	0.132	0.064
	1000	0.019	0.327	0.133	0.034
walnut sawdust	700	0.007	0.346	0.141	0.046
Dp<0.4 mm	800	0.009	0.439	0.145	0.093
	900	0.017	0.538	0.174	0.068
	1000	0.024	0.569	0.158	0.047
bagasse	800	0.029	0.408	0.217	0.085
Dp =0.50 - 0.86 mm	900	0.031	0.515	0.153	0.058
	1000	0.051	0.717	0.009	0.025
straw	800	0.022	0.245	0.326	0.075
Dp = 0.50-1.00 mm	1000	0.04	0.593	0.101	0.035
olive waste	800	0.007	0.251	0.158	0.114
Dp = 0.50 - 0.80 mm	1000	0.022	0.432	0.129	0.078
birch wood	800	0.011	0.445	0.114	0.137
Dp =0.80 - 1.00mm	1000	0.03	0.571	0.112	0.09

The effect of ash, sulphur, nitrogen and chlorine in the feed influences plant Opex via disposal or mitigation costs; here it is assumed a level of syngas clean-up suitable for pipelines/turbines. To obtain the base Opex, the following assumptions are used:

- Base values of the impurity values as per the assumed feedstock composition base cases
- Insurance was assumed at 1% of TIC
- Other Opex, e.g. labour, fuels, reagents are process specific and determined on a process by process basis

For the base Opex, the O&M, labour, fuels and reagents costs are from[138]. These incremental effects on Opex are illustrated below, showing how different amounts of biomass constituents in the feedstock alter the total plant Opex. Ash handling stands out as the largest cost to mitigate, followed in order by nitrogen, sulphur and chlorine contents.

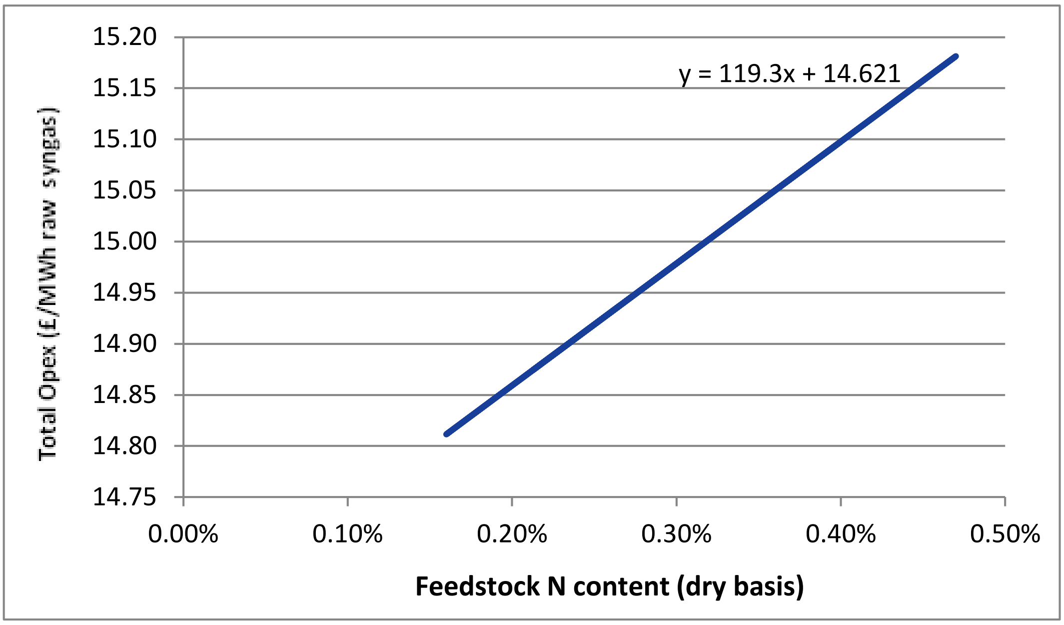

Figure 124: Entrained flow gasifier annual total opex vs. feedstock N content (labour, O&M, fuels and reagents from[138], calculation based on section 2.4.2.4)

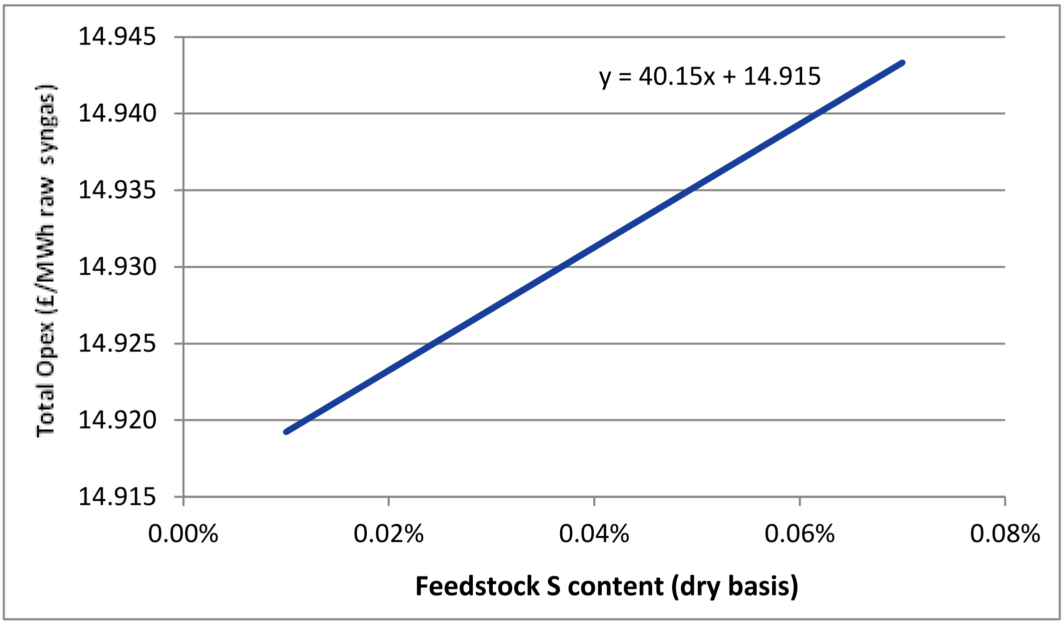

Figure 125: Entrained flow gasifier annual total opex vs. feedstock S content (labour, O&M, fuels and reagents from[138], calculation based on section 2.4.2.4)

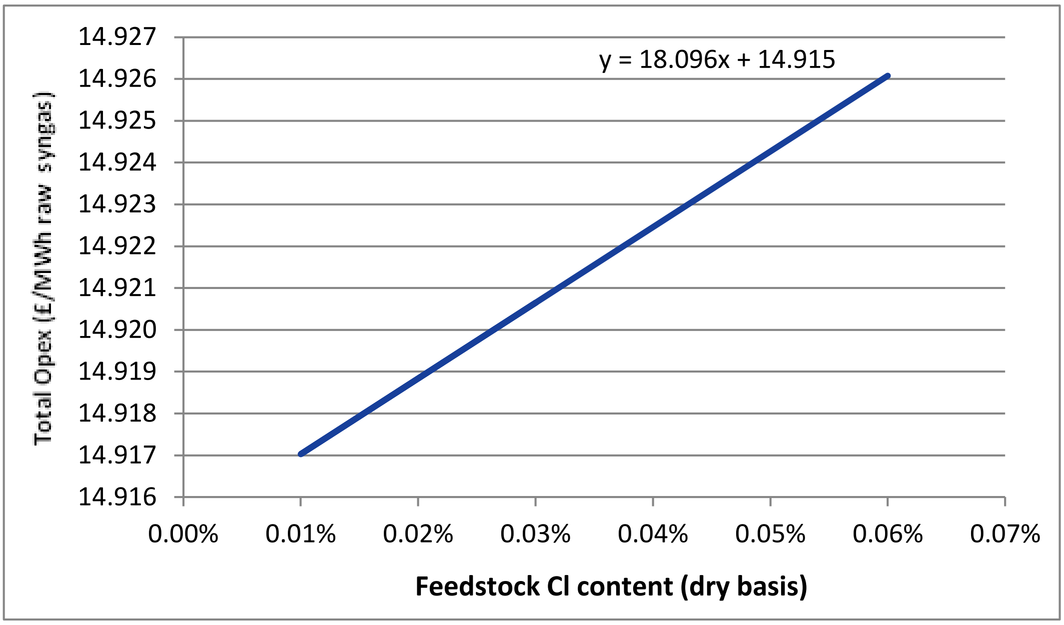

Figure 126: Entrained flow gasifier annual total opex vs. feedstock Cl content (labour, O&M, fuels and reagents from[138], calculation based on section 2.4.2.4)

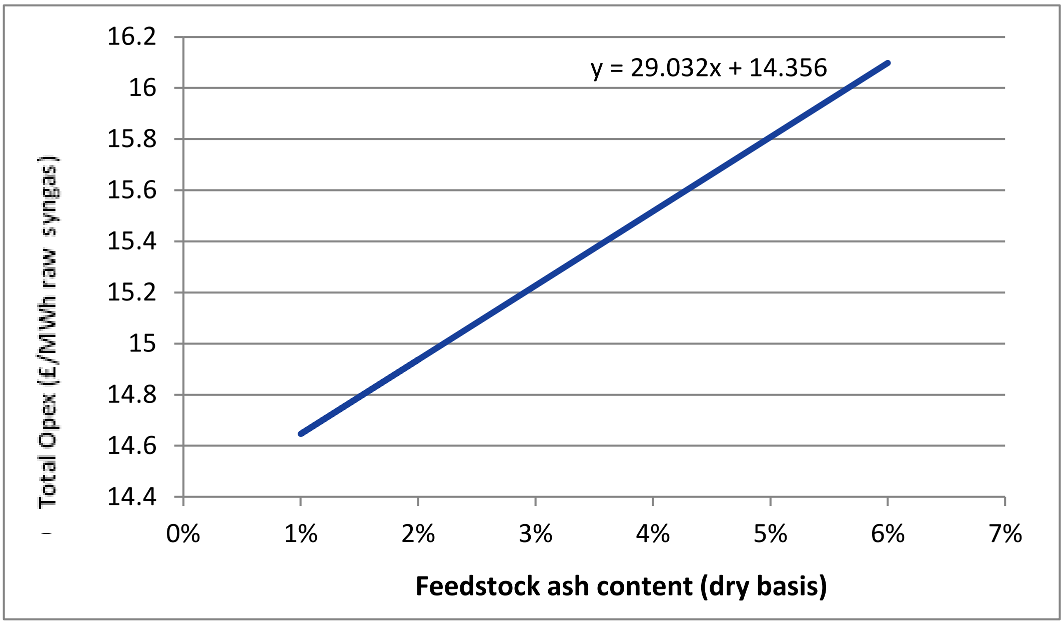

Figure 127: Entrained flow gasifier annual total opex vs. feedstock ash content (labour, O&M, fuels and reagents from[138], calculation based on section 2.4.2.3)

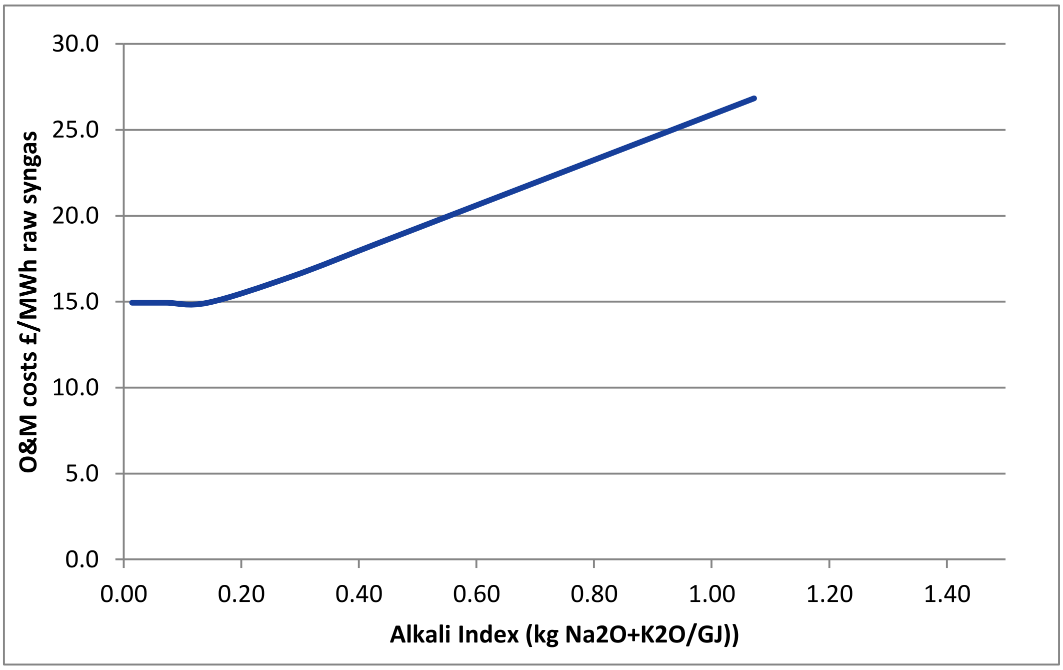

Figure 128: Entrained flow gasifier annual total opex vs. feedstock alkali index (labour, O&M, fuels and reagents from[138], formula based on [139] and section 2.4.2.5)

4.6.4 Available options for improvement

The issues discussed for entrained flow reactors are similar to other gasifiers: tar should be minimised, char gasification maximised, unburned hydrocarbons reduced and fouling minimised. An entrained flow reactor has a very short residence time and there is no potential for a catalytic bed material to be added, hence any catalytic improvements need to be done by feedstock pre-processing (potentially by washing).

Biomass gasification in entrained flow gasifiers is a technology in the process of development, so focus is mainly on the deployment of the actual process and not so much on improving the existing technology. Almost all the entrained gasifiers in the world run on coal today but there still some interest in biomass gasification. Developers of biomass gasifiers can learn from coal gasification, but there are significant differences in fuel properties between coal and biomass. For example, biomass ash has a relatively low melting temperature but is also more aggressive against refractory materials in the gasifier than coal ash. The reactivity of most biomass fuels is higher than most coal types, but is harder to grind biomass to the small particle size needed. One problem is also that more tars are formed during biomass gasification compared with coal gasification.

Torrefaction is one promising method to transform biomass to a material which is more similar to coal and can be used more readily in an entrained flow reactor. This is, essentially, low temperature pyrolysis to drive off a proportion of the volatiles and produce a more friable material (hence one that can be ground smaller for an equivalent energy demand and has a more rapid burnout).

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4.7 Syngas Boiler

4.7.1 Technology description

Syngas boilers are essentially the same as natural gas boilers but with modifications to the burner design to accommodate the different gas characteristics.

There are two main families of design: water-tube boilers and fire-tube boilers. In water-tube boilers, tubes containing water are heated by combustion gases flowing outside the tubes, while in the fire-tube boilers, hot combustion gases flow inside the tubes and water (and steam) flows outside. These two designs are illustrated below.

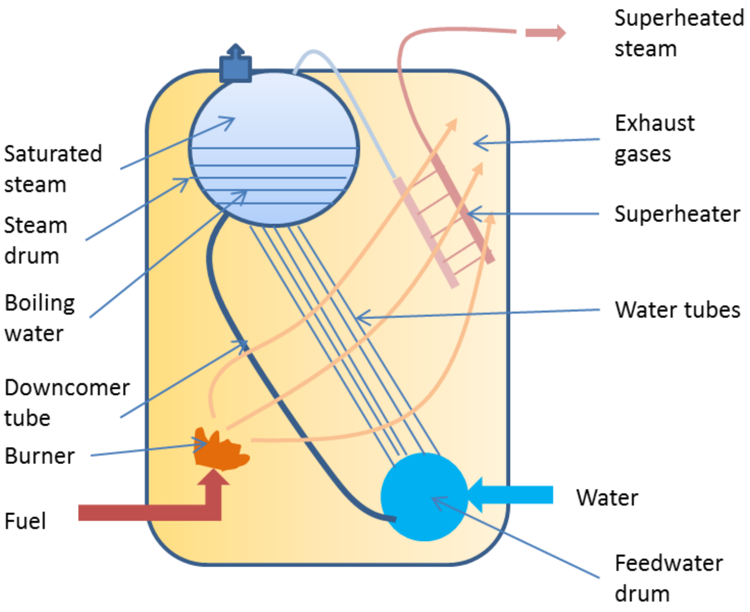

Figure 129: Illustration of a water-tube boiler, derived from[143]

Water-tube boilers use a large number of tubes in the boiler drum. They are normally offer better heat transfer than fire tube boilers. On the other hand they are more sophisticated and expensive units, and due to the narrow tube diameter are more susceptible to fouling problems when low quality water is used. They are suited to more demanding applications (steam flowrate and pressure).

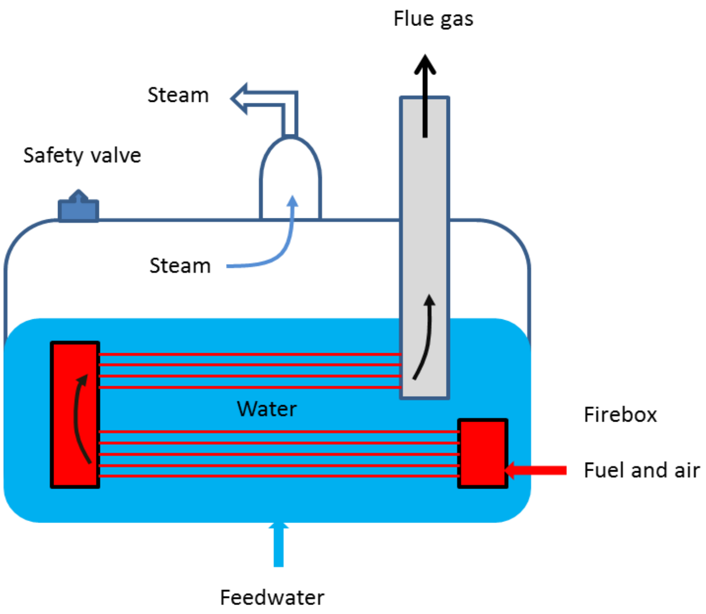

Figure 130: Illustration of a fire-tube boiler, derived from[143]

Fire-tube boilers feed hot gases from the firebox source through tubes. These tubes are positioned inside a water filled drum in order to heat and boil the water. Being of relatively simple design, fire-tube boilers are more straightforward to build and to operate than water-tube boilers. They are normally suited to less demanding and lower pressure applications due to the need for the drum to withstand the steam pressure.

Water tube boilers can operate from 10 GJ/hr to over 250 GJ/hr with pressures beyond supercritical, e.g. 125 bar and above. Fire tube boilers require the boiler to be designed as a pressure vessel itself, so become expensive at higher pressures. Typically, 20 bar would be a sensible upper limit[140].

4.7.2 Development status and timescales

There are a large number of syngas boilers in operation, running off syngas generated from biomass, waste, petcoke and coal for example. These can be seen to be at TRL 9. According to Rezaiyan and Cheremisinoff[141], "combustion of coal-based syngas in boilers is a fully-developed technology". Boilers tend to be tolerant of different syngas composition (and adjustments can be made to burner design if necessary – for example, see Aecometric[142]) and so this also holds true for syngas from other sources.

It is worth noting that the existing fossil fuel boiler fleet is aging - in the industrialised countries, more than 50% of industrial boilers use natural gas as the primary fuel, and about 76% of the total boiler population is older than 30 years. Boiler sales in 2002 amounted to only 1/6th of the total sales in 1967[143].

4.7.3 Impact of different feedstock parameters on operation and cost

The technology is robust to variation in feedstock parameters. Most of the Opex parameters have been dealt with under the "Gasification" sections and will not be repeated here. However, in assessing the various mitigation costs (tar, ash, other constituents) in the gasification to end use technology calculations, the costs are calibrated for high quality syngas (e.g. for turbine or pipeline applications). It is estimated that a reduction of 5% of the annual non-fuel Opex (associated primarily with reduced requirements for wastewater generation and treatment, and less equipment to maintain) is possible against these whole system costs when the application is a boiler. These are taken into account in the whole system LCOE calculations, along with the removal of the syngas clean-up capital costs (as these are not required for close coupled boiler applications, which can generally take tars in the syngas in order to maximise energy production).

The other key sensitivity is the Capex vs scale, which is illustrated below. Syngas boilers, like natural gas boilers, are a very inexpensive conversion technology.

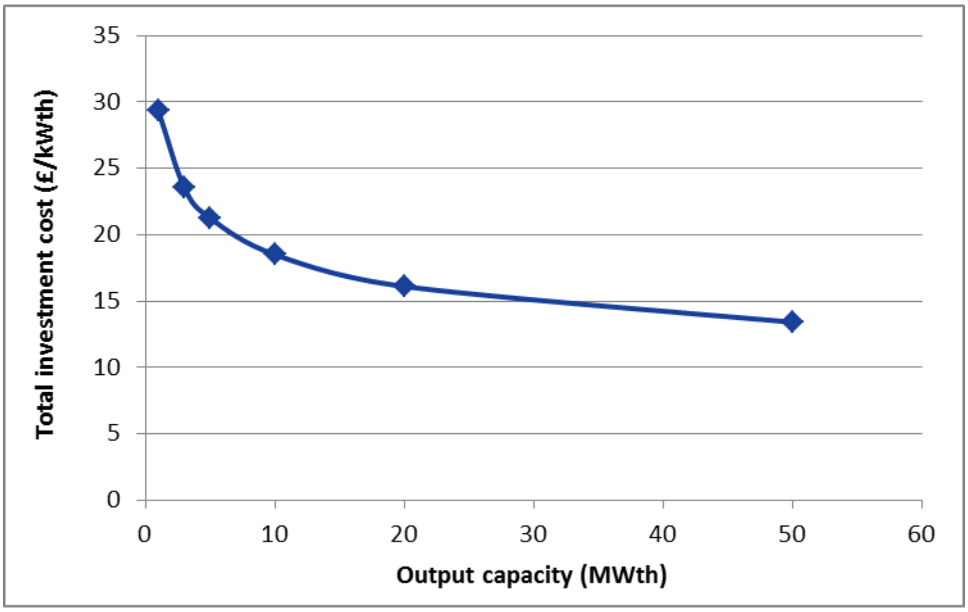

Figure 131: Syngas boiler total investment cost vs. output heat capacity (base cost data from [143] and assuming a 20% uplift on natural gas, and engineering scale factor of 0.8)

4.7.4 Available options for improvement

According to ETSAP[143], the main areas for improvement in industrial boilers are in:

- Efficiency – moving towards a target of 94%
- Construction cost
- Emissions

Each of these involves incremental improvements in burner design, heat exchanger design, fabrication techniques and systems integration, rather than any breakthrough technologies.

Considering the relationship between the original feedstock and syngas in particular, the burner design area has scope for improvement because it will be the case that the gas composition (H_2/CO ratio, CO_2 content, etc.) and associated derived properties (e.g. Wobbe number) will most likely be more variable and feedstock dependent than natural gas or coal based syngas. Furthermore, syngas combustor design (for boiler and turbine applications) is an area of ongoing research. This research will manifest itself in both efficiency improvements and in increased feedstock flexibility.

Regarding materials and corrosion, it is estimated that boiler Capex is a very small proportion of the LCOE (of order 3%[143]). This indicates that the use of more corrosion resistant materials when moving from natural gas to syngas, accounting for the carry-over of potentially aggressive species (e.g. S, N, Cl and alkali metals) which may lead to a doubling or trebling of Capex while reducing the need for pre-treatment may well be worthwhile as the effect on LCOE for the syngas to heat system would only be of the order of 6-9%. Note that post combustion emissions will nevertheless need to be mitigated before release.

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4.8 Syngas engine/CHP

4.8.1 Technology description

Syngas engines are modifications to gas engines. Because of its different composition, and particularly due to the presence of hydrogen which has fast ignition properties, a degree of de-rating and lean burn control is required to avoid pre-ignition and knocking.

The engine can run in power only mode or in CHP mode. For the latter, additional equipment is required for heat recovery, pumping etc. and an estimated 20% additional Capex is required for CHP mode[144]. A typical internal combustion system configuration is shown below.

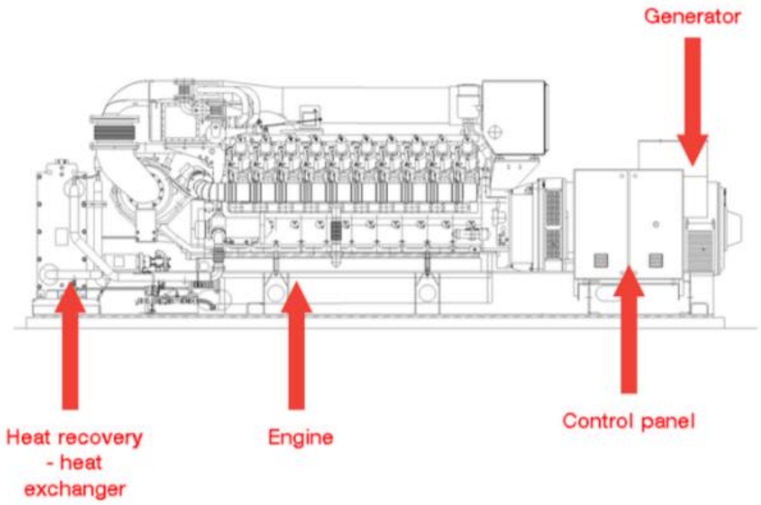

Figure 132: Gas engine layout (Clarke Energy)[145]

The system has four main components:

- **The internal combustion engine**. This is a multi-cylinder engine within which the gas is burnt. At the moment engines are designed for natural gas and modified but in the longer term syngas offers opportunities for optimising the compression ratio[146]
- **The cylinders** cause rotation of the crankshaft – typical RPM is 1000-1500
- **The crank shaft** in turn rotates the alternator which results in the generation of electricity
- **The waste heat** can either be recovered (in the CHP configuration) or dumped via radiators.

Typical efficiencies are 38-48% power and 40-46% heat; giving a total efficiency of around 85-90%.

4.8.2 Development status and timescales

There are a reasonable number of syngas engines in operation, running off syngas generated from a variety of feedstocks. Despite the differences with natural gas, there is increasing use of syngas in engines, with over 30 GE Jenbacher engines in use. Typical capacities per engine are 300kW to 10 MW, with examples of installations described by Obernberger and Thek[147]. Advanced Plasma Power and Broadcrown have also proposed 5 MW_e and 3 MW_e projects for waste syngas feeding gas engines in the UK.

Syngas engines are therefore judged to be at around TRL 8, as deployment is not completely widespread and best practice not fully disseminated. Similarly, larger scale deployment of biomass gasification more widely will allow engine customisation to biomass-derived syngas. Given the smaller applicable scales, more robust operation and lower cost of this technology, compared to others (such as CCGT), further deployment to reach TRL 9 could happen relatively rapidly.

Commercialisation of small-scale and micro-scale biomass-fuelled CHP systems is yet to happen, despite the successful commercial operation of large-scale and medium-scale CHP systems[148]. The application of micro (<15 kW_e) or small-scale (<100 kW_e) biomass-fired CHP systems have significant market potential in both UK and the rest of the world, but the research and development required for these systems is still at a relatively early stage. The relevant technologies currently cannot meet all the demands from different industrial sectors.

4.8.3 Impact of different feedstock parameters on operation and cost

The technology is sensitive to variation in feedstock parameters, in particular tars and alkali metals. The associated Opex parameters have been dealt with under the “Gasification” sections and will not be repeated here.

The other key sensitivity is the Capex vs scale, which is illustrated below. The CHP case has lower Capex, as the output is power + heat, not just power.

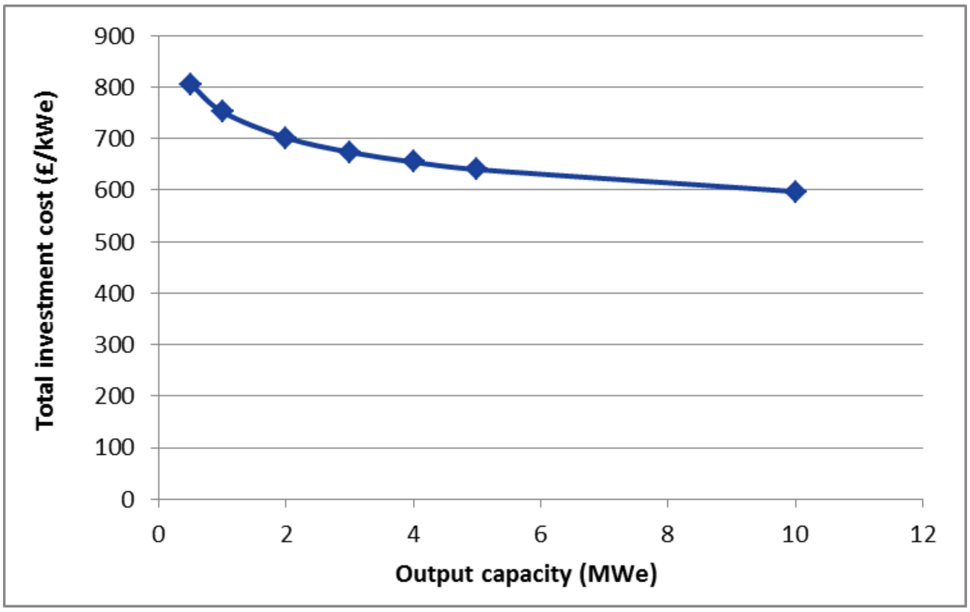

Figure 133: Syngas engine total investment cost vs. MW_e output power capacity – power only case (data from [147])

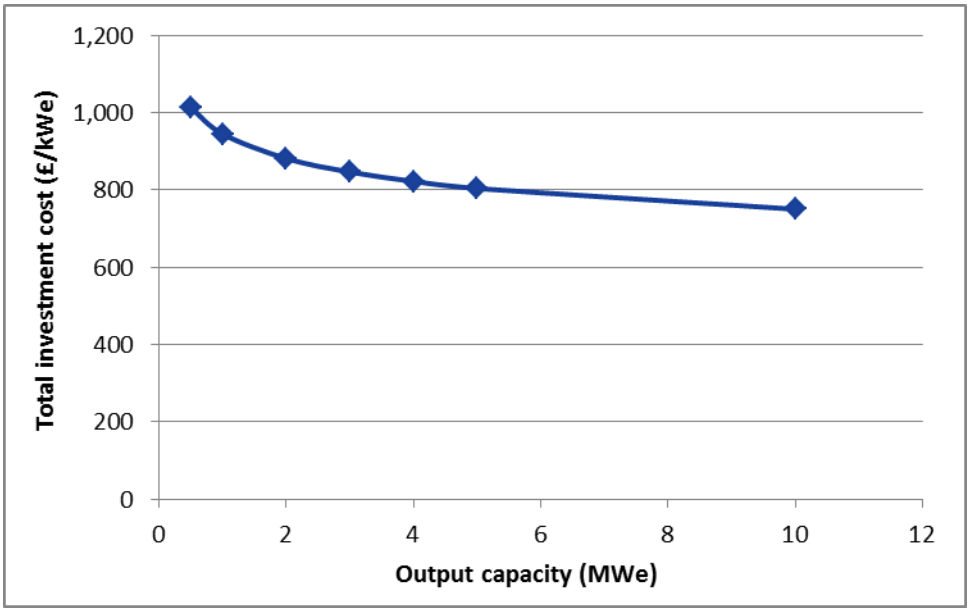

Figure 134: Syngas engine total investment cost vs. MW_e output power capacity – CHP case (data from [147] and cross-checked against quotations)

4.8.4 Available options for improvement

Although the current TRL is around 8, there is still considerable scope for system optimisation and fine-tuning. This includes:

- Dealing with high hydrogen content fuel
- Optimising the physical configuration to maximise the efficiency and robustness based on syngas fuel
- Advanced control to handle varying syngas composition (especially the H_2:CO ratio) given that the original biomass feedstock will vary

The last point is probably the most important in relation to the TEABPP because of the inevitability of needing to deal with varying syngas composition arising from different types of biomass and even different batches of the same biomass. Unlike the syngas boilers, the Capex is a more significant element of the cost. As the technology depends on production against defined standards (not quite mass production but along those lines) materials of construction will be unlikely to differ from those of natural gas engines. As a result, issues relating to trace impurities will have to be dealt with by upstream technologies.

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4.9 Syngas CCGT

4.9.1 Technology description

The combined cycle gas turbine (CCGT) is a high efficiency (>55%) power generation technology which combines a gas turbine (Brayton cycle) with a steam turbine (Rankine cycle). A high level schematic is shown in Figure 135 below.

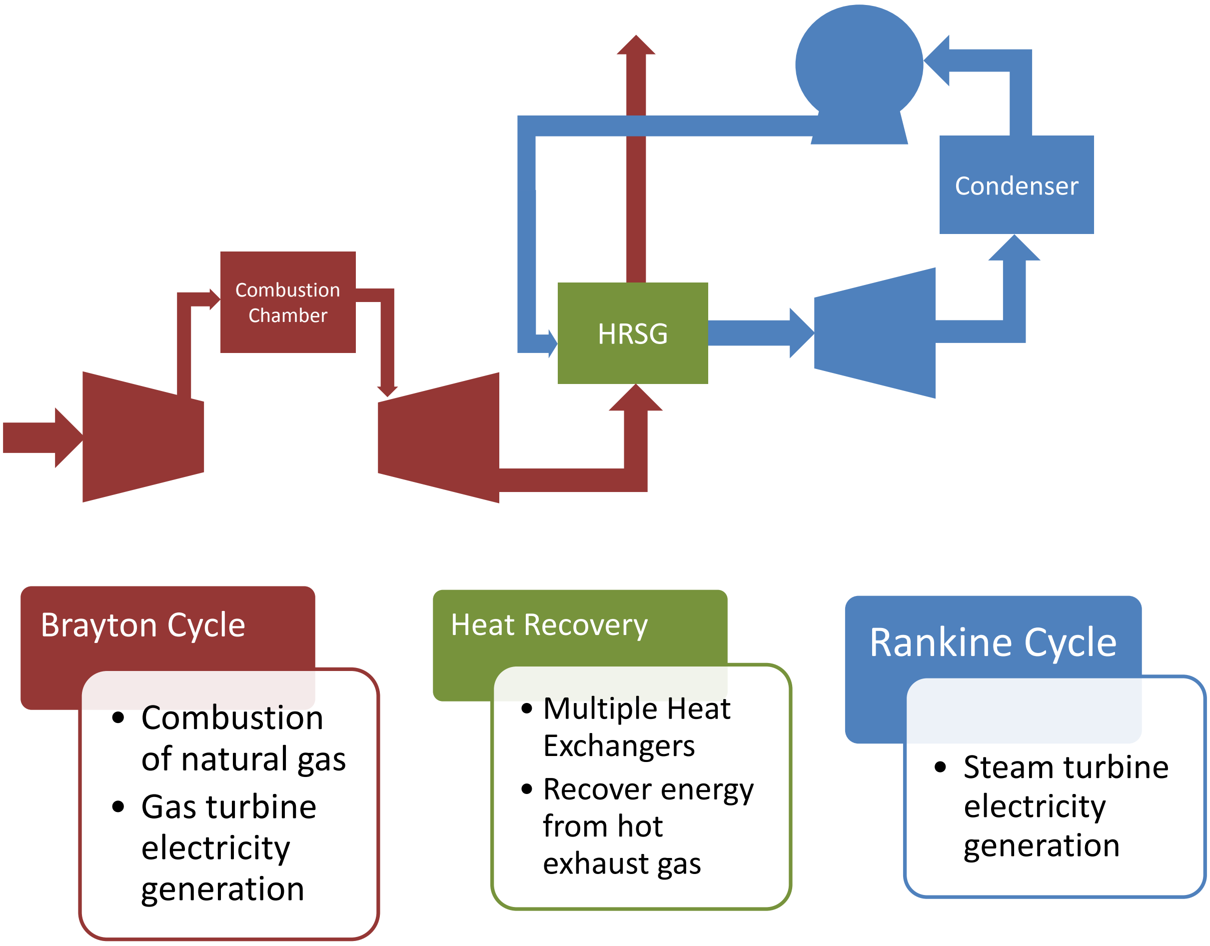

Figure 135: High level schematic of CCGT plant

The gas turbine section includes the compressor, combustor, turbine and power generator. The gas turbine is usually made of high performance metals as the turbine inlet temperature is high (up to 1400°C). However the mass of metal is small relative to the amount of power produced. The flue gas exits at around 500-700°C and is used to raise steam in the HRSG (heat recovery steam generator), at temperatures of 450-600°C. The Rankine cycle uses a steam turbine; following expansion through the turbine the steam is condensed and cooled via heat exchange with a water body or by evaporative cooling.

The HRSG is a sophisticated piece of equipment and serves to increase the overall efficiency by maximising the amount of useful steam that can be raised. A typical configuration is depicted below.

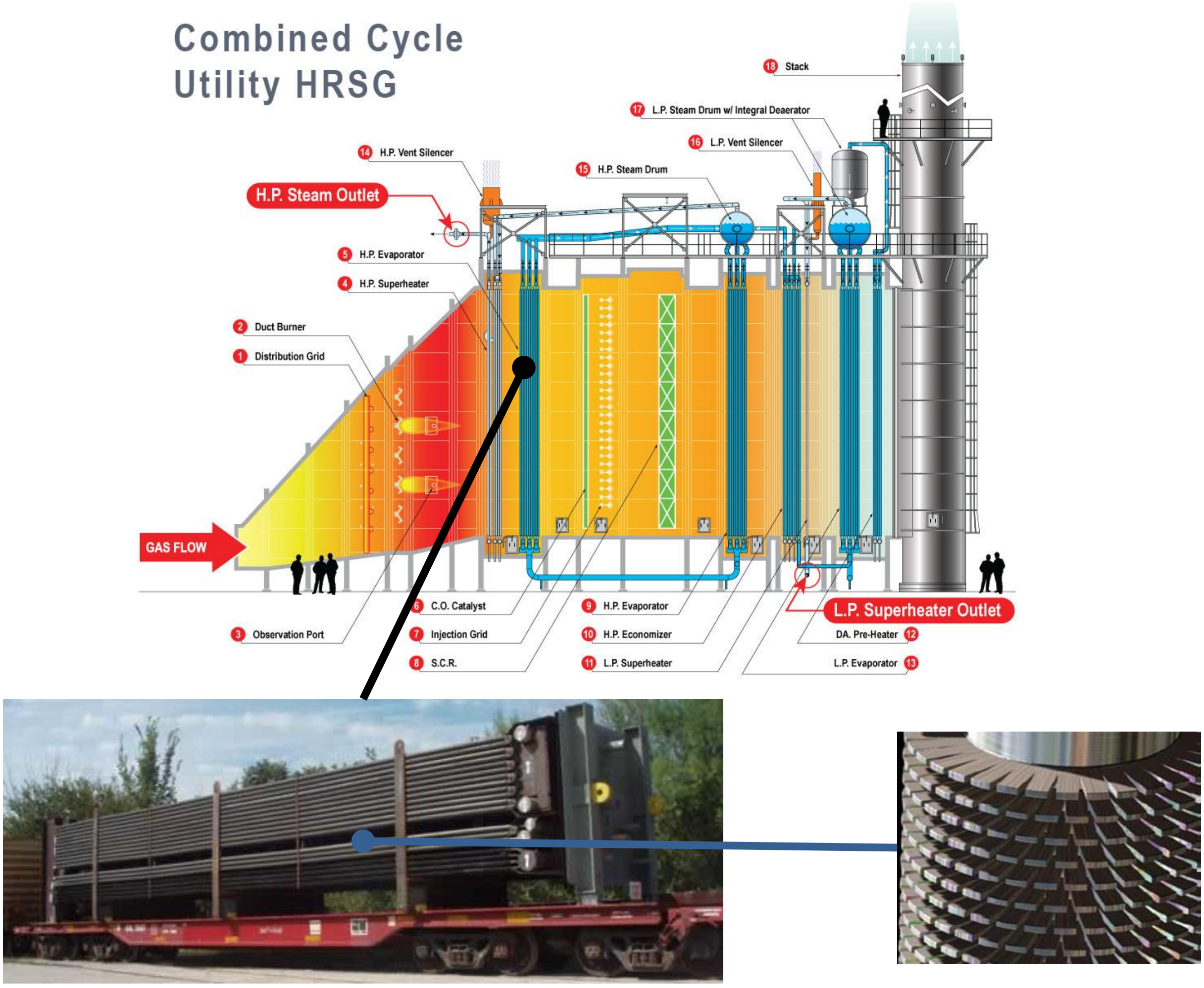

Figure 136: HRSG of a CCGT power plant[149]

The vast majority of CCGT plants use natural gas. However, there is growing experience of using syngas in such plants, not least from integrated gasification combined cycle (IGCC) processes. Modifications may need to be made regarding the combustor design, gas turbine and HRSG (see next section). Regarding power production from plant, an example of the IGCC plant process are depicted in Figure 137.

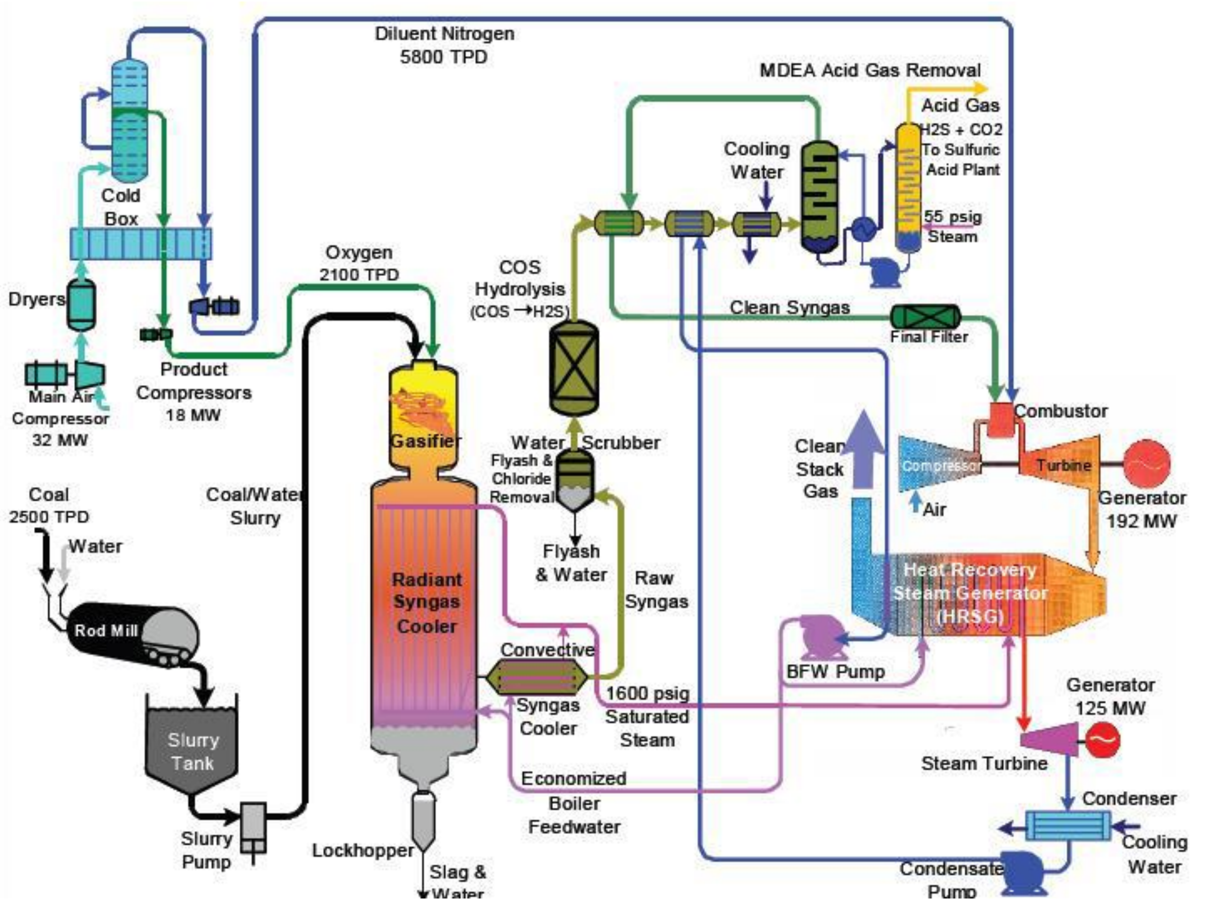

Figure 137: IGGC plant configuration[150]

4.9.2 Development status and timescales

There are a considerable number of syngas CCGTs in operation. The area that has received the most attention is the design and operation of the syngas turbine.

By 2003, GE had 17 relevant installations in operation[151], running off syngas generated from a variety of feedstocks – although mostly coal, oil and other fossil sources, not biomass. The turbine availability has been over 90% in these trials. Syngas composition is dependent on the upstream process, the oxidant and the biomass itself. The variations in flammability limits and calorific value require design modifications to the combustor and turbine. Efficiencies comparable to those of gas turbines should be achievable, and more deployment at larger scales will allow more turbine customisation. Developments are already ongoing in combustion, turbine design, materials and coatings (e.g. thermal barrier coatings).

As seen in Table 21, the experience with biomass-derived syngas is much less extensive than fossil syngas. Although there have been demonstration biomass IGCC plants built in the 1990s and early 2000s, these were smaller scale units, with several unsuccessful due to problems with the gasifier and syngas clean-up steps. On the other hand, progress in R&D on combustion and turbines is ongoing with considerable focus on details (see Figure 138).

The Värnamo Bioflow 6 MW_e plant ran on wood from 1993 to 2000, but the UK's ARBRE project failed in commissioning their hot tar clean-up system, and the Bioelettrica project and TPS'

Aerimpianti plant in Italy both failed due to slag accumulation[152]. There are no IGCC plants currently known to be running using any biomass, and the potential waste-fed Royal Dahlman/ECN project in Grimsby would only be 7 MW_e. The 253 MW_e coal IGCC plant at Buggenum, the Netherlands had recently been co-firing at up to 70% biomass, but shut down in 2013.

Table 21: GE IGCC experience (to 2003)[154]; the "Power block" section indicates the particular GE turbine technology used

Customer	Location	COD	MW	Power Block	Application	Integration	Gasifier	Fuel
Cool Water IGCC	Barslow CA	1984	120	107E	Power	Steam	Texaco	Coal
PSI Wabash River	Terre Haute IN	1996	262	7FA	Power	Steam	Destec	Coal
Tampa Electric	Polk FL	1996	250	107FA	Power	Steam/N_2	Texaco	Coal
Pinon Pine	Sparks NV	1996	100	106FA	Power	Steam/Air	KRW	Coal
Texaco El Dorado	El Dorado KS	1996	40	6B	Cogen	Steam/Air/N_2	Texaco	Pet Coke
Ilva ISE	Taranto, Italy	1996	520	3x109E	Cogen	none	Stell Mill	COG
SUV Vresova	Vresova, Czech Rep.	1996	350	209E	Cogen	Steam	ZVU	Coal
SVZ	Schware Pumpe, Germany	1996	40	6B	Cogen/MeOH	Steam	GSP	Coal/Waste
Shell Pernis	Pernis, Netherlands	1997	120	206NB	Cogen/H_2	Steam	Shell/Lurgi	Oil
Fife Energy	Fife, Scotland	1999	109	106FA	Power	none	Lurgi	Coal/Waste
Motiva	Delaware City, DE	1999	180	2-6FA	Cogen	Steam/N_2	Texaco	Pet Coke
Sarlux	Sarroch, Italy	2000	550	3x109E	Cogen	Steam	Texaco	Oil
Fife Electric	Fife, Scotland	2000	350	109FA	Power	none	Lurgi	Coal/Waste
Exxon Singapore	Jurong Island, Singapore	2000	173	2-6FA	Cogen	none	Texaco	Oil
IBIL Sanghi	Gujarat, India	2001	53	106B	Cogen	Steam/Air	Carbona	Coal
Boilelectrica TEF	Cascina, Italy	2001	12	1-PGT10B/1	Power	Steam	Lurgi	Wood/Waste
EDF-Total	Gardanne, France	2003	400	2x9E	Cogen/H_2	Steam	Texaco	Oil

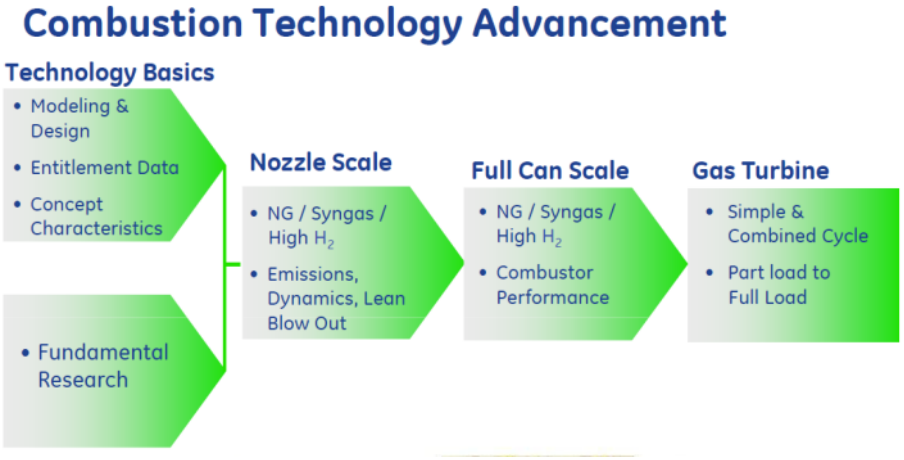

Figure 138: GE combustion development roadmap[153]

Siemens report similar experience, using their heavy duty turbines – again mostly on fossil fuel derived syngas and steel mill gases[154]. A list of projects is in Table . Siemens note that only minor modifications are required to move to combustion of biomass derived syngas, with new developments in catalytic combustion. Whilst experience with biomass derived syngas is currently limited (with Buggenum being the only large-scale demonstration of co-fired IGCC), the deployment of fossil syngas currently places syngas CCGT technology at TRL 8. A TRL of 9 is therefore feasible within 5-10 years, although this will depend heavily on the industrial demand for IGCC plant – and any specific adaptation for biomass will depend on biomass and co-fired IGCC gaining considerably more interest than present today.

Table 22: List of Siemens turbine projects[153]

Gas Turbine	Plant Location	Main Features	Fuel	Start-up year
SGT-200	Many locations		80-85% H_2	
SGT-500/600	Many locations		20-90% H_2	
VM5	Dortmund, Germany	Compressor Drive GT	Blast-furnace Gas	1960
VM5	Handan, China	Compressor Drive GT	Blast-furnace Gas	2000
CW201	Chicago, USA		Blast-furnace Gas	1960
V93	Luenem, Germany	First CC plant in the world with integrated goal gasification	Syngas	1987
2XSGT6-3000E	Plaquemine, USA	CC plant with integrated DOW coal gasification	Syngas	1972
2XSGT6-3000E	Sweeney Cogeneration L.P., USA	CC plant with integrated DOW coal gasification	0-30% H_2	1998
SGT5-2000E	Buggernum	CC plant integrated with coal gasification (hard coal and biomass blend)	Syngas	1994/5
V94.3	Puertollano, Spain	CC plant integrated PRENFLO coal gasification (coal and petroleum coke blend) s	Syngas	1997/8
2XGT5-2000E	Priolo Gargallo, Italy	CC plant with integrated GE heavy-oil (asphalt) gasification	Syngas	1998/9
SGT5-2000E	Servola, Italy	CC plant with steel-making recovery gas	Steel-making recovery gas	2000
SGT5-2000E	Sannazzarro, Italy	CC plant with integration SHELL heavy-oil gasification	Syngas	2005

The HRSG, although complicated, is not expected to require major modifications, although designs must take account of different flue gas flowrates and temperatures in syngas CCGTs. NEM designed the three HRSG systems for a 1300 MW_e natural gas CCGT plant in Eemshaven, Netherlands (operational from 2013), with the capability and provision for the plant switching over to coal/biomass IGCC + CCS (and hence they would therefore also work with syngas turbine flue gas). However, Nuon has abandoned these conversion plans[155].

4.9.3 Impact of different feedstock parameters on operation and cost

The technology is sensitive to variation in the syngas feedstock parameters, in particular tars, S, Cl and alkali metals. The associated Opex parameters associated with the mitigation of these have been dealt with under the “Gasification” sections and will not be repeated here.

The Siemens and GE trials discussed above indicate good flexibility with respect to syngas composition.

The other key sensitivity is the plant Capex vs scale, which is illustrated below.

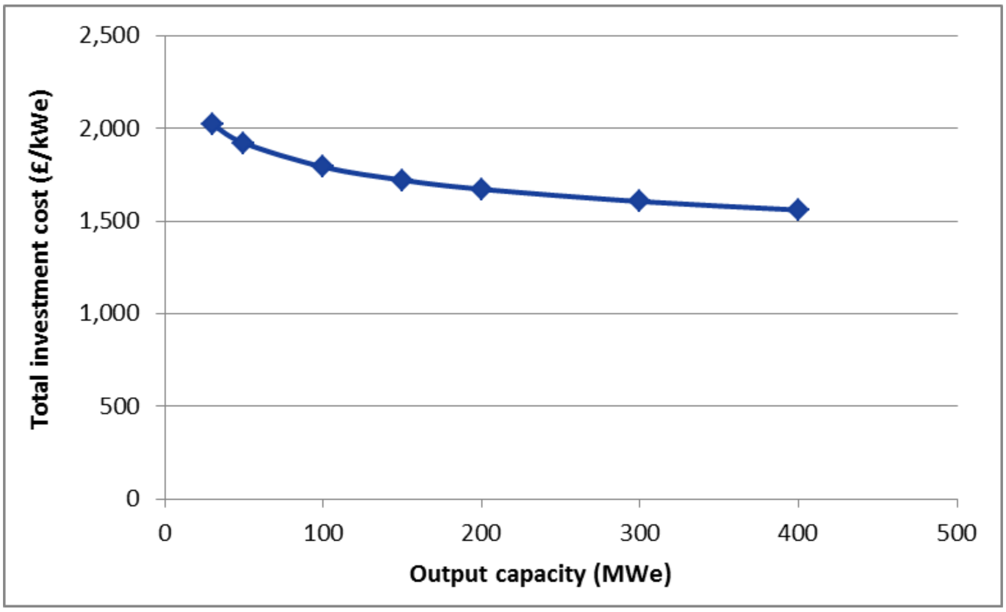

Figure 139: Syngas CCGT total investment cost vs. MW_e output power capacity – power only case (data from [156] and DECC scenarios)

4.9.4 Available options for improvement

Despite the current TRL of 8, there is still considerable scope for system optimisation and fine-tuning because the systems are less mature than those based on natural gas. For the use of syngas, this includes:

- Optimising combustor design, including catalytic combustion and NOx abatement
- Optimising turbine design (including cooling) and materials of construction (including coatings which reduce thermal conductivity)
- Dealing with faster flame speeds and wider combustion limits

Another important challenge, mentioned in the boiler section as well, and relating back to the varying nature of feedstock is the varying nature of the syngas, both in terms of combustion details and calorific value. This requires sophistication in combustion section design, to enable a broad range of inputs and potentially reduce processing of the syngas before the CCGT e.g. shift reactions to balance H_2/CO ratios.

In terms of impurities, the turbines are high specification equipment and becoming increasingly sophisticated as OEMs target higher and higher efficiencies. It will continue to be the case that clean-up of the incoming gas will be required and that stringent quality thresholds on impurities will remain in place. This implies post- and possibly pre-gasification processing will be always be needed.

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[150] NETL, Power, Typical IGCC Configuration http://www.netl.doe.gov/research/coal/energy-systems/gasification/gasifipedia/igcc-config

[151] Brdar and Jones "GE IGCC Technology and Experience with Advanced Gas Turbines". Schenectady, NY, Doc GER-4207.

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[155] NEM "The Magnum Project". Available at: http://www.nem-group.com/EN/projects/hrsgs/2/magnum_-_the_netherlands/18/

[156] US EPA, "Biomass Combined Heat and Power Catalog of Technologies", 2007, www.epa.gov/chp/documents/biomass_chp_catalog.pdf

5 Pre-processing Technologies

5.1 Water washing & Chemical washing

5.1.1 Technology description

Water washing

Biomass is increasingly being used to produce electricity and heat. However, it may contain various components that detract from its use as a fuel. Therefore, considerable interest has arisen in pre-treatment techniques to improve fuel quality. Potassium and sodium, along with sulphur and chlorine, influence ash chemistry and the behaviour of the fuel in terms of its tendency to corrode equipment and cause slagging and fouling. The main causes of these undesirable slagging effects are attributed to the reactions of alkali metals with silica to form alkali silicates, which melt or soften at temperatures as low as 700°C, and the reactions of alkali metals with sulphur, which form alkali sulphates on heat transfer and combustor surfaces. When high-alkali biomass is used in high-temperature furnaces and boilers, additives are required to help ameliorate the associated problems, sometimes with considerable additional expense. The presence of chlorine and sulphur in the biomass can also increase the corrosion potential on the boiler walls. Demineralization of the fuel, by water washing, is a simple and effective method for removing a large fraction of the fuel mineral content, thereby reducing the above-mentioned problems[157].

Main process steps

Biomass enters the plant with a wide size range and moisture content, via delivery trucks directly into a ground hopper which has a belt conveyor in its base to a rotary screen to remove stone and dirt. If required, chipping is then used to reduce the size of the biomass to 20-40 mm chips. The chips are conveyed into a magnetic drum to remove metal impurities from the fuel[158]. A belt conveyer is then used to convey the chips to the washing machine, which mixes the biomass with water at room temperature in a certain mixing ratio, typically 5:1.

The produced chips have a maximum moisture content of 50%. Depending on the conversion technology, after filtering the washed material, a further drying step can be added to the process to decrease the moisture content to the required limits.

The main washing process steps are:

- Screening to remove stone impurities
- Chipping to a size of 20 – 40 mm (if required)
- Magnetic drum screening
- Water washing
- Chemical washing (if required, uses similar vessels to water washing)
- Filtering
- Drying (if required)

Chemical washing

The process of chemical washing is not a mature technology. Currently, there are no known plants using this technology on an industrial scale. The only attempts found are the research experiments on treating biomass with three steps of washing starting with water, then with 1.0 molar solution of ammonium acetate, then with 1.0 molar solution of hydrochloric acid. The impact of the second and third chemical treatment steps is to further remove the residue of alkali and earth alkaline metals up to 90-100%, in addition to further 30% removal of Al, 60% of Fe, 20% of S, and 20% of Si[159]. As an overall process, the treated biomass should finally be rinsed to remove the chemical traces and pressed or filtered by gravity in a filter belt to remove excess water.

The impact of leaching biomass with chemicals on the pyrolysis and combustion behaviour of the fuel has been also studied. Figure illustrates the effects observed, with the highest benefits shown when treating high alkali content feedstocks such as straw. Improvements of the yield of pyrolysis oil is also possible, increasing from ~42 % to ~55 % as a function of HCl concentration from zero to 7 %[160].

Benefits of washing technologies

It has been proven that water washing can remove 10 - 90% of the ash content, 45 - 75% of the alkali metal content, 5 - 50% of Ca, Mg and Mn content, and 5 - 10% heavy metals, for various biomass species depending on the initial elemental content, washing method and time of mixing (Saddawi, 2012; Gudka 2015). However, washing at higher temperatures (still lower than water boiling temperature) such as 60 - 90°C can increase the removal efficiency up to 100% for many mineral elements. A list of experimental analysis of removal % for various biomass species in different washing methods is shown in Table .

Table 23: Elemental Removal of water washing testing on different biomass types[161]

Fuel	Species	Solvent	Temp, Time	Size	% removed
Range of urban wastes, agricultural residues, wood fuels, grasses and straws	Si, Ti, Al	water	Hot plate, not boiling, 16 h, soaked	<60 mesh	Very low or not soluble
Range of urban wastes, agricultural residues, wood fuels, grasses and straws	K, Na, Ca, Mg	water	Hot plate, not boiling, 16 h, soaked	<60 mesh	50-90% K, 10-90% Na, 5-35% Ca, 5-60% Mg
Range of urban wastes, agricultural residues, wood fuels, grasses and straws	Cl, S, P	water	Hot plate, not boiling, 16 h, soaked	<60 mesh	30-100% Cl, 0-95% S, 10-80% P
Range of urban wastes, agricultural residues, wood fuels, grasses and straws	Fe	water	Hot plate, not boiling, 16 h, soaked	<60 mesh	0-30% Fe
SRC Willow	Na, K, Ca, Mg, P, S, Cl	Water	Room temp, 20h stirred	2-4 cm chip	30% Na, 46% K, 14% Mg, 3% Ca, 56% PO_4^{3-}, 10% SO_4^{2-}, up to 100% Cl
Eucalyptus	Na, K, Ca, Mg, P, S, Cl	Water	Room temp, 20h stirred	2-4 cm chip	60% Na, 45% K, 18% Mg, 4% Ca, 27% PO_4^{3-}, 8% SO_4^{2-}, up to 100% Cl
Miscanthus	Na, K, Ca, Mg, P, S, Cl	Water	Room temp, 20h stirred	1-3 cm chopped	53% Na, 62% K, 56% Mg, 19% Ca, 49% PO_4^{3-}, 33% SO_4^{2-}, 84% Cl
Wheat straw	Na, K, Ca, Mg, P, S, Cl	Water	Room temp, 20h stirred	1-3 cm chopped	92% Na, 54% K, 32% Mg, 12% Ca,0.2% PO_4^{3-}, 21% SO_4^{2-}, 100% Cl
Peach Stones	K, Na, Cl, Mg, Fe	Water	20°C, 8 h, soaked	?	70% Cl, 30% K, 40% Na, 50% Ca, 70% Mg,30% Fe
Wood, forest residue, bark, waste wood	Si, Al, Fe, Ca, Mg, Na, K, P, S	Water	Room temp, soaked	< 5 mm	<10% Si, 4-15% Al, ~5% Fe, 5-15% Ca, 15-20% Mg, 35-45% Na, 50-65% K, 25-50% P, 10-30% S.
Straw	Si, Al, Fe, Ca, Mg, Na, K, P, S	Water	Room temp, soaked	< 5 mm	~5% Si, ~5% Al, ~5% Fe, ~25%Ca, ~40% Mg, ~60%Na, ~80%K, ~65%P,~60% S
Spruce, Pine, Birch, Aspen (stem, bark, twigs, branch woods)	K, Na, Mg, Mn, Ca, Al, Fe, Cl, P, S, Si	Water	Room temp. 24 h stirred	< 1 mm	75% K, 65% Na, 40% Mg, 25% Mn (bark, twigs, foliage all species), 25% Ca (stem wood), 10% Al (spruce, birch, aspen), 5% Fe, 90% Cl, 65% P, 15% S, 5% Si
Fuel	**Species**	**Solvent**	**Temp, Time**	**Size**	**% removed**

Wheat straw	Al, Si, Ti, Fe, Ca, Mg, Na, K,, S, P, Cl	Water	30, 60, 90 °C, 3h	280-450µm	Ash removal increases with temperature of the water (from 55-75% reduction). SiO2 removal efficiency increases with water temperature (25-65%). A small benefit for K removal with T, but >90% removed at all temps. S 85-95% removed, Cl 70-85% removed
Rice straw	Al, Si, Ti, Fe, Ca, Mg, Na, K,, S, P, Cl	Water	30, 60, 90 °C 3h	280-450µm	Ash removal increases with temperature of the water (from 20-40% reduction). SiO2 removal efficiency increases with water temperature (2-30%). A small benefit for K removal with T, but >86% removed at all temps. S 90-95% removed, Cl >90% removed at all temps.
Corn stalk	Al, Si, Ti, Fe, Ca, Mg, Na, K,, S, P, Cl	Water	30, 60, 90 °C 3h	280-450µm	Ash removal increases with temperature of the water (from 60-70% reduction). SiO2 removal efficiency increases with water temperature (65-75%). A benefit is seen for K removal with T, but >85% removed at all temps. Small benefit of water T for S: >90% removed, Cl 35-58% removed and a negative effect of T seen.
Cotton stalk	Al, Si, Ti, Fe, Ca, Mg, Na, K,, S, P, Cl	Water	30, 60, 90 °C 3h	280-450µm	Ash removal increases with temperature of the water (from 20-40% reduction). SiO2 removal efficiency increases with water temperature (2-30%). A small benefit for K removal with T, but >86% removed at all temps. S 90-95% removed, Cl also showed a benefit for hot water 85-95% removed.
Candlenut wood	Al, Si, Ti, Fe, Ca, Mg, Na, K,, S, P, Cl	Water	30, 60, 90 °C 3h	280-450µm	Ash removal increases with temperature of the water (from 10-30% reduction). SiO2 removal efficiency increases with water temperature (10-30%). A large benefit for K removal with T from 70-85% removal. S 75% removed at all temps. Small benefit from hot water for Cl 50-55% removed.
Rice Hull	Al, Si, Ti, Fe, Ca, Mg, Na, K,, S, P, Cl	Water	30, 60, 90 °C 3h	280-450µm	Ash removal slightly increases with temperature of the water (from 5-10% reduction). SiO2 removal efficiency slightly increases with water temperature (5-10%). A benefit is seen for K removal with T, from 85% to >90%. S 60-80% removed, Cl 68-85% removed, 60°C was best.
Rice straw,	Al, Si, Ti, Fe, Ca, Mg, Na, K,, S, P, Cl	Water	20-25°C spray, pouring, soaking 24h	< 90 mm	Spraying: All species except Si, Ti, Ca, Mg decreased significantly: Cl 50% reduction, Na 20% reduction, K and P 40% reductions, S 10% reduction. Soaking: All species except Si, Ti, Ca decreased significantly: Cl 90%, Mg 55%, Na 90%, K 80%, P 80%, S 70% Rain washed: All species decrease except Si, Ti, Ca. Similar results to soaking observed.
wheat straw	Al, Si, Ti, Fe, Ca, Mg, Na, K,, S, P, Cl	Water	20-25°C spray, pouring, soaking 24h	< 90 mm	Soaking results reported. Greater than 80% removal of Cl, Na, K and S. 37% and 53% removal of Ca and Mg respectively,
Eucalyptus Loxophleba (mallee)	Na, K, Mg, Ca, Cl	Water	Room Temp. Sampled with time	150-250 µm	Na, K, 80-95% removal; Cl 100% removal; Mg, Ca 10-30% removal. Differences seen between batch and semi-continuous operation.
Type C Red Pine	As, Cr, Cu	0.19N H_2SO_4	75 °C, soaked	<12 mm	1st leaching step: 71±19% As, 69±21% Cr, 77±15% Cu 2nd leaching step: 92±+6% As, 87±9% Cr, 94±5% Cu 3rd leaching step: 98±+1% As, 94±4% Cr, 99±1% Cu

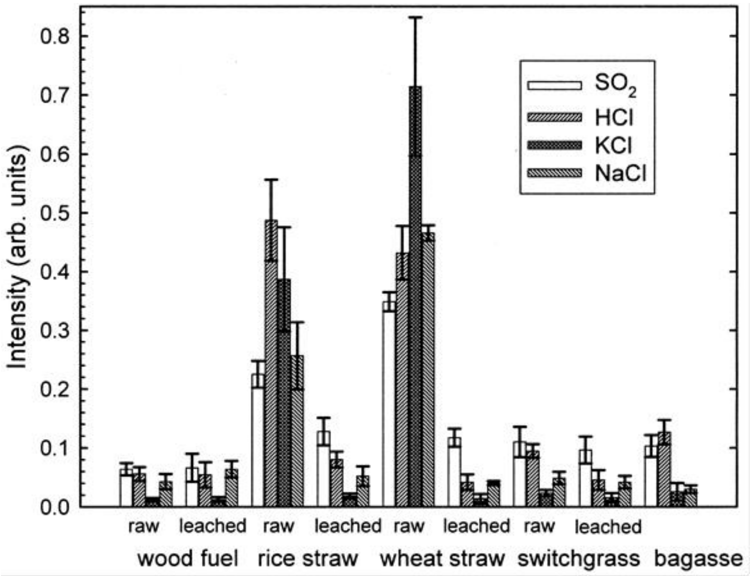

Figure 140: Relative amounts of SO_2(g), HCl(g), KCl(g) and NaCl(g) detected during the combustion of untreated and treated biomass, with various treatment methods, at 1100°C (data from [162])

In terms of the benefits of water washing, Figure 141 and Figure 142 indicate the impact that a percentage removal of alkali content can have on down-stream combustion. A removal of c. 55% of potassium and sodium will be sufficient to place many feedstocks at only a low risk of causing fouling, whereas as the untreated biomass would have been much more likely to cause issues.

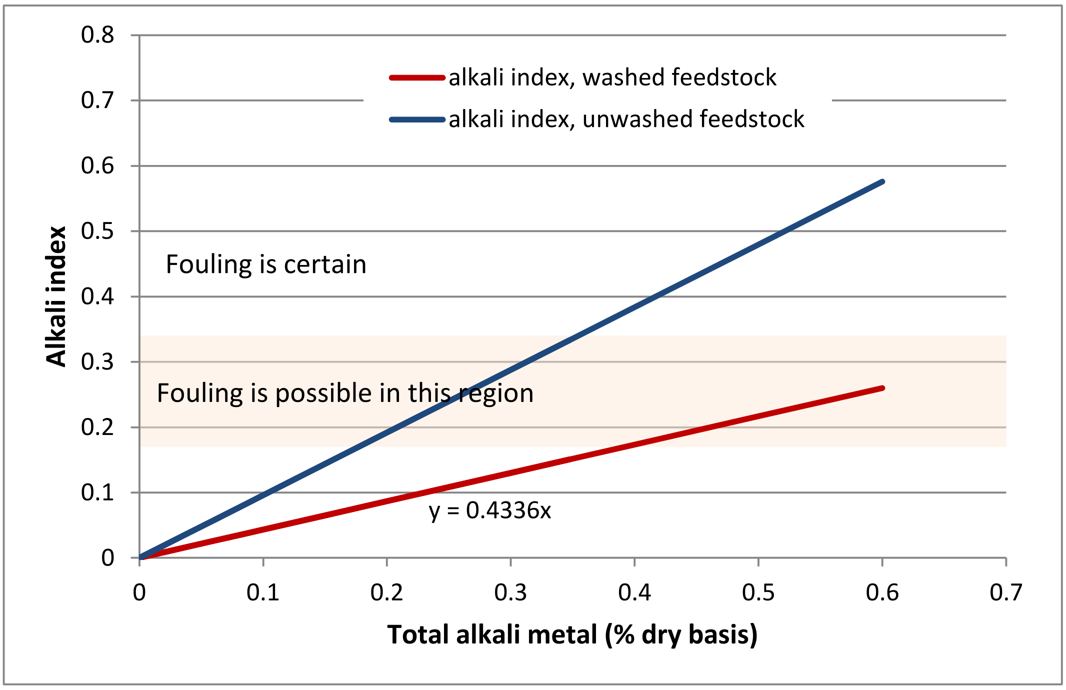

Figure 141: Alkali index, with and without water washing, vs. input alkaline content (based on [179] and Sheffield Alkali Index calculations from section 2.4.2.5)

Slagging in combustion is very likely when using biomass fuels containing more than 2-4% of ash. Table 24 shows indicative values for the total ash content of the feedstocks covered in this report.

Table 24: Total ash content of biomass[163]

Total ash content (wt.% dry)	Mean	Min	Max
Miscanthus	3.75%	1.10%	9.3%
SRC Willow	1.96%	0.45%	4.59%
SRF coniferous	1.07%	0.07%	10.7%
SRF deciduous	1.45%	0.15%	11.2%

Washing gives rise to significant improvements in the slagging index $R_{B/A}$, and although the benefits of water washing are limited for low alkaline content biomass, at higher alkaline contents, the reduction in slagging risk can be significant.

$$R_{B/A} = \frac{\%\left(Fe_2O_3 + CaO + MgO + K_2O + Na_2O\right)}{\%\left(SiO_2 + TiO_2 + Al_2O_3\right)}$$

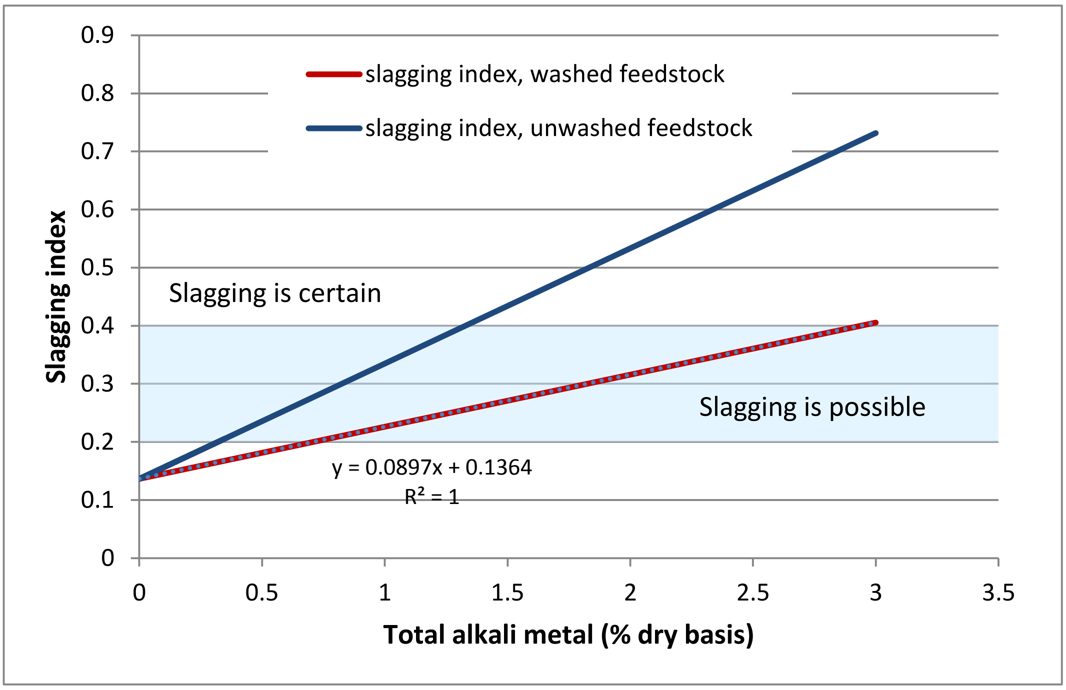

Figure 142: Slagging index, with and without water washing, vs. input alkaline content (based on [161] and Sheffield Slagging Index calculations)

For chemical washing, up to 100% removal of alkali metals can be achieved (hence the product Alkali Index would equal zero), which could also lead to avoidance of all slagging risks, as shown in Figure 143.

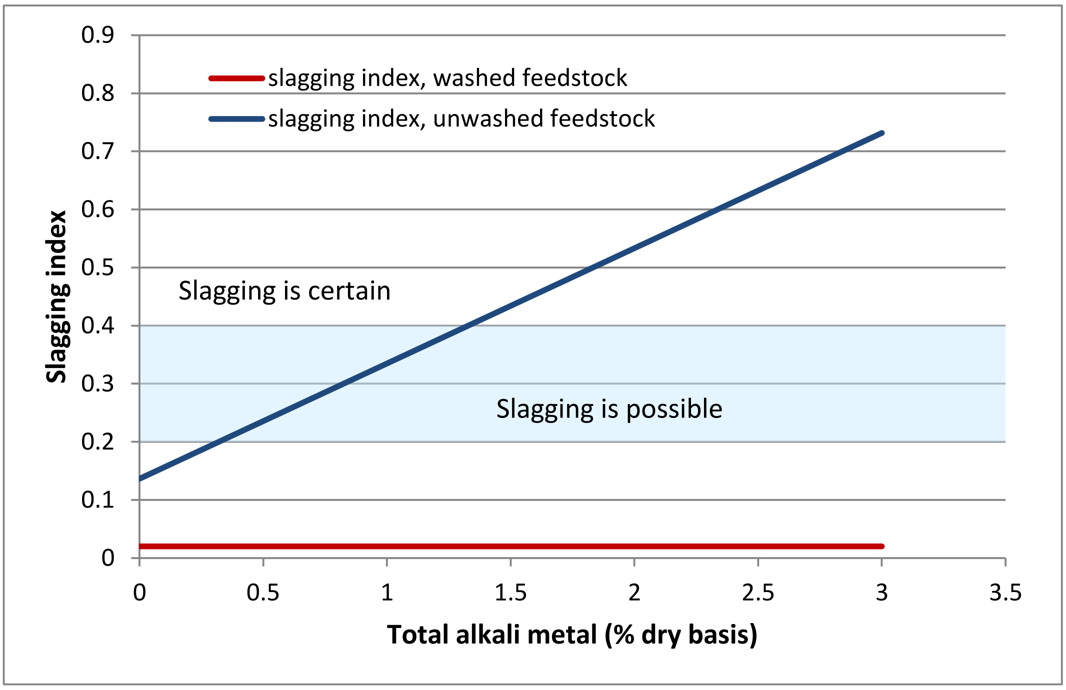

Figure 143: Slagging index, with and without chemical washing, vs. input alkaline content (based on [161] and Sheffield Slagging Index calculations)

Waste Water Treatment

The treatment of waste water discharged to sewers is governed by the European Union Urban Waste Water Treatment Directive[164]. Up to four stages of treatment are possible, a preliminary treatment to filter out large solids, a primary treatment to settle larger suspended organic matter, a secondary treatment to biologically breakdown residual organic matter and tertiary treatments that cover a variety of methods to address specific pollutants. A guide to trade effluent control[165] lists specific limits for various components. In the context of water washing, the most relevant limit is sulphate at a guideline of 1000 mg/l. Given the percentage removal of sulphate from SRC Willow water washing in table 13 at 10 %, along with the sulphur content of the feedstock of ~0.1%, and assuming a 5:1 by weight water:fuel ratio; it can be calculated that the sulphate concentration in this case is well below this limit at ~60 mg/l. Another potential issue are phosphates that could be produced from ash removal by washing, by a similar procedure of taking the percentage phosphate removal for SRC Willow water washing given in table 13, combined with a measurement of P_2O_5 in the ash composition of torrefied willow[157] a phosphate concentration in the effluent of ~40 mg/l can be calculated. This level may present a problem in the context of the Urban Waste Water Treatment Directive, which specifies a total phosphorus limit of 2 mg/l for discharge from large population equivalent agglomerations into sensitive areas subject to eutrophication[166].

In the case of chemical washing an additional complication will be the pH level. The normal range of pH allowed is between 6 and 10[165], which may necessitate an additional treatment step either by a continuous flow through or a batch pH adjustment system.

Drying

A disadvantage of the washing process concerns the trade-off between the level of removal of undesirable components, which is dependent on the biomass particle size, and the fact that the product of the washing process could be a water saturated biomass product. This may necessitate two stages of drying, firstly to allow the efficient grinding of the feedstock to a size range of mm or less at which it has been shown that washing techniques can be effective[167,171], and then a subsequent drying process to reduce the moisture content back down to a suitable level for the following application. To our knowledge, no-one has tried to quantify the effectiveness of water washing against particle size along with the subsequent drying requirement, but it will no doubt represent an additional level of complexity and cost in the overall process.

5.1.2 Development status and timescales

Water washing of biomass is developed at a TRL level of 7, with some commercial solutions available[168,169,170], although designed primarily for the agricultural produce sector (e.g. potato, sugar beet washers). The current maximum available scale is 80 tonnes/hr, based on the capacity of the washing machine equipment, but commercial plants are offered for sale from 1.0 tonne/hr upwards. Given the technology is relatively simple and can be operated on a small-scale, adaptation to biomass chips and reaching mass deployment of TRL 9 should be possible within the next 10 years, if there is sufficient industry interest. A potential driver that may accelerate this is the application to feedstocks in which it may be particularly beneficial such as straws and grasses discussed in section 5.1.3.

Chemical washing is at a much lower TRL of 4. It has been demonstrated in the lab[171] but not yet piloted or demonstrated on a larger scale. No future scale-up plans, or involvement in chemical washing from industrial actors in the biomass supply chain were found. Given that there are significant added costs with multiple washing steps, and water washing is already effective at removing a number of key biomass minerals, it remains to be seen if the technology will be demonstrated and taken up. TRL progression to reach TRL 9 is therefore highly uncertain, despite the components and chemicals used being commonplace.

5.1.3 Impact of different feedstock parameters on operation and cost

Biomass from different sources including forest residues, wheat/rice straw, corn stover and Miscanthus can become a higher quality fuel for CHP, power combustion, pyrolysis or gasification processes, if first pre-treated with water washing.

Virgin woody biomass has more desirable fuel properties for electricity production in furnaces and boilers than other biomass species, because of the generally lower ash concentrations. Although water washing can remove 50-95% of alkali metals and ash, it is also important to remove dirt and sand from wood sources. Agricultural residues, such as straws, and fast growing biomass crops, such as grasses, tend to have high potassium and chlorine contents that make them less desirable biomass fuels. Water washing provides a technical solution to their use in extant facilities such as gasifiers, without the need to incorporate new technology.

The washing technology performance and economics depend on the plant size and the feedstock properties, as well as the ratio of water:feedstock used in the process. In addition, a wastewater treatment plant is necessary to be added to the technology in order to comply with the European Union Urban Waste Water Treatment Directive.

Capex versus plant size

It is found that the Capex increases with a 0.53 power law based on increasing plant size. Figure 144 and Figure 145 show the Capex as a function of the output capacity in MW of water and chemical washed feedstock respectively. In each case, the feedstock is SRC willow chips with the base case representing 10 tonne/hr wet feedstock capacity, equivalent to an output of 32 MW with water washing and 31 MW with chemical washing.

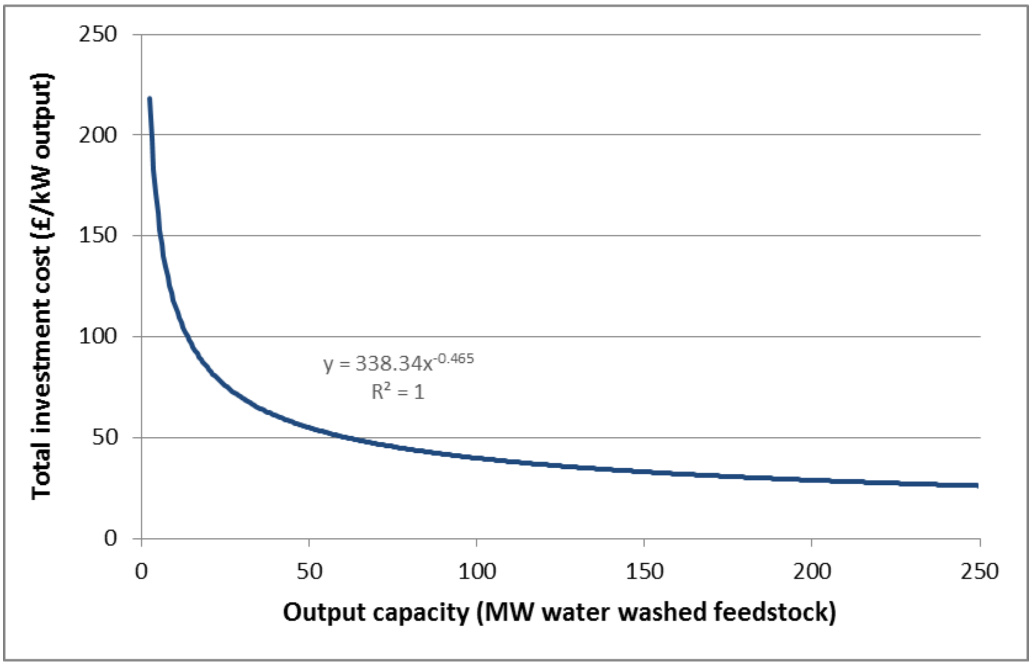

Figure 144: Water washing total investment cost vs. plant output capacity (based on data from [172] and quotes from Blue Machinery (Group) plc, applying a 0.53 power law)

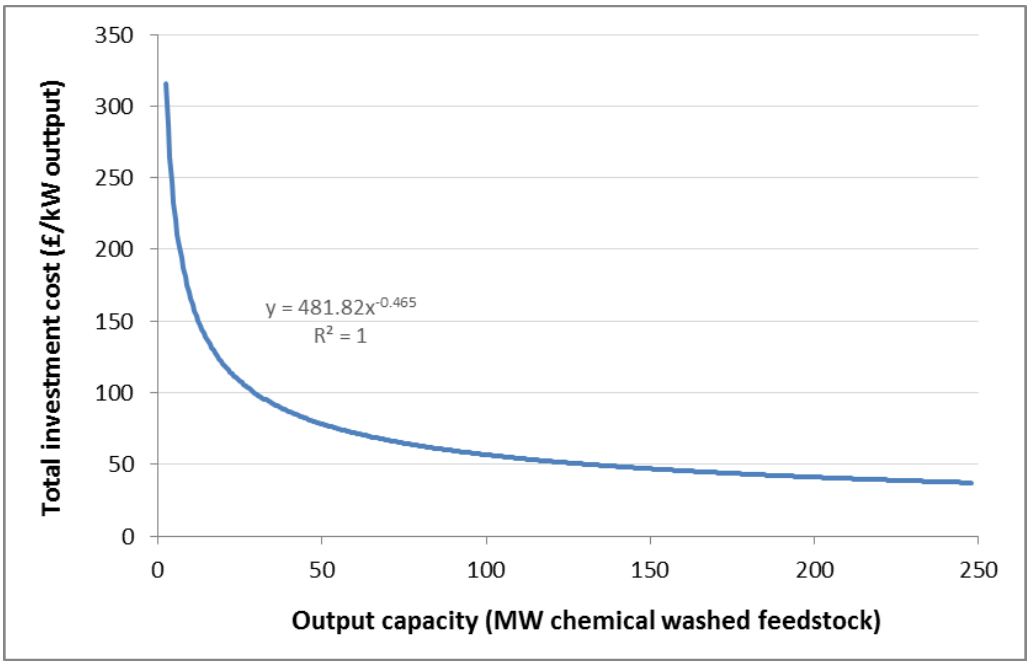

Figure 145: Chemical washing total investment cost vs. plant output capacity (based on data from [181,173] and quotes from Blue Machinery (Group) plc, applying a 0.53 power law)

It can be seen from above results that the Capex for chemical washing is about 45% higher than the Capex of water washing, due to the increased numbers of process steps and reactor vessels.

Opex vs. water:fuel ratio consumed for washing

The water washing process is more effective with demineralization of the biomass when more water is used, until a limit is reached whereby the additional water is not effectively improving the mineral removal (i.e. the gains become marginal, and vessels costs and energy use are increasing). In industry applications, there is typically a 5:1 water to fuel ratio applied, although different washers can take from 0.2:1 up to 10:1 water to fuel ratio. There is no data on the optimized mixing ratio yet.

The amount of water used and the wastewater treatment Opex will increase with the increase in water:fuel ratio, especially for larger plant sizes as illustrated in the following figures. The base case OPEX consists of: insurance (1.5% of TIC), maintenance – parts (2% of TIC), maintenance – labour (60% of operations labour), operations labour – UK labour rates[174] and from [176], and waste water disposal charges[172, 175].

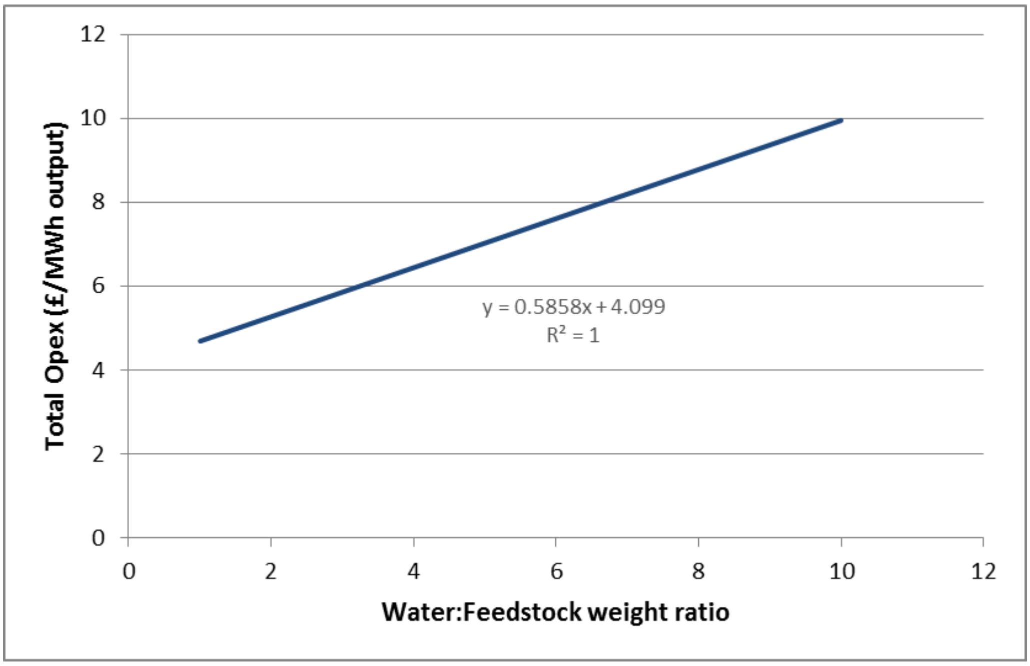

Figure 146: Water washing total Opex vs. water:feedstock weight ratio (calculation conducted by Sheffield based on water input and waste water treatment, cost data from [172, 175, 176])

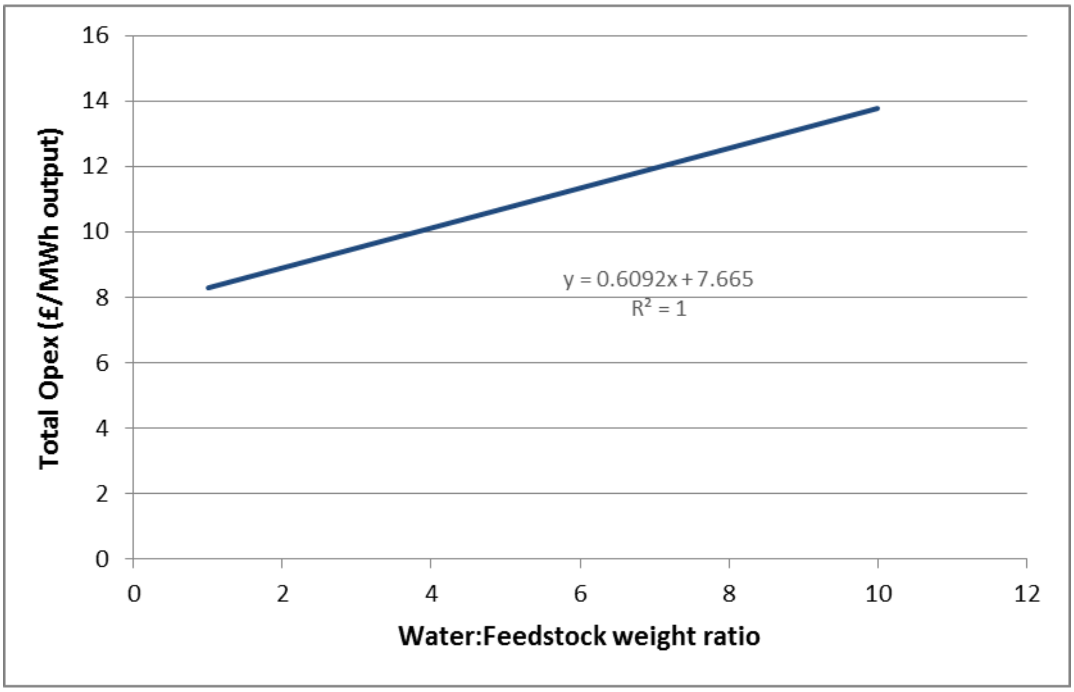

Figure 147: Chemical washing total Opex vs. water:feedstock weight ratio (calculation conducted by Sheffield based on data from [172, 175, 176, 177])

Opex vs. the input Alkali Index

The lower alkali content in the biomass means lower wastewater treatment costs, because of the lower cost of remediating alkali metals in the waste water. A feedstock having no alkali content (and hence no alkali metals present in the waste water) is estimated by Sheffield to save 15% of the base case total opex. The common measure of alkali content in the biomass is the Alkali Index (AI), as defined in section 2.3.2.3. The relationship between the total waste water treatment Opex and the AI for water washing is therefore displayed in Figure 148, assuming a SRC willow feedstock with an AI = 0.13 kg/GJ[178], and where 46% of the alkali metals in the feedstock end up being washed out into the waste water[157], and requiring treatment.

Chemical washing effectively removes all the alkali metals from the feedstock, and hence the waste water ends up having a higher concentration of alkali metals than when using water washing, which results in a greater expensive to remediate. This is demonstrated in Figure 149 by the steeper gradient and higher overall total opex.

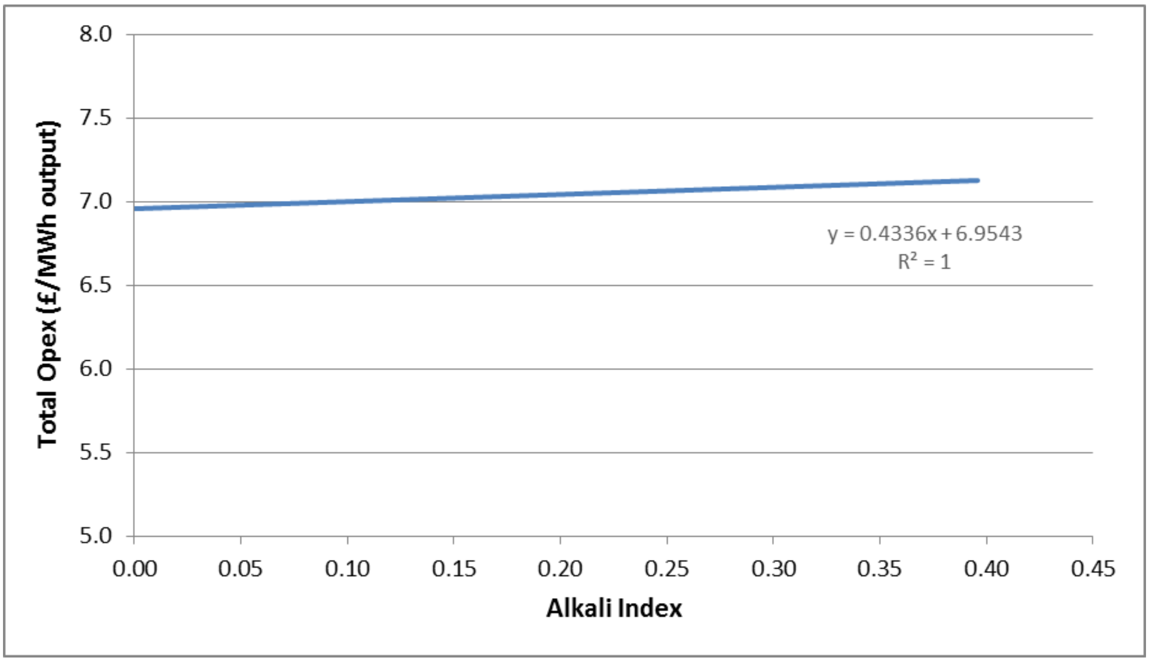

Figure 148: Water washing total Opex vs. input Alkali index (calculations by Sheffield, AI formula based on [179] and section 2.4.2.5)

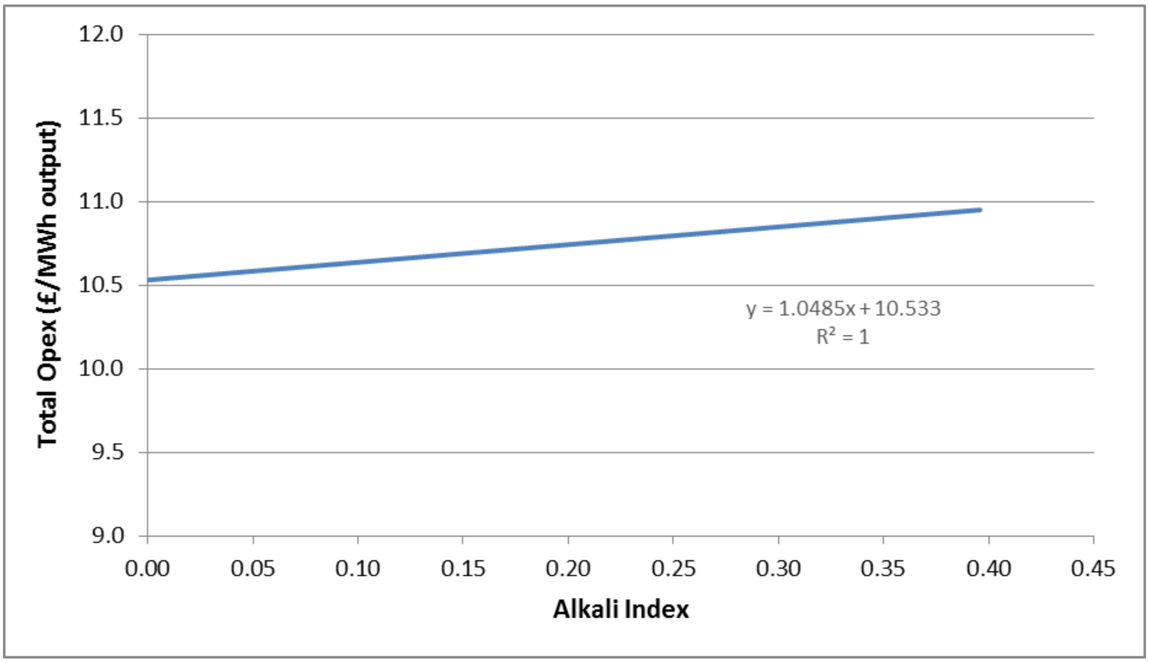

Figure 149: Chemical washing total Opex vs. input Alkali index (calculations by Sheffield, AI formula based on [179] and section 2.4.2.5)

5.1.4 SWOB analysis

The business case will make sense when demand for the improved properties of the biomass make it worthwhile to install at the stage after the biomass has been ground or chopped (when the benefit of the washing is greatest), but at the same time it causes further difficulties, given that the waste water has to be treated and the wet ground biomass re-dried.

Strengths	Can be effective on any natural biomass species Washing with water at ambient temperature reduces the ash content of the fuel up to 95% Water-washing reduces alkali release during combustion by up to 32% Chemical washing reduces alkali release during combustion by 90-100% Reduction in Cl and S contents significantly affects formation of acid gases and consequently acid corrosion on boiler components
Weaknesses	The impact of washing is usually detrimental to the Lower Heating Value (LHV), due to the wetter output product Consumes large amounts of clean water Waste water treatment Forced drying back to a low moisture content after water washing would entail significant energy input
Opportunities	Greatest benefits and hence willingness to pay for treatment for high alkali content biomass, such as straw and Miscanthus Water washing can be more effective with higher temperature of the slurry, up to 90°C Improvement of the washer design can reduce the water consumed. Components and chemicals used in chemical washing are commonplace and industrially available, and the full process is not yet optimised
Barriers	Chemical washing in particular is expensive and there is less experience with testing on different feedstocks The higher moisture content of the input biomass, the harder it is to grind the chips (due to the incompressibility of water) Chemical washing has not yet been proven on an industrial scale Demand for washing technologies is somewhat at odds with the industry trend towards ever drier and higher energy density products such as pellets and torrefied pellets. Uptake could be limited because of increased transport costs Likely to be limited demand for washing for cleaner virgin wood feedstocks that already have lower mineral contents

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5.2 Forced Drying

5.2.1 Technology description

If biomass were to be allowed to dry naturally in the open air then the water content of woody biomass would reduce from c. 50 - 60% when freshly harvested to c. 25 - 30% moisture content; this would occur over a period of 6 months to 2 years depending on climatic conditions and protection. If lower water content is required, then it is generally necessary to dry the biomass by thermal methods but this process is energy intensive[180]. Drying is a common step in many pre-processing plants, such as pelleting, briquetting and steam explosion.

The drying technologies considered in this report are drum dryers using direct heating and belt dryers using direct or indirect heating to give dried product at 10% w/w moisture content using feedstock with typically 60% to 100% of the particle size in the range of 6 to 32mm. For both types of dryer there is a risk of a dust explosion and therefore the equipment located in any hazardous area would need to be ATEX classified, for instance inside the drum or local to the belt conveyor where a dust laden atmosphere could be created (from fans).

For biomass fuels the emission limits depend upon the cumulative total of the thermal input from combustion plant on a site, and depending upon the value there are different emission limits. For thermal drying using either direct or indirect drying the technology required to comply with the current emission requirements is well developed and should not inhibit technology selection.

Drum dryers

Drum dryers operate at higher temperatures (than belt dryers) and are heated directly by hot gas and the biomass is fed into a rotating cylinder. The inlet temperatures can be up to 600°C. However, these dryers may release wood particles during the drying process and volatile organic compounds arising from the biomass due to the higher operating temperatures. Drum dryers need to be provided with a flue gas cleaning system due to these issues; the complexity of which will be dependent upon the operating legislative environment. Flue gas cleaning for a drum dryer may consist of a bag filter and/or cyclones and, due to the water evaporation, a moisture control system to mitigate the extent of vapour plume.

Drum dryers are the most common type for biomass applications as they are of simple and robust construction with corresponding low maintenance costs and more tolerant to feedstock size variation. However, due to the high operating temperature, a biomass fire is more of a hazard than with a belt dryer. To control any fires arising, a water sprinkler system would also normally be provided.

Figure 150: Example of drum dryer[181]

Belt dryers

The biomass is spread on a moving perforated conveyor to dry the material in a continuous process with fans blowing the drying medium through the belt and biomass material. The drying medium can be low temperature hot air arising from direct firing or is more normally from a heat source arising from waste process heat. The belt dryer operates at lower temperatures (than drum dryers) with drying temperatures of 100 to 200°C and with corresponding lower gas outlet temperatures, but usually above 100°C to reduce the effects of flue gas corrosion. Therefore, problems regarding fire hazard from biomass combustion, emissions arising from the biomass, and flue gas corrosion, can be reduced due to the lower temperatures. Flue gas cleaning on a belt dryer would normally consist of cyclones and, due to the water evaporation, a moisture control system to mitigate the extent of vapour plume, but this would be dependent on the operating legislative environment.

Figure 151: Example of belt dryer arrangement[182]

Fluidised bed drying

There is one other type of industrial drier technology, using a fluidised bed reactor with heated bed material. However, fluidised bed drying tends to be used for sludges and for fine pharmaceuticals. Enquiries were sent to a number of dryer manufacturers to obtain details of any biomass fluidised bed dryers, but it was not possible to obtain a quotation for a fluidised bed dryer for biomass. In particular, one of the major suppliers, from whom quotes were obtained for both belt drying and drum drying, advised that they would not offer fluidised bed drying of biomass because the alternative technologies were much more cost effective.

5.2.2 Development status and timescales

The demand for UK sourced wood fuel of all types has been rising rapidly in recent years. UK sourced wood fuel is being used successfully for generating heat and electricity using wood fuelled boilers. This has mainly been due to the introduction of the Renewables Obligation and Renewable Heat Incentive. Drying has seen uptake as a component within other pre-processing plants, mainly pelleting.

The provision of machinery for drying is a mature market (TRL 9), particularly on the European continent, though its deployment in the UK has historically been limited. The suppliers of proven drying equipment are from the continent and range from small specialist companies to large industrial companies.

Dryers of both the belt and drum dryer designs can be obtained in a wide range of capacities. As an indication of scale, the drum dryer can be built in capacities (feedstock c. 50%w/w moisture to 10% output) of less than 1 tonne/hr of feedstock and belt dryer capacity starts at c. 1 tonne/hr. The economics improve with scale and single units in the order of 70 tonnes/hr for drum dryers and c. 30 tonne/hr for belt dryers are available. However, both technologies can be utilised on a multi-stream basis to give the desired capacity if in excess the capability of a single unit.

In order to construct a complete dryer system, the overall construction period to full operation would be 6 months to a year for capacities in the order of 1 tonne/hr to 10 tonnes/hr, and a period of 1 to 1.5 years for a plant capable of the higher capacities indicated above.

5.2.3 Impact of different feedstock parameters on operation and cost

None of SRC Willow, LRF or SRF are particularly friable, primarily because of their high moisture content. The primary difference between the different feedstocks is the amount of moisture level in the biomass as it is received. With the exception of Miscanthus, the moisture content when harvested will be in the range of 45 to 60%. Miscanthus grass contains a narrower range of moisture levels when harvested in the spring, ranging from 14% to 23% and is a similar crop to straw.

Availability of dryers varies depending on the moisture content and the particle size distribution of the feedstock, rather than on the type of feedstock inherently. There is no difference in plant lifetime between different feedstocks.

Moisture Content

The primary purpose of a dryer is to evaporate water as efficiently as possible with no physical product degradation or burning. Therefore, the overall cost structure (Capex and Opex) of a drying system will be dependent upon the quantity of moisture removed from the biomass as this will influence the energy consumption, dryer size and system efficiency.

Figure 152 below gives an indication of the energy demand required at different feedstock moisture contents for a belt dryer, and Figure 153 shows a similar chart for a drum dryer.

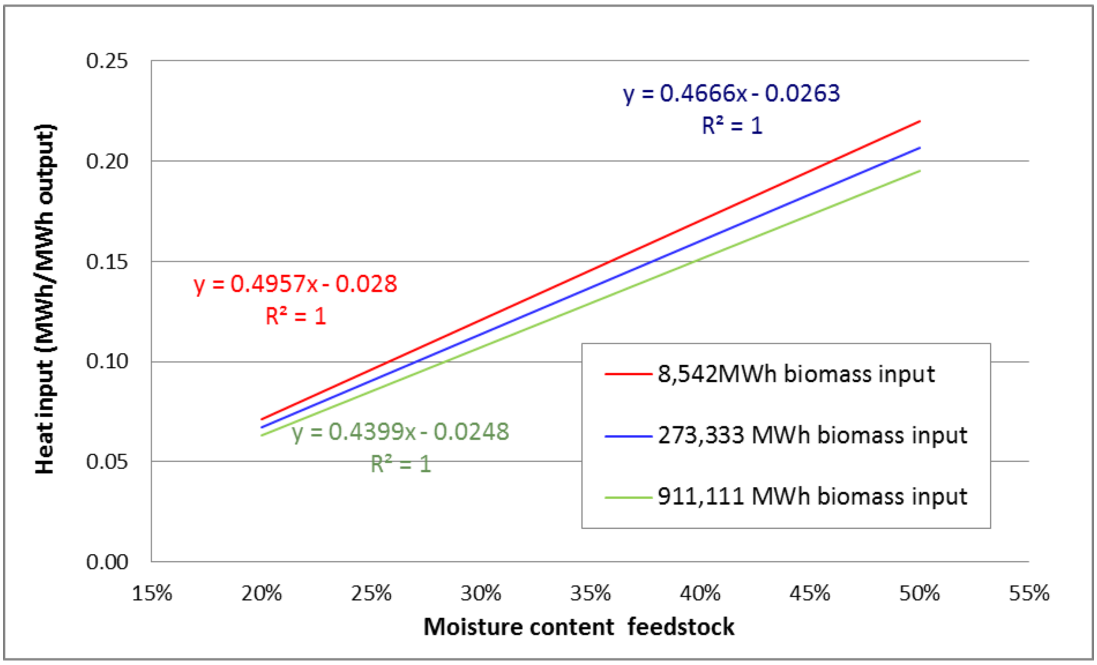

Figure 152: Belt dryer heat input vs. feedstock moisture content, shown for different biomass annual input scales (derived by B&V based upon 2015 supplier data)

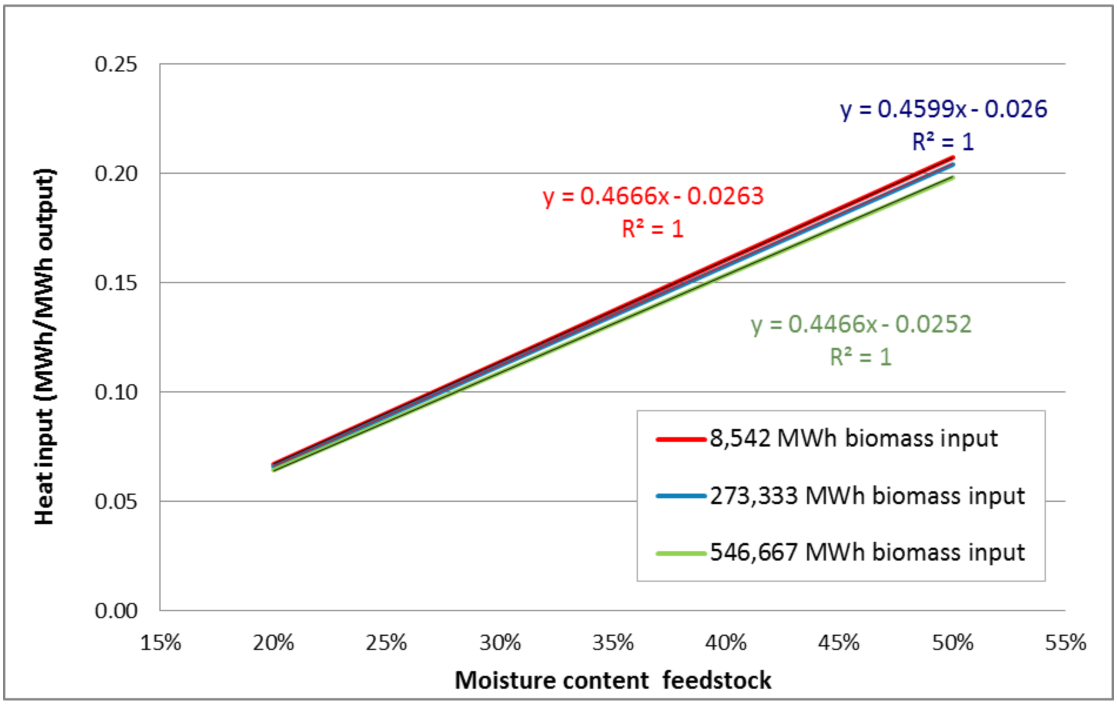

Figure 153: Drum dryer heat input vs. moisture content, shown for different biomass annual input scales (derived by B&V based upon 2015 supplier data)

Plant size

The variation in Total Investment Cost of a belt dryer and a drum dryer with varying biomass input are given below in Figure 154. This curve includes the costs of chipping, screening and drying the feedstock to reduce the moisture content from 50% to 10% and it has been derived from quotes provided by suppliers. It includes capital costs for chipping and screening from mobile plants and drying the feedstock from 50% to 10% moisture content. These total investment costs are divided by the plant output capacity in kW – these are not levelised costs across the plant lifetime.

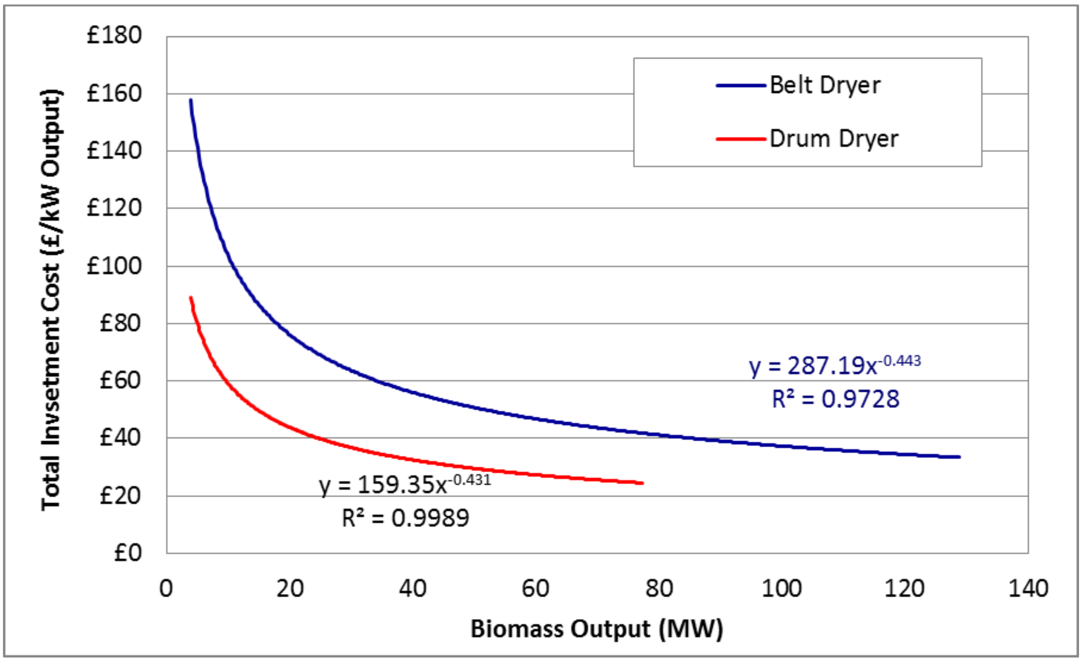

Figure 154: Total investment cost for belt and drum dryers (derived from 2015 supplier data and B&V estimates for all-in overnight EPC costs)

Opex

The variation in Total Opex of a belt dryer and a drum dryer with varying biomass input at 50% moisture feedstock is given below in Figure 155. The curve is derived based on data provided by suppliers with assumptions made on labour requirements for the whole system (chipping, screening and drying). It includes operating costs for chipping and screening from mobile plants. A 1% of total installed costs has been assumed as insurance cost.

The major contributor to decreased Opex/MWh is the increased operating hours that would be used for larger drying plant economics. The heat provided to both belt and drum dryer has been costed on the basis of using natural gas.

The drum dryer needs to use high grade heat (derived from natural gas in the case considered) whereas the belt dryer could use low grade heat from a waste process stream. If the belt dryer were to use low grade waste heat from a process, as would normally be the case (no fuel cost), then the Opex could be significantly less.

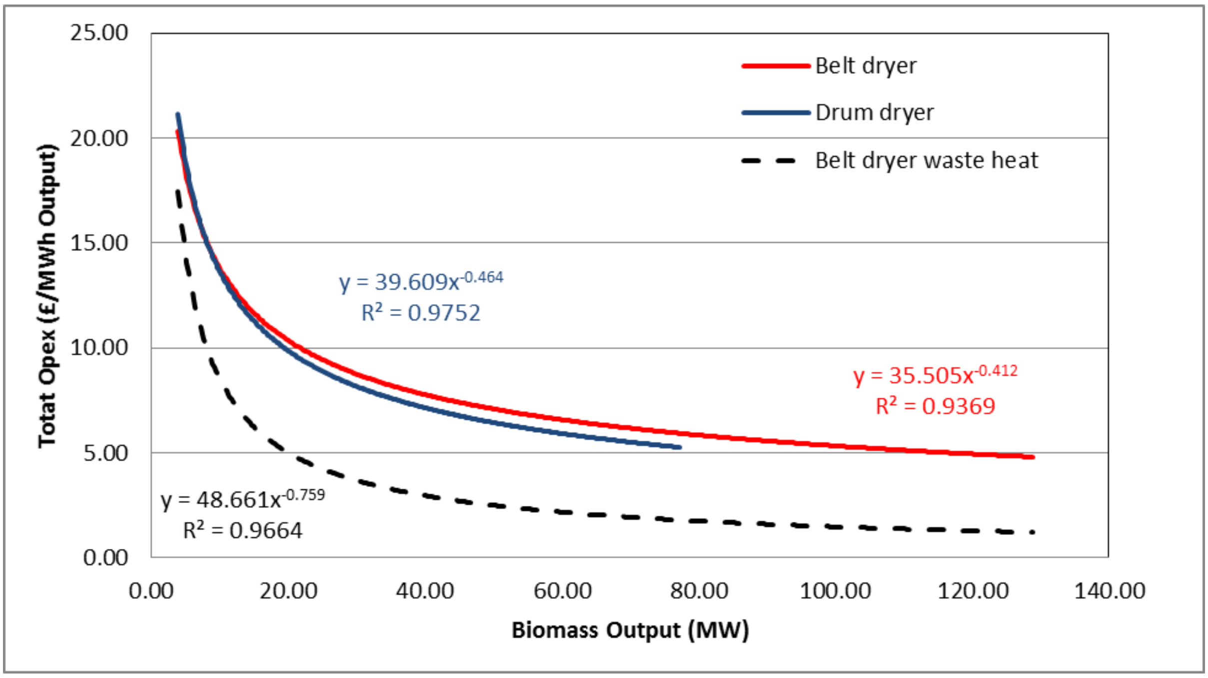

Figure 155: Total Opex for belt and drum dryers using feedstock at 50% moisture content based upon natural gas, with waste heat option for belt dryer (derived from 2015 supplier data and B&V estimates)

5.2.4 SWOB analysis

Thermal drying is an important pre-processing process to improve the energy density of the biomass feedstock. Thermal drying also has the effect of reducing transportation costs as the bulk density will increase. Dried biomass results in more compact and cost effective mechanical handling systems. Combustion processes using non dried biomass would be physically larger to accept the higher volume for a given output.

Thermal drying is essential for downstream processes such as pelletising and briquetting which require wood moisture content lower than that which could be achieved by natural open air drying.

Combustion technologies can accept 'non dried' wood but at reduced boiler efficiency and with poorer emissions to air: it is generally more difficult to maintain consistent results over the operating range. If non dried biomass is used then the exhaust plume would also be more visible in colder conditions due to the condensation of the moisture in flue gases in the atmosphere. At very high moisture content, wet biomass requires very high energy inputs to evaporate contained moisture and subsequently to heat the water vapour, to the extent that combustion temperatures may be reduced below the minimum temperature required to sustain combustion.

Strengths	Energy densification, as LHV of the product improves by removing water Multitude of suppliers for drum and belt dryers Lower transportation costs following drying, as less water being transported Less problematic storage problems, with reduced fungal spore build-up, reduced loss of dry matter through microbial action and reduced self-heating (all issues with long-term storage of wet biomass)
Weaknesses	Energy intensive process. Energy input provided from either use of fossil fuel (typically natural gas) or process waste heat. If the latter, then the drying site would be required to be integrated into another production facility, for instance a sawmill or CHP[183] Hot biomass can be prone to fire hazards and flue gas cleaning is required to deal with volatile gases and steam released.
Opportunities	In order to maximise the overall system economics, the dryer system should be integrated into a downstream facility, for instance pelletising. torrefaction or briquetting plant, or be able to access waste heat The new Contracts for Difference (CfD) regime is forcing power plant projects to find CHP applications for a proportion of their heat, because the CfDs are not available to new dedicated biomass power-only plants. Wood chip drying on site is therefore being proposed for several projects as the use of the heat
Barriers	Wood supply chains are relatively small and fragmented in the UK which makes it more difficult to provide secure long term “bankable” supply contracts for new ‘greenfield’ projects, particularly if the projects are large Market for thermally dried biomass is limited, as any major end user requiring dried biomass for pelletising or torrefaction would provide a fully integrated production facility with drying Use of batches of overly dried biomass in biomass combustion can lead to issues of inconsistent combustion and resultant emission deviations from design.

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5.3 Briquetting

5.3.1 Technology description

Briquetting is a conventional process for biomass densification that uses high pressure to increase the particle size and bulk density of the feedstock, producing a cylinder of compressed biomass.

Prior to briquetting, the biomass will need preparation, as requirements for briquetting generally include a moisture content between 10% and 15% and a particle size of less than 25 mm. The size reduction of the feedstock can be achieved by chipping or grinding and the required moisture content levels can be achieved via natural and forced drying (belt or drum dryers). In addition, the feedstock should be cleaned of any foreign material that can cause excessive wear to the press.

Briquetting of woody biomass takes advantage of the presence of natural binding materials such as protein, starch and lignin. A combination of high pressures and temperature are required to destroy the cell walls and make the lignin available. Depending on the feedstock, the use of binders may be required.

The technologies used in the briquetting process are well developed. Any resultant emissions can be controlled to meet the legislative requirements and should not inhibit technology selection.

The three main briquetting technologies are piston screw, screw press and roller press, as outlined below.

Piston Press

In piston presses, pressure is applied discontinuously by the action of a piston on material packed into a cylinder. They may have a mechanical coupling or utilise hydraulic action on the piston. The compressed material is heated by frictional forces as it is pushed through the die. The lignin contained in the woody-cellulose acts as a natural binder to the compressed material. As the material exits the machine, cooling may need to be provided for the lignin to solidify and hold together to form the briquettes. Production capacities can be up to 3,000kg/hr, producing a pellet up to 120mm in diameter.

Screw Press

In screw-presses, material is fed continuously into a screw which forces the material into a cylindrical die. Pressure builds up along the screw rather than discontinuously under the impact of a piston. The die is often heated to raise the temperature to the point where lignin flow occurs. The temperature is normally raised to 250-300°C; below this a binding material may have to be added. Production capacities vary from around 75 to 400kg/hr.

Roller Press

In the roller press, ground biomass is forced through the gap between two cylindrical rollers of the same diameter. The rollers rotate horizontally in opposite directions on parallel axes, allowing feed to be drawn in one end and briquettes to be discharged at the opposite. The distance between the rollers depends on the type of biomass, the particle size and the moisture content among others[184].

Briquettes are used not only for domestic heating but also in industrial boilers. Table 25 shows a comparison of different briquette presses. The presence of hard materials like sand and grit reduces the life of the die and hence the operating cost. Note that maintenance will strongly depend on the silica, ash, sand and/or soil/stone in the feedstock. The abrasiveness of the cellulosic material also affects die life.

Table 25: Comparison of briquetting equipment (Idaho National Laboratory, 2010)

	Piston Press	Screw Press	Roller Press
Optimum moisture content of raw material	10-15%	8-9%	10-15%
Wear of contact parts	Low	High	High
Output from the machine	In strokes	Continuous	Continuous
Power consumption (kWh/tonne)	37.4 to 77	36.8 to 150	30 to 83
Density of briquette	1-1.2 g/cm^3	1-1.4 g/cm^3	0.6 to 0.7 g/cm^3
Maintenance	High	Low	Low
Combustion performance of briquettes	Moderate	Very good	Moderate

5.3.2 Development status and timescales

The production of briquettes is a well-established technology (TRL 9). Industrial methods of briquetting date back to the second part of the 19th century[185]. Briquetting plants are provided by a large number of European companies supplying the UK market. Technology providers include RUF, C.F. Nielsen, Di Piu, Sahut Conreur, etc.

Briquettes produced in the UK and Europe are mainly sold to consumers within their own country, with little international trade occurring. The main producer in Europe is Germany. The briquette market in Asia and Africa is growing, and briquettes are becoming an alternative to replace charcoal and wood consumption for cooking. In the UK, briquettes are mainly made from agricultural wastes. Probably the largest plant built in the world is located in Germany, with a production capacity of 100,000 tonnes/yr. The briquettes are produced from pallet production residues by 12 piston presses each producing 1 to 1.5 tonnes/hr[186].

In general, commercial plant scales range from 0.075 – 3.0 tonnes/hr for each press. Delivery time could be between 2 and 4 months with typically 3-6 months lead time depending on project scale.

The briquetting market is expanding and adapting to the market needs. For example, torrefied wood briquettes are being piloted as an alternative to torrefied pellets. Similarly, straw briquettes are co-digested with manure for the generation of biogas.

5.3.3 Impact of different feedstock parameters on operation and cost

Briquettes can be made from almost any type of biomass with low moisture content and uniform size, shape and material properties. Availability of the plant will be dependent on the life of dies for each feedstock, depending on the grit content. Lifetime of the plant as a whole will not vary between feedstocks as dies are changed frequently.

Capex

The cost curve in Figure 156 corresponds to the total investment cost for a briquetting system, including chipping, screening, drying (via drum dryer) and briquetting press. It has been derived from quotes provided by suppliers. It assumes the utilization of chipping and screening mobile plants driven on diesel, drying via a drum dryer to reduce the moisture content from 50% to 10% and a stationary briquette press processing wood chips with a particle size <25mm.

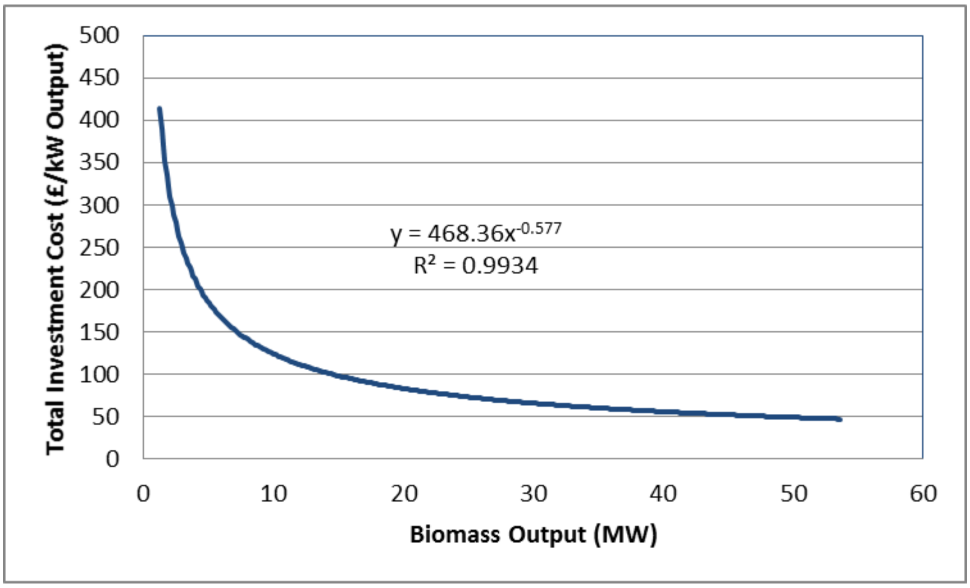

Figure 156: Briquetting total investment cost vs. output capacity (derived by B&V based upon 2015 quotes from 3 suppliers and B&V estimates for all-in overnight EPC costs)

Moisture Content

As mentioned above, moisture content should be between 10% and 15%. If the moisture content is too high or too low, the briquette will not maintain its shape. The moisture content for as-received SRC Willow and SRF will be significantly higher than 20% and therefore forced drying is always required. This will result in an increase in the total investment cost and the operating costs. In the case of Miscanthus, less drying may be required, as the moisture content of Miscanthus harvested in spring will typically vary between 14-23% moisture.

The curves in Figure 157 show the variation in operational costs to produce briquettes with a moisture content of 10%. The curves assume drying via a drum dryer of feedstocks with different moisture content ranging from 50% to 30% (as received).

The curves in Figure 157 have been derived based on information provided by suppliers and assumptions made on labour requirements for the whole system (chipping, screening, drying and briquetting). A 1% of total installed costs has been assumed as insurance cost.

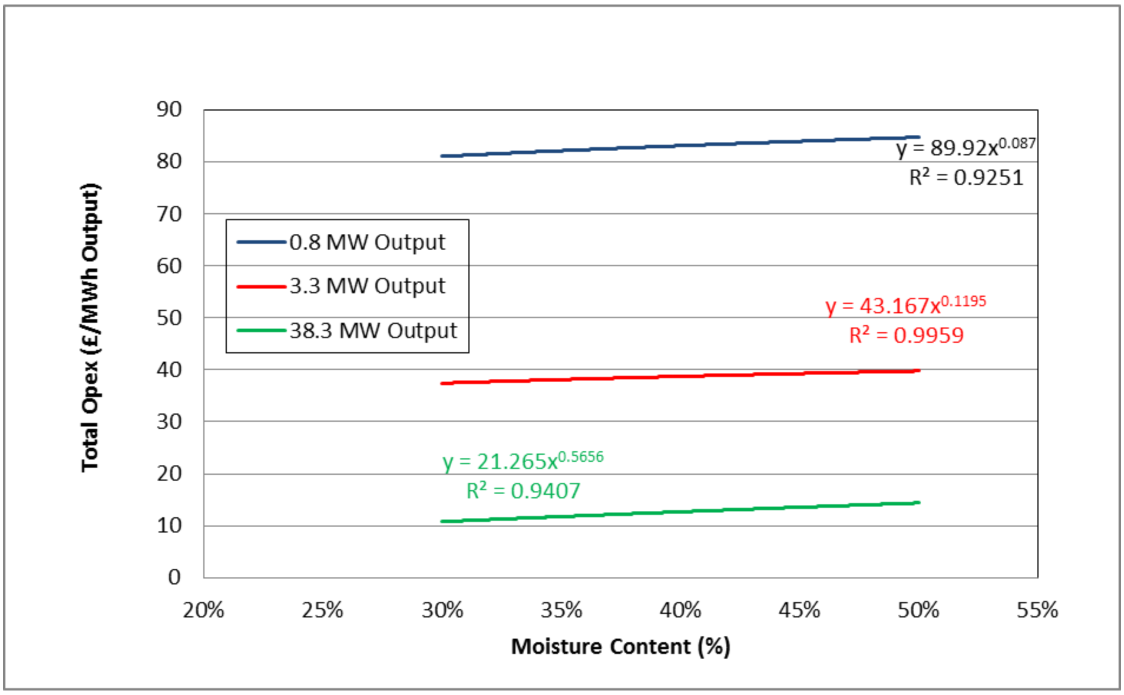

Figure 157: Total Opex vs. feedstock moisture content, given at different output capacities (derived from 2015 quotes provided by suppliers and B&V estimates)

Ash content

In addition to silica and/or glass, metal and stone content, ash content of biomass can increase wear on vital parts of the briquetting press. The maintenance cost (wear parts/service parts) for a briquette plant when using clean wood is c. £0.70/tonne of briquette produced. However, when using more abrasive material with higher ash content the maintenance cost is c. £1.4/tonne of briquette produced.

Particle size

Smaller particle size offers a greater surface area for biomass densification technologies. However, a mixed particle size distribution will result in better binding of the briquettes.

Biomass composition

Lignocellulose is the main component of woody biomass. Lignin can act as a natural binding agent. With moisture content in the range of 8% to 15%, the lignin will be more available at temperatures between 100° to 130°C (Idaho National Laboratories, 2010). Table 26 shows typical lignin quantities in the different feedstocks. If a binding agent is required, such a maize starch, this may cost ~£320/tonne[187].

Table 26: Typical biomass lignocellulosic composition (mean values wt% (dry))[188]

Feedstock	Cellulose	Hemicellulose	Lignin
Miscanthus	31-55	25-38	6-13
SRC willow	30	45	25
SRF coniferous	40-45	25-30	25-35
SRF deciduous	40-50	25-35	20-25

5.3.4 SWOB analysis

Briquetting is often overshadowed by pelletising, as briquetting is only beginning to emerge in the market place as a viable alternative to pelletising. Briquetting has a lower investment and operation cost in comparison with pelleting, and is also being investigated for use with torrefied chips. In addition, portable briquetting systems facilitate transport of feedstock to processing plants, accessing biomass that it is too expensive to transport.

Strengths	Briquetting generates a product with an increased energy density, through increasing bulk density as well as through the combination with drying; Briquetting provides better and more consistent thermal and physical qualities allowing improved combustion characteristics Briquettes are of uniform size and shape and therefore are easier to transport, store and mechanically handle Briquetting preserves the biomass quality over time with minimum degradation Lower investment and operation cost in comparison with pelleting production Briquettes can be produced from a variety of raw materials
Weaknesses	Briquettes have a lower bulk density in comparison with pellets, hence higher transport costs Higher ash content feedstocks lead to increased abrasion and operating costs, and less robust briquettes
Opportunities	Producing higher density briquettes which will result in improved calorific values and combustion Introducing more efficient extrusion methods Reducing production costs
Barriers	Few combustion plants in the UK currently take in briquettes. As a result the market volume is much lower and the unit price offered by supplier is not as attractive as pellets Wood supply chains are relatively small and fragmented in the UK, making difficult to provide long term “bankable” supply contracts

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[184] Idaho National Laboratory (2010) “A review on biomass densification technologies for Energy application”.

[185] FAO “Chapter 4.Technical aspects of briquetting”. Available at: http://www.fao.org/docrep/t0275e/T0275E03.htm#

[186] Bioenergy International No 66, 4 (2013). Available at: http://www.cfnielsen.com/files/uploads/1372413804_cfn_article_bioenergi_juni_2013.pdf

[187] Münch Edelsthal GmbH (2015) “Personal communication, 2015 prices”.

[188] NNFCC (2014) “Lignocellulosic feedstock in the UK”, Available at: http://lb-net.net/wp-content/uploads/2015/04/LBNet-Lignocellulosic-feedstock-in-the-UK.pdf

5.4 Screening

5.4.1 Technology description

Screening is a size separation process in which the feedstock is separated into at least two fractions: oversize (material remaining on the screen) and undersize (material passing through the screen).

The main factors to be considered in the selection of screening equipment include[189]:

- Particle size, particle size distribution, bulk density, moisture content, particle shape and flow characteristics (potential for the material to stick together or entangle);
- Screen design characteristics, such as size of screen openings, shape of screen openings, total surface screening area;
- Operational characteristics.

Biomass does not require preparation prior to screening: screens can take feedstock with a moisture content as high as 60% and with a range of size and shapes.

Screening is a well-developed technology and any resultant emissions can be controlled to meet the legislative requirements and should not inhibit technology selection.

The two main types of screens used on woody biomass applications are vibrating and disc screens, as outlined below.

Vibrating Screen

The vibrating screen consists of an inclined box frame fitted with one or more wire mesh plates mounted one above the other. The screen frame is driven by an electric motor or diesel engine, producing vertical oscillations.

As the material vibrates the particles become airborne and fall further down the inclined screen mesh, providing opportunities for the material to pass through an opening.

Some of the advantages & disadvantages of the vibrating screens are mentioned below:

- High efficiency of separation;
- Relatively inexpensive;
- Can be noisy; and
- Can become blocked easily

Disc Screen

A disc screen consists of evenly spaced shafts rotating in parallel that are arranged perpendicular to the material flow. The shafts are fitted with discs in an interlocking pattern which forms the opening of the screen. These screens can have multiple opening sizes along the length of the screen.

Star screens are a type of disc screen. The stars are made of polyurethane and the shafts rotate at higher speed than a conventional steel disc screen.

Some of the advantages & disadvantages of the disc screens are mentioned below:

- In comparison with vibrating screens, disc screens run quietly;
- Disc screens tend to jam less than vibrating screens;
- Disc screens are more mechanically complex;
- Disc screens are easier to maintain.

5.4.2 Development status and timescales

Biomass can be a difficult material to screen: it binds together; it does not flow well; it can vary in moisture content and density; and it can freeze or catch fire easily. Most of the technology involved in biomass screening has been borrowed from screening of MSW and aggregates.

Biomass screening has developed out of the forest products industry around the world, chiefly where there are extensive pine forests such as Scandinavia, Canada and Russia. In this industry there have been many applications; from cleaning up brash and logs from the forest to removal of contraries in the paper making process. Such screening activities have been carried out since the beginning of the 20th century. When bark or woodchips arrive in a power station, the bulk deliveries are typically passed over a very coarse screen or grizzly. With more refined biomass products, for example at a paper mill, feedstock would most likely be passed over a disc screen with finer screening apertures.

Biomass screening is a mature technology (TRL 9). Both vibrating screens and disc screens are available in a whole range of applications and sizes depending on the application and the feedstock. This technology is available to all aspects of woody material screening, such as compost preparation and also inorganic material screening such as construction and demolition waste.

Some of the leading developers include Lubo Systems Screening & Recycling, Vecoplan and Komptech. Screening can achieve different fractions fine (0 to 25mm), medium (0 to 60mm) and coarse (0 to 150mm), with throughputs typically ranging from c. 100m^3/hr to 500m^3/hr. Equipment delivery time could be c. 3 months.

5.4.3 Impact of different feedstock parameters on operation and cost

Availability of screening varies depending on the content of contrary matter, and on the particle size distribution of the feedstock and product required, rather than on the type of feedstock inherently. There is no difference in plant lifetime between different feedstocks.

Capex

The cost curve in Figure 158 corresponds to the total investment cost for a screen system. It has been derived from quotes provided by suppliers for mobile plants driven on diesel with the possibility of being connected to the mains. It includes chipping and screening from mobile plants capital costs. These total investment costs are divided by the plant output capacity in kW – these are not levelised costs across the plant lifetime.

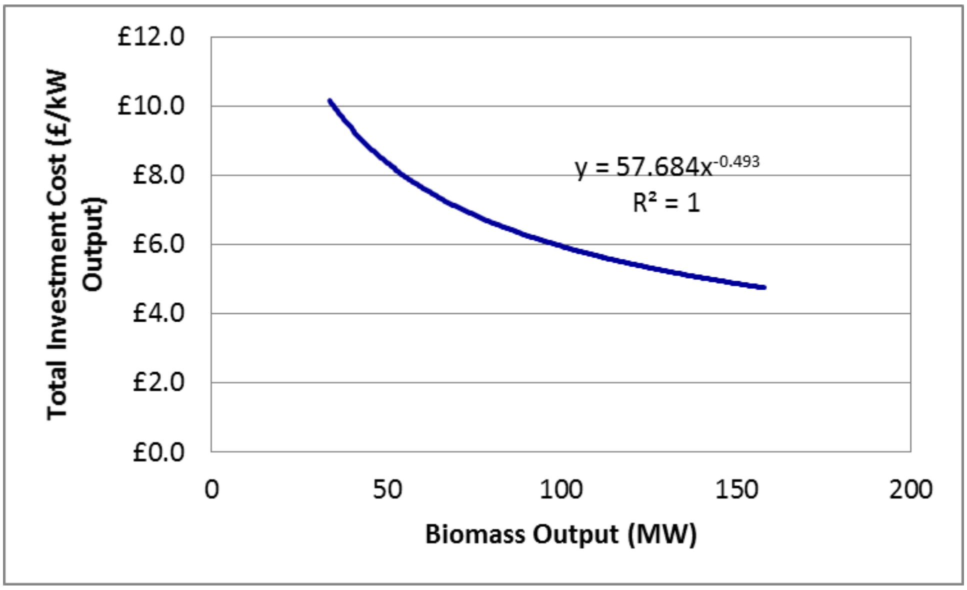

Figure 158: Screening total investment cost vs. biomass output capacity (based upon 2015 supplier data and B&V estimates for all-in overnight EPC costs, and power law scaling)

Opex

Opex for screening includes fixed and variable Opex (fuel costs, operations labour, insurance, maintenance parts and maintenance labour) and Figure 159 is for screens processing wood chips with 50% moisture content and screening oversize material larger than 50mm. It includes chipping and screening from mobile plants operating costs. The typical assumption is that ~1% of the input feedstock will be removed as undersize material and oversized rejects – but this is entirely dependent on the quality of the feedstock. Feedstocks with a much greater spread of particle sizes will have much more material screened out. A 1% of total installed costs have been assumed as insurance cost.

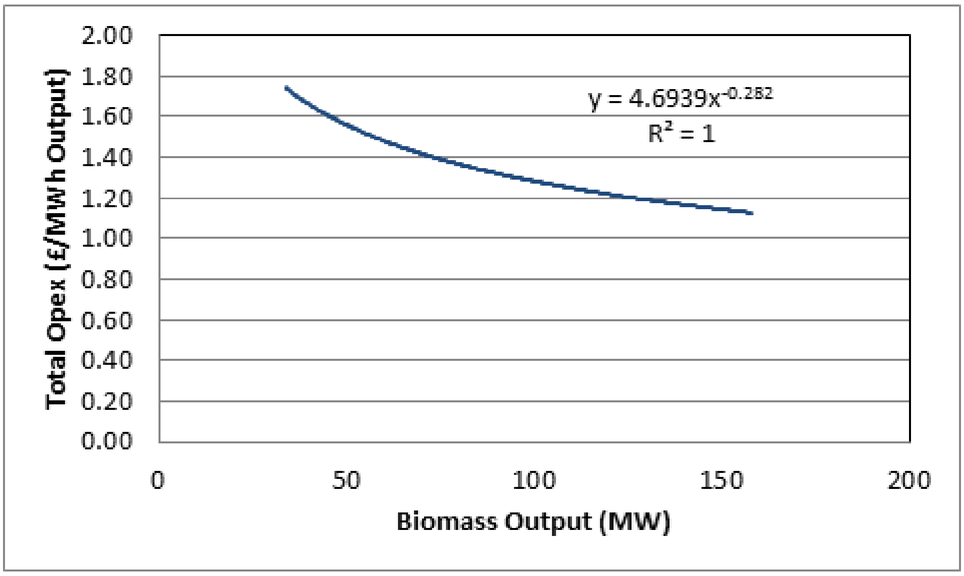

Figure 159: Screening total opex vs. biomass output capacity (derived from 2015 supplier data and B&V estimates and power law scaling)

Moisture content

There is no limit on the moisture content of the feed, as such. However, the wetter the material the lower the energetic throughput, and hence more energy consumed per MWh of screened product. However, initial discussions with suppliers have indicated that this energy consumption does not vary significantly. In addition, the screen may tend to block when processing high moisture content feedstock (some sources suggest above ~50%), as smaller particles will tend to clump together or stick to larger particles reducing the efficiency of the screening.

5.4.4 SWOB analysis

Screening is an important pre-processing process when processing green cuttings, forestry residues, bark, chips and shredded waste wood. When dealing with clean wood chips screening is not so relevant. The added value of the screen is generally that it protects the equipment downstream. By removing oversize material, the screen may prevent blockages and avoid costly downtime. By removing fines, the screen may prevent clogging of piping, deposition on tubes, loss of unburnt material through a grate and loss of ignition (and therefore efficiency) of fine material elutriated from a fluidised bed.

Strengths	Removes oversize particles that could damage or block downstream equipment, thereby improving conversion plant availability Removes fine particles that would cause a dust problem, thereby reducing explosion/fire risks Produces a feed of uniform size with slightly improved density Increases the overall heating value of the material by removing inerts, such as stones Low energy input required
Weaknesses	Typically a very noisy process Space requirements can be significant Disposal of undersize adds to the overall plant operating costs
Opportunities	New screen materials to improve overall separation efficiency and at the same time lower costs Variable speed hydraulic vibrator is being developed and used to optimise the performance for vibrating screens Technology could become more important in the future if high pressure gasification market expands, as feedstock blockages have historically been a major reason for gasifier downtime Can be integrated onsite at a conversion plant as a first filtering step Likely to continue to have a role to play in dealing with waste wood, which has very different sizes

Barriers	Lack of standardisation in chip supply and heterogeneous sizing. Until the market starts to demand more tightly specified clean wood chips, technology will not have a major role to play

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[189] Tchobanoglous,G., Kreith, F. (2002) “Handbook of solid waste management”. Second Edition.

5.5 Chipping

5.5.1 Technology description

Chipping is the process of reducing log wood and brash to chips. This is achieved with a machine called a chipper, which can be either stationary or mobile. The latter can be mounted on a trailer, a truck or a tractor. Chippers can have their own engine or use the power take off from a tractor. Any resultant emissions can be controlled to meet the legislative requirements and therefore this factor does not inhibit technology selection.

There are three basic generic types of chipper[190], based on the chipping mechanism available: disc, drum and screw.

Disc chipper

The disc chipper is the most common type of chipper. The mechanism is based on a spinning disc with various blades mounted at set angles on the face of the disc. Material to be chipped is fed towards the blades at a certain infeed angle. The rotating knives cut woody material into chips as they pass an anvil or a fixed knife. Blower paddles on the back of the disc accelerate the chips up to a spout where they are discharged.

The size of the chips is influenced by several factors:

- The speed of infeed: the faster the infeed, the larger the piece size of the average chip, while a slower infeed may produce a smaller, more consistent chip;
- The number of blades: fewer blades produce larger chips;
- The speed of disc rotation: the higher the speed, the smaller the chips. However, a minimum speed of rotation must be maintained to ensure enough air movement for chip outflow;
- Angle of infeed: an acute angle produces shorter chips.

Drum chipper

In a drum chipper, knives are mounted on a rotating drum. Wood to be chipped is fed in and cut against an anvil. The chip size can be adjusted. Separate blowing provision is required.

Similar to disc chippers, factors affecting the size of chips are:

- Infeed speed;
- Number of blades and their angle with the drum surface;
- Speed of drum rotation.
- Drum chippers can produce more heterogeneous chips compared to disc chippers.

Screw chipper

The screw chipper is the least common in the UK. It consists of a spinning conical screw with cutting edges. Blade alignment is fixed as the cutting edges are integral to the screw. The only method for adjusting chip size is by adjusting the screw pitch. Screw chippers are particularly suitable for processing full trees and logs and produce larger chips compared to disc and drum chippers.

5.5.2 Development status and timescales

The provision of machinery for chipping is a very mature market, having been developed in the early part of the 20th century with the development of the pulp and paper industry. The technology is therefore seen to be TRL 9. Countries with well-developed forest products industries such as Scandinavia, USA and Russia have all developed chippers for use in the forest and in the pulp mills. Latterly, development of chipping equipment has continued in Europe, in countries such as Germany, Switzerland and Austria. Deployment in the UK has historically been more limited and so development of industrial chipping equipment has not taken place. Industrial chippers can be obtained in a wide range of capacities. As an indication of scale, chippers are available in capacities ranging from about 1 tonne/hr to 70 tonne/hr and chipper economics improve with increased capacity and utilisation factor.

Industrial chipper delivery times can typically be between 1 month and 6 months depending upon size, with longer delivery periods required for units with higher capacities.

5.5.3 Impact of different feedstock parameters on operation and cost

The key parameters affecting chipping are the size, particularly the diameter, of log to be processed, and the type of feedstock. The moisture content of the feedstock does not affect the feasibility and the economics of the technology. Each chipper is designed to process up to or within a range of log diameters. Smaller chippers can typically process up to 20cm diameter and larger machines can chip trees with log diameter above 30cm.

Feedstock diameter determines the power requirement and the output of the chipper. To optimise output, the average size of the material to be chipped should be used to decide the power requirement of the chipper. Output is drastically reduced towards the higher ends of a chipper capability, suggesting that sizing the chipper based on the largest piece to be chipped is a cost-effective approach. Output increases and cost/m^3 decreases with increased chipper scale/power.

Miscanthus is a thin-stemmed plant not suitable for conventional chippers and therefore relies on separate processes such as adaptations to combine harvesting, specialised rotary crop headers[191], or baled shredding, in order to achieve size reduction.

However, SRC Willow, SRF and LRF can all be chipped using a conventional chipper. Chipping also plays a crucial role in the harvesting process. In particular, the biomass can be harvested with a forage harvester and then conventionally chipped with chips being blown into an accompanying tractor trailer which transports the chips to an interim or final storage site.

The cost curve in Figure 160 corresponds to the total investment cost for a chipper. It has been derived from three quotes provided by suppliers for mobile machines run on a diesel engine (hence there is no curve displayed here – but the underlying dataset has derived a cost curve). It is worth mentioning that the operating cost of producing clean chips via mobile equipment (i.e mobile chipper vs stationary chipper) is potentially c.30% higher than when using stationary equipment due to higher cost of logistics.

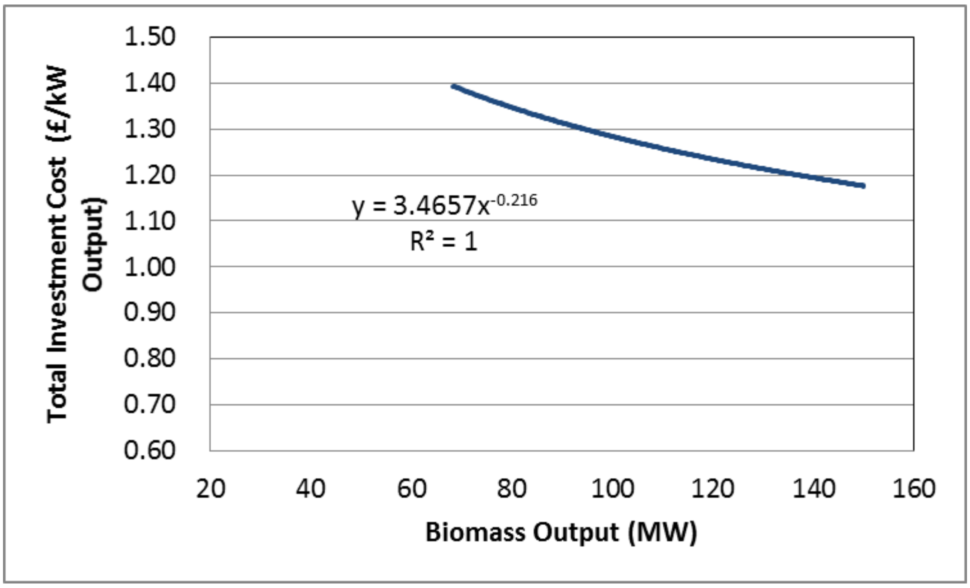

Figure 160: Chipping total investment cost vs. output capacity (based upon 2015 supplier data for mobile plants run on red diesel and B&V estimates for all-in overnight EPC costs, and power law scaling)

Operational cost for chipping includes fixed costs (insurance, maintenance parts and labour) and variable costs (diesel, operations labour). The curve in Figure 161 shows the total operational cost for a chipper. The curve has been derived based on information provided by suppliers and assumptions made on labour requirements. A 1% of total installed costs has been assumed as insurance cost.

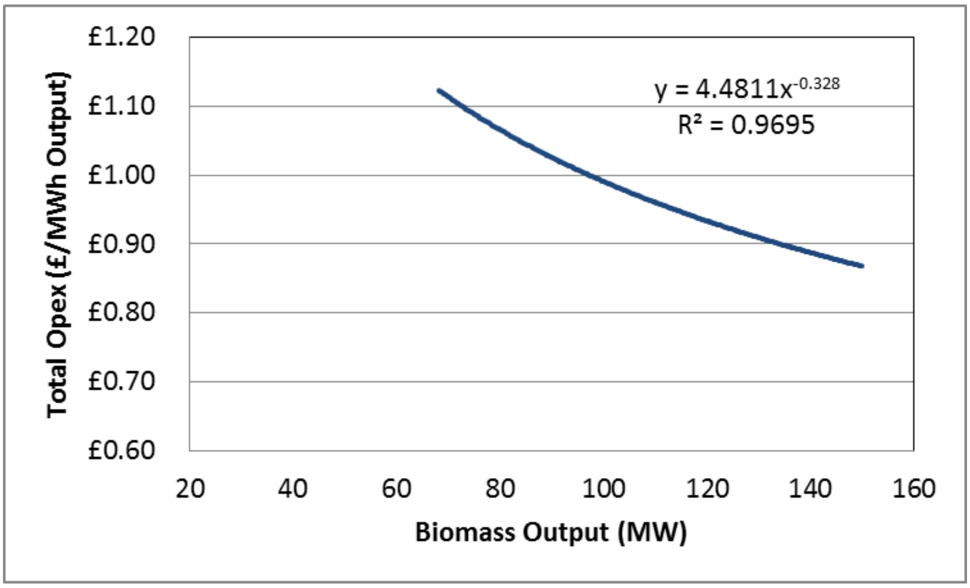

Figure 161: Chipping total opex vs. output capacity (based upon 2015 supplier data for mobile plants run on red diesel and B&V estimates and power law scaling)

5.5.4 SWOB analysis

Chippers facilitate material handling and provide a biomass feedstock that can comply with the requirements of boiler technology, especially small size biomass boilers which require a high degree of consistency of the feed.

Strengths	Fundamental technology in the supply chain of wood fuel for energy conversion technologies, both in terms of forming a part of many harvesting operations, and as a first step for other pre-processing technologies (for example pelleting) Mature, well-understood technology, with multiple suppliers Ease of handling, transporting and storing wood chips rather than wood logs Large surface to volume ratio means wood chips can be burnt efficiently
Weaknesses	Low bulk density of chips compared to other pre-treated forms like pellets leads to expensive transport Overall non-homogeneous wood chip sizes, which may require further grinding before a down-stream wood conversion process (for example pelleting). A sieve can be installed in the chipper to select the chips based on their size, but this lowers the throughput, hence the productivity of the machine is decreased as a result Not particularly well suited to thin-stemmed plants, for example Miscanthus and grasses
Opportunities	Flexible technology that can be adapted and integrated into a number of processing opportunities
Barriers	Main barrier is economic: growth of capacity in the UK is dependent upon the growth of woodfuel heating to allow investments to be made in chipping capacity Wood supply chains are relatively small and fragmented in the UK, making difficult to provide long term “bankable” supply contracts. This has led to significant amount of wood chips/pellets being imported from overseas into the UK

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[190] Forest Research (1998), “Woodfuel Chipping: Field Trials”.

[191] Kemper (2015), “Seasonal Tips and Tricks”, Available at: http://www.kemper-stadtlohn.de/en/home/service-and-parts/seasonal-tips-tricks.html

5.6 Pelleting

5.6.1 Technology description

Pelleting is the process of compressing fine wood particles into pellets to produce a homogenous fuel with high energy density. Pellets are characterised by uniform size and chemical composition. They are used as a combustion fuel for power, CHP and heat purposes.

The main production steps are described below[192]:

Chipping. The feedstock is reduced to chips. Chippers can be set to achieve the desired particle size, typically below 30mm.

Screening. Feedstock chips are screened to filter out oversized particles.

Drying. The moisture content of the harvested and chipped feedstock needs to be reduced to around 10%. If the material is too dry, the surface of the particles may carbonize in the pelleting stage. Conversely, if the feedstock is too wet, moisture contained in the particles cannot escape, making them mechanically weak.

Grinding. Feedstock chips need to be ground down with a hammer mill to reduce and homogenize their size. The size of the raw material that needs to be provided to the pellet mill depends on the required diameter of pellets. In general, the particle size should be below the diameter of the hole in the die, hence the diameter of the resulting pellet.

Conditioning. Steam is used to condition the particles and soften the lignin, which aids in the binding of the particles.

Pelleting. The conditioned particles are fed to the pellet mill. Inside the mill, rollers extrude the mix through a perforated flat or ring die, which effectively compresses the product into pellet form. During this process the lignin, naturally contained in the biomass, melts and acts as a further 'binder' sticking the fibres together. In the case of lower lignin feedstocks, a binder, typically starch, is added to facilitate this process.

Cooling. The temperature of the wood increases during the pelleting step. Pellets are cooled with air to guarantee high durability.

The technologies used in the pelleting process are well developed and any resultant emissions can be controlled to meet the legislative requirements and should not inhibit technology selection.

5.6.2 Development status and timescales

Pelleting is a mature and well-established technology (TRL 9). Production capacity is either large scale (feeding power plants for electricity generation and other uses) or small scale (feeding domestic applications). The demand for wood pellets is set to increase, both in the UK and globally, due to renewable energy generation targets, with world production of wood pellets currently (2015) less than c. 25 million tonne per annum.

In Europe, the first wood pellet plant was built in Sweden in 1982 and currently Sweden is the largest producer of wood pellets in Europe, followed by Germany and Austria.

The UK relies heavily on imported pellets, mostly from the US. Wood pellet production started in the US in 2002, with the largest plant built by the year 2005 only having a capacity of 55,000 tonne per annum.

Globally, the first very large scale wood pellet plants were built in Canada in 2001, followed by installations in the US and Australia in 2008. Today, the world's largest pellet plant is in the US state of Georgia with a processing capacity of 750,000 tonne per annum, having started operations in 2011. Drax has also invested in two 750,000 tonne per annum pellet plants in Louisiana and Mississippi.

Pelleting of agricultural residues and Miscanthus is a more recent development, with only a handful of smaller plants in existence. One example is the 90,000 tonne per annum Goole pellet mill that feeds Drax, which was designed for straw, but has also been incorporating volumes of Miscanthus. Terravesta also run a smaller pellet mill for Miscanthus in Kimbolton.

Commercial scale wood pellet plants can be built in the range starting from 800 tonne per annum, although there are some suppliers offering smaller mobile or containerised mills. A typical industrial plant is sized to produce around 200,000 tonne per annum and takes around 18 months to develop and build[193].

5.6.3 Impact of different feedstock parameters on operation and cost

All the feedstocks considered (Miscanthus, short rotation coppice Willow, short rotation forestry coniferous and deciduous) are suitable for pelleting. The feedstock-dependent variables that affect pelleting process efficiency and pellet quality are discussed below.

The moisture content of the feed affects pellet quality and the process productivity. In particular, a raw material with moisture content c. 1% away from the ideal can force the pellet mill to use up to c. 20% more power per tonne during production. It is crucial to dry the feedstock to the accurate moisture content (usually around 10%) to reduce energy costs.

The lignin content in biomass varies considerably between the different feedstocks. A higher lignin content translates into a malleable and more binding feed into the pellet mill. This in turn determines the amount of binder needed and the temperature of conditioning. In particular, binders need to be added at specific percentages to reduce the load on the pellet mill motor. The right amount of binder and the correct temperature of steam conditioning are fundamental to ensure good product flow with good compression and to avoid stalling the pellet mill drive motor.

The level of pressure during compression depends, among other things, on the type of feedstock. In general, harder woods require higher pelleting pressures. Underestimating the pressure needed can lead to "clogging" of the holes in the die and to interruptions in the pelleting process.

The distance between the roller and the die affects the quality of pellet, the wear on machine parts, and the energy consumption. In particular, trials have shown that an increase in the distance between 1 and 10 mm can cause an increase in energy consumption of c. 20%, although it also reduces the fine particles size (dust) by c. 30%.

Pelleting plant availability will be dependent on life of the die for each feedstock, which depends on the grit content. Lifetime will not vary between feedstocks as rollers and dies are changed frequently.

In summary, different feedstocks will require different binder quantities, conditioning temperatures and compression pressures. Different feedstock will also require different drying requirements, distance between roller and die, size of die and rollers, and milling speed. Each of these variables need to be tailored accurately to the specific feedstock treated to ensure quality of the pellets and to optimize the cost-effectiveness of the pelleting process in terms of energy costs, maintenance costs and productivity.

The cost curve in Figure 162 shows the total levelised capital cost for the pelleting process, including the preparatory processes of chipping and screening. The curve has been derived from quotes provided by suppliers and includes capital costs for chipping and screening from mobile plants, drying from 50% to 10% moisture content via drum dryer and pelleting equipment.

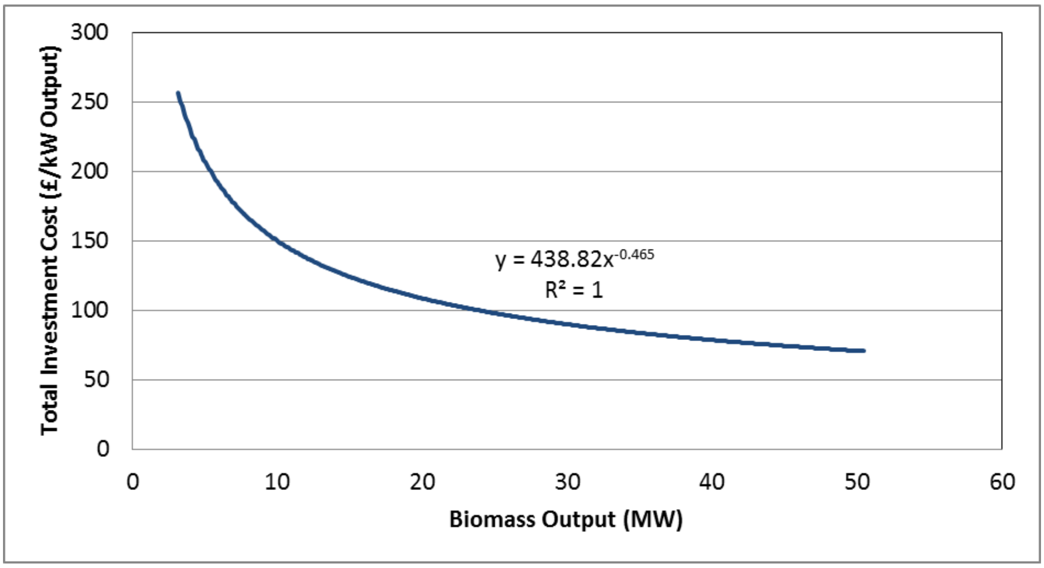

Figure 162: Pelleting total investment cost vs. output capacity (derived by B&V from 2006 supplier data corrected to 2015 prices and B&V estimates for all-in overnight EPC costs)

Total operational costs for a pelleting process include fixed costs (insurance, maintenance parts and labour) and variable costs (diesel for chipping and screening, electricity, natural gas for drying, operations labour, binder, waste disposal). The curve in Figure 163 shows total operational costs for the pelleting process assuming an initial (prior to chipping) feedstock moisture content of 50%, 40% and 30% respectively. The curve has been derived based on information provided by suppliers and assumptions made by B&V on labour requirements. A 1% of total installed costs has been assumed as insurance cost.

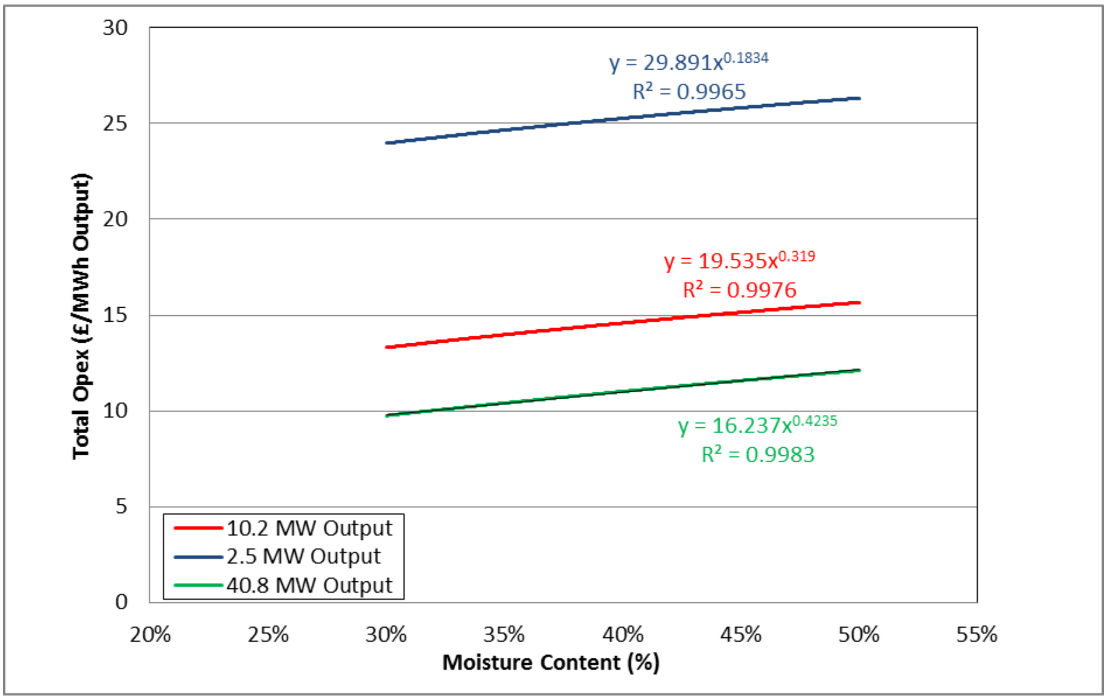

Figure 163: Pelleting total opex vs. feedstock moisture content for a given biomass output (derived by B&V based upon 2015 supplier data and B&V estimates)

5.6.4 SWOB analysis

Pelleting allows the use or residues from other processes, producing products with higher energy densities. In addition, it generates a product with standard size and composition that allows automatic feeding.

Strengths	High energy density pellets that can be easily handled, cheaply transported and stored Well-established and competitive technology Highly standardised, uniform product that is increasingly commoditised and widely traded
Weaknesses	Energy intensive, with largest load constituted by drying Pellet durability is typically achieved at the cost of throughput. Depending on the use requirement of the pellets and the quality needs (long distance transport and handling as opposed to direct combustion), a choice must be made between quality of pellet with a high percentage of durability and the throughput achieved during the pelleting process High sensitivities of pellet quality and operation costs to the type of feedstock processed and the need to accurately tailor process variables to each feedstock Wood pellets cannot be stored outdoors as they absorb moisture and disintegrate Poor durability leads to increased dusting levels and fire risks. In addition, the decomposition of pellets during bulk storage can lead to self heating, self ignition

	and off-gassing resulting in increased fire risks and fatalities due to the presence of CO, CO_2 and CH_4 in non-ventilated storage units
Opportunities	Stringent targets on renewable energy generation, both electricity and heat, which will increase the global demand for wood pellets – not just in Europe and the US, but also increasingly Asia (e.g. South Korea) Closure of numerous paper and pulp mills in forested areas (e.g. SE USA) as paper demand falls, or diseased forestry, is leading to undermanaged forestry and available resources Large number of coal conversion projects ongoing to make use of wood pellets in mass volume Use of additional biomass for drying leads to much lower GHG pellet production than using natural gas
Barriers	To ensure sustainability, new forestry either needs to be planted at the same rate as wood is harvested for pelleting, or the forest carbon stock levels managed to ensure falls are avoided – highly complex area with unclear methodologies and counterfactual uses in play. Carbon debt and forestry management issues are being debated at present in policy circles (with considerable NGO pressure on the pellet industry), industry currently uncertain if additional GHG accounting rules might be introduced. However, this issue is not specific to large pellet plants, it applies to any large scale biomass project. Pellets remain more expensive than coal and even many heating oils, particularly in the current low commodity price environment There is a limit to the quantity of virgin wood available for pelleting in the UK, as well as currently a very limited supply of perennial energy crop

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[192] Van Loo; Koppejan (2008), "The handbook of Biomass Combustion & Co-firing".

[193] Laborelec GDf Suez (2012), "Industrial Wood Pellets Report".

5.7 Pyrolysis

5.7.1 Technology description

Pyrolysis can convert biomass feedstocks into three product streams: liquid (bio-oil), gas and solid product (biochar). The distributions and properties of these product streams are strongly dependent on the pyrolysis conditions and the feedstock composition[194].

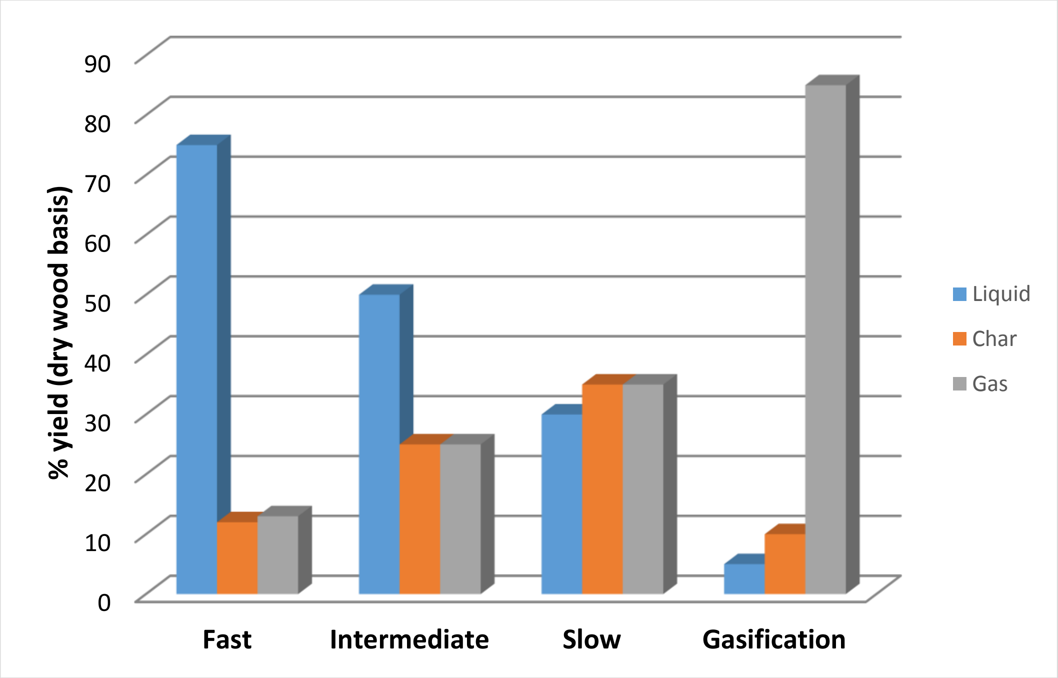

Figure 164: Typical product yields from different modes of wood pyrolysis[195]

Many variables control the outcome of the pyrolysis process, the most important consisting of moisture content, particle size, the pyrolysis temperature, the heating rate and the residence time at this higher temperature. As illustrated in figure 165, fast pyrolysis leads to the highest yield of a liquid bio-oil product, with any gaseous or solid by-product potentially of use as an energy source for the process itself. In this process, vapour residence times are short, of the order of seconds and the temperature is moderate, of the order of 500°C[195]. The main steps to be considered in a fast pyrolysis process are as follows[196]: Drying is desirable to reduce the water content which otherwise ends up as a component of the produced bio-oil. An optimal level of between 10 to 15 % has been reported[197] as lower values increase the viscosity of the resulting bio-oil. Grinding of the feedstock to a small particle size of 1 to 10 mm is necessary in order to maximise the efficiency of the pyrolysis process in the conditions of a very short residence time at the moderate temperature. Rapid quenching of vapours is important to minimise secondary cracking reactions, and after the pyrolysis process, an efficient means of liquid recovery and separation of solid by-products are required.

Reactor Types

A variety of reactor types have been employed in fast pyrolysis processes[195],[197]. The main large scale systems are typically either Bubbling Fluid Beds or Circulating/transported Fluid Beds (as illustrated in Figure 166), along with Rotating Cone reactors.

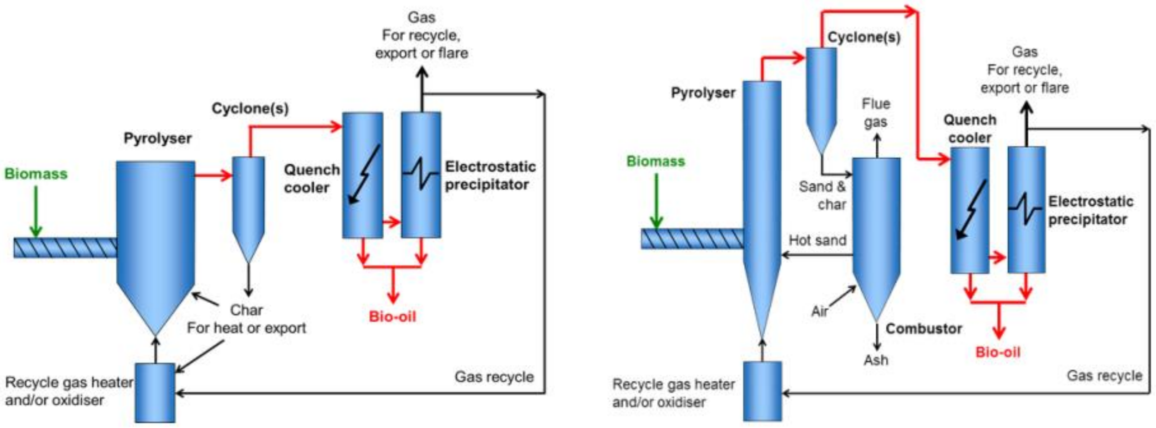

Figure 166: Bubbling Fluidized Bed Reactor (left), Circulating Fluidized Bed Reactor (right)[198]

Bubbling fluidized beds are a well understood technology with established scaling laws, good temperature control and efficient heat transfer to biomass particles, capable of liquid yields typically of 70 – 75% by weight. In the context of fast pyrolysis, suitable particle sizes are limited, of the order of 1 to 5 mm[197]. Circulating fluidized beds share many features, the main difference being the char residence time being shorter (of a similar order to the vapours and gases). Rotating Cone reactors originated from research work at Twente University[199], with similarities to Circulating Fluidized Bed reactors, but different in that biomass and sand particles are transported through centrifugal forces in a rotating cone without the use of a gas. Due to developments in the technology, rotating cone reactors can accommodate larger particle sizes, up to 10 mm[200].

Other types of reactor considered for fast pyrolysis include Ablative Pyrolysis reactors in which the biomass feedstock is in direct contact under a high contact pressure against a hot moving surface. This allows larger particle sizes to be used, and has no inert gas requirement thus allowing smaller processing equipment and more efficient product collection. However, contact surface area controls the process and so scaling is problematic, and the mechanically driven reactor increases the complexity. Entrained flow reactors are relatively simple, but are at a disadvantage because of their lower heat transfer rate, requiring large gas flows to compensate, thus leading to large plant size and a less effective liquid collection. Vacuum pyrolysis reactors have an advantage in being able to handle larger particle sizes, with reduced char in the bio-oil product, but they are complex and costly as a result of the high vacuum requirement. Screw and Auger kilns involve the mechanical movement of biomass through a hot reactor, however they are not suitable to achieve the very short residence times comparable to fluid and circulating beds. Feedstock boundary conditions are summarised in Table 27.

Table 27: Fast Pyrolysis feedstock limitations

Parameter	Value
Size range (mm)	< 10 (rotating cone), < 5 (BFB)
Moisture content fed to reactor (%)	10 - 15

5.7.2 Development status and timescales

While conventional pyrolysis has a long history, the technological development of fast pyrolysis began in the 1980's. Miguel et al.[197] discuss the commercial status of fast pyrolysis and summarise the status of 26 demonstration and commercial plants constructed up to 2012. Some notable examples are Dynamotive Energy Systems, who constructed two BFB based commercial plants in 2005 in West Lorne, Ontario and 2006 in Guelph, Ontario, although both plants have now ceased operation. BTG, employing rotating cone technology, commissioned a 50 ton/day plant in Malaysia processing palm oil empty fruit bunches in 2005. In collaboration with a number of partners, BTG are also involved in the EMPYRO project[201], which opened a 120 tonne/day capacity plant in May 2015. Ensyn use a transported bed technology, with a plant at Renfrew in Canada that was recently upgraded to a capacity of 45 tonnes/day, and are in a joint partnership with Fibria that plans to locate a 249 tonnes/day fast pyrolysis plant in Brazil directly linked to a refinery co-processing operation for the production of transport fuels. Ensyn also has a 4 barrel/day pilot operating in Hawaii, with UOP carrying out the subsequent upgrading step. In addition, Ensyn has been involved in the design of commercial plants based on transported fluid bed technology[197] used by Red Arrow in the production of food flavourings.

So in summary, whilst many demonstration and first commercial (30 – 192 tonnes/day) plants have been constructed, predominantly in the USA and Canada so far, the technology has still to be fully commercialised and is at a current TRL of 8. Based on industry experience and project construction timelines, it could be expected that commercial scale plants will be replicated within the next 10 years, in order to progress to TRL 9 by around 2025.

The fast pyrolysis bio-oil product is proven for use in combustion heating applications, with Ensyn claiming to have sold more than 20 million gallons into the industrial heating fuels market. Whilst there is considerable interest in upgrading bio-oil to transport fuel applications in the industry, this is yet to happen at commercial scale and is out of scope of this study.

5.7.3 Impact of different feedstock parameters on operation and cost

Bio-Oil Properties

The liquid product of the fast pyrolysis process, commonly referred to as bio-oil, is very different to conventional hydrocarbon fuels. It consists of a mixture of complex carbohydrates and lignin-derived aromatics in water. The organic fraction has a much higher oxygen content than mineral oils (comparable to that of the biomass feedstock) and high water content along with a greater acidity. Stewart[202], from a compilation of the data that are available in the Phyllis2 database[203], produced a summary of the ultimate analyses of bio-oils and for comparison included the ASTM D7544 specification of pyrolysis liquid biofuel, as given in Table 28. While the heating value of the bio oil is relatively low on a mass basis (similar to solid biomass), the volumetric energy content increases markedly.

Table 28: Ultimate analyses of pyrolysis oil and other oils[202]

Component	Pyrolysis Oils (mean values)	ASTM D7544	Fuel Oil	Gas Oil
Water content wt.%)	18.8	≤30	0.02	<0.5
Volatiles (wt.% daf)	66.8			
Ash (wt.% dry)	0.1	≤0.25	<0.05	<0.08
LHV (kJ/kg)	18320		40800	42900
C (wt.% daf)	56.2		87.7	87.2
H (wt.% daf)	6.65		10.3	12.8
O (wt.% daf)	36.2		1.2	<0.1
N (wt.% daf)	0.57		0.48	<0.01
S (wt.% daf)	0.05	≤0.05	0.7 (≤1)	0.1 (≤0.1
Cl (wt.% daf)	0.1			
Pyrolysis solids		≤2.5		

In principle, there are no restrictions on feedstock type. However, to optimise the pyrolysis oil yield requires that the feedstock be dried to about 10% moisture and be ground to particle sizes of 1 to 10 mm, dependent on reactor technology as discussed earlier. Ash content is desired to be minimised as it impacts the efficiency of the conversion process, reducing the bio-oil yield at the expense of increased char formation. Different feedstocks have an impact on the concentrations of the minor components of ash, nitrogen, sulphur and chlorine, and water content may vary substantially. Stewart[202] tabulates examples extracted from the Phyllis2 database based on a wood or herbaceous crop feedstock. Consequences of this different composition are that the pyrolysis oil could be around 40% of the heating value of fuel oil, it will be immiscible with hydrocarbon fuels and the stability will be lower, both physically in terms of a tendency to separate into phases and chemically it will be prone to autoxidation and polymerisation.

Technology Improvements

As discussed previously, the technology itself is insensitive to feedstock, in that provided the feedstock is pre-processed to a certain size range and moisture content, it can be used in the fast pyrolysis process. The aspect of the feedstock that is of most concern is the ash content as this directly impacts on the relative yields of bio-oil and char. It is relatively straightforward to control the feedstock particle size and moisture content, but significant reduction of ash would require an additional washing of the ground material followed by a re-drying back to the 10-15 % moisture content level. Unlike a standard BFB reactor, that may have very high operating temperatures, the intermediate range of temperatures involved in fast pyrolysis of the order of 500°C mean that the impact of the ash fusion on the plant operation and lifetime is not so severe.

The most promising technology improvements that might be possible revolve around the improvement of the properties of the produced bio-oil as discussed by Venderbosch and Prins[200]. They noted that the bio-oil composition is extremely complex, is difficult to analyse quantitatively, and contains a range of all types of possible oxygenated compounds that lead to some of the undesirable properties of the bio-oil such as its poor stability and low pH. If the role of the various oxygen functionalities in the oil could be clarified, and control of the pyrolysis process could be improved, for instance by the use of catalysts and also the tuning of the condensation process, then the bio-oil properties may be significantly enhanced.

Plant size

Plant size strongly affects the Capex, as shown in Figure 167.

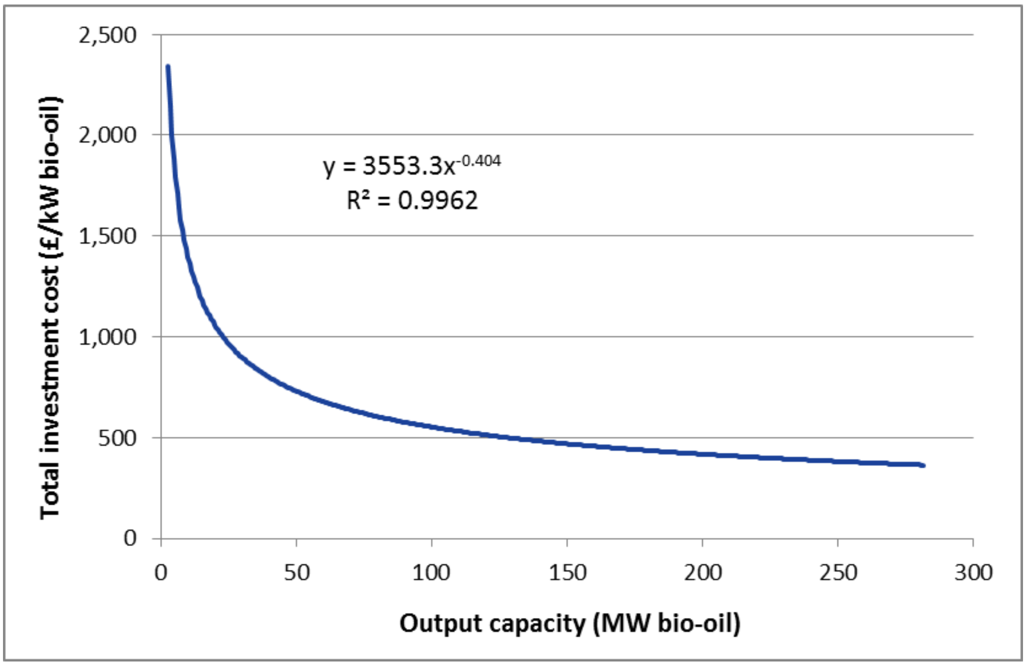

Figure 167: Pyrolysis total investment cost vs. plant output capacity (based on [204])

This data originates from a study by Braimakis et al.[204] who investigate a range of plant size from 1 to 80 dry tonnes/hr of input biomass feed rate. In that study, they proposed a number of discrete equations describing the costs of the pretreatment, pyrolysis, steam and power generation unit, and cooling tower unit and other utilities. The base case scenario in this calculation represents an output capacity of 28 MW consistent with the largest currently operational plants at about 10 dry tonnes/hr input biomass feed rate, with an arbitrary range of a factor of ten smaller or larger.

Feedstock Conditions

Moisture Content

The moisture content affects the efficiency due to the latent heat of vaporisation of the water in the biomass. The relationship is illustrated in Figure 168. The base case efficiency of 65% at an ash content of 2 % comes from a study of Rogers and Brammer[205] who reported correlations of bio-oil yield as a function of ash content.

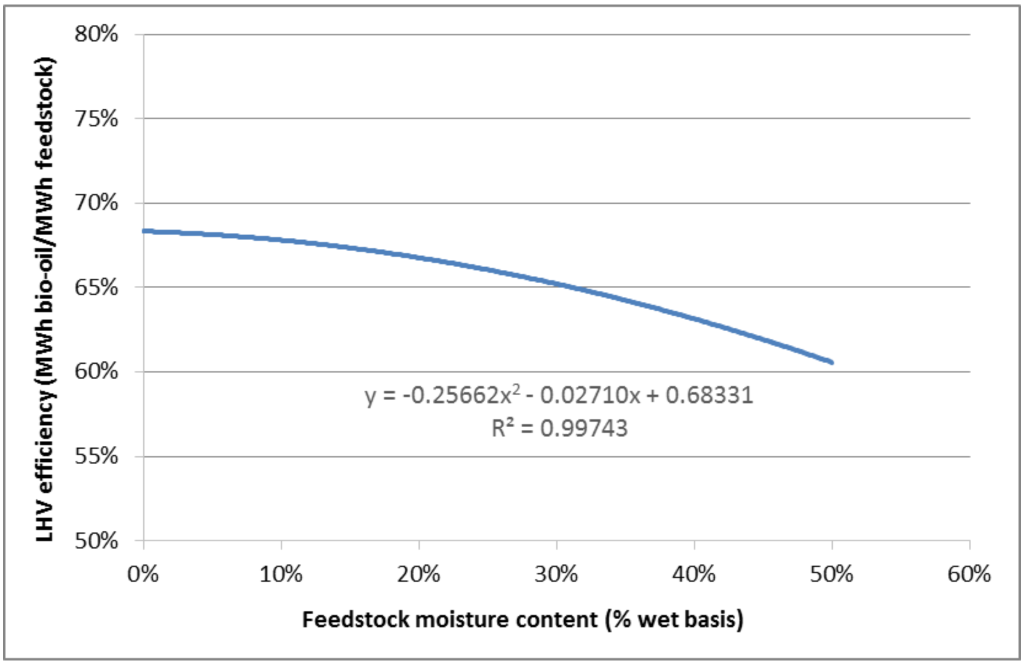

Figure 168: Pyrolysis efficiency vs. feedstock moisture content (based on ICON energy balance calculations)

Feedstock Composition

Ash Content

The ash content affects the bio-oil yield, with increasing ash favouring char production over bio-oil. Using a correlation based on work by Rogers and Brammer[205], the effect on efficiency of bio-oil production as a function of ash content can be calculated as shown in Figure 169.

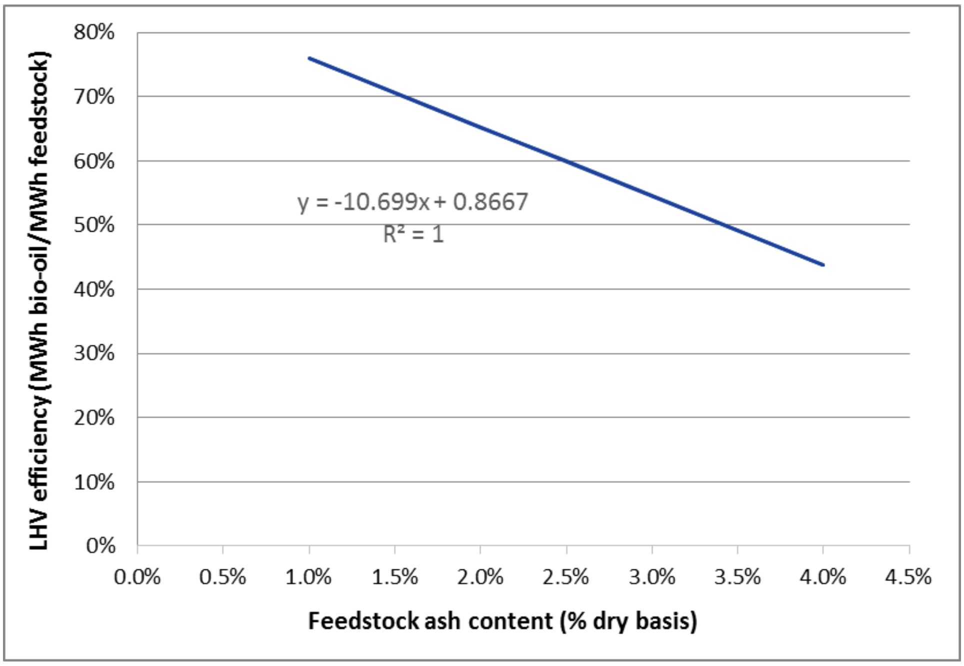

Figure 169: Pyrolysis efficiency vs. feedstock ash content (based on [205])

The other effect of ash content is on the total Opex. The base Opex consists of: insurance (1% of TIC), maintenance – parts (2% of TIC), maintenance – labour 1% of TIC[205], operations (labour) and reagents/materials[206]. Assuming an ash disposal cost of £120/tonne, ash disposal costs increase linearly with feedstock ash content. However, by considering the simultaneous effect of the biomass ash content on the pyrolysis system efficiency, a non-linear function can be derived, as shown in Figure 170. The cubic function shown is a reasonable fit over the given range.

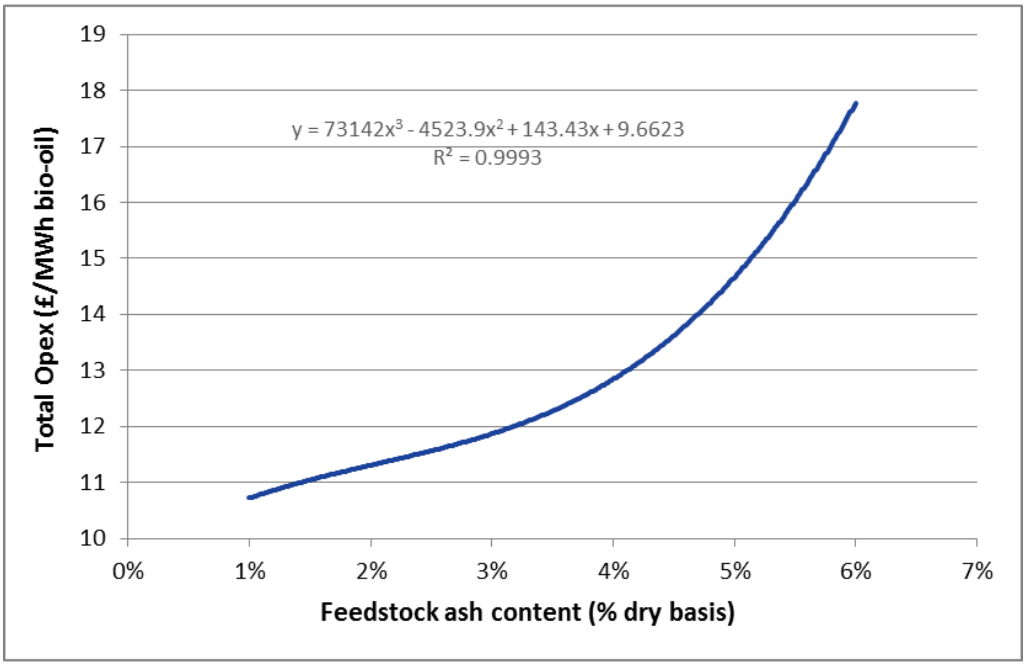

Figure 170: Pyrolysis total Opex vs. feedstock ash content, including the impact of ash on efficiency (based on [205, 206] and calculations from section 2.4.2.3)

Potential emissions of NO_x, sulphur and chlorine into the atmosphere are accounted for by a procedure of calculating the Opex for the mitigation of each component. Applying a urea based NO_x mitigation cost, the effect on total Opex of the level of nitrogen present in the feedstock can be calculated, and is shown in Figure 171. In this case, the base nitrogen content is taken as 0.265% and is less important than the ash effect.

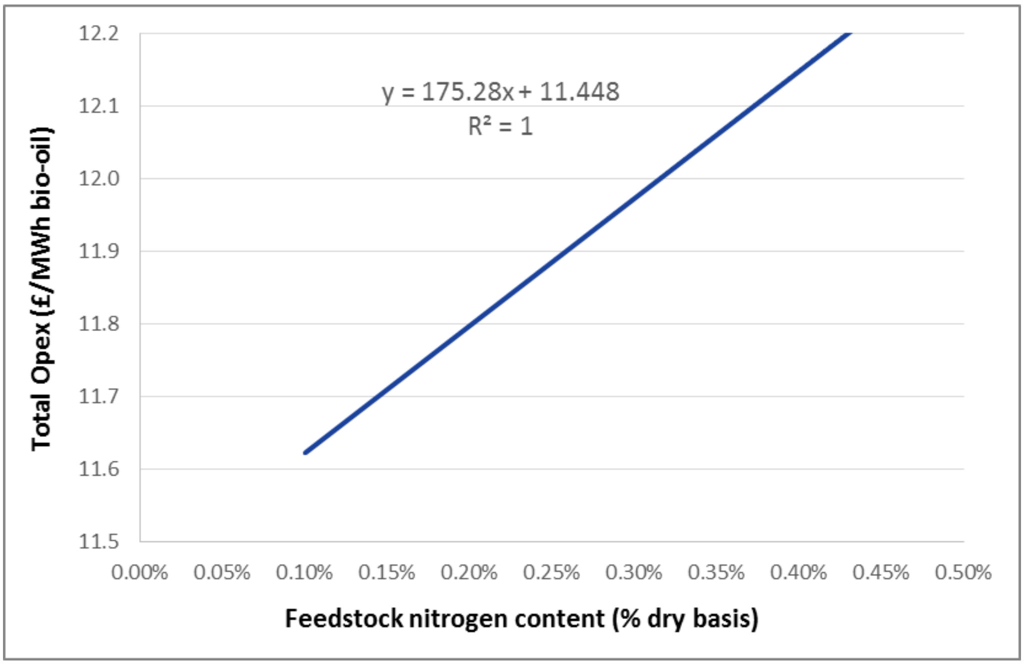

Figure 171: Pyrolysis total Opex vs. feedstock N content (based on [205, 206] & ICON calculation based on section 2.4.2.4)

Following a similar principle based on the use of lime to mitigate sulphur and chlorine in the feedstock, the effect on total Opex of the level of sulphur (base case 0.04 %) and chlorine (base case 0.03 %) present in the feedstock can be calculated and is shown in Figure 172 and Figure 173. These in turn are less significant than ash or nitrogen impacts.

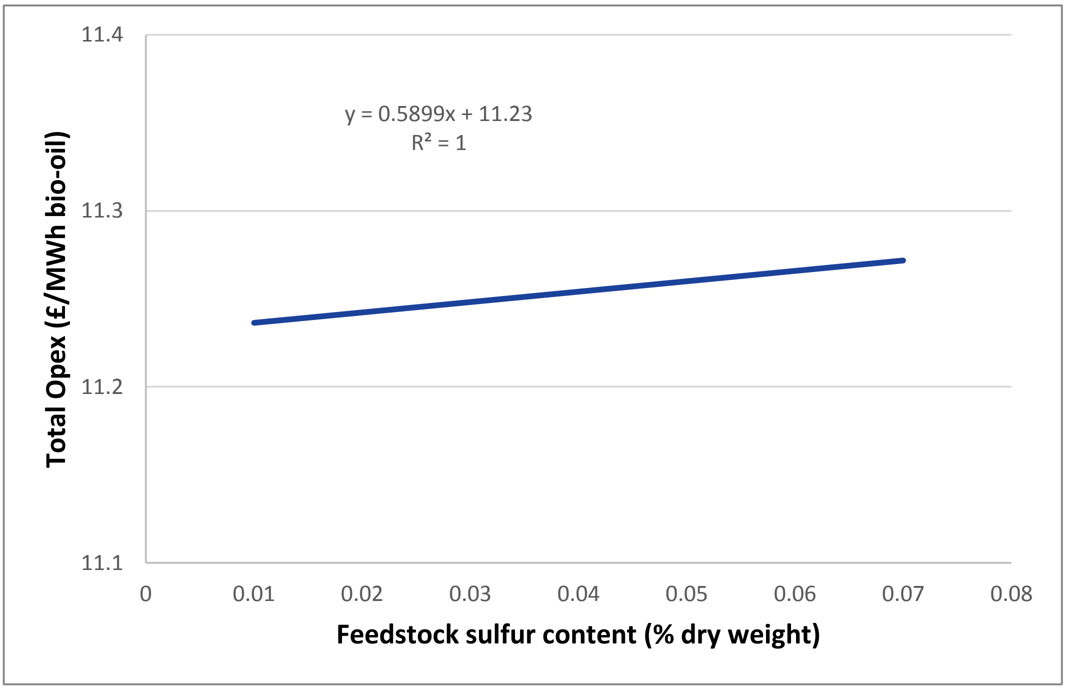

Figure 172: Pyrolysis total Opex vs. feedstock S content (based on [205, 206] and ICON calculation based on section 2.4.2.4)

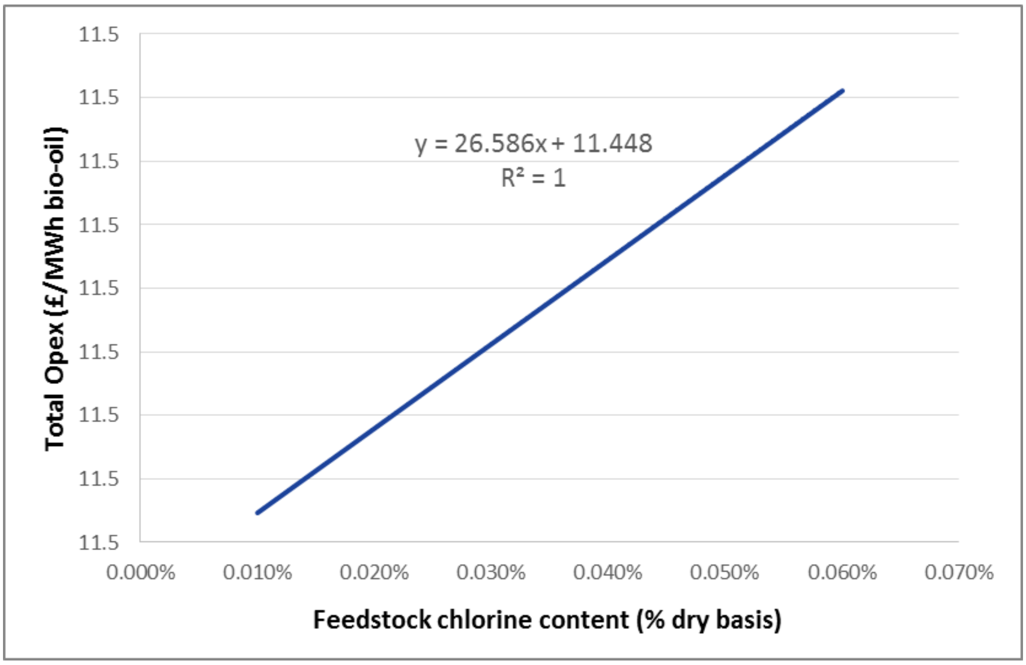

Figure 173: Pyrolysis total Opex vs. feedstock Cl content (based on [205, 206] and ICON calculation based on section 2.4.2.4)

5.7.4 SWOB analysis

Fast pyrolysis is unique in the biomass pre-treatment technologies in that the product is a liquid and of higher energy density than the input feedstock. This gives it advantages in terms of handling, storage and transportation over solid and gaseous products, and it has potential to be used in heat, power, transport fuel and chemicals sectors.

Strengths	A liquid product, storable, pumpable and transportable Increased energy density product Proven technologies for production Mild operating conditions allow exploitation of high ash content biomass
Weaknesses	Lower quality than conventional liquid fossil fuels; viscous, acidic, higher oxygen, solids, ash and water content. Not able to drop-into existing fossil fuel supply chain infrastructure Some protection against corrosion required in pumping and tanks Low pH value of 2 to 3 requires additional safety measures in the handling and storage. Unstable, more prone to autoxidation/polymerisation and phase separation (particularly with changes in temperature) Long term operation of large scale reactor systems is yet to be proven

	Feedstock requirement of small particle size and low moisture content
Opportunities	Upgrading potential of bio-oil to diesel/gasoline fuel, or other valuable chemicals, potentially with an integrated process Build on demonstration scale plants Potential for building a network of small and medium scale decentralised plants, using local feedstock (of low energy density), before transporting the bio-oil to a centralised conversion plant Most entrained flow gasifiers able to take pumped liquid, and slagging design will cope with the high ash contents
Barriers	Expensive in comparison to conventional fossil fuels Process implementation by developers and application development by users Improving product quality, product regulation and standards Health and safety issues in handling, transport and usage Not straightforward to upgrade the bio-oil if trying to produce a transport biofuel (via catalysis with hydrogen, or potential for crude oil refinery integration), and the transport biofuel market was expected to be one of the main markets for pyrolysis technology As a pre-processing technology, this may be excessively complicated and costly compared to alternative pre-processing options.

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5.8 Torrefaction

5.8.1 Technology description

Torrefaction is a thermal process that involves heating biomass (typically in chip form) to temperatures between 250 and 300°C in an inert atmosphere (limited oxygen). When biomass is heated at such temperatures, the moisture evaporates and various volatile low-calorific components contained in the biomass are driven out. Between 80% - 95% of the original heating value is retained by the biomass after torrefaction, depending on the process severity. Torrefied biomass chips can then be more easily ground into powder, before pelletisation, or alternatively can be made straight into briquettes.

The technologies used in the torrefaction process are well developed in other thermal applications. Any resultant emissions can be controlled to meet the legislative requirements and should not inhibit technology selection.

Torrefaction Equipment

The torrefaction technology providers offer a variety of physical configurations with vertically upward, vertically downward and horizontal movement of the raw feedstock in both batch and continuous processes. Most of the providers utilise some form of heat integration where the volatile gases driven from the biomass along with some biomass fuel are used to generate the heat needed for pre-drying and torrefaction. The styles of equipment used to torrefy biomass are numerous and include (somewhat in order of popularity):

- Rotary horizontal drum;
- Screw conveyor;
- Multiple hearth furnace;
- Compact moving bed;
- Fluidised bed;
- Belt conveyor;
- Combinations of the above.

The torrefaction process itself usually includes drying of the biomass feedstock. However, prior to this the feedstock will need to be chipped, and if not using clean wood, screening will be also required. The torrefied powder produced can be converted into pellets which will reduce storage and transportation costs.

Torrefaction with Pelletising

Most torrefaction facilities will pelletise their product for ease of handling and shipping. Pelletising requires that the torrefied biomass is pulverized before being introduced to the pelleting equipment. Binders are often used to provide a means to keep the particles together in a mechanically strong lump, able to withstand shipping and handling without significant mechanical failure.

The torrefaction with pelletising process has a significant densifying effect resulting in a product with a density of between 650 - 800 kg/m^3, resulting in a very high energy density compared to wood chips.

Torrefaction with Briquetting

Briquetting is often overshadowed by pelletising, as briquetting is only beginning to emerge in the market place as a viable alternative to pelletising. It offers the advantage of eliminating the pulverizing step prior to pelletising and overall requires less energy than the pelletising process. Briquetting has the disadvantage of slightly lower density, with most briquetting operations resulting in briquettes between 20 – 30 mm in diameter. Torrefied products (pellets and briquettes) are by their very nature very brittle and will break on impact. It is generally accepted that when a briquette begins to break or mechanically fail, it will break into the original torrefied wood chips rather than into the much smaller pulverized torrefied wood particles. Because of this, and because briquettes tend to be much larger than pellets, then the weight percentage of dust generated tends to be less than pellets (where dust is generated through attrition). The resulting chips may stay in a configuration that can still be handled and utilized with the briquettes. However, this concept of briquetting torrefied material has not been extensively field tested.

5.8.2 Development Status and Timescales

This industry has been expanding and developing very rapidly in the last few years, with high levels of industry interest, especially because of the high demand for co-firing biomass in EU and UK power plants. Several ports in the US have begun or are planning major expansions to handle the anticipated torrefied biomass pellet and black pellet (generated via steam explosion) market (although they are mostly set up to handle conventional white pellets at this point).

There are 40 or more entities worldwide that purport to be a torrefaction equipment or product provider. Many of these are undercapitalized and should be considered start-up companies but the industry is growing and maturing quite rapidly.

Several vendors of torrefaction equipment and/or torrefied product continue to build pilot plants, demonstration plants and commercial plants. Some of the leading developers include Andritz, Topell, TSI, Vega Fuels, Solvay, New Biomass Energy and Earth Care Products. There are at least six torrefaction operators that have commercial plants built and operating, with four of these providers operating commercial plants that also include integrated pre-drying of the feedstock. There are multiple other providers that have successful pilot plants in operation.

Probably the biggest torrefaction plant is the Quitman Plant in USA, with a production capacity of 80,000 tonne per annum. The plant was built and developed by New Biomass Energy. Currently the plant is being extended by Solvay to c. 250,000 tonne per annum.

Commercial plants are considered to be those above a scale of 50,000 tonne per annum tonnes output per year. An integrated plant including drying and densifying would take 18-24 months to build. Given the first commercial plants operating, the TRL of torrefaction technology is judged to be TRL 8 when operating on forestry and sawmill residues. As pelleting is a fully commercialised process, the TRL of torrefaction + pelleting is also judged to be TRL 8. It is expected that the industry will reach TRL 9 within the next 5 years, as fully integrated commercial plant designs are replicated.

Commercial scale plants using straw and Miscanthus are yet to be established. Briquetting of torrefied material is also at an earlier stage of development, nearer TRL 5 at present (one Danish funded pilot project by CF Nielson), but with industrial interest from the likes of DONG Energy.

5.8.3 Impact of different feedstock parameters on operation and cost

Calorific Value

The lower heating value of torrefied biomass is between 19 MJ/kg and 23 MJ/kg. The resulting calorific value depends on the feedstock and the degree of torrefaction employed. This increase in calorific value (compared to natural biomass forms) occurs because more oxygen and hydrogen is driven from the biomass than carbon. The oxygen to carbon ratio of dry wood is between 0.6 – 0.75 and the hydrogen to carbon ratio is 1.4 – 1.6 but, once torrefied, these values drop to 0.4 – 0.6 and 0.9 – 1.4 respectively.

Appearance

The dimensions of wood chips used for torrefaction generally range between 10 mm and 40 mm for two dimensions with the third dimension not exceeding 2 mm, depending on the technology utilised for torrefaction. Once torrefied, the biomass chips are considerably darker in colour depending on the level of torrefaction applied, ranging from dark tan to black.

Grindability (ability to be ground or pulverized)

Proprietary information of trials run at commercial scale facilities concludes that the entire pelletizing line (milling, screening, pelletizing, de dusting, cooling and storage) can be up to 20% smaller capacity than for comparable white pellets because of the more brittle properties of torrefied pellets and the loss of lignin’s binding forces in torrefied pellets.

The same source has also reported on attempts to co-fire coal with white pellets using coal mills: there were problems with the mills when the co-firing level reached 8-10%. Whereas, torrefied wood has been co-fired coal with no problem using coal mills.

SGS North America and Wyoming Analytical Laboratories tested the grindability of torrefied pellets and compared the results to the coal industry specification. The specification requires a Hargrove Grindability Equivalent value of >40. The torrefied pellets tests resulted in an equivalent Grindability Equivalent value of >40 matching the value in the specification.

A confidential report advises that there is a substantial reduction in milling energy required when milling torrefied wood and coal as compared with conventional wood pellets with coal. This is one of the reasons that torrefied product is used to co-fire with coal and allows the torrefied product to be pulverised using the coal mills rather than dedicated mills. This will also be very attractive for dedicated biomass plants using pulverised dust combustion technology, or entrained flow gasification, that both rely on powdered material. This low energy input is a significant advantage when grinding the torrefied material in order to make torrefied pellets.

Storage

Trial runs at commercial scale facilities have demonstrated that torrefied biomass has characteristics very similar to coal and therefore can be mixed with the coal (typically in pellet or briquette form) and milled in the coal mills with no limit as to percentage of the fuel mix (100% torrefied fuel has been achieved). Laboratory tests have shown that torrefied particles are hydrophobic (will not readily absorb moisture), but the spaces between the particles of the pellets or briquettes may absorb some moisture and eventually cause the pellet or briquette to begin to deteriorate. Torrefied wood has a high degree of energy densification when pelleting, producing a drier product with a higher calorific value in comparison with conventional pellets. Torrefied pellets have a lower transportation cost than wood chips or conventional pellets per MJ of energy.

Explosion Hazard

Biomass dust is an explosion hazard. Unprocessed torrefied powder is a major explosion risk, and torrefied chips are also brittle, leading to major dust risks. Though durable, torrefied pellets or briquettes should avoid excessive mechanical handling as this may generate dust. All dust generated should be contained and evacuated via air filter.

The available data on dust explosivity is mixed. To be safe, it is necessary to carry out specific dust explosivity tests on the product. Note that the dust generated in mechanically handling either torrefied wood pellets or conventional wood pellets is a function of the degree of mechanical handling and on the particle size distribution itself. It is therefore not possible to state that a product is inherently safer than another without carrying out such tests.

Torrefied Characteristics Summary and Uses

Because wood chips are not very friable (not able to be milled or pulverized easily), primarily because of their typically high moisture content and lignin structure, the most common feedstocks when co-firing in a pulverised coal facility are “white” (standard) pellets or torrefied pellets.

The increase in energy density makes torrefied pellets ideal for applications where significant transportation distances are required, due to their reduced transportation cost. The improved friability makes it an excellent candidate for co-firing because of the reduced energy for pulverising and the ability to mill the torrefied product with the coal. The minimal energy loss experienced when processed is an additional benefit. These improved fuel characteristics must be offset against the increased cost of torrefied biomass.

Besides co-firing, torrefied biomass is suitable for almost any application where biomass is being utilised. This might also include straight combustion and gasification (where significant transportation distances but a powdered feed are required).

The cost to torrefy the various feedstocks in the scope of TEABPP will be very similar. The primary difference between Willow, conifer and deciduous forestry will be the amount of moisture level in the biomass as it is received. If the torrefaction facility has an integrated feedstock dryer, then additional energy will be required to dry the feedstock to the required level for entrance into the torrefaction reactor, which ranges from 10% to 20% on a wet basis depending on the technology utilised. Some technology developers, i.e. TSI claim that their process is largely autothermal, therefore the energy input mentioned above is required for start-up purposes only. These three species will typically be received with between 35% – 50% moisture. Miscanthus generally contains less moisture and hence a narrower range of moisture levels when harvested, ranging from 14% to 23% resulting in a more predictable cost to process. All these feedstocks can and have been torrefied successfully. There will be slightly different equipment necessary to prepare the Miscanthus than required for the other feedstocks, but this difference is not expected to have any significant impact on the total capital costs. Bales of Miscanthus would need to be scarified whereas woodchips would need no further preparation. Availability and lifetime are unlikely to vary between feedstocks.

Every input and output of a torrefaction facility will affect the costs associated with the torrefaction process. Earth Care Products estimate that the Opex (Including electricity, labour, fossil fuel and maintenance wear, parts and spares), not including feedstock costs, will result in a cost to torrefy biomass of approximately £28/tonne of product.

One should also remember that about 30% of the input feedstock is required to be used as a fuel to allow the torrefaction process to proceed (depending on the technology utilised), with the remaining biomass gaining in LHV (GJ/tonne) due to the removal of low calorific value volatiles that become the heating fuel. This 30% along with the general reduction in weight through loss of moisture will require that approximately 2.7 tonnes of green feedstock at 50% moisture content is provided for every 1.0 tonne of torrefied product required. However, the actual feedstock energy lost on an input LHV basis is typically only 5-10%.

The cost curves Figure 174 correspond to the total investment cost for torrefaction systems and have been derived from quotes provided by suppliers. The total investment cost for torrefaction includes drying. For torrefaction and pelleting the total investment cost includes chipping, drying, thermal treatment and pelleting. The total investment cost for the torrefaction and briquetting system includes chipping, screening, drying, thermal treatment and briquetting press. It assumes the utilisation of chipping and screening mobile plants driven on diesel, drying via a drum dryer to reduce the moisture content from 50% to 10%. These total investment costs shown below are divided by the plant capacity in kW– these are not levelised costs across the plant lifetime.

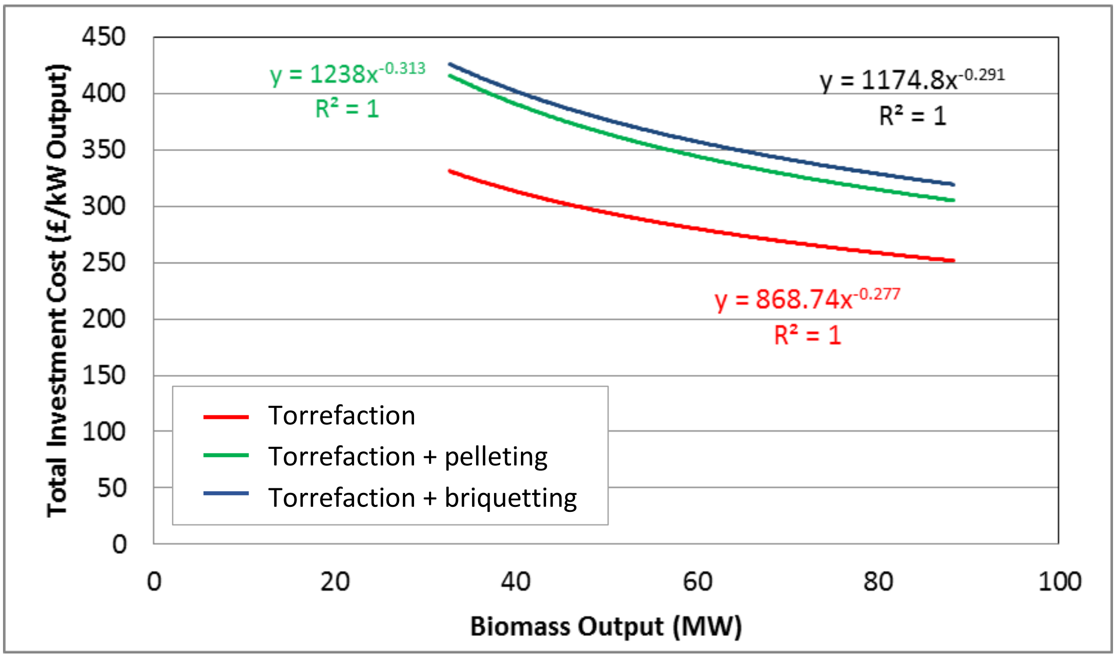

Figure 174: Total investment costs for torrefaction alone, torrefaction + pelleting and torrefaction + briquetting (derived by B&V based upon 3 supplier quotes from 2012 (corrected to 2015), 2015 and B&V estimates for all-in overnight EPC costs)

Operating costs in Figure 175 include fixed and variable Opex (fuel costs, operations labour, insurance, maintenance parts and maintenance labour), processing wood chips (50% moisture content) into torrefied powder, torrefied pellets and torrefied briquettes respectively. 1% of total investment cost has been assumed as insurance cost.

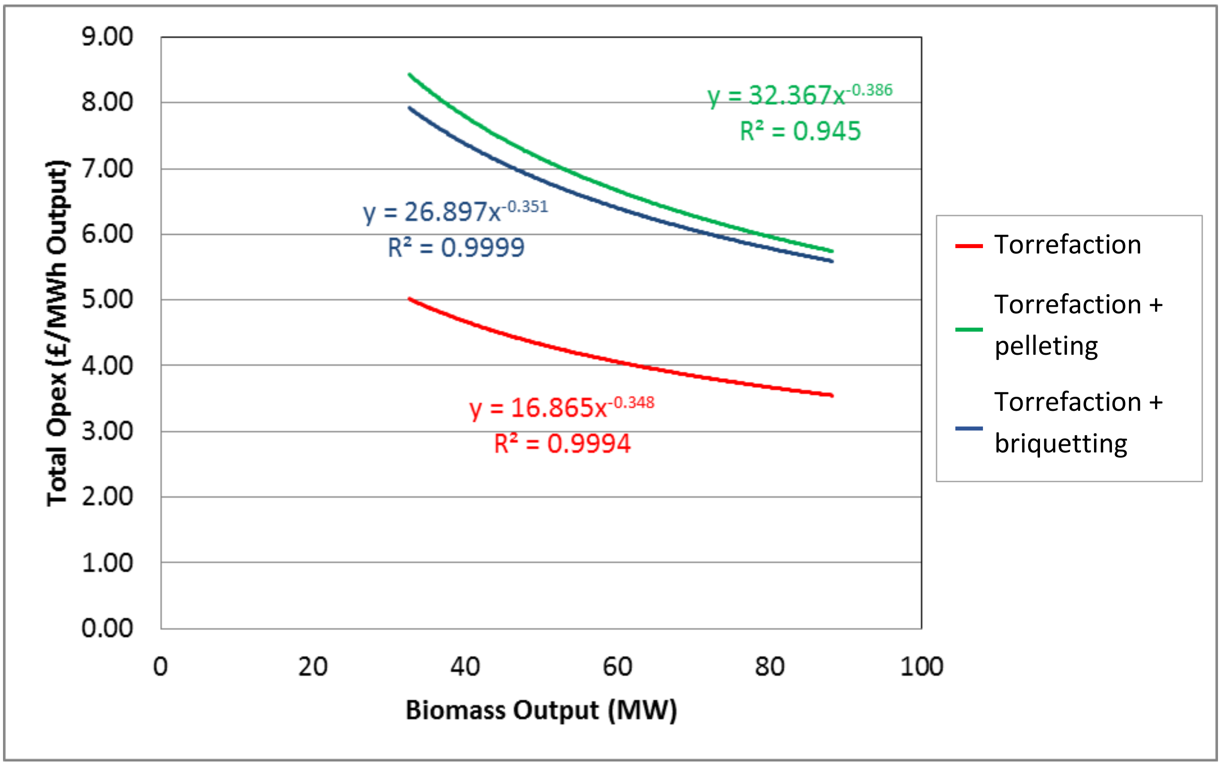

Figure 175: Operating costs for torrefaction alone, torrefaction + pelleting and torrefaction + briquetting (derived by B&V based upon 3 supplier quotes from 2012 (corrected to 2015), 2015 and B&V estimates)

5.8.4 SWOB analysis

Torrefaction produces a more homogeneous fuel with an increased energy density with higher cost savings in transport[207]. When combined with pelleting, torrefaction provides a better product that is less sensitive to degradation. The use of torrefied pellets allows a higher percentage co-firing rate with coal.

Strengths	High degree of energy densification when pelleting, producing drier and higher LHV than white pellets, with less oxygen content Torrefied pellets have lower transportation costs than white pellets or wood chips per MJ of energy Torrefied pellets have characteristics very similar to coal, including ability to withstand outdoor storage, and low grinding costs Durability improved substantially, with lower dust risk for torrefied pellets Uniform product that helps reduce any natural variability in the feedstock, for

	example, differences in biomass species, climate and seasonality that give rise to different chemical compositions. The uniformity of the product is dependent on the quality of the feedstock. Miscanthus and willow have been successfully torrefied.
Weaknesses	Torrefaction and torrefaction with pelleting are relatively energy intensive processes Torrefaction on its own, without pelleting, produces a low density powder or chip product, that is difficult to transport and poses an explosion risk in storage. Torrefied chips are also brittle and present dusting issues Torrefaction has only recently become commercial, so some technology risks are still present for a number of designs and long-term plant operation over many years still remains to be proven
Opportunities	Ability to integrate within existing coal supply chains, with port facilities and feedstock handling at power plants. The big opportunity is the replacement of coal Developer or owner of an existing power plant that desires torrefied fuel could form a consortium with a torrefaction provider to share the technology risk, while providing the long term contract needed by torrefaction provider to fund its scale up Torrefaction with pelleting will expand the biomass market for a given location because it reduces transportation costs allowing the sales to occur further distances from the biomass source
Barriers	The main barrier is associated with financing. Most torrefaction developers are small companies with a limited financial base, which face the challenge of convincing investors to finance the necessary R&D and up-scaling The lack of a product standardisation may delay the maturity of the market There is a variety of torrefaction concepts and a dominant one is likely to emerge. This may impose R&D and scaling up risks onto those companies whose torrefaction concept is significantly different than the dominant one Current technologies work with a narrow range of particle sizes. Increasing this range (and proving technology for other feedstocks such as agricultural residues) is a challenge Not acknowledged by the UK Biomass Supplier List of feedstocks

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5.9 Steam Explosion

5.9.1 Technology description

Uncatalysed steam explosion is already used industrially for the production of fibreboard by the hydrolysis of hemicellulose via the Masonite process[208]. Steam explosion works by treating the biomass with high pressure steam, holding it at high pressure and temperature for a period of time, and then exposing the biomass to atmospheric pressure again. In the case of biomass pre-processing the process is typically carried out at around 160-260°C and a pressure of 0.69-4.83MPa for several minutes before the pressure is explosively released[209].

By treating the pellets with steam explosion the lignin in the biomass is “freed”, acting almost like natural waterproof glue, which melts at about 125°C. Therefore the treated biomass product is a tightly bonded, hard pellet. The pellets tend to be much darker in colour and are therefore often referred to as black pellets. The advantage of these steam exploded pellets (sometimes called “black pellets”) over conventional white pellets is that they contain more energy, are more durable, not self-heating, with minimal off gassing and improved safety due to less dust. The higher energy density and bulk density also means they are cheaper to ship[210] .

An example layout of a steam explosion pellet plant is shown in Figure 176, where a standard pellet plant was adapted to produce steam exploded pellets. The biomass is dried before being sent to the steam explosion reactors (thermal conditioning). The treated biomass is then pelleted and sent on to the coolers and storage. Figure 177 shows a schematic of a process layout for steam explosion connected to a pelleting plant. As can be seen in both examples, steam explosion can be easily added to existing pelleting plants, thereby offering inexpensive retrofit opportunities.

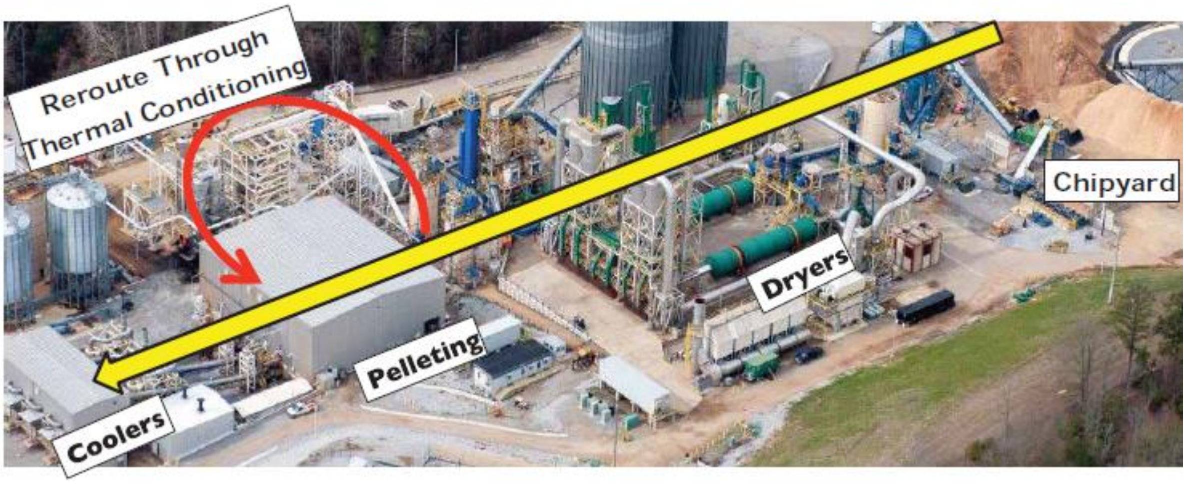


Figure 176: Overview of Zilkha's plant in Selma[211]

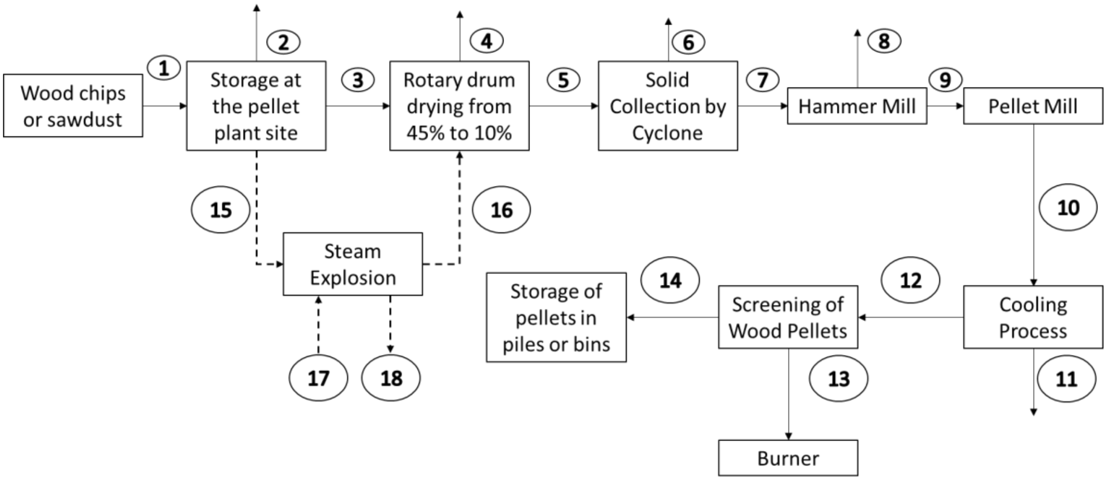


Figure 177: Example layout for steam exploded pellets[212]

Multi-Stage Energy Analysis of Steam Explosion

The energy consumption of steam explosion results from three main stages; the pressure boost, the pressure behind held, and the instantaneous decompression. Sui and Chen have developed a multistage heat transfer model of the steam explosion process and used it to evaluate moisture content of the feedstock. From their model they deduced that the main contributing factors to the heat transfer of the process are the charging coefficient which determines the amount of materials, holding temperature, retention time, moisture content and ratio of height to diameter of the reactor (determines the weight). Holding temperature and ratio of height to diameter impact all 3 stages, charging coefficient and moisture content affect the pressure boost and decompression stage and retention time only affects the holding pressure stage.

They found that moisture content had the largest impact on the energy consumption of the process with a steam consumption/unit dry mass feedstock increasing from 0.53kg/kg to 3.1kg/kg when going from 10% to 90% initial moisture content, with 10% approaching a minimum. They also note that high moisture content can lead to uneven cooking of biomass and has negative impacts on the heat transfer of the steam by blocking its penetration into the porous structures.

Increased holding temperature led to increased energy consumption. They also note that retention time is linked with temperature and that excessively high temperatures can actually reduce the efficiency of the process, with energy consumption hydrothermal effects and mechanical forces all needing to be considered when finding the optimal temperature.

It was found that the energy consumed in the pressure boost stage accounts for nearly all of the energy consumed (98.6% at their optimum conditions) showing the importance of this stage in designing an efficient process. They suggest that moisture content should not exceed 40% as past this point leads to a large waste of energy going towards heating the liquid water in the materials.

Pressure Boost Stage

Materials are heated to the desired temperature by high pressure saturated steam. The steam has multiple roles. It heats the reactor, air and materials to the desired temperature, which causes the steam to condense into water. The other is to raise the pressure in the reactor and penetrate into the porous structure of the feedstock.

Holding Pressure Stage

The reactor is held at these constant conditions for a set period of time. The energy consumption at this point is heat radiation from the reactor.

Instantaneous Decompression Stage

Once the desired time has been reached the pressure is releases explosively and the materials are sent into a buffer tank along with internal superheated water flashing into steam – causing the expansion in the feedstock.[213]

Improvements to product quality

Steam explosion causes several improvements to the product quality over standard white pellets. Zilkha reports that the durability increases from 96.5% to 98% PDI, the pellets grind more easily with an increase from 15-20 HGI to 25-35 HGI and an energy density increase from 10,880mJ/m^3 to 14,5540 mJ/m^3 [213]. The steam exploded pellets are harder than the white equivalents due to the lignin melting and acting as a water resistant glue for the pellets. This also leads to reduced fines and a lower dust content relative to the standard white pellets.

5.9.2 Development status and timescales

Steam explosion is one of the more commonly applied biomass pre-processing technologies, being used both for ethanol fuel production, as well as heat and power applications. The steam explosion technology for cellulose release is typically more severe (hence expensive) than that for producing pellets[214]. Scales of commercial operation are expected to range from 50,000 to 450,000 tonnes (as received) of biomass per year.

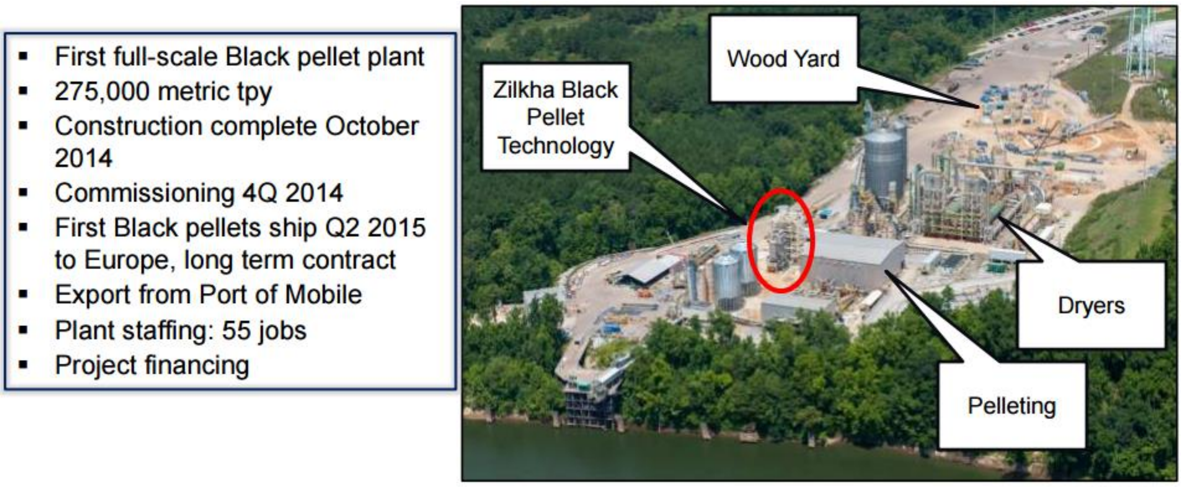
In the case of ethanol production, the goal of the pre-processing is to maximise cellulose yields for downstream hydrolysis and fermentation. Iogen and Abengoa have been operating demonstration plants in Canada and Spain respectively for the past few years. Abengoa has also recently opened a full scale biofuels plant using steam explosion in Hugoton, Kansas, which began operation in September 2014. This plant has a capacity of 25million US gallons of ethanol per year and also produces 21 MW from the residual biomass solids from the ethanol conversion process[215]. Beta Renewables are operating a plant in Crescentino, Italy. They use steam explosion to treat their biomass before breaking down the sugars and converting them to ethanol. The technology is licensed to Biochemtex and licensed to other companies under the name PROESA.

The use of steam explosion for heat and power products is less widespread. However, Zilkha has recently opened a plant in Selma, Alabama with capacity to produce 275,000 tonnes of steam exploded pellets per year. These are being sold in Europe on a long term contract. Zilkha also has plans to open a second larger $90million facility in Monticello, Arkansas producing 450,000 tonnes per year[216]. The technology is also to be licenced to Thermogen Industries[217], who plan to produce 300,000 tonnes/yr of black pellets in a new $70million facility in Millinocket, Maine[218]. Zilkha also has a 40,000 demonstration plant in Crockett, Texas that has been operating since 2010, after converting an existing white wood pellet mill.

Whilst there does appear to be increasing interest in treated biomass pellets, other than Zilkha and Arbaflame who have a full scale demonstration plant, the majority of the focus appears to be on torrefaction with companies such as Andritz, Topell, TSI, Vega Fuels and Solvay testing the technology with pilot and demonstration scale plants.

Based upon the existence of a first commercial plant in Selma, with the plans for another two plants, the **TRL is rated at 8** on woody feedstocks. Given the amount of industry activity, achieving TRL 9 does not appear to be more than a few years away, with limiting factors being the adoption of steam exploded pellets over white pellets, and the competition from torrefaction equivalents. The possibility of retrofitting existing white pellet mills may also be another route to faster deployment of the technology.

For other, non-forestry material, there is now some first commercial plant experience using steam explosion on wheat straw and corn stover, but optimised for downstream ethanol production. The TRL for steam explosion of non-forestry feedstocks (optimised for pellet production) is therefore at an earlier stage, in the 6-7 range.



- First full-scale Black pellet plant
- 275,000 metric tpy
- Construction complete October 2014
- Commissioning 4Q 2014
- First Black pellets ship Q2 2015 to Europe, long term contract
- Export from Port of Mobile
- Plant staffing: 55 jobs
- Project financing

Figure 178: Zilkha's first commercial 275,000 tonnes/year plant in Selma, USA[219]

5.9.3 Impact of different feedstock parameters on operation and cost

The factors that affect steam explosion are steam loading, residence time, temperature, particle size and moisture content. Various combinations of temperature and residence time can be used, essentially simplifying down to either high temperature, low residence time or low temperature, high residence time. Typical parameter values are[220]:

- Temperature: 180-234°C
- Residence time: 3-15 min
- Steam loading: 0.3 to 1.5 kg/ dry kg
- Moisture content: 10% to 50%
- A higher steam loading (i.e. higher heat input) is required with higher moisture content, as shown in Figure 179 below.

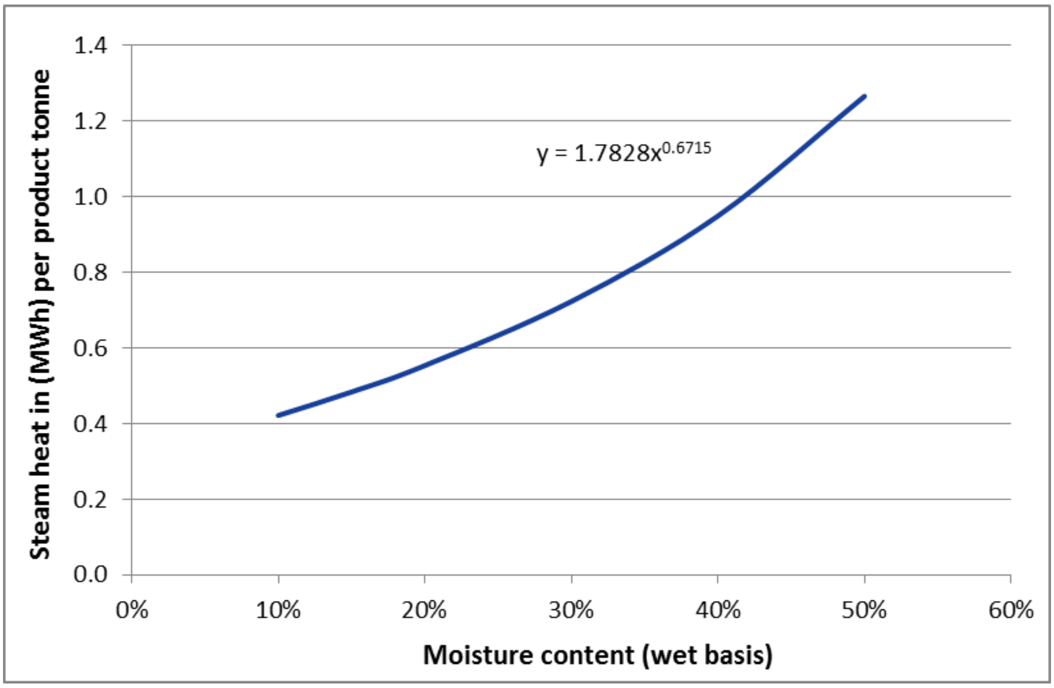


Figure 179: Steam loading vs. feedstock moisture content (based on ICON energy balance equations)

There is a non-linearity as the amount of steam required to dry the biomass is proportional to the moisture content, but the amount of product produced is not constant because the pellets have almost no moisture, so the mass yield of pellets is lower for wetter biomass.

The energy efficiency related to the steam loading is as below:

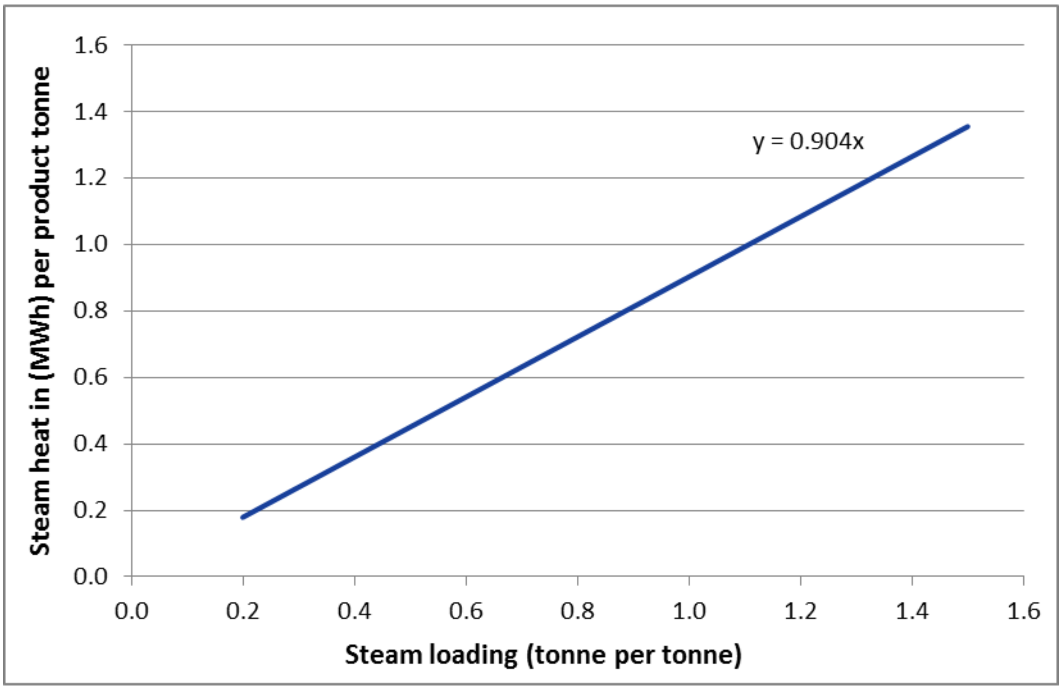


Figure 180: Steam energy input to mass output ratio vs. steam loading (based on ICON energy balance equations)

The relationship is linear, because here the base moisture content of 30% is assumed. It is clear that potential process improvements would be to reduce the steam loading to the lowest value possible.

Feedstock Suitability

Steam explosion has been shown to work on a wide range of biomass feedstocks and is therefore expected to be able to work for all of the TEABPP feedstocks. The basic front processing e.g. screening also ensures robustness, and overall the sensitivity of availability to feedstock variation is small. Table 29 summarises some of the feedstocks that steam explosion has been tested on and the optimal conditions found. Many of these optimal conditions are from a cellulose yield perspective however the conditions should be similar for creating pellets for heat and power.

Table 29: Summary of optimal conditions for various feedstocks for steam explosion pre-processing[221]

Species	Temp (°C)	Time (min)	Particle Size	Liquid/Solid Ratio (g/g)
Woods				
Eucalyptus	230	2	4 mm	10
Poplar	230	4	4 mm	10
Poplar	215 – 225	3 – 7	1 mm	-
Populus Tremoloides	180 – 230	0.7 – 4	20 mm	-
Populus Deltoides	220 – 245	0.42 – 1.47	20 – 40 mesh	7.1 – 10
Populus Deltoides	180 – 235	2	-	-
Pine (Pinus Silvestris)	210	4	4 mm	10
Pine (Pinus Pinaster)	190	8	4 mm	10
Mixed Hardwoods (Oak and Gum)	160 - 280	1	6 mm	-
Oak	230	4	4 mm	10
Mixed Hardwoods (Oak and Gum)	200 – 230	2.08 – 2.23	0.5 mm	20
Aspen	187 – 240	0.5 – 5	2 mm	-
Agricultural Residues				
Corn Cobs	180 – 223	3 – 5		-
Corn Stover	120 – 190	15 – 120	35 – 200 mesh	2 – 10
Sunflower Seed Hulls	200	5	1 mm	-
Sugar Cane bagasse	185 – 208	20 – 29		4
Wheat Straw	190	8	4 mm	10
Vine Shoots	190	8	4 mm	10
Thistle Biomass	210	2	4 mm	10
Sweet Sorghum Bagasse	230	0.5	4 mm	10
Almond Shells	180 – 200	1 – 68	60 mesh	3
Corn Stalk	190 – 230	10 -20	0.5 mm	25
Sugar Cane Bagasse	150 – 170	-	0.8 mm	5 – 10
Wheat Straw	120	15 – 300	60 mesh	28
Bamboo Grass	169.6 – 206.2	10	0.5 mm	-
Wheat Straw	205 – 230	2	-	-
Almond Shells	180 – 240	12 – 25	60 mesh	14.3

Particle Size

Larger particles may result in uneven biomass cooking, with the surface being overcooked and the interior being incompletely treated. However, steam explosion pre-treatment has been carried out on a wide range of particles size so it is not expected to require much processing past basic milling of the feedstock post-harvest. More research would need to be done to find economic maximums for particle size for specific goals and feedstocks.

Steam explosion causes a reduction in the particle size, with the more severe the treatment is the greater the particle size reduction. Boussaid et al, found that the percentage of particles less than 0.074mm increased from 5% to 95% (wet basis) by increasing the severity of the treatment for Douglas Fir[222]. It is important to note that it has been reported that conventional mechanical methods of particle size reduction require about 70% more energy to achieve the same effect as steam explosion for poplar and aspen wood[223]. The particle reduction is caused by the explosive depressurisation, causing the water in the core of the wood to vaporize and rapidly expand. This rapid expansion causes the disintegration of the chips. An example of the reduction in particle size can be seen in Table 30.

Table 30: Summary of the particle size reduction by various severity of steam explosion treatments[224]

Steam Explosion conditions and feedstock variables					Pressure (MPa)		Geometric mean diameter (mm)	
Temperature (^{0}C)	Time (min)	Severity	Particle Size (mm)	Moisture content (w.b.)	Average	SD	Average	SD
Untreated			0.9	15%	-	-	0.90	0.06
			0.4	15%	-	-	0.42	0.03
200	5	3.64	0.9	10%	1.17	0.07	0.83	0.05
				15%	1.18	0.03	0.92	0.01
			0.4	10%	1.24	0.02	0.45	0.01
				15%	1.21	0.15	0.44	0.01
	10	3.94	0.9	10%	1.18	0.03	0.81	0.00
				15%	1.31	0.06	0.86	0.03
			0.4	10%	1.24	0.10	0.40	0.02
				15%	1.32	0.07	0.45	0.05
220	5	4.23	0.9	10%	1.73	0.12	0.74	0.02
				15%	1.81	0.11	0.82	0.03
			0.4	10%	1.78	0.02	0.38	0.02
				15%	1.75	0.08	0.44	0.07
	10	4.53	0.9	10%	1.91	0.07	0.69	0.03
				15%	1.84	0.15	0.78	0.04
			0.4	10%	1.94	0.14	0.35	0.04
				15%	1.92	0.10	0.37	0.04

Moisture Content

Low moisture content chips heat rapidly by steam penetration and condensation inside the chips, whereas large water filled chips heat slowly by conduction and after short times in temperature steam can be unevenly cooked. However, if the chips are small this issue is negated and they are evenly heated very rapidly[225]. Despite this Cullis et al have shown that increase in chip size and moisture content improved the removal of lignin[226]. It is not expected that moisture content would have a major effect on the outcome of the pre-treatment, as drying takes place before steam explosion; however there may be an increase in steam costs from the poorer heat transfer and the extra heat requirement, as in Figure 179 above. More research in this area is needed to find out exactly how it will affect the process from a heat and power point of view.

Impact of plant scale on cost

The plant cost is based on the reference indicated in the caption and is a direct figure for a particular scale. The plant scale is assumed to follow a 0.62 economy of scale power, being typical of a process plant. This gives rise to the following cost curve:

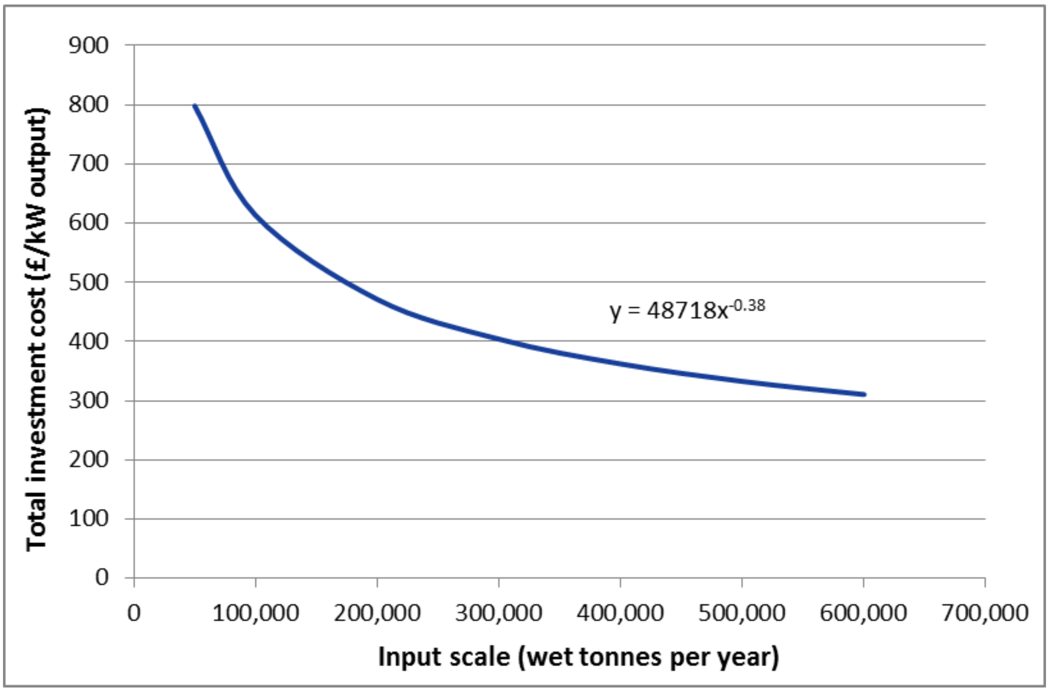


Figure 181: Steam explosion + pelleting total investment cost vs. plant scale (based on data from[210])

For the benchmarking techno-economic analysis, the maintenance costs are assumed to be 2% of the Total installed costs, and the plant labour, fuels and reagents costs are taken from [214].

5.9.4 SWOB analysis

The business case for steam explosion is fairly strong, with general improvements in pellet quality, existing commercial scale plants and a growing market for biomass pellets. There are some issues, such as the competition with torrefied pellets, lack of process information for a varied selection of feedstocks and that the majority of the research into the process is aimed at cellulose release for ethanol production. The core benefits will arise from: (i) densification and reduced transport costs; (ii) storability and stability and reduced spoilage and seasonality issues and (iii) better thermal performance in operation due to higher LHV and lower moisture content.

Strengths	Flexible on feedstocks Relatively simple process, no complicated equipment needed Produces pellets without needing additional binders due to lignin Resulting pellets are energy dense (19.5 GJ/tonne) – cheaper transport High bulk density (750 kg/m^3) – cheaper transport Low dust for safer storage, lower explosion risk No self-heating/minimal off-gassing More durable, robust pellets – able to withstand rain without disintegrating and overcome any seasonality aspects of biomass supply Reduces costs of milling before pelleting and makes the resulting pellet easier to grind (Zilkha claim 40% less power use than grinding white pellets) Low moisture content (1.5%)
Weaknesses	Lack of optimisation (economic/technical) from a heat and power view point Requires relatively large amounts of energy for steam, which either adds cost or reduces yield. Energy cost can be up to 15% of output energy Due to there being large amounts of water, there will be large amounts of waste water to be treated. This would be reduced depending on the degree of treatment and recycling, which is eminently possible; it just requires additional energy.
Opportunities	Only 1 company currently doing this – Zilkha so not a saturated area Growing market for biomass pellets. Black pellets are claimed to have higher density (by 16%), higher energy density (by 15%) and better moisture resistance than standard pellets, so could capture a large share of a growing market Potentially major opportunity to retrofit existing white pellet plants with steam explosion technology, at low cost Pellet product has a good degree of fit with existing coal supply chains and logistics infrastructure – the stability and durability means more standard storage solutions can be used. 100% milling in coal power stations has already been successfully tested Pellets are claimed to have higher density (by 16%), higher energy density (by 15%) and better moisture resistance than standard pellets
Barriers	Limited research done on optimising the process for various feedstocks from a heat and power perspective Lack of information on the effects of certain parameters from a heat and power perspective Lack of technical specifications / product standardisation

	Only an emerging technology with few suppliers (e.g. Zilkha), or optimised for cellulose production (e.g. ethanol producers) Not acknowledged in the UK Biomass Supplier List of feedstocks

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5.10 Ammonia Fibre Expansion

5.10.1 Technology description

AFEX (ammonia fibre expansion) is a pre-treatment method that uses aqueous ammonia at elevated temperatures and pressures. Similar to steam explosion, the process relies on a rapid depressurisation (expansion) phase that ruptures the biomass structure, however it is a somewhat wetter process because the ammonia must penetrate the biomass for effective operation. The treated biomass can then be pelletised using a typical pelletising process (see section 5.6). The AFEX process reduces lignin and removes some hemicellulose while decrystallising cellulose in the biomass, and significantly reduces average particle size. A high level process flow diagram of the AFEX + pelletising process is shown below in Figure 182. The rest of this section will primarily focus on the AFEX part of the process, as pelleting is a commercialised step.

The important process variables in AFEX pre-treatment are the ammonia loading (ratio to biomass), water loading (ratio to biomass), reaction temperature and residence time. As with many other pre-treatment methods, the important feedstock variables include moisture content, particle size and feedstock composition.

Typical values for the key process variables for the AFEX process are as follows:

- Pressure: 20-30 bar
- Temperature: 70-140°C
- Residence time: 5-10 minutes
- Ammonia: dry biomass loading, 0.2–2.0 to 1 (w/w)
- Water: dry biomass content, 0.2-2.5 to 1 (w/w)

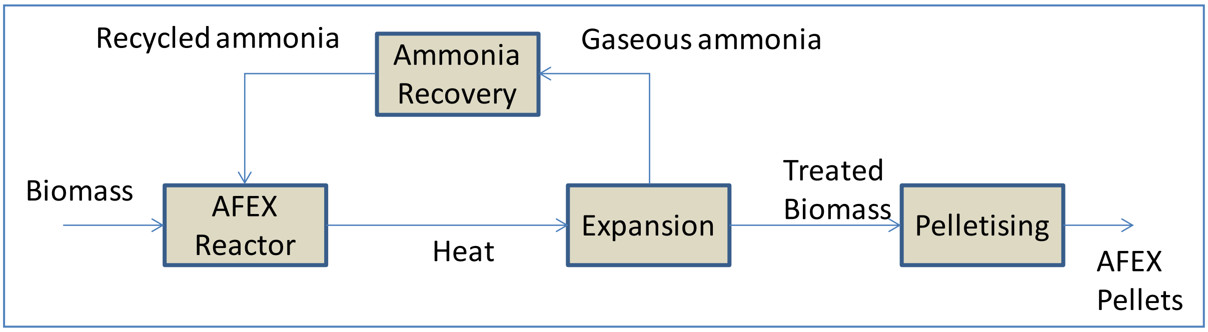

Figure 182: High level overview of AFEX process[227]

During AFEX pre-treatment, 99% of the ammonia is recovered and reused. The remainder remains in the biomass. This is an advantage for biochemical processing downstream (the principal use for AFEX pellets at the moment), but less so for thermal processes where the N will add load to pollution mitigation technologies. Similarly, nearly all of the water in the process is recycled, with a small amount requiring treatment.

AFEX pre-treatment has currently mainly been used in the USA for corn stover pre-treatment and densification prior to biochemical processing (e.g. fermentation to ethanol). It is particularly suited to this downstream operation because it retains most of the hemicellulose and cellulose sugars in their polymeric forms after pre-treatment and there is little biomass degradation. AFEX causes some physical and chemical changes in the biomass, but with little hydrolysis of hemicellulose to sugar monomers and minimal effects on cellulose.

The low water content and protective coating of lignin, together with the lack of accessible sugars means that AFEX pellets biomass are a stable intermediate that can be stored and transported over long distances.

According to INL[227] *"AFEX pre-treatment of biomass offers significant advantages for densification, storage, and transportation. Because AFEX-treated biomass is relatively dry and inert, it is more easily stored, transported, and densified to further improve bulk handling properties. These unique features allow biomass to be pre-processed and treated at a site close to the biomass source and then shipped to a centralized biorefinery location. AFEX treatment transfers some lignin and hemicellulose oligomers to the surface of biomass fibers where it can act as a binding agent."*

Figure 183: Morphology of untreated corn stover bales (left) and AFEX pellet product (right)[228]

Research and scale up activities have indicated that densification of 3-5 times the original biomass density is possible after pelletising, and that the pellets produced were hard and durable with a specific gravity of up to 1.16, without using any binders.

Hoover et al. (2014) describe how the process conditions affect pellet properties; using lab scale equipment they were able to produce pellets with a density of over 600 kg/m^3 and a durability of over 99%, above the standard white pellet durability for long-distance transport of 97%. The moisture content was found to be less than 9%.

5.10.2 Development status and timescales

AFEX was invented by Bruce Dale and co-workers at Michigan State University, and is being commercialised by the associated venture division, MBI (www.mbi.org). MBI built a 1 tonne per day pilot plant in 2013 (see Figure 184) and has completed successful trials, using the products for both downstream biochemical processing and animal feed. They are now in the process of running the system on different feedstocks and have developed a design for up to 200 tonnes per day (tpd). They are seeking partners for broad deployment from 2016. In October 2014, MBI conducted a testing campaign for Deinove[229], looking at the benefits of AFEX for ethanol production, but there have been no announcements since regarding future collaboration or plants.

The current status of the AFEX part of the process is large scale pilot, with a plan to move to a larger commercial scale plant within 3 years. The pelletising part of the process is mature technology. The overall TRL of the system is therefore around 5, although noting that the technology is currently being optimised for downstream ethanol and feed applications, rather than optimising to obtain the highest LHV pellets for heat & power applications (e.g. through better control of nitrogen).

The largest lignocellulosic ethanol facilities today (first commercial plants) use of the order of 700 – 1,200 tpd of feedstock However, AFEX is considered by the industry as potentially being suitable for smaller decentralised plants, in order to locally aggregate low density agricultural residues before transport to a centralised plant[230]. MBI are proposing that designs of 50 - 200 tpd could therefore be an economically viable first commercial plant.

Reaching this scale in one step from 1 tpd is fairly ambitious and could still take a while to achieve, as none of the main lignocellulosic ethanol developers are currently using the technology. Assuming the first commercial plant is successful, it may still take at least 10-15 years from today to reach TRL 9 (commercial scale replication).

Figure 184: MBI's 1 tpd AFEX pilot plant[228]

5.10.3 Impact of different feedstock parameters on operation and cost

Feedstock suitability

The pilot scale technology at MBI has been tested on the following feedstocks:

- Corn stover
- Switchgrass
- Sugar cane bagasse
- Rice straw
- Miscanthus
- DDGS

This is indicative of the likely suitability of AFEX to fibrous rather than woody biomass; indeed Mosier et al (2005)[231] state that AFEX will not work on softwood feedstocks and performs poorly on hardwood feedstocks. For the TEABPP feedstocks, AFEX is only recommended here for Miscanthus as it is unlikely to work effectively on SRF, LRF and SRC feedstocks. The system availability to other variations in feedstock parameters is minimal.

Feedstock conditions

Moisture content

Because the process takes place using an aqueous reagent, the moisture content of the feedstock is not an issue, and raw biomass can be treated by adjusting the water input to the moisture content to meet the desired loading ratio. Hence moisture content does not affect the process significantly. Overall, the heat input (assuming some degree of energy integration) can be estimated from the energy required for ammonia recovery (based on ammonia loading), and then the difference between moisture content of feedstock and product (the balance requiring heat to evaporate the water; note that this can be relatively low grade heat).

The relationships between moisture content/ammonia loading and heat required are illustrated in Figure 185.

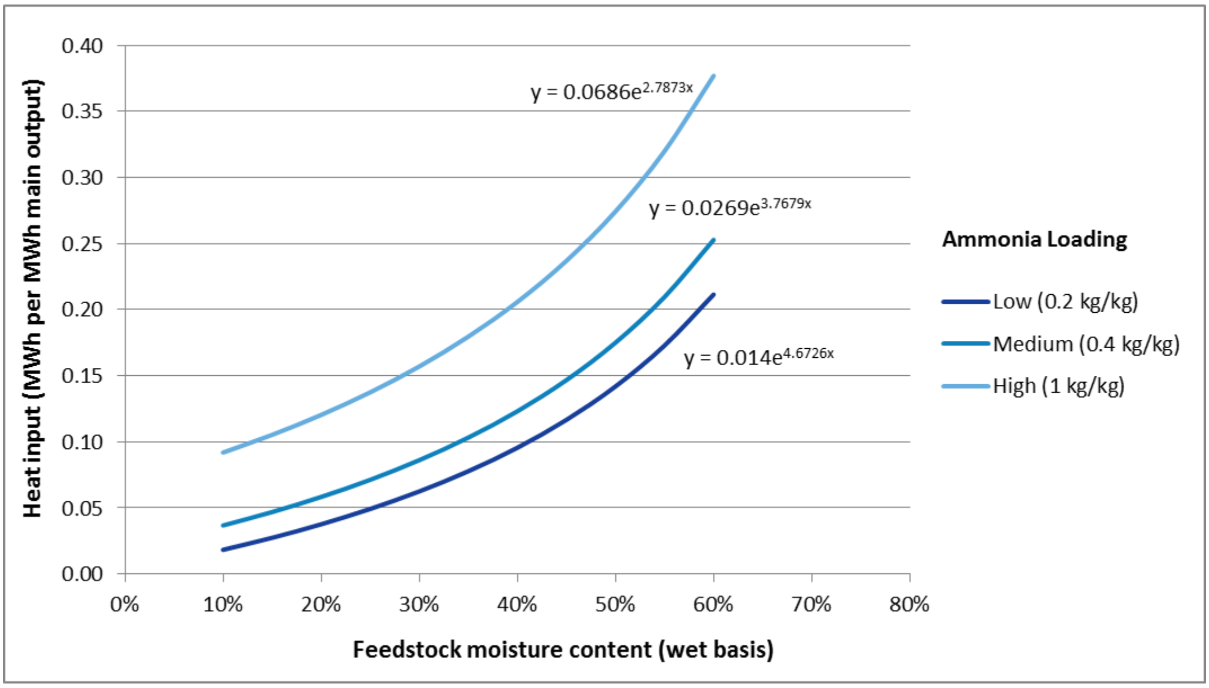

Figure 185: Heat input vs feedstock moisture content for low (0.2 kg/kg), medium (0.4 kg/kg) and high (1 kg/kg) ammonia loadings (based on [228] and ICON energy balance equations)

Particle size

The typical feedstock is a coarse cut/chipped feedstock of particle size 20-70mm; hence fairly basic processing of the feedstock post-harvest is required (e.g. bale chopping) before feeding into the system. AFEX is effective at reducing average particle size during the process, which is beneficial for the final production of pellets from the treated biomass.

Feedstock composition

There is no evidence of any constraints on inorganic components nor any indication that these are extracted to any significant extent into the process wastewater (Small amounts of lignin and hemicellulose are removed, leading to a degree of wastewater treatment requirement (included in the Opex). For the base OPEX, the O&M, labour, fuels and reagents costs are from[228]. Hence it is reasonable to assume that the inorganic species remain in the pellets.

Plant size

The plant size primarily affects the Capex, which is taken directly from the reference in the caption. A 0.75 power law is estimated given the mixed nature of the equipment. This gives rise to a cost curve as in Figure 186, which compares well with MBI's estimate of $12-17 per tonne[228].

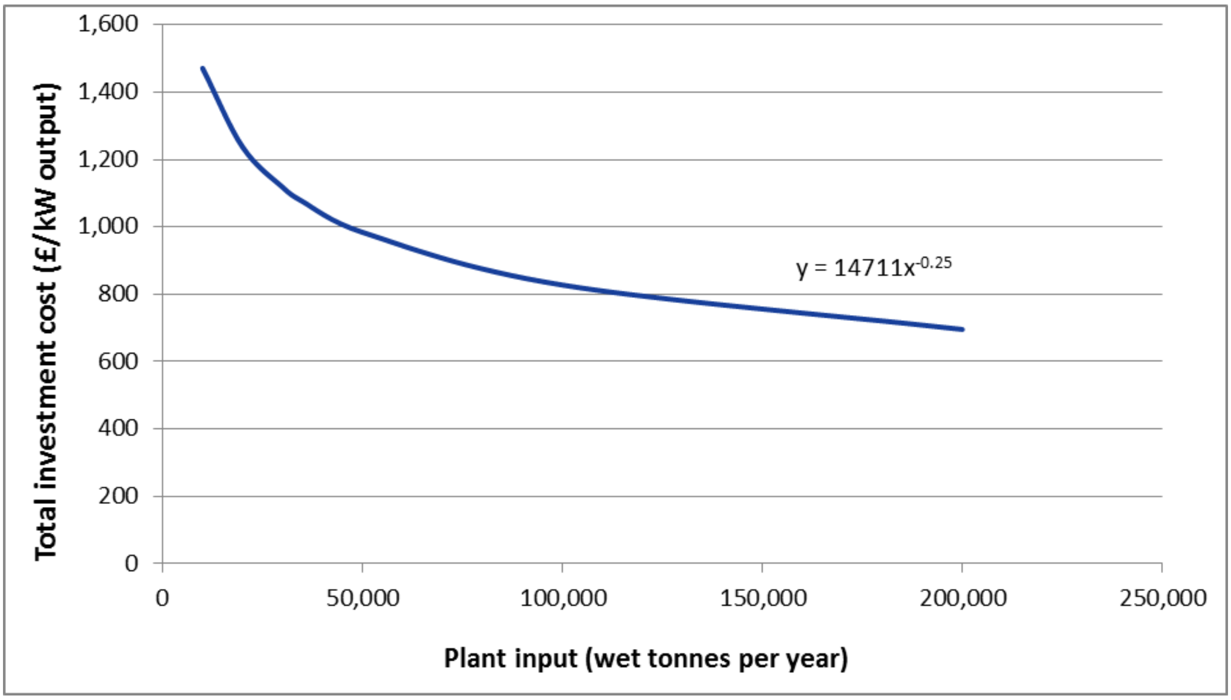

Figure 186: Capex (£/kW) versus plant scale (based on data from[228]) based on 0.75 power law

A further sensitivity, unique to this process and important for downstream processing, is the nitrogen content of the biomass product, which is a function of the ammonia loading used:

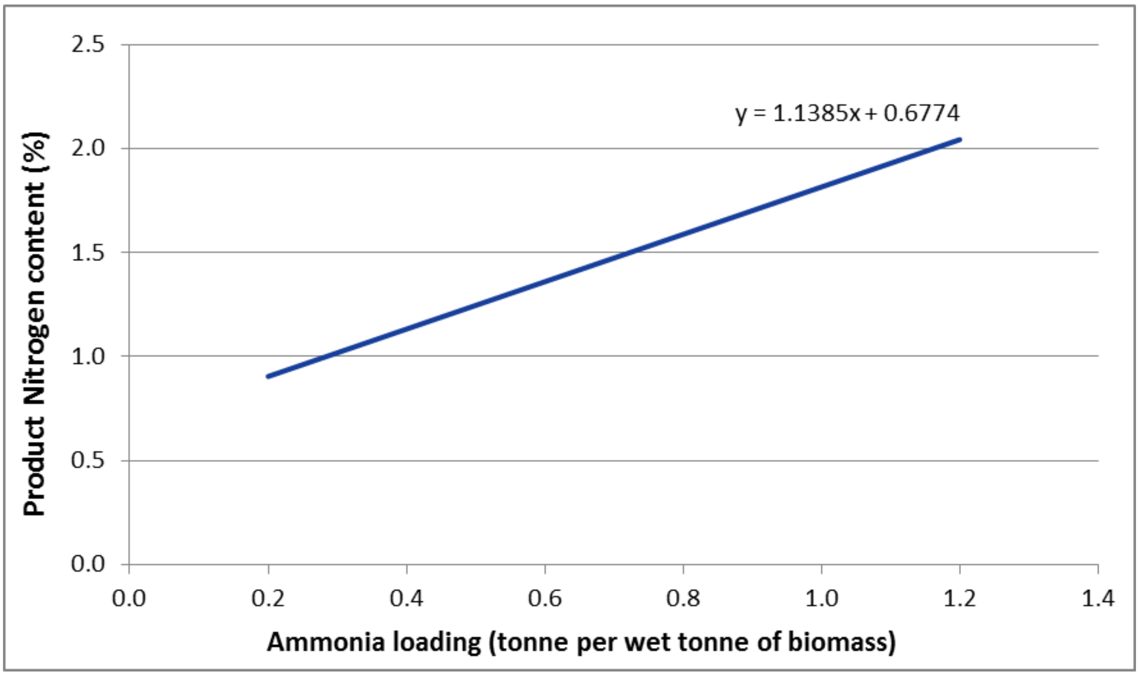

Figure 187: Product N content versus ammonia loading (based on [228,232] and ICON energy balance equations)

For the benchmarking techno-economic analysis, the operating costs of O&M, labour, fuels and reagents costs are taken from [228].

5.10.4 SWOB analysis

The process can contribute to improved bioenergy chains because it offers a distributed processing option which will (a) reduce transport costs once the material is densified (both in terms of mass and energy density) and (b) reduce spoilage and increase stability/storability of the processed biomass. A more uniform (composition and form factor) product may also bring advantages. However, it is not suitable for woody biomass.

Strengths	Can produce pellets without additional binders due to lignin re-arrangement Good increase in biomass density; should reach at least 500 kg/m^3 if not considerably higher Produces stable, transportable pellets that can be stored for long durations and handled cheaply High durability reduces dust and hence fire risks Product can be suitable for multiple purposes – thermal, biochemical and animal feed (depending on the feedstock) Produces a uniform product at <10% moisture
Weaknesses	Not proven at commercial scale and market volumes uncertain Lack of optimisation (economic/technical) from a heat and power view point Not suitable for woody biomass NH3 content of pellets – expected to be in the order of 0.3% - increases NO_x emissions of downstream thermal processing
Opportunities	Scale up to next level and evaluate the performance Enables densification of agricultural residues[233], hence step change in transport costs of straw and Miscanthus Enables hybrid decentralised/centralised processing – for example, a future network of decentralised pre-treatment plants, transporting treated biomass to a centralised conversion plant
Barriers	Pellets have mainly been tested for biochemical processing rather than thermal technologies. Currently optimised for availability of cellulose (e.g. ethanol producers) Limited research done on optimising the process for various feedstocks from a heat and power perspective Primarily tested on corn stover; most of the literature is on this feedstock only Lack of technical specifications / product standardisation Pellets not acknowledged on the UK Biomass Supplier List of feedstocks Only an early stage technology, without any suppliers or technology guarantees

AFEX appears to be an attractive technology for agricultural residues, switchgrass, Miscanthus and similar feedstocks which are difficult to use in bundle form and which are expensive to ship over long distances without densification. Also makes them storable over a longer period, getting away from seasonality issues.

The key supply chain challenge would developing an appropriately sized AFEX facility to be shared by several farms from which a homogeneous, tradeable product can be produced, as well as ensuring that the pellets become an acceptable feedstock meeting standards in a way similar to other pellets.

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[227] Tumuluru et al, “A Review on Biomass Densification Technologies for Energy Application”, IDL, 2010. INL/EXT-10-18420

[228] A. Julian, “AFEX®: A Densified, Cost-Effective Biomass Solution”, Bioplastek 2013. Available at: http://web.archive.org/web/20140703132632/http://www.mbi.org/wp-content/uploads/2014/02/AFEX-BioPlastek-Conf-6-3.pdf

[229] Deinove (2014) "DEINOVE teams up with MBI, pioneer of AFEX technology, to evaluate its process on industrial biomass”. Available at: http://www.deinove.com/en/news/all-press-releases/deinove-teams-mbi-pioneer-afex-technology-evaluate-its-process-industrial-biomass

[230] E4tech, RE-CORD & WUR (2014) “From the sugar platform to biofuels and biochemical”. Available at: https://ec.europa.eu/energy/sites/ener/files/documents/EC%20Sugar%20Platform%20final%20report.pdf

[231] Mosier, N. et al. (2005) Features of Promising Technologies for Biomass Pretreatment. Bioresource Technology; 96 (6): pp 673-686

[232] Tumuluru et al, “A Review on Biomass Densification Technologies for Energy Application”, IDL, 2010. INL/EXT-10-18420

[233] Hoover et al, “Effect of pelleting process variables on physical properties and sugar yields of ammonia fiber expansion pretreated corn stover”, Bioresource Technology, 164 (2014) 128–135

6 Horizon Scanning

6.1 Introduction

The combustion, gasification and pre-processing technologies described above are the technologies in the scope of the TEABPP project, and by generally being at the top end of the commercialisation status scale, are most likely to have credible, robust techno-economic data that would be suitable for the more detailed analysis and modelling in the latter WPs of the TEABPP project.

However, there are a very large number of other conversion technologies that utilise biomass feedstock in order to produce heat, power, syngas and/or other intermediate products that have not been mentioned so far in this report. The purpose of this Horizon Scanning chapter is therefore to highlight some of these other conversion and pre-processing options, writing up short technology descriptions and their likely TRL development, as many of these options have a potentially promising future, but do not have the available data to be included in the rest of the project.

These technologies have been grouped into sub-type categories, as detailed in Table 31. Noticeably, there are no heat-only technologies on the list – other than the pre-processing technologies, there is a strong focus in Table 31 on electricity (and potentially CHP) generation, or the components that could form part of a new power generation plant. Several of the technologies are also focused on or have been applied to waste or liquid feedstocks, as opposed to woody biomass.

Table 31: List of horizon scan technologies and their categorisation

Type	Sub-type	Technology
Conversion technologies	Gasification	Microwave Induced Plasma Gasification
	Gasification	Super-critical steam gasification
	Gasification	CO_2 gasification
	Syngas clean-up	Plasma clean-up
	Syngas clean-up	OLGA clean-up
	Syngas clean-up	Catalytic tar removal
	Pyrolysis	Pyrolysis (gas + biochar)
	Other combustion	Externally fired (biomass direct combustion) gas turbine
	Other combustion	Chemical Looping Combustion and Gasification
Downstream power generation	Alternative cycle	Stirling Engines
	Alternative cycle	Organic Rankine Cycle
	Alternative cycle	Entropic Cycle
	Fuel cell	Direct syngas Solid Oxide Fuel Cells
	Fuel cell	Bio-hydrogen Alkaline Fuel Cells
	Fuel cell	Fuel cell gas turbine hybrid
Pre-treatment	Thermochemical	Aqueous Phase Reforming
	Thermochemical	Hydro-Thermal Upgrading
	Thermochemical	Microwave Pyrolysis
	Thermochemical	Hydrothermal carbonisation

There are several other technologies that were considered for inclusion in the horizon scanning section, but that were not included as a result of a lack of focus on biomass to heat and power (or syngas):

- Ionic liquids are generally considered for pre-treatment of biomass for biological routes to fuels, rather than for thermochemical processes. There are a number of papers and a couple of patents on pyrolysis in ionic liquids, including microwave pyrolysis. However these appear to mostly be focused on routes to synthesis of chemicals rather than energy.
- Molten metal gasification involves gasifying carbon-containing feedstocks in a bath of molten metal. Two companies have worked on developing technologies and a demonstration plant in this area, but neither of these are currently active, and neither focused on biomass (only wastes and other fossil feedstocks).
- Sorbent-based syngas clean-up techniques use various solid powders to remove tars, sulphur and other contaminants in a regeneration cycle. These are potentially interesting due to less need to quench to low temperatures, however, they have to date only been investigated in pilot applications where syngas is derived from coal[234]. These technologies also have links to development of carbon capture and storage technologies, with solid sorbents able to capture CO_2.

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[234] Gupta., Turk B., Lesemann M (2009) "RTI/Eastman Warm Syngas Cleanup technology: Integration with Carbon Capture". 2009 Gasification Technologies Conference. Available at: http://www.netl.doe.gov/File%20Library/Events/2009/37GUPTA.pdf

6.2 Technologies with lower Technology Readiness Levels

This section presents technologies that have been excluded from benchmarking due to their early-stage development and the resulting lack of data at commercial scales.

6.2.1 Gasification technologies

Here some more novel gasification approaches are discussed, including microwave-induced plasma gasification, super-critical water gasification and CO_2 gasification.

6.2.1.1 Microwave Induced Plasma Gasification

Technology description

In plasma gasification, biomass is converted into syngas using high temperature plasma, and inorganic material present in the feedstock is vitrified into slag. Syngas can then go through the process of cleaning and conditioning before use in an engine or turbine to produce electricity (and heat)[235].

Conventionally, plasma is generated by passing a high voltage electric current between two electrodes, forming an electric arc (Figure 188). The electric arc constitutes a part of a plasma torch, in which an inert gas is ionised by the plasma and used to vaporise the biomass at the bottom of a reactor vessel in order to produce syngas at very high temperatures (Figure 189).

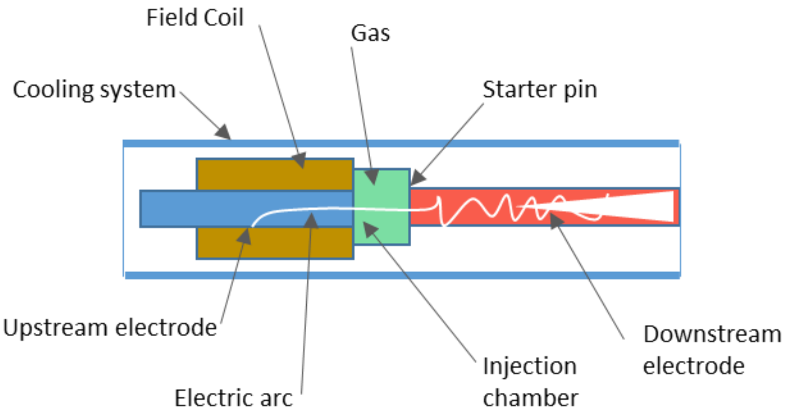

Figure 188: Conventional plasma generator using an electric arc[236]

An alternative way of generating plasma is through microwave induction (Figure 190). Microwave plasma generators are usually microwave coupled reactors, where microwave power is fed into a tapered waveguide resonator (applicator), surrounding a dielectric tube (usually quartz), filled with the working gas. Intense electric fields in the applicator cause the gas to break down and maintain the plasma (see Figure 190). Microwave plasmas have a very high electron temperature (exceeding 5000°C) and can operate in a variety of gas pressures, ranging from 0.1mPa to 10^5 Pa, depending on the application. Due to their high electron densities, the working gas in microwave plasmas is highly dissociated and therefore chemically very reactive. The microwave plasma discharge is induced without the need for electrodes, reducing contaminants and electrode ion sputtering, as well as ensuring a long service interval and reduced consumables costs[237]. Developers have claimed that up to significantly less electricity input is required compared to conventional plasma torches, with a microwave-induced plasma torches only consuming 10-20% of the electricity the power plant produces[238], compared to 25-80% in the case of conventional plasma electrode arcs[238,239].

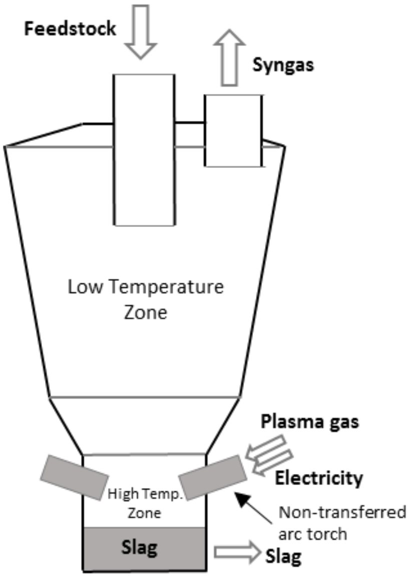

Figure 189: Biomass gasification using plasma as the gasifying agent[241]

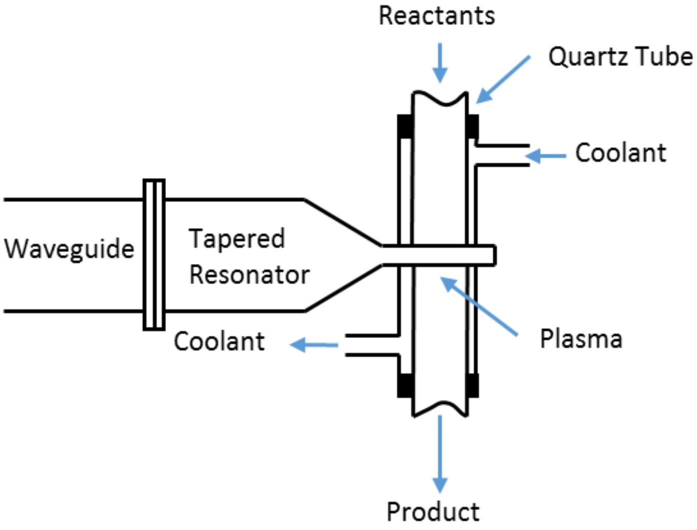

Figure 190: Microwave plasma generator[240]

Similar to conventional plasma gasification, microwave induced plasma gasifiers also produce a very high quality syngas, because the high temperatures obtained boosts the rate of gasification and leads to more complete reactions, compared to conventional lower temperature gasification (e.g. updraft gasifiers). This leads to cost savings, more valuable raw products as well as a syngas with lower tar content, which requires less downstream clean-up before use[240]. Microwave plasma gasifiers can also operate under atmospheric conditions, without needing pressurisation and the associated costs[241].

Note that these bulk gasification reactions that directly use plasma on the feedstock are not to be confused with plasma syngas cleaning using a plasma torch (see Section 6.3.1.1), which relies on a previous gasification step to produce syngas, with the plasma only acting on the tars in the syngas (rather than the whole of the biomass).

Development status and timescales

To date, microwave-induced plasma gasification has only been demonstrated at pilot scale by Stopford Energy and Environment, in partnership with the researchers at Liverpool John Moores University[242]. Plasma2Energy also built a 10 tonnes per day batch reactor in 2007, which was being tested by the Monterrey-Tech University in Mexico. They attempted to construct a 150 tonnes per day (27 MW_e) plant in McAllen, Texas[243], but the project was terminated due to insufficient capital having been raised[244]. The current status of microwave-induced plasma gasification is therefore judged to be **TRL 5**.

Over the past decade, a number of pilot and demonstration facilities have been built for waste plasma gasification, focusing on conversion of MSW to electricity (and effective waste destruction)[238]. There are a number of first commercial plants in operation, including a 50 MW_e plant in Tees Valley, UK[245] (built by Air Products[246] and Alter NRG). However, these are all based on conventional plasma generators using plasma arc electrodes. There would however be potential to transfer much of the useful learnings from these plants, as the only difference is the method for generating the plasma – not how the plasma is applied, or the feedstocks, reactor or downstream steps required.

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[235] Herbert B. (2014) "Advanced Plasma Gasification Systems – Current & Emerging Technologies", The 14th Annual APGTF Workshop. Available at: http://www.apgtf-uk.com/files/workshops/14thWorkshop2014/210BenHerbert.pdf

[236] Themelis N. J., Vardelle A. M. (2012) "Plasma-Assisted Waste-to-Energy Processes", Encyclopaedia of Sustainability Science and Technology pg. 8097-8112. Available at: http://link.springer.com/referenceworkentry/10.1007%2F978-1-4419-0851-3_407

[237] Ambrozic K. (2015) "Microwave Steam Plasma Gasification", University of Ljubljana. Available at: http://mafija.fmf.uni-lj.si/seminar/files/2014_2015/Klemen_Ambrozic-Seminar_1b_ver2.pdf

[238] Waste Management World "Microwave Plasma Gasification Heats up in the US". Available at: http://www.waste-management-world.com/articles/print/volume-12/issue-6/features/microwave-plasma-gasification-heats-up-in-the-us.html

[239] Circeo, L. J. (2008) "Plasma Arc Gasification of Municipal Solid Waste", Available at: http://www.energy.ca.gov/proceedings/2008-ALT-1/documents/2009-02-17_workshop/presentations/Louis_Circeo-Georgia_Tech_Research_Institute.pdf Indicates that St Louis plant would be 160 MW_e gross, and 120 MW_e net, hence 25% consumed in parasitic loads. Assumed majority is for the plasma torches

[240] Ismail N., Ho So., Amin A. S., Ani N. (2015) "Microwave Plasma Gasification Of Oil Palm Biochar, Universiti Teknologi Malaysia. Available at: http://www.jurnalteknologi.utm.my/index.php/jurnalteknologi/article/view/4827/3341

[241] Heidenreich S., Foscolo P. U. (2014) "New Concepts in Biomass Gasification", Progress in Energy and Combustion Science Volume 46 Pg. 72-95. Available at: http://www.sciencedirect.com/science/article/pii/S0360128514000495

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[243] Sims B. (2011) "Microwave-induced plasma gasification technology makes headway". Biomass Magazine. Available at: http://biomassmagazine.com/articles/6814/microwave-induced-plasma-gasification-technology-makes-headway/?ref=brm

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[245] Waste Management World (2012) "50 MW Plasma Gasification Facility to Treat Waste in Tees Valley". Available at: http://www.waste-management-world.com/articles/2012/08/50-mw-plasma-gasification-facility-to-treat-waste-in-tees-valley.html

[246] Air Products "Air Products' Tees Valley Renewable Energy Facilities". Available at: http://www.airproducts.co.uk/microsite/uk/teesvalley/index.htm

6.2.1.2 Super-critical Water Gasification

Technology description

Super-critical water/steam has a dual role as a reactant and heating medium in the gasification of biomass. Using super-critical water as the gasifying agent produces syngas with a high hydrogen content due to steam reforming, where water is a hydrogen-providing participant in the overall reaction. It is an efficient method to treat biomass with very high moisture contents (at least >30%)[247]. The syngas is also generated at high pressure, which allows further utilisation without compression.

Organic compounds, including lignocellulosic material such as solid biomass, will readily dissolve in supercritical water[248]. Once dissolved, supercritical water will efficiently break cellulose bonds. The reactions are generally not selective and result in the rapid formation of gaseous products. Furthermore, many organic chemicals that typically do not react in water without the presence of strong acid or base catalysts will readily react under these conditions (at temperatures over 374°C and pressure over 22.1MPa)[249]. Furthermore, unlike traditional gasification options, super-critical water gasification claims to be able to demonstrate an energy balance that can yield self-sufficient processing[250].

The disadvantages of super-critical water gasification reactions include the need to use expensive materials of construction that are resistant to the aggressive environment of supercritical water[251] (e.g. Hastelloy®), as well as the energy requirements and costs of raising super-critical steam – both are significant obstacles to further development. One way to solve these issues is through introducing a hydrothermal catalyst, which is currently an important research direction[252].

Development status and timescales

The earliest research goes back as far as the 1970s and since then, supercritical water gasification has been the subject of considerable research, mainly in the field of thermochemical conversion of wet biomass (i.e. using feedstocks that are outside of the TEABPP project scope)[253]. A number of research institutions are interested in the technology, such as Lassonde School of Engineering in Canada and TU Delft in Netherlands.

However, despite this interest, the supercritical water gasification of biomass has only been tested on a lab-scale, mainly due to the estimated low cost-effectiveness of the process to date. There have been no pilot or demonstration projects to date, hence the current status of the technology is judged to be **TRL 4**.

The range of applicable scales would depend on the gasification reactor vessel used.

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[247] Reddy S. N., Nanda S., Dalai A. K., Kozinska J. A. (2014) "Supercritical water gasification of biomass for hydrogen production". International Journal of Hydrogen Energy, Volume 39, Pg. 6912-6926. Available at: http://www.sciencedirect.com/science/article/pii/S0360319914005230

[248] Iowa Energy Center (2007) "Supercritical Water Gasification of Biomass". Available at: www.iowaenergycenter.org/supercritical-water-gasification-of-biomass/

[249] Boukis N., Galla U., Muller H., Dinjus E (2007) "Biomass gasification in supercritical water. Experimental progress achieved with the Verena Pilot Plant". 15th European Biomass Conference & Exhibition. Available at: http://www.ikft.kit.edu/downloads/boukis-pub6.pdf

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[251] Yakaboylu O., Harinck J., Smit K. G., de Jong W. (2015) “Supercritical Water Gasification of Biomass: A Literature and Technology Overview”. Energies 2015, Volume 8, Pg. 859-894. Available at: http://www.mdpi.com/journal/energies

[252] Guo Y., Wang S. Z., Xu D. H., Gong Y. M., Ma H. H., Tang X. Y. (2010) “Review of catalytic supercritical water gasification for hydrogen production from biomass. Renewable and Sustainable Energy Review, Volume 14, Pg. 334-343. Available at: http://www.sciencedirect.com/science/article/pii/S1364032109002123

[253] Mobius A., Boukis N., Sauer J. (2013) “Gasification of biomass in supercritical water (SCWG)”. Institute of Catalysis Research and Technology, Karlsruhe Institute of Technology (KIT). Available at: http://www.formatex.info/energymaterialsbook/book/264-268.pdf

6.2.1.3 CO_2 Gasification

Technology description

CO_2 has been proposed as a candidate for a gasifying agent since its utilization has the potential to further reduce CO_2 emissions, by finding a suitable CO_2 stream to be consumed. Furthermore, CO_2 has been observed to be a more efficient gasification environment than steam, resulting in improved bio-char conversion during gasification. The chemical effect is similar to that of steam, as CO_2 will act as a chemical quench due to the fact that CO_2 reactions are endothermic[254]. Introduction of CO_2 as a co-feed into the gasification environment results in higher CO and lower H_2 and CH_4 concentrations enabling greater control of the syngas ratio produced, by adjustment of the level of CO_2 introduced[255]. Moreover, CO_2 is a less corrosive gasification medium compared with steam[256].

Using CO_2 as a co-reactant can enhance the char burnout rates that will result in either increased throughputs or the need for smaller units. Additionally, the injection of CO_2 and H_2O together increases the char reactivity that results in more efficient use of the feedstock with less residue to be post-processed[257]. The energy inputs in heating the CO_2 are still significant, however.

It has to be noted, that using CO_2 derived from the combustion of fossil fuels to produce syngas leads to the fact that the resulting product cannot be considered as entirely renewable and sustainable.

Many researchers have performed experimental studies on gasification in a CO_2 atmosphere with various feedstocks. Overall, reaction rates increase with temperature and vary among different feedstocks. It has been reported that CO_2 gasification generally has lower reaction rates than steam gasification[257].

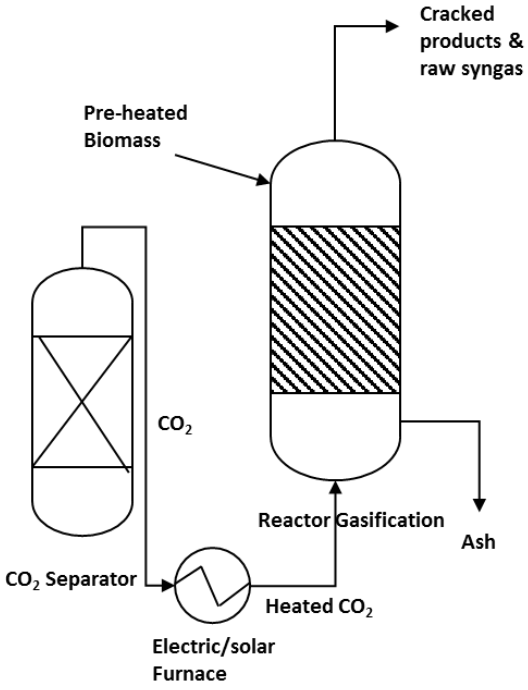

Figure 191: CO_2 gasification of biomass process diagram[258]

Development status and timescales

Although the technology has been known for over a decade, research to date has mainly been based on thermodynamic analysis of different CO_2 gasification processes followed by a number of small-scale lab tests.

The Department of Earth and Environmental Engineering at Columbia University is very active in this field, alongside a handful of researchers in Southern Asia. However, there have been no pilot or demonstration projects to date, hence the current status of the technology is judged to be **TRL 4**.

The range of applicable scales would depend on the gasification reactor vessel used.

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6.2.2 Chemical Looping Combustion and Gasification

Technology description

In chemical looping, biomass is reacted with an oxygen-bearing solid material, producing heat. This heat is then used to raise steam and generate electricity. The CO_2 produced from the reaction is highly concentrated compared with standard combustion in air, making chemical looping of interest for biomass CCS systems.

Chemical looping involves a high temperature cycle, transferring the solid oxide carrier material from one reactor to another. Within one reactor (the “fuel”, or “reduction” reactor, see Figure 192) the solid is reduced from an oxidised form (such as Fe_2O_3) to a reduced form e.g. Fe_3O_4, by reaction with the feedstock. In general, it is easy to use a gaseous fuel such as CH_4 and harder to use a solid fuel such as coal or biomass directly. In a second reactor (the “air”, or “oxidation” reactor) the reduced oxide is oxidised using air and heat is given out. This heat can then be utilised in a steam cycle to generate power. The reduction step can be exothermic or endothermic, depending on the oxygen carrier. The majority of research and development so far proposes that circulating fluidised beds (CFBs) are used to carry out the reactions (similar to a Dual fluidised bed gasifier configuration), though recent work from Ohio State University has suggested a pressurised moving bed reactor is used[259]. An example flowsheet is given in Figure 192.

One alternative configuration is that instead of oxidising using air, steam could be used, which would still generate heat, but also generate a relatively pure stream of H_2 out of the left-hand cyclone that could then be used in a fuel cell to generate more power. Furthermore, by limiting the input of oxide material into the reduction reactor, gasification conditions could be established to produce syngas instead of a CO_2 rich flow via combustion.

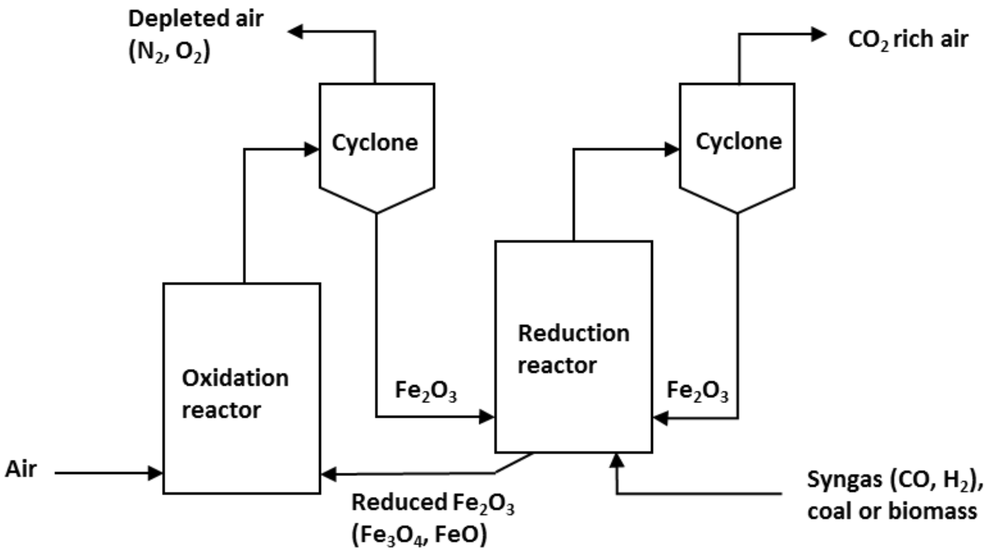

Figure 192: Chemical looping process utilising coal[259]

One major advantage to solid looping processes is that they can piggy-back on developments in CFB design and deployment. Such CFBs, which have up to 800 MW_e designs available, and plants deployed at up to 460 MW_e (for fossil feedstocks), with several larger units being built, are clearly commercial technologies – hence the process of deploying chemical looping is significantly de-risked.

A number of demonstrations of chemical looping combustion are being made at the MW_{th} input scale and above. Alstom are demonstrating $CaSO_4$-based chemical looping technology in the Alstom Windsor lab in the USA[260]. Chemical looping is being demonstrated at the scale of 1 MW_{th} in Darmstadt, Germany[261]. Chemical looping was also found to be a promising technology for the combustion of biomass in the ETI TESBIC project, a joint industrial/academic project[262].

Development status and timescales

Table 32 is a summary of recent trials of CLC with biomass[263]. It is clear that few large-scale tests have been conducted of CLC with biomass. Abad[264] has reviewed the literature and identified promising results with Fe and Ni-based oxygen carriers during steam gasification of solid fuels. It is further stated[265] that Alstom has reported results from a CLG process using $CaSO_4$ as an oxygen carrier.

Table 32: Operation of chemical-looping combustors for biomass – studies sorted by (a) location, (b) year and (c) size of unit[263]

Location	Year	Size (kW)	Oxidation chemical	P_{fuel} (kW)	T_{FR} (°C)	Reported operational hours to date	Reference
CSIC Consejo Superior de Investigaciones Científicas, Spain	2013-2014	1.5	Spray-dried, 60% CuO	0.5-1.2	860-950	65	266,267,268
IFP Institut Français du Pétrole, France	2012	10	Natural ore, BMP	-	850-940	52	269
SU Southeast University, Nanjing, China	2009	10	Fe_2O_3	12	740-920	30	270
SU Southeast University, Nanjing, China	2009	10	NiO /Al_2O_3 (impreg.)	8.3	720-980	230	271,272,273
SU Southeast University, Nanjing, China	2011	1	Iron ore	-	720-930	-	274
OS Ohio State University	2015	1000	Iron ore	-	-	-	-

Based on ICON's expertise, CLC for biomass is at **TRL 4**, compared with CLC for coal, which is at around 5 (based on the experience at Darmstadt[275]), where a 1 MW_{th} reactor has been constructed. Since CLC is by nature a power generation technology using a basic steam cycle, the TRL is similar for CHP, heat and power applications. For the production of syngas, the technology is around **TRL 3**, with a very limited number of laboratory tests conducted.

In terms of developing a commercial plant, based on the experience of the Caoling consortium, 3 years would be sufficient for design and construction of a 50 MW_{th} plant. This is because a CFB is a standard piece of equipment and two large scale pilots (at Darmstadt and La Pereda) for either CLC or Ca-looping are already operational.

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6.2.3 Alternative cycles

Once steam or syngas has been generated from biomass, there are then a variety of options for generating power and/or heat with the intermediate product. The traditional approaches of steam engines, steam turbines, syngas boiler, gas engines/CHP and CCGT plants have been considered in earlier sections, hence this horizon scanning chapter looks at some of the alternative generation technologies that utilise different thermodynamic cycles to generate power (the remainder of the technologies can be found in Section 6.3.4).

6.2.3.1 Entropic Cycle Generation

Technology description

Entropic cycle systems are a combined heat and power (CHP) cycle quite similar to the Organic Rankine Cycle (ORC) but using higher temperature differentials to provide an increase in thermal efficiency. The entropic cycle may prove to be a cost effective cycle with high efficiencies in comparison to similar cycles, as seen in Table 33. With its relatively high efficiency and ability to be used at small scale it may have a high potential to be used for small biomass based systems, and is expected to compete primarily with the ORC and Stirling engine systems. The entropic cycle uses a closed loop fluid flow with a non-steam fluid which contains 4 times the enthalpy of organic fluids. Using this fluid allows for latent heat recuperation providing pre-vaporised fluid to the heater[276]. This potentially allows for the one of the highest overall CHP efficiencies.

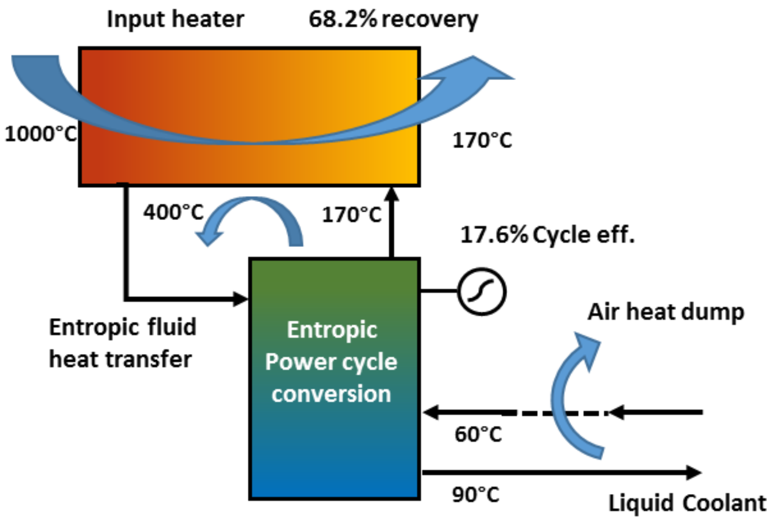

Figure 193: Entropic Cycle Power System[269]

Table 33: Efficiencies of Modular Biomass Systems, based on Conversion of Switchgrass at 20 Percent Moisture[277]

System Type	Electric Efficiency	Thermal Energy Delivered	Overall CHP Efficiency
Small steam	6%	59%	65%
Air Brayton	8%	41%	49%
Organic Rankine	11%	56%	67%
Entropic	13%	63%	79%
Stirling	13%	64%	77%
Modular Gasifier	16 - 22%	29 – 53%	55 – 75%
Hybrid gasifier/combustor	<15%	45 – 55%	60 – 70%

Development status and timescales

This cycle is still within the pilot stages of development and so not much information is currently available on it[278]. The current status is therefore judged to be at **TRL 4-5**. It will be at least 10 years before the cycle has completed the development phase and it becomes first commercially available, although this will depend on willingness of industry or research spin-outs to take the technology on and find sufficient value in the higher efficiency process.

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6.2.4 Fuel cells

Once syngas has been generated from biomass, there are then a variety of options for generating power. The traditional approaches of syngas boiler, gas engines/CHP and CCGT plants have been considered in earlier Sections, hence this horizon scanning sub-chapter looks specifically at high efficiency routes via fuel cells. The available options are:

- Bio-hydrogen use in Alkaline fuel cells
- Direct use of syngas in Solid Oxide fuel cells
- Direct use of syngas in a SOFC gas turbine hybrid

6.2.4.1 Bio-hydrogen Alkaline Fuel Cells

Technology description

Alkaline Fuel Cells (AFCs) were one of the first fuel cell technologies. However, in the 1970s and 80s as new chemistries and catalysts emerged and the alkaline FC chemicals remained expensive, research efforts shifted to other fuel cell technologies. However, as the technologies progressed it became clear that different chemistries were appropriate for specific applications. The barriers around AFCs were overcome through novel catalysts and nanomaterials. AFCs can be coupled with a biomass gasifier and a syngas conditioning unit that outputs pure hydrogen, efficiently generating electricity from biomass[279].

AFCs use potassium hydroxide as the electrolyte and non-precious metals as catalyst at the anode and cathode. Novel AFCs operate at temperatures ranging from 23 to 70°C. The high rate at which electro-chemical reactions take place in the cell lead to relatively high performance compared with other fuel cell technologies, in excess of 60% efficiency. The key disadvantage of AFCs is the susceptibility to CO_2 poisoning, making it necessary to use pure hydrogen and oxygen to extend the operating lifetime. This leads to the requirement of very durable materials, which makes AFCs less cost-effective than other technologies. Problems with durability prevent AFCs from achieving a lifetime of 40,000 cycles, which is necessary for economic viability in large-scale utility applications[280].

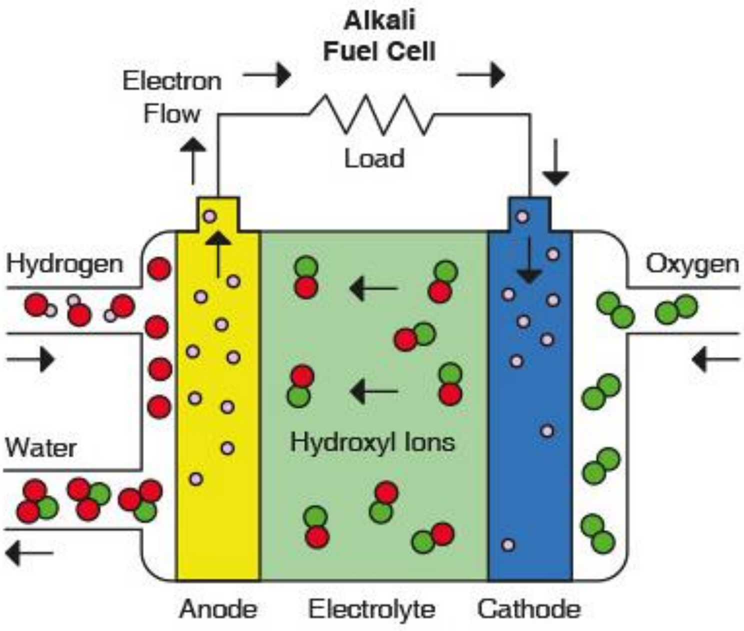

Figure 194: Operation principle of an Alkaline Fuel Cell[281]

Development status and timescales

According to the E4tech Fuel Cell Industry Review, there were no AFC units shipped in 2014[282]. However, in 2014 PowerHouse Energy (PHE) bought 2 Beta+ units from AFC Energy in order to develop a system that uses biomass as a feedstock and output electricity through a fuel cell[283]. Also in 2014, Waste2Tricity partnered with AFC energy to develop solutions to utilise their syngas derived from waste plasma gasification[284]. However, there has been no public news about the development status of those partnerships or prospective demonstration units. The current status of the use of biomass derived hydrogen in an alkaline fuel cell is therefore currently judged to be **TRL 4-5**.

Fraunhofer has also utilised the hydrogen-rich syngas from biomass pyrolysis to obtain hydrogen and use it in fuel cells at lab scale, testing out PEM, SOFC and high temperature PEM systems, as well as direct use of the pyrolysis gas in fuel cells. This route to producing power was demonstrated in 2011 but further information on progress is not available in the public domain[285], therefore the status of PEM fuel cells using bio-hydrogen is judged to be around TRL 4.

Due to the mobile nature of AFCs, it can be expected that commercial units would be rolled out at the kW scale, rather than at multi-MW scale.

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6.2.4.2 Direct syngas Solid Oxide Fuel Cells

Technology description

A Solid Oxide Fuel Cell (SOFC) produces electricity directly from oxidising a fuel in an electrochemical conversion. The key characteristic of SOFC is that it has a solid oxide or ceramic electrolyte. The main advantages of SOFC compared with other types of fuel cells include efficiency, long-term stability, fuel flexibility, low emissions and relatively low cost. A drawback of the technology is high operating temperature, which leads to longer start-up times as well as mechanical and chemical compatibility issues[286].

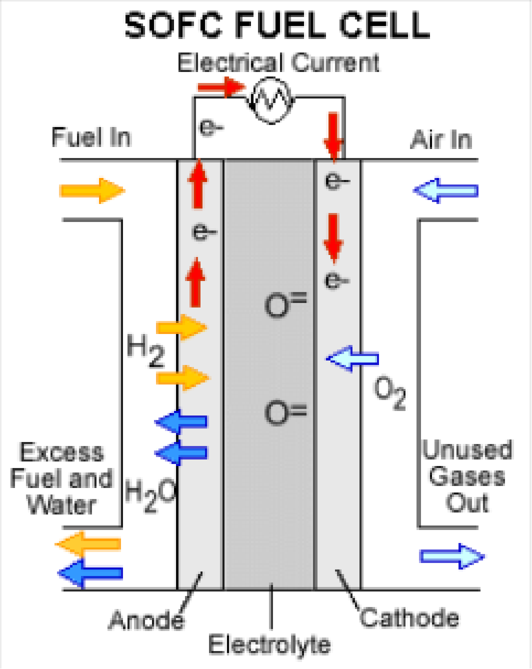

Figure 195: Operation principle of SOFCs[286]

In order to use syngas in most fuel cells, pure hydrogen needs to be isolated and used as a fuel, which adds cost and complexity of methane reforming, water gas shift and hydrogen separation operations. However, SOFCs can be fuelled with syngas directly due to higher resistance to gas impurities than other fuel cells, good matching in temperatures with common gasifiers (downdraft and fluidised bed) as well as potentially a good matching in scales. Another advantage is that the heat of combustion is lower than in the case of pure hydrogen, which makes thermal management easier. Theoretical studies on performance analysis showed potential CHP efficiencies of a gasifier and SOFC of around 80%. Such installations are termed Integrated Gasifier Fuel Cells (IGFC)[287].

However, using syngas as a fuel has a number of challenges, the key one being the presence of tars. Although they can act as a fuel, fouling due to carbon deposition on the anode reduces its performance[288].

Development status and timescales

In 2014, estimated global SOFC capacity shipped amounted to 32.3 MW. The technology can be considered mature and commercially available, but using natural gas or other fossil fuels[289]. IGFC is much less developed but there are a number of R&D projects taking place worldwide. A hybrid SOFC-MGT 600 kW_{th} turbine has been demonstrated by VIKING, as described in section 6.2.4.3. The current status for direct use of syngas in a SOFC is therefore judged to be **TRL 5** for the overall IGFC concept.

The development of IGFC tends to be driven by the search for novel uses of coal[290]. Siemens is particularly active in this field, working with the US DOE on feasibility studies of multi-MW_e applications[291]. The US Department of Energy's development timeline for IGFC (for any feedstock) is the following[292]:

- 125 kW_e module test in 2016
- 250 kW_e Power System Field Test in 2018
- Utility-scale IGFC central station post 2020

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6.2.4.3 Fuel cell gas turbine hybrid

Technology description

Hybrid fuel cell gas turbine systems comprise of two major components – a fuel cell that generates electricity and heat in the process, and a heat engine that uses the regenerated heat to obtain additional electrical energy[293]. It has been proposed to supply hydrogen fuel produced from the gasification of biomass. Several heat engines are feasible for this type of system including gas turbines, steam turbines and reciprocating engines. However, the only conversion device that has been tested in this role to-date is a micro-gas turbine (MGT), which is well matched to the high temperature requirements of a fuel cell in a hybrid system[294] [295]. As fuel cells advance and scale-up and pressurization of SOFC technology becomes viable, larger and more sophisticated gas turbine engines (e.g. axial compressors and turbines, higher pressure ratios, high turbine inlet temperature) will be required[296].

The fuel cell and gas turbine of a hybrid system can be configured in several different ways with many potential cycle configurations. For example, a fuel cell can be operated in between the compressor and turbine of the gas turbine engine. Another configuration is when the turbine operates on air that does not come into contact with the fuel cell exhaust, but rather receives heat through a heat exchanger

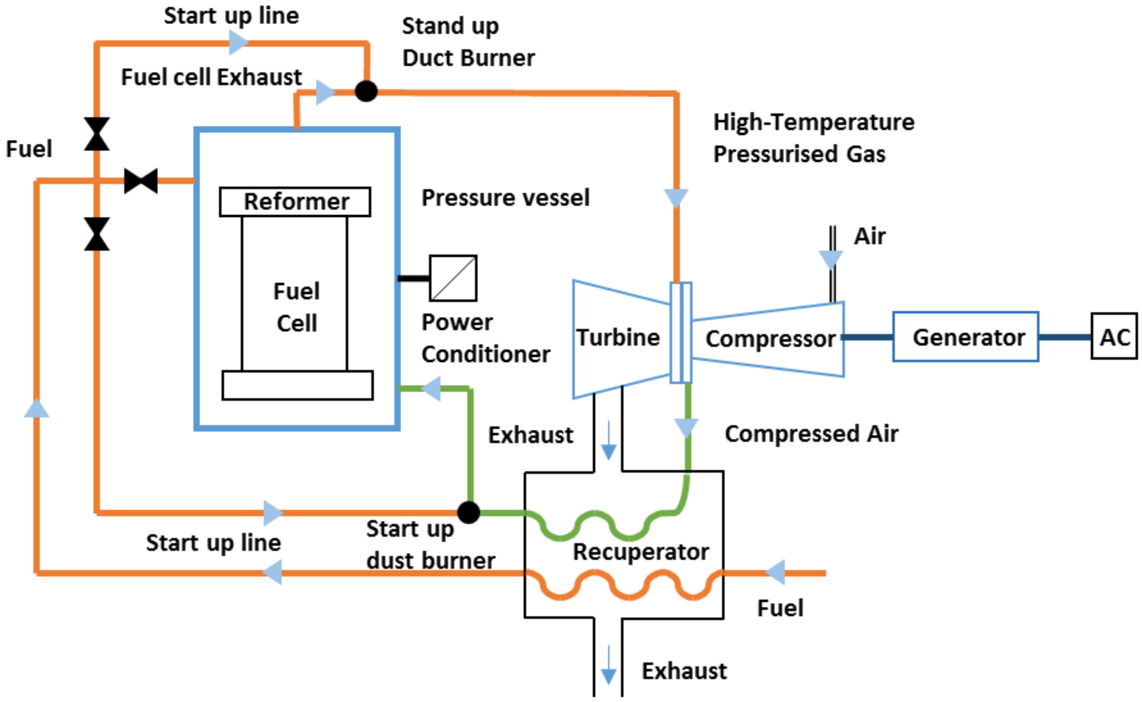

Figure 196: Schematic of a direct hybrid gas turbine fuel cell topping cycle[294]

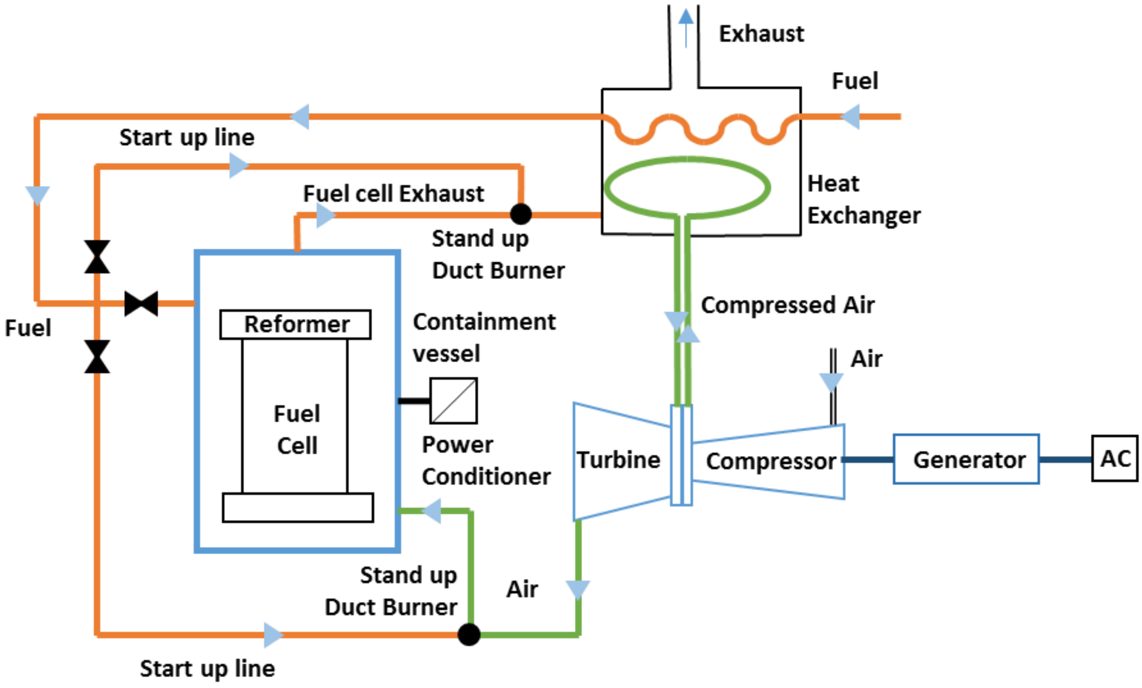

Figure 197: Schematic of an indirect hybrid gas turbine fuel cell bottoming cycle[294]

Development status and timescales

In 2000 the first tests and demonstrations of hybrid gas turbine fuel cell systems began with efforts in the U.S. and Japan. SOFC hybrid systems have been built and tested each proving the potential for such systems to achieve high efficiency and low emissions production of electricity from natural gas. To-date five types of hybrid gas turbine fuel cell systems have been tested, each using a different design concept.

A number of experimental studies have been conducted on biomass gasification coupled with SOFC and gas turbine hybrid systems[297]. A concept of coupling an IGFC with a fluidised bed gasifier has shown potential scalability up to the range of 3-10 MW[298]. However, the largest demonstration (built by VIKING) to date has been 0.6 MW_e[299]. No commercial schemes have been developed to date, but the current status is judged to be **TRL 5**.

In 2002 a consortium of energy companies including EDF, EBW and Siemens, under funding from the European Commission, completed a feasibility study of constructing a MEGASOFC demonstration project in Marbach Germany in the range of 250-350 kW. However, the project has not materialised[300]. An interesting recent development is the MHI “triple combined cycle”, a hybrid of SOFC and gas combined cycle power plant which has a target electrical efficiency of 75%.[301]

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6.2.5 Aqueous Phase Reforming

Technology description

Aqueous phase reforming (APR) is a technology to process biomass to produce hydrogen, chemicals and/or liquid fuels[302]. Before undergoing reforming, lignocellulosic biomass needs to be converted into a liquid state, which can be done through pyrolysis (most common approach), cellulosic hydrolysis to sugars[303] or biomass deconstruction[304].

APR enables processing of wet biomass resources without energy intensive drying, leading to significant energy savings. The process is unique in that the reforming is done in a liquid phase and it does not require non-renewable inputs (e.g. no need for external hydrogen for deoxygenation of the products). Furthermore, it occurs at temperatures of 220–450°C and pressures up to 200 bar where the water-gas shift reaction is favourable[305,306], making it possible to generate hydrogen with low amounts of CO in a single chemical reactor. By taking place at low temperatures, the process also minimizes undesirable decomposition reactions typically encountered when carbohydrates are heated to elevated temperatures[307]. However, most of the routes being developed are reliant on the production of intermediate sugars (whether from food or lignocellulosic crops), which is not compatible with heat and power applications.

Virent Energy Systems has exclusive rights to the APR process. However, other chemical companies use derivatives of the process[308]. For example, BTG developed Liquid Phase Processing (LPP), which is the same as APR when water is used as a solvent.

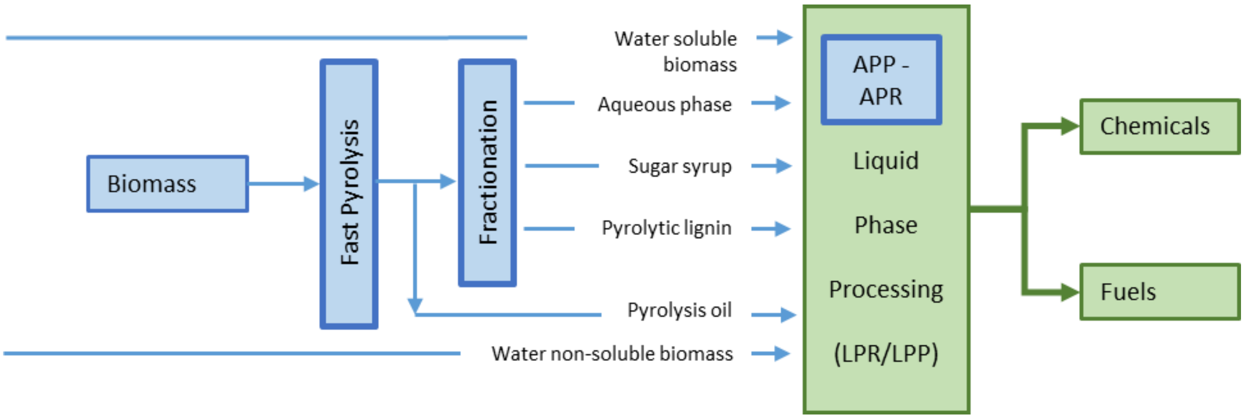

Figure 198: APR process, shown as a part of Liquid Phase Processing technology developed by BTG[309]

Development status and timescales

APR was discovered in 2001 and first developed at lab-scale in 2002 by J.A. Dumesic at the University of Wisconsin[310]. Virent initially were looking at hydrogen and natural gas[311], but have switched to focus to diesel, gasoline and jet fuels, plus chemicals such as para-xylene. Their 10,000 gallon per year pilot plant in Madison, USA has been operational since 2010, and was replicated at Shell's site in Houston in 2012[312]. Both are small units, i.e. not demonstration facilities.

In the EU-funded project "SusFuelCat", started in 2013, six universities and research institutions, three SMEs and Johnson Matthey (from Germany, Finland, the Netherlands, UK, Spain, Italy and

Russia) are collaborating to pilot hydrogen production using BTG's Liquid Phase Processing technology[313].

Further work on effective and low cost catalysts to reduce the cost and produce more value-added liquid fuel products is still necessary[314]. As well as readily accessible sugars from sugarcane, sugarbeet and corn syrup, Virent's Madison pilot plant has also been tested on hydrolysates from cellulosic biomass including pine residues, sugarcane bagasse and corn stover. An end-to-end woody biomass to fuels plant is yet to be established though, and the end product (biofuels and high value chemicals) are out of scope of the TEABPP project.

The current status of the technology is **TRL 5**, given the pilot plant activities. However, the commercialisation timescales and potential commercial plant sizes are unknown – Virent's ongoing collaboration with Shell on liquid fuels and Coca-Cola on para-xylene is yet to produce any scale-up plans for the future[315]. It is unlikely that the APR technology would be adapted to convert sugars into heat and power, as these are lower value uses compared to uses in transport fuels and chemicals for a valuable intermediate such as sugar.

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6.2.6 Hydro-Thermal Upgrading

Technology description

The Hydro-Thermal Upgrading (HTU) process is the hydrothermal conversion of biomass into an organic crude oil (bio-crude). This liquefaction process was developed to convert biomass with high moisture content into a product with high energy density[316].

In the HTU process, the reaction temperature ranges from 300 to 360°C. The pressure is 17-27 MPa and the residence time 5-20 min. The bio-crude produced is not miscible with water and the heating value is 30-35 MJ/kg. During this process, the oxygen content of the organic material is claimed to be reduced from about 40% to between 10 and 15%. The removed oxygen ends up mainly in CO_2 and water[317].

The resulting product is then more efficient to transport than raw biomass and is flexible in terms of end-uses. It can be directly combusted as a liquid, co-fired with a solid fuel; it can replace charcoal or could be upgraded into products such as biofuels or chemicals[318].

Due to the severe process conditions, industrial application of these processes suffers from various challenges. Corrosion requires the use of expensive alloys and the high operation pressures put tough requirements on process components such as feed pumps. The high investment cost is also a considerable hurdle for commercialization[319].

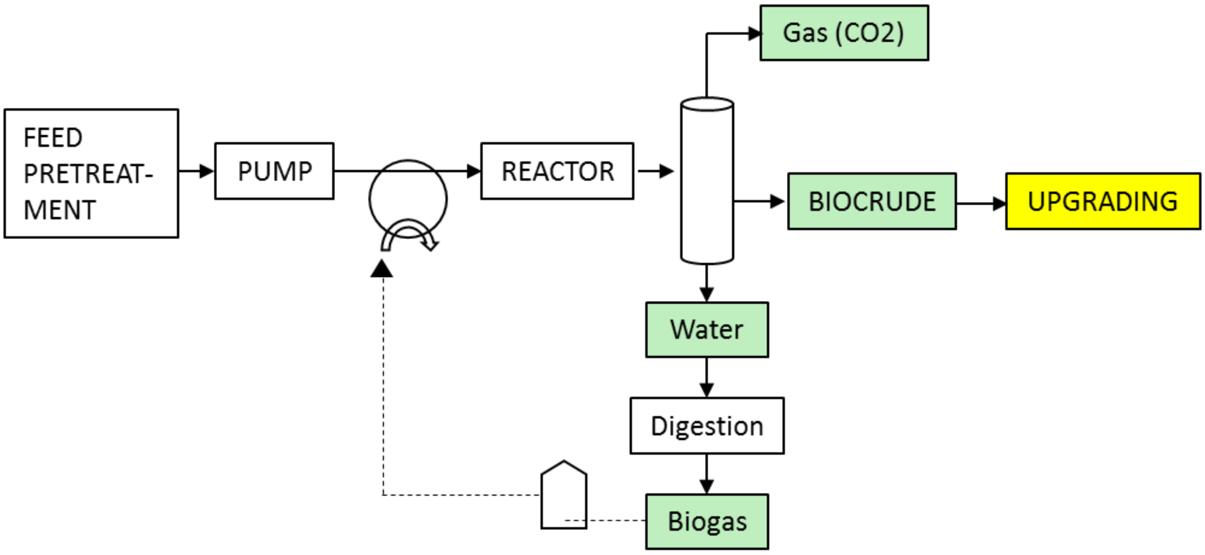

Figure 199: Block scheme of HTU process[320]

Development status and timescales

HTU was invented a few decades ago and by the end of the 1970's to early 1980's several laboratories, mainly at universities, were working on the technology. Most work on hydrothermal liquefaction has so far been carried out at lab- or bench-scale. Shell was one of the original developers, but halted work in 1988 and in 1993 due to low oil prices.

Together with TNO and the Biomass Technology Group (BTG), Shell and Biofuel B.V. set up a 100kg/hr feedstock input pilot plant in Apeldoorn, Netherlands[321]. Although the pilot plant was completed, running from 1999 to 2000, further development work in scaling up has not been achieved to date[322]. The current status of the technology is **TRL 4**, given that the only pilot plant that has shut down and industry interest has stopped (e.g. Shell has now invested in APR). The potential capacities of commercial plants and development timescales are unknown, as the technology may never be commercialised.

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6.2.7 Microwave Pyrolysis

Technology description

In microwave pyrolysis processes, the biomass feedstock is mixed with a highly microwave-absorbent material such as particulate-carbon[323]. This material absorbs microwaves to generate sufficient thermal energy to achieve the temperatures required for extensive pyrolysis to occur. As a result of microwave heating, the biomass material is thermally cracked in the absence of oxygen into smaller molecules. The resulting volatile products are either re-condensed into an oil product (pyrolysis oil) or collected as incondensable gaseous products (pyrolysis gases) of different compositions depending on the reaction conditions[324].

Microwave pyrolysis has a number of advantages over conventional pyrolysis processes. Firstly, microwaves achieve a rapid heating rate and make uniform temperature distribution throughout the fuel more readily achievable. Secondly, the products' quality are improved, due to the fact that the contact between the pyrolytic vapours released and the hot char is reduced along with char-catalysed secondary cracking in the fuel bed[325,326]. Microwave pre-treatment is a highly effective method for biomass. In a recent study done by BDC York they found a 340% increase in biogas yield when maize silage was pre-treated using microwaves[327].

However, there are many challenges associated with the reactor durability that hinder the proposed advantages at present. Those include deposits of volatile vapours and fine carbon particles on the vessel walls, as well as achieving homogeneity in the microwave field in order to avoid hotspots[328, 329].

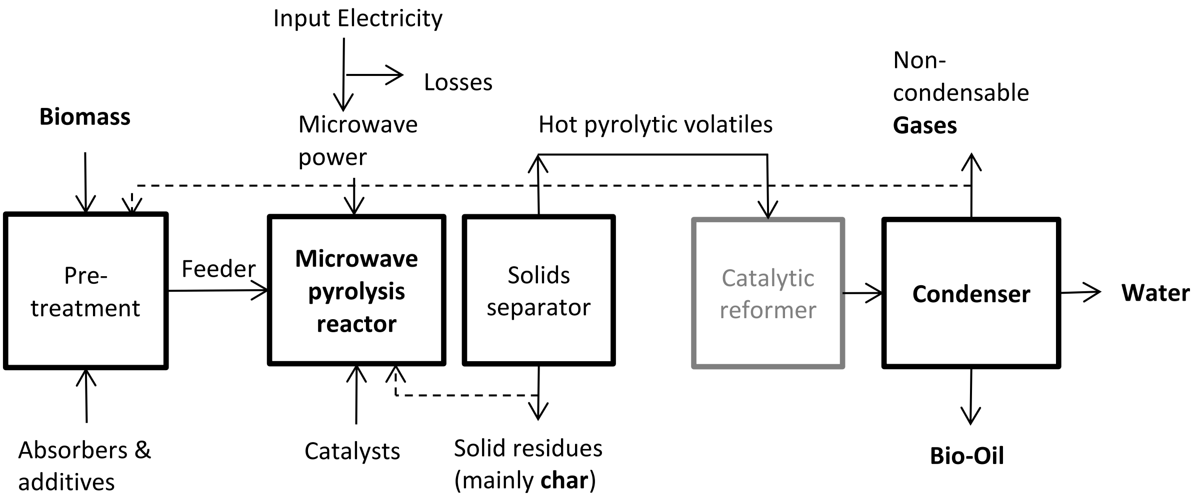

Figure 200: Microwave pyrolysis of biomass[304]

Development status and timescales

There are a number of commercial advancements made in microwave pyrolysis of wastes and plastics, however, primarily focused on waste plastics and car tyres, not biomass. For example, Enval developed a modular process, through which the plant can be economically operated at a variety of scales, which allows for local treatment of wastes. A typical Enval plant operates at a feed rate of up to 350 kg per hour, which equates to a nominal input capacity of 2,000 tonnes per year[330]. Their demonstration plant is claimed to be able to produce up to 1,200 tonnes per year of bio-oil and 250 kW_e output from burning the pyrolysis gases.

Scandinavian Biofuel Alliance and AMT have recently patented a technology based on microwave pyrolysis that allows recycling any material of hydrocarbon origin and have plans to pursue the industrialisation and commercialisation of it[331]. Even though, microwave pyrolysis of woody biomass remains a lab-scale process. Another example is the Green Chemistry Centre of Excellence/Biorenewables Development Centre at the University of York, which has facilities enabling to conduct research on the controlled microwave decomposition of biomass[332] [333]. The current TRL is therefore only **4**.

The timescales to commercialise the technology are currently unknown.

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6.2.8 Hydrothermal Carbonisation

Technology description

The Hydrothermal Carbonisation (HTC) process involves chemical and physical dehydration of biomass in order to transform the biomass feedstock into an energy dense material with similar properties to coal (the heating value is approximate the same as that of dry, high quality brown coal). The resulting powder can be used directly for combustion or further refined through pelletizing or gasification to produce syngas[334].

The first step in the process is preparation of biomass by washing to remove impurities, mashing as well as pressurizing and pre-heating with steam. Secondly, the resulting slurry is fed into the HTC cooker for carbonisation. A mechanical press is then used to physically dehydrate the slurry before entering the thermal drying stage, where dispersed biomass is heated under weakly acidic conditions in a closed vessel to temperatures of around 200°C[335].

The process is claimed to be relatively cheap and scalable, mainly due to the low temperatures and pressures required (up to 25 bar). HTC inherently requires wet liquid biomass and the final carbon needs filtering from the reaction solution, with waste water a main by-product. Under acidic conditions and below 200°C, most of the original carbon stays bound to the final structure (developer claim of 90%). The output product (carbon powder) produced by this technology can replace coal and be hence used for bioenergy production, soil sequestration or materials use[336].

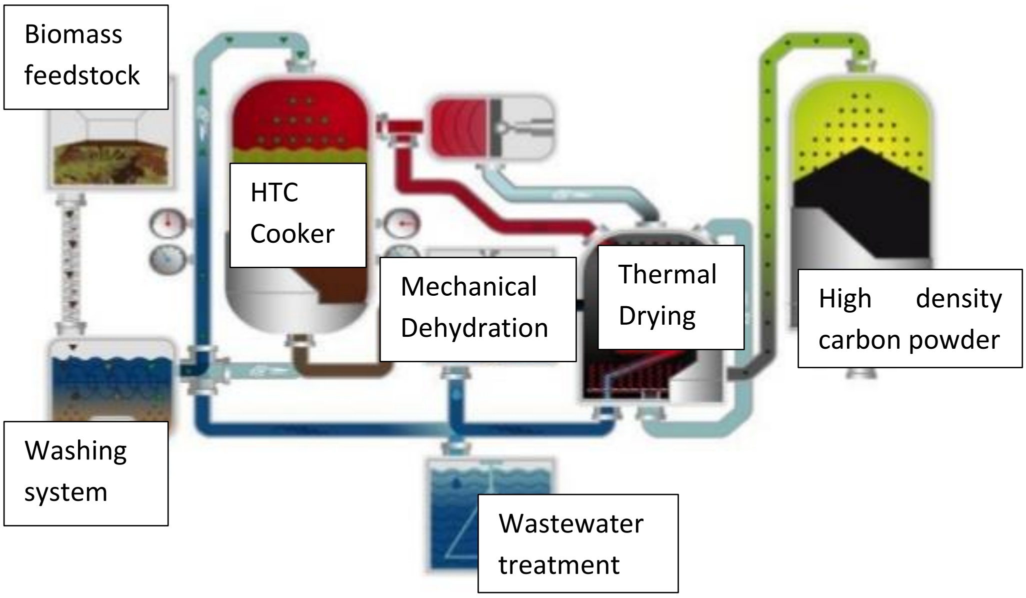

Figure 201: Hydrothermal Carbonization process[337]

Development status and timescales

Hydrothermal carbonisation was first described in 1913, but revived in 2006 by the Max-Planck Institute of Colloids and Interfaces in Golm/Potsdam (MPI). In 2007, in co-operation with the MPI, the University of Applied Sciences Ostwestfalen-Lippe completed the first feasibility study on the HTC process using organic waste, funded by the DBU (Deutsche Bundesstiftung Umwelt)[338].

In 2012 SunCoal successfully completed an installation HTC pilot plant in Ludwigsfelde, Germany. However, there is no more recent available information on whether they have managed to attract clients to take up their CarboREN technology[339,340].

Separately, in 2014 Antaco received a £858k grant from DECC to develop a 500 tonnes of wastes per annum HTC plant in the UK, although it appears nothing has happened to date[341].

The current status of the technology is judged to be around **TRL 5**.

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6.3 Technologies with higher Technology Readiness Levels

This section presents technologies that are at a late-development stage, however not enough data is available in the public domain to conduct the benchmarking.

6.3.1 Syngas clean-up technologies

The present tar reduction or destruction methods can be divided into five main processes as described in section 2.5: mechanical methods, self-modification, thermal cracking, catalytic cracking and plasma methods.

The syngas tar removal technologies discussed in this horizon scanning chapter are three additional or innovative variations of methods for handling gas phase tars after their production:

- Tar cracking using plasma, which would replace thermal cracking
- Multi-stage oil scrubbing, which would replace wet scrubbing
- Catalytic tar removal through part-combustion, which would replace thermal cracking

6.3.1.1 Plasma Clean-up

Technology description

Plasma clean-up is typically implemented straight after gasification, and before final syngas scrubbers and the downstream conversion technology. The technology is a variation of thermal cracking.

Crude syngas exiting the gasifier is passed into the plasma conversion unit, where the heat from an electrically generated plasma arc results in cracking of tars and the breakdown of char (in some cases, accelerated by ultraviolet light). The cracking creates a clean syngas, whilst the inorganic elements in the ash carried over from the gasifier are vitrified. The clean syngas exiting the plasma converter is then cooled and further conditioned through wet and dry scrubbers before being used directly to generate electricity[342].

It is claimed that the process is capable of achieving energy conversion rates of around 90% in terms of syngas production from the starting waste feedstock – although having knowledge of the different gasifier designs, this will be an upper bound on the syngas cold gas efficiency, and will not consider parasitic power and heat loads, or further downstream syngas clean-up[342]. The net exportable power generation efficiency for a commercial scale plant is claimed to be over 25% when using waste feedstocks[343].

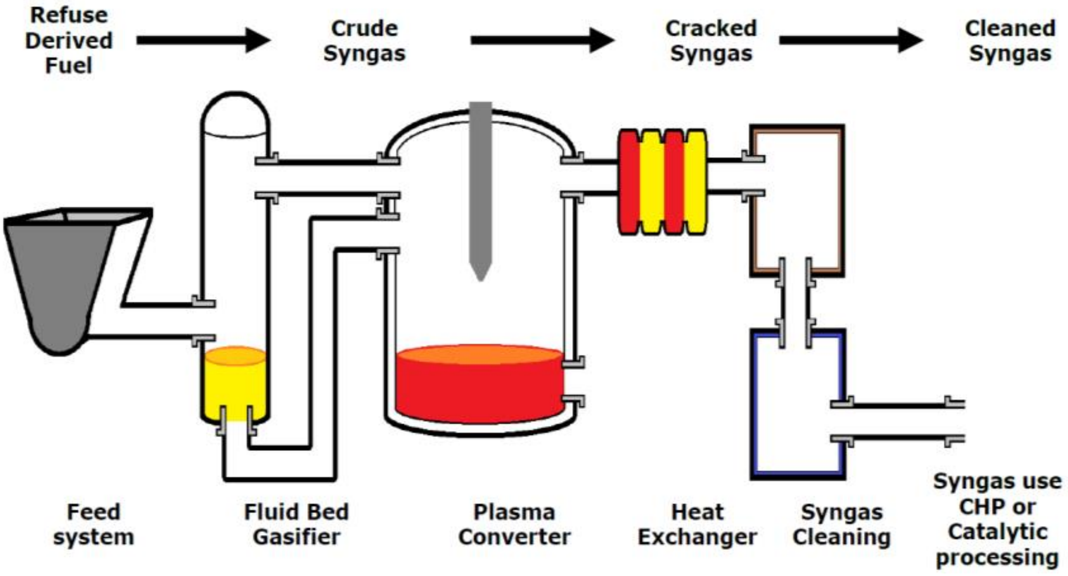


Figure 202: Gasplasma process[342]

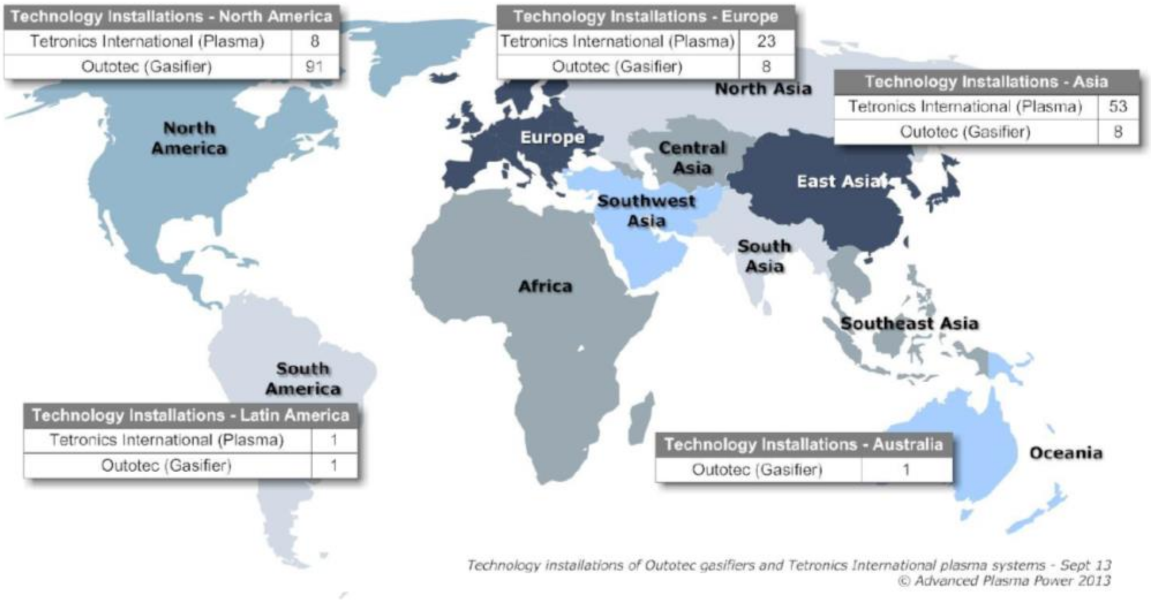
Development status and timescales

Tetronics is the main actor responsible for commercialising the plasma cleaning technology, to date they installed over 80 commercial waste treatment facilities using high temperature DC Plasma Arc Technology in industrial and metallurgical applications[344]. The consortium of Tetronics and Advanced Plasma Power (APP) offers a combined gasifier + clean-up system under the Gasplasma brand[345]. Tetronics has Development is predominately focusing on Municipal Solid Waste (MSW) and Refuse

Derived Fuel (RDF)[346], due to the benefits of removal of harmful chemicals and feedstock components in the inert slag, which can be sold as aggregate material into the building sector. However, there is little evidence of the number or effectiveness of these systems, how long they have been operating, or if they are still operational. There is also no evidence presented to date on the use of woody biomass feedstocks.

In terms of other developers, these are at demonstration stage. Plasco has been operating a 70 odt/day MSW demonstration plant since February 2008 in Ottawa, Canada, exporting 4.2 MW_e of power. There is also a 12 MW_e demonstration plant in Morcenx, France[347] owned by CHO-Power that started up in 2012 (using a Europlasma plasma torch for syngas clean-up). Although MSW remains a dominant feedstock for all these gasification plants, woody biomass also can used - for example, Morcenx treats 37,000t of MSW and 15,000t of wood chips annually[348]. Plasma syngas clean-up is therefore judged to be at **TRL 8**.

There is no quoted maximum or minimum size for the technology available.



Technology Installations - North America	
Tetronics International (Plasma)	8
Outotec (Gasifier)	91

Technology Installations - Europe	
Tetronics International (Plasma)	23
Outotec (Gasifier)	8

Technology Installations - Asia	
Tetronics International (Plasma)	53
Outotec (Gasifier)	8

Technology Installations - Latin America	
Tetronics International (Plasma)	1
Outotec (Gasifier)	1

Technology Installations - Australia	
Outotec (Gasifier)	1

Technology installations of Outotec gasifiers and Tetronics International plasma systems - Sept 13
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Figure 203: APP's and Tetronics' experience in gasification and plasma conversion installations worldwide[349]

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6.3.1.2 Solvent Scrubbing

Technology description

In solvent scrubbing technology, tar removal is based on a multiple stage scrubber in which syngas is cleaned by a solvent. The method provides high removal efficiency, particularly for heavy and heterocyclic tar compounds[350]. The main advantage of the solvent-based scrubbers is that due the recycling, they achieve higher efficiency than traditional water-based scrubbers, even when a wet electrostatic precipitator is used[351]. They also eliminate the issues involved with tar condensation and water solubility (a particular problem in wastewater treatment, where small non-polar components may dissolve in small amounts that can exceed allowable concentrations).

The most commercially developed solvent scrubbing technology for applications in the bioenergy sector is OLGA, which uses a proprietary oil as the scrubbing solvent. Figure 204 shows the OLGA process steps in detail. In the first section, syngas is cooled down by the scrubbing oil. Heavy tar particles condense and are collected, after which they are separated from the scrubbing oil in the stripper and can be recycled to the gasifier. In the second section, lighter gaseous tars are absorbed by the scrubbing oil, which becomes saturated when the mixture reaches the absorber column. The saturated oil is then passed through a stripper, where it gets regenerated by stripping the tars using hot air or steam. All heavy and light tars can be recycled to the gasifier and contribute to the overall energy efficiency.

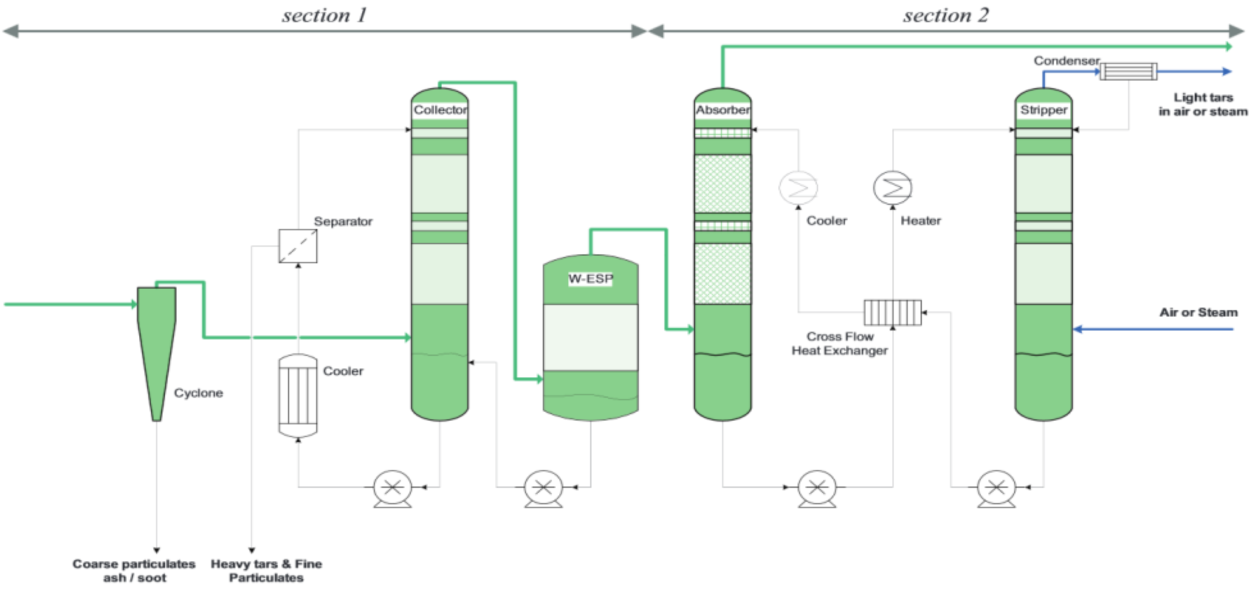

Figure 204: OLGA clean-up process[350]

Development status and timescales

A number of solvents have been tested on a lab scale[352], however OLGA remains the major technology that has been demonstrated in biomass applications and has significant commercialisation plans.

The OLGA tar removal technology is a patented invention of the Energy Research Centre of the Netherlands (ECN). One of the earliest pilot-scale deployments of oil-based tar scrubbers was demonstrated at ECN in 2004 (at a scale of 25 $\mathrm{kW_{th}}$). Since then, ECN has partnered with an engineering company Royal Dahlman and developed a number of demonstration plants utilising the OLGA technology, in France, India and Portugal[353]. In 2014 a 1 MW plant was completed in India at the RSIL site in Mahashtra, India[354], with a similar 1 $\mathrm{MW_e}$ plant started up in Portugal[355]. These demonstrations place the technology currently at **TRL 7**.

In 2014 ECN and Royal Dahlman signed a deal to develop a 6 $\mathrm{MW_e}$ biomass plant in China[356], which is planned for commissioning in 2016. A 7 $\mathrm{MW_e}$ plant to be built in Grimsby, UK has been granted planning permission from the North East Lincolnshire Council[357]. In the pipeline there are also biomass plants in Korea and the Netherlands (the much delayed 12 $\mathrm{MW_{th}}$ HVC project).

A likely scale for commercial CHP units based on OLGA technology is expected to be around 10 $\mathrm{MW_{th}}$, although ECN did have much larger plans for single-train methanation facilities of 50-500 $\mathrm{MW_{th}}$[358].

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6.3.1.3 Catalytic Tar Removal

Technology description

Tar cracking is used to break down tars into simpler and lighter molecules by the action of heat and in the presence of a catalyst (but without the addition of hydrogen)[359]. Many different catalysts have been used to eliminate the tars in the product gas from gasification. The two most researched groups are Ni-based catalysts and dolomites[360].

When Ni-based catalysts are used, tar concentrations in the product gas can be reduced significantly by means of reforming but since the process is endothermic, a part of the chemically bound energy of the gas has to be burned to sustain it, reducing the overall efficiency of the gasification process.

In contrast, when tar cracking catalysts such as dolomite are used, the only thing that is reformed is the tar itself while low molecular weight hydrocarbons e.g. methane, ethane and propane are left intact. The process cracks the high-boiling, high molecular weight oxygenates, phenolics, ethers, poly-aromatic hydrocarbons (PAH) and large PAHs into lower boiling-point, lower molecular weight olefinic and aromatic hydrocarbons and then hydrogenates them.

Catalytic tar cleaning has been proven to be an effective method at high temperatures, as it has no additional energy input, efficiency and heating value losses are kept at a minimum, and no tarry waste streams are generated that need to be disposed of or recycled to the gasifier. At the same time, the need for collection, disposal, or returning of the tar (containing solvents) is also eliminated[361]. Catalytic cracking, unlike thermal cracking, has the benefit of being able to take place at gasification temperatures of 600 – 800°C. Nevertheless this method also has its shortcomings as the catalysts tend to be rendered inactive after a period of use. For example, with prolonged use, commercial alkali metal and Ni-based catalysts are rendered inactive by carbon and hydrogen sulphide deposition, while dolomite catalysts are rendered inactive by broken particles.

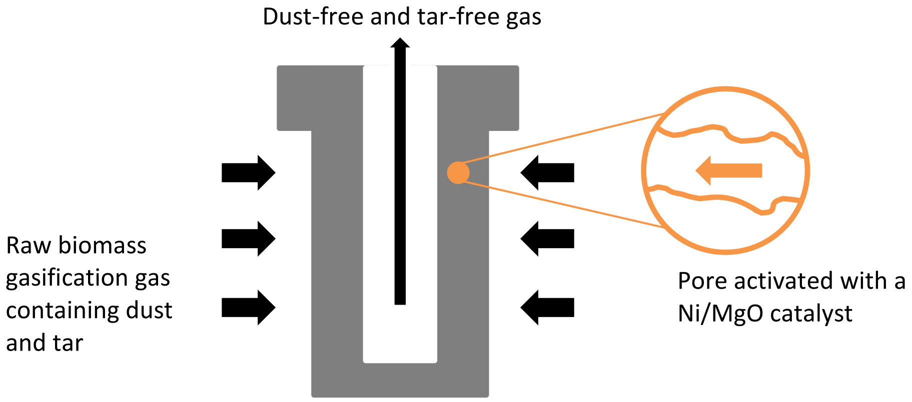

Figure 205: Tar removal using Ni-based catalyst

Development status and timescales

In 2008, Nexterra began a collaboration with GE Energy to develop a syngas conditioning system. Nexterra now believes that their technology is a commercial proposition as a result of[362]:

- Over 5,000 hours of operation of the conditioning system at their site development centre
- Over 3,000 hours of operation on a 239 kW_e GE Jenbacher Engine on the conditioned syngas
- Installation of a 2 MW_e CHP system at the University of British Columbia which started up in 2012.

Dahlman also has a simple TARA tar reduction system, using a fixed bed with catalytically active material to lower (but not complete removal of) the dust and tar contents, and dew point. This appears to have been tested at pilot scale.

Presuming that the Nexterra demonstration plant has operated successfully, the technology is judged to be at **TRL 7**. Nexterra are also due to be providing the gasification technology for Birmingham Bio Power's proposed 10 MW_e plant in Tyseley, UK, which will gasify waste wood to generate steam for a turbine – it is unclear what the syngas clean-up step will entail at this stage[363].

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[362] Nexterra "Syngas Conditioning". Available at: http://www.nexterra.ca/files/syngas-conditioning.php

[363] http://www.nexterra.ca/files/BBPL-project.php

6.3.2 Pyrolysis focused on gas and biochar products

Technology description

Fast pyrolysis pre-treatment and gasification conversion technologies were discussed in early chapters. However, research on pyrolysis also focuses on intermediate and slow modes, which produce significantly less bio-oil, but produce much more gas and biochar. This slow pyrolysis is therefore interesting to the horizon scan chapter, not as a pre-treatment option (as with fast pyrolysis bio-oil), but as a potential thermal conversion technology to heat and power with co-product solid biochar.

Biochar (i.e. charcoal) is already being used widely in various industries for producing silicon, carbon bisulphite or metals[364]. Even though biochar could be used as a direct replacement of coal, use of biochar for energy applications is uneconomic (due to the low process yields from biomass), and therefore uncommon. In the process of slow pyrolysis, gases can be either recirculated and used for internal heat provision, or cleaned and combusted to produce heat to external heating applications. Table 34 presents a list of various slow pyrolysis plants, which differ according to their heat source. Product yields are dependent on pyrolysis type[365].

Table 34: Pyrolysis plants producing heat and charcoal[366]

System	Heating efficiency (%)	Capital costs per tonne of charcoal (k€/t)	Production per unit reaction volume (t/year/m^3)
Direct Heating			
JCKB retort	23%	0.18	12.6
University of Hawaii	50%	0.18	594
Indirect Heating			
Twin Retort Carboniser	33%	0.38	70
Policor (Ecolon System)	25%	0.06	71
Enviro Carboniser	53%	0.14	192
LSIWS Carboniser	n.d.	0.27	63
Heating with recirculating gases			
Reichert	34%	n.d.	34
Lambiotte CISR	30%	0.36	16
Lurgi process	35%	0.32	10
Rheinbraun process	n.d.	n.d.	265

For pyrolysis of most biomass feedstocks, the resulting gas is typically hydrogen-rich (in excess of 80%), which makes it a valuable resource in downstream applications such as fuel cells. Fraunhofer claims that pyrolysis has a number of advantages over conventional reforming, including less expensive catalysts without precious metals, a simpler construction and the lack of process water[367].

Development status and timescales

Most of the development of slow pyrolysis to date has either been for the low efficiency production of charcoal for non-energy uses, or as a result of treating wastes and biomass to produce biochar for use in enhancing soil quality. Companies involved in this area of land and soil remediation and carbon storage include Balt Carbon[368], Cool Planet[369], Carbon Terra[370], Pacific Pyrolysis[371] and Carbon Gold[372]. These companies are not currently focusing on optimising gas yields and energy production; rather, they are concerned with maximising the value of the biochar product.

As shown in Table 34, a number of plants producing biochar/charcoal have been developed worldwide. The status of the technology can therefore be judged as **TRL 8-9**, but the economic drivers are poor, and many of the charcoal plants are old, closed down or only in partial operation. The technology is well established, but has not yet been fully optimised to maximise the gas and biochar yields from specific feedstocks such as perennial energy crops.

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[370] Carbon Terra Company Website. Available at: http://www.carbon-terra.eu/en/biochar

[371] Pacific Pyrolysis Company Brochure. Available at: http://pacificpyrolysis.com/images/PacPyro%20Brochure%20-%20Email%20Version.pdf

[372] Carbon Gold Case Studies. Available at: http://www.carbongold.com/kilns-biochar-production/case-studies/

6.3.3 Externally Fired (biomass direct combustion) Gas Turbine

Technology description

The process known as the Brayton Cycle is carried out by initially compressing the working fluid (usually air) then transferring it through a constant pressure combustor where it is used to burn a fuel, typically natural gas or biogas; the increase in temperature at a constant pressure provides an increase in velocity of the gas to allow effective expansion over the turbine stage of the cycle. The expansion of gases over the turbine produces work for the compressor and the remainder is used by the generator to produce power.

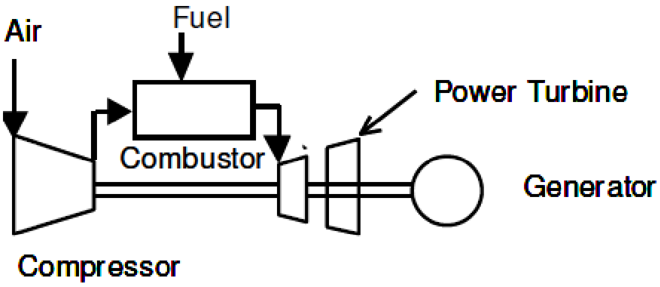

Figure 206: Components of a simple cycle gas turbine[373]

However, the heat input can be biomass, in order to externally fire the gas turbine. Zilkha is one of the main developers of this direct biomass combustion turbine technology. Their “Zilkha Biomass Unit” generates electricity and heat by burning pellets or residual biomass from the wood processing industry. The unit consists of pressurized combustor, a cyclonic separator and a gas turbine. They state that different configurations can produce between 1.5 and 20 MW_e of electrical power, with multiple units able to be installed in parallel to serve larger energy demands.

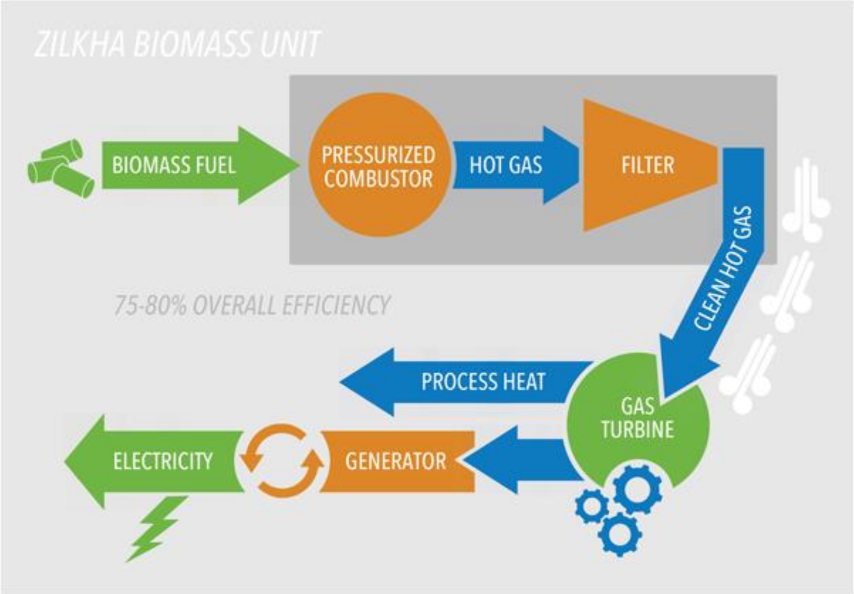

Figure 207: Zilkha unit configuration[374]

The technology combusts wood at high pressure and sends the gases produced by combustion into a gas turbine. Energy is then generated via two methods. The gas turbine's compressor section creates high pressure air used as both conveying air and combustion air. Then the gas turbine turns a generator to create electricity. The exhaust gas from the gas turbine is in the range of 525°C and is used for drying products and making steam and/or additional electricity. When producing both electricity and hot exhaust gas for direct heating, Zilkha claim that their technology can operate at above 80% overall CHP efficiency. If a heat recovery steam generator (HRSG) is used to provide steam for use by the host, the system operates at approximately 70% overall CHP efficiency. If an HRSG is instead used to provide steam to a steam generator to make additional electricity, the system is claimed to be able to achieve 50% electrical efficiency, approximately twice the efficiency of a traditional biomass combustion and steam turbine configuration. These claims will of course depend on the level of gas cleaning that needs to occur from the feedstock combustion.

Work by University of Manitoba in 2003 highlighted the low efficiency (typically <10%), larger equipment sizes and very high parasitic load consumed by the air compressor as the main disadvantages of the approach[375].

Development status and timescales

In 2006, Zilkha installed what it claims was its "first commercial" 1.5 MW plant in the city of Jaffrey, USA – however, this is small even for an open cycle gas turbine, hence more likely to be a demonstration plant. Zilkha claim that a number of commercial and industrial facilities have been powered with the technology since, but no details are in the public domain.

Another developer of the same technology is Bio-Power Technologies Ltd. BioPower Technologies Ltd is a UK company developing biomass power solutions for steam and power generation using soft biomass, i.e. straw and husks. The technology does not process harder woody materials but may be applied to Miscanthus. BioPower Technologies Ltd is working with EGOVITA SP Zoo, which is a Polish company developing biomass processing solutions for both the food industry and as a fuel. The companies are jointly developing a 2 MW CHP gas turbine demonstration, funded by the European Union FP7 Life+ More energy programme. Demonstration is expected during 2015.

Whilst the gas turbine or micro-turbine is well developed for gaseous and liquid fuels at TRL9, the status for biomass fired gas turbines is currently **TRL 6-7**.

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[373] EPA (2007) "Biomass combined heat and power catalog of technologies". Available at: http://www.epa.gov/chp/documents/biomass_chp_catalog.pdf

[374] Zilkha "Our Biomass Turbine". Company Website. Available at: http://zilkha.com/our-biomass-turbine

[375] E. Bibeau (2003) "Decentralised biomass power production", Available at; http://home.cc.umanitoba.ca/~bibeauel/research/papers/2003_Bibeau_eerc.pdf

6.3.4 Alternative cycles

6.3.4.1 Stirling Engines

Technology description

A Stirling engine is a piston-cylinder arrangement that functions through the expansion and contraction of a working fluid when exposed to temperature differentials. In comparison to internal combustion cylinder systems that are used regularly in today's automobiles, Stirling engines have the opportunity to be more efficient, and they have the opportunity to operate efficiently at much smaller scales < 100 kW_e, than other forms of biomass-based electricity generation technologies, including in CHP systems. With working fluids chosen to best suit the application, there are also various different component set-ups that can be chosen and these engines can vary quite significantly.

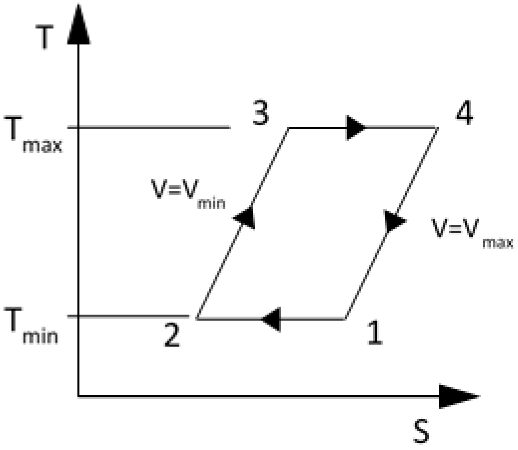

Figure 208: Temperature-entropy diagram for a Stirling Engine (T - Temperature, s - Entropy, V - Volume, 1-2-3-4 - Stages of the cycle)[376]

As Stirling engines require an external heat source, they are often designed, produced and installed together with the heat source. This often ensures, amongst other things, better heat integration and the combined production of heat and power (CHP). A schematic of a CHP system with a solid biomass boiler furnace is shown in Figure 209. Air pre-heater and economisers are included with the scope of increasing the system efficiency.

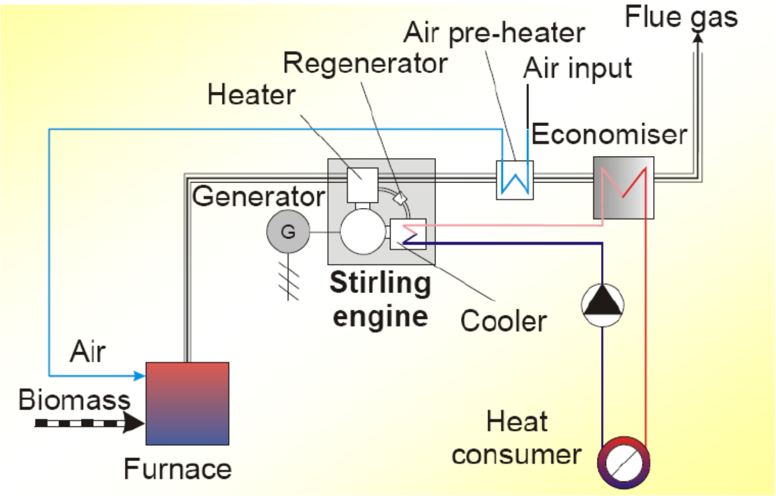

Figure 209: Schematic of a biomass CHP plant based on Stirling engine[377]

Development status and timescales

Use of a biomass powered Stirling engine was first recorded in 2007 when announced by the Precer Group. With companies such as Stirling DK Types, Biomass Authority and Bioenergiesysteme GmbH developing these engines, in varied forms, there is competition surrounding them. However, none of the companies have gone into full commercial manufacturing. It is clear from current developments, that the Stirling engine's potential for use in CHP processes is generally receiving far more attention than using them as a standalone generation system, which is understandable when Bioenergiesysteme's standalone engine only produces an electrical output at a range of 35 to 70 kW.

The majority of Stirling engines have been used in fossil fuel CHP, but Stirling engines currently under operation on biomass sources include:

- A direct biogas combustion system with 35 kW_e Stirling engine and 165 kW_{th} output is in operation since 2010, fuelled on landfill gas[378]
- An updraft gasification unit with 2 35 kW_e Stirling engines and 240 kW_{th} output is in operation since 2009, fuelled on woodchips[379]
- A system using pyrolysis oil as fuel in the combustor is in operation since 2008 in Denmark, with 35 kW_e and 110 kW_{th} of electricity and heat respectively. The system is also producing char for agricultural uses[380].

This current deployment places the technology at around **TRL 7**.

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[376] van Loo et al. (2008) "Handbook of biomass combustion and co-firing".
[377] Obernberger I., (2010) "Innovative biomass CHP technologies based on biomass combustion – ORC process and Stirling engines", Bios Bioenergiesysteme GmbH.
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[380] BlackCarbon (2011) Company website. Available at: http://www.blackcarbon.dk/

6.3.4.2 Organic Rankine Cycle

Technology description

Organic Rankine Cycle (ORC) systems generally utilise waste heat from various processes to generate electricity through a turbine system with an organic fluid as the working medium. The ORC system is very similar to the standard Rankine Cycle first originated by William Rankine in the 19th Century although it has been altered to handle much lower temperatures with fluids other than water.

Typical heat sources that this ORC would operate on would be: biomass, biogas, exhaust gas from gas turbines or engines, process steam and even waste heat from industrial processes[381].

ORC systems operate using an organic working fluid, specifically chosen to best suit the application. This fluid should be the most suitable fluid with critical temperatures suitable for the generation process. The working fluid initially is evaporated by the waste heat input until superheated and then processed through an integrated turbine to produce electricity at the generator. The working fluid is then passed through a recuperator, into a condenser and then is pumped up to evaporator pressure to start the cycle again[382].

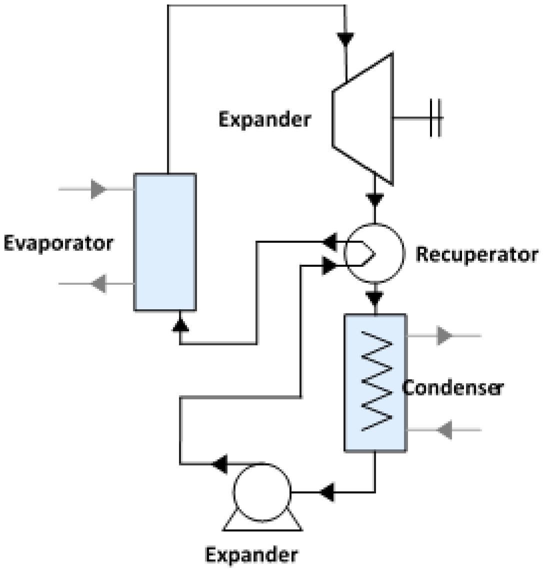

Figure 210: ORC System with recuperator[382]

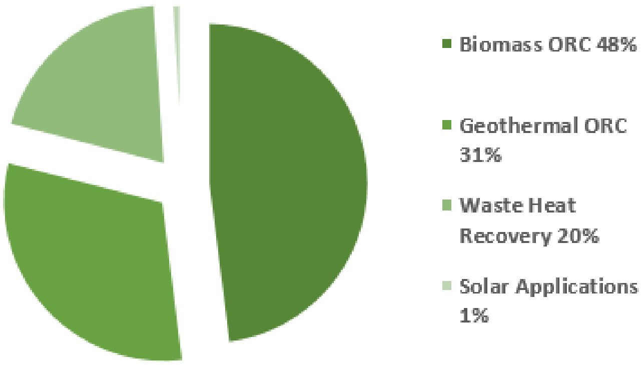

Figure 211: ORC heat sources[381]

The Siemens' ORC module works with a silicon fluid at a medium temperature of 300°C. There is a quoted output range of roughly 0.3 – 2 MW_e. Infinity Turbine's ORC system works with low temperatures from 80-140°C and alternating turbine diameters from 5 to 24 inches. There is a quoted output from as little as 2 kW_e all the way up to an output of 3 MW_e. The electrical efficiency of reference units is between 18-19.5%[383].

Development Status and Timescales

Since the demonstration unit by Turboden in 1980, there have been over 290 installations of ORC systems globally with the most suitable applications being shown in Figure 211 in relation to the total number of ORC systems in place. A substantial number of biomass-driven ORC systems have been deployed, which places the technology at a current **TRL of 8-9**.

Turboden has developed and produced over 220 biomass plants all over the world, with around 200 of them being in operation to date, with a particular focus on Italy, Germany and Austria. Their power ranges from 200 kW_e to 15 MW_e[384] and main feedstocks used are wood biomass (sawdust, chips, bark or treated wood), other biomass (straw, rice husks or bio-sludge) and high organic content waste. Siemens produces medium-size ORC turbines up to 2 MW_e[385]. Another player is ElectraTherm, who produce small ORC systems in the range 35 kW_e-110 kW_e[386]. As an example, ElectraTherm installed a 65 kW_e unit a Mill in Oregon where wood waste is used as feedstock[387].

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[381] Rowshanzadeh, R. Performance and cost evaluation of Organic Rankine Cycle at different technologies. Available at: http://kth.diva-portal.org/smash/get/diva2:410363/FULLTEXT01

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[386] ElectraTherm Products. Company Website. Available at: https://electratherm.com/products/

[387] ElectraTherm "ElectraTherm Power+ Generators generate electricity from wood waste at mill in Oregon". Available at: https://electratherm.com/case-studies/biomass-in-oregon/

6.4 Summary of TRLs

Table 35, below, gives a summary of the TRLs for each of the technologies covered in this horizon scanning section, and an estimate of the time that each might take to reach TRL9. The methodology for determining the number of years that it may take to reach commercial scale and mass deployment at TRL 9 (i.e. repeated plant designs) is based on assessing the realistic chances of each technology successfully gaining TRL levels based on current activity, and following roll-out and construction times. This approach does not assume a fixed TRL progression over time. The following sets of qualitative indicators were reviewed in order to gauge interest in each horizon scan technology, based on the information given in the sections above:

- Assessing the number of developers, industrial partners and academics actively working on the technology (to the best of the available data for each technology). Having at least two developers, with involvement of at least one strong industrial partner is much more likely to lead to demonstration and commercial uptake, i.e. those technologies without this are likely to take longer to develop, if they develop at all
- Assessing if there are multiple planned projects announced publically for the technology, or if there are no plans. Having projects already in the pipeline can save several years
- Assessing which end use applications are attracting most attention, and whether these are in scope. Aqueous Phase Reforming is likely to be ruled out for this reason, due to only being developed for biofuels/biochemicals. If the end use applications are small (e.g. alkaline fuel cells) these plants can be rolled out faster than multi-year large construction projects
- Reviewing the focus on biomass, wastes or wet feedstocks. The use of very wet feedstocks (such as slurry) is not relevant to TEABPP, nor is the use of MSW and RDF – hence the focus on activities using woody biomass and energy crops
- Reviewing if any other similar but more developed technologies or components could offer a “piggy-back” or faster track to commercialisation. This could vary depending if the whole plant is the same (with only one component improved), to the reactor design already being commercially available

The combination of these key factors is summarised in Table 35, giving the explanation for the likely commercialisation timescales. As previously, the symbol “~” indicates a more approximate estimate has been made, due to a greater than average level of uncertainty regarding the specific technology’s TRL progression, typically due to a lack of developers currently working on the technology.

Table 35: TRL progression assumptions for the Horizon scanning technologies

Horizon Scan technologies	Current TRL	Years to TRL 9	Rationale
Lower TRL			
Microwave Plasma Gasification	5	15	Two developers. Could piggy-back off standard plasma gasification once microwave resonator proven
Super-critical Water Gasification	4	~ 25+	No developers. Several academic groups still interested. Wet feedstock
CO_2 Gasification	4	~ 25+	No developers. But recent academic research, at only one group. Focus may be on char gasification, not biomass
Chemical Looping Combustion	4	25	No developers, but several academic groups, and uses standard CFB equipment. Depends on coal CCS development and competing tech
Chemical Looping Gasification	3	~ 35	No developers and little academic activity
Entropic Cycle	4-5	~ 20+	No developers and little sign of academic activity. May not be developed
Direct Syngas SOFCs	5	15-20	One developer, few academics, but US DOE and industry is driving a program of work in coal
Bio-H_2 Alkaline Fuel Cells	4-5	20	One main developer looking at wastes/biomass angle. WGS presents additional challenge. But small modules lend themselves to roll-out
Fuel cell Gas Turbine Hybrid	5	15-20	One developer, few academics, but US DOE and industry is driving a program of work in coal
Aqueous Phase Reforming	5	Not applicable	One developer, major industry presence. Liquid sugars at present, unlikely to apply to heat and power. For biofuels/chemicals only
Hydro-Thermal Upgrading	4	~ 30+	No developer, industry pulled out long time ago. Wet biomass. May not be developed, too similar to pyrolysis
Microwave Pyrolysis	4	25	Multiple developers for wastes, but only academics for biomass. Not likely to piggy-back, but substitute conventional fast pyrolysis
Hydrothermal Carbonisation	5	20-25	Two developers, a few academics, but little info. Wet feedstock
Higher TRL			
Plasma Clean-up	8	<5	Three developers. Extensive deployment of clean-up torches already, although very waste focused
Solvent Scrubbing	7	5-10	One academic developer, but strong industrial partner with demos, and multiple plans at scale
Catalytic Tar Removal	7	5-10	Two developers and strong industrial partner. Very waste focused. One planned plant
Pyrolysis (gas and biochar)	8-9	5+	Significant history. Industry currently not engaged in developing new projects, though there is interest related to biochar soil developers
Externally Fired Gas Turbine	6-7	10	Two developers. Standard components being used and variety of relevant feedstocks
Stirling Engine	7	5-10	Multiple developers and examples, including some on biomass. CHP gaining attention
Organic Rankine Cycle	8-9	<5	Multiple developers and examples, including on biomass

7 Benchmarking

7.1 Introduction

Now that each of the combustion, gasification and pre-treatment technologies within the TEABPP project scope have been introduced, this section of the report focuses on comparing these technologies based on a set of key metrics and features:

- Levelised cost of energy output, utilising the Base Case techno-economic data gathered in Excel for each technology during WP1
- Current TRL and potential time to reach TRL 9
- Range of commercial plant scales
- Applicable end use applications (syngas, heat and/or power) for the conversion processes
- A summary of major strengths for the pre-treatment processes

This analysis will enable the ETI to understand which technologies are nearest/furthest to being commercialised, and which are cheapest on a stand-alone basis.

7.2 Levelised cost of energy

The Levelised Cost of Energy (LCOE) is a measure of the total costs (in GBP 2015) that have been incurred to produce one MWh of the “main output” from each conversion or pre-processing plant over its lifetime. In the case of the conversion technologies, this “main output” is either electricity (for the power and CHP applications), hot water (for the heating applications), or cleaned syngas (for the syngas applications). For the pre-processing technologies, the “main output” is the new form of the processed biomass (e.g. washed chips, dried chips, pellets, pyrolysis oil etc).

These LCOE calculations include:

- Levelised capital costs, taking the total investment cost and applying a common discount rate of 10% across the plant’s technical lifetime. Common assumptions were also utilised when deriving the total investment cost from the total installed equipment cost, including “operations and utilities”, “civils and land”, “project development” and “contingency”
- Fixed operating costs. A common insurance assumption of 1%/year of the total investment costs was assumed, alongside technology-specific data on maintenance costs (in many cases, a fraction of the total investment costs)
- Variable operating costs. These include operations staff, costs of imported energy (such as natural gas, diesel and electricity), chemicals and water inputs, and the costs of disposing of waste water, ash etc. Please see Section 2.6 for a more detailed breakdown.
- Feedstock costs, taking a common biomass price of £14.8/MWh, based on a recent NW European spot price for industrial wood chips[388], and dividing by the plant LHV efficiency. Note that this is a basic price for long rotation forestry, with little transport and processing included, and is only used in D1 to act as a fair basis for all the pre-processing and conversion technologies. The feedstock prices used at the start of the chains in D2 and D4 will be higher, because Miscanthus, SRC willow and SRF grown in the UK are considerably more expensive than LRF chips, and then the chains add transport, pre-processing and storage costs before arriving at an even higher delivered feedstock price (which could also

be a densified form, such as pellets). It is therefore worth bearing in mind that lower efficiency technologies will be penalised more once the whole-chain delivered feedstock prices in D2, starting with energy crops and SRF, are implemented.

- Co-product credits (if any). For the CHP technologies, a common heat price of £32.3/MWh was taken, based on recent large industrial natural gas consumer prices in the UK and an assumed counterfactual 85% boiler efficiency – i.e. what industrial users would currently be willing to pay to for a supply of hot water[389]

Note that all these costs are given in today's GBP 2015, so there is no cash flow modelling out to the 2030s relying on assumptions regarding inflation. All costs and efficiencies are based on current best in class (or in some cases, the only available) plant – but there is no assumption regarding efficiency or cost improvements over time, or future projections. All the total investment costs are for new build, dedicated biomass conversion and pre-treatment plants – retrofit or conversion modelling is out of scope, as is carbon capture and storage.

It is not possible to scale all the technologies to the same MW output scale in order to conduct the WP1 benchmarking analysis at a homogenised scale, as this will lie outside of the allowable scale ranges for at least a few of the smaller-scale or larger-scale technologies. For example, downdraft gasifiers are only applicable up to around 2 MW_{th} input, whereas entrained flow gasifiers start at 100 MW_{th} upwards. The benchmarking analysis is therefore conducted for each technology at the scale given in the Base Case in the Excel sheets, i.e. within the applicable range of commercial scales given in the earlier report sections (exact numbers given in tables later in this section). This WP1 cost analysis is therefore representative of the costs of the technologies as at the scales they would be installed, and not homogenised to an infeasible or unrepresentative scale. The full modelling exercise in WP3 will be able to use different allowable scales from the Excel input files, including data uncertainties and the sensitivities already presented in earlier sections.

Lower Heating Value (LHV) is the basis used for renewable energy deployment and GHG emissions accounting within Europe. Please remember that feedstock moisture content has a very strong impact on the amount of energy that the feedstock contains (on an LHV basis). Higher moisture content feedstocks require additional energy to evaporate the water, hence have lower LHV values than just scaling with the moisture content. Because of this, if imported energy (e.g. natural gas) is expended in drying a feedstock, then the resulting LHV efficiency of a dryer (measured as dry biomass GJ output/wet biomass GJ input) can easily exceed 100%. In other words, LHVs are always applied to the biomass at the given moisture content, and never to the bone dry biomass tonnages.

Most technologies are comprised of combinations of multiple components, in order to present techno-economic data for a standalone plant. Many different technologies use the same component, e.g. a drying step is included within briquetting, pelleting, pyrolysis, torrefaction, steam explosion + pelleting, AFEX + pelleting, as well as fluidised bed and entrained flow gasification. The benchmarking results should therefore not be directly used to add e.g. water washing to pelleting, as this will double-count the cost of several components. For information on the steps included in each of the technologies shown in the following graphs (and gathered in WP1), see Table 36.

This table also lists the base case feedstock, upon which the benchmarking analysis is based. For the pre-processing technologies, this is typically the least processed form of the feedstock, as the form data from the ETI "Characterisation of Feedstocks" project is not yet known. However, LCOE values would change if lower moisture or chipped material is used instead, for example, due to less natural gas or avoiding capex for a chipper. These relationships between feedstock form, physical & chemical characteristics and the plant costs & efficiency will be included within the D4 process modelling – this D1 report benchmarking only permits an LCOE snapshot for one configuration.

Table 36: Component steps in each of the technologies presented, and the Base case feedstock

Technology	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**Base case feedstock**
Pre-processing technologies								
Water washing	Stone removal	Chipping	Magnetic screening	Water washing	Filtering			SRC willow billets
Chemical washing	Stone removal	Chipping	Magnetic screening	Water washing	Chemical washing	Filtering		SRC willow billets
Drum drying	Chipping	Screening	Drying					LRF logs
Belt drying	Chipping	Screening	Drying					LRF logs
Briquetting	Chipping	Screening	Drying	Press	Cooling			LRF logs
Screening	Chipping	Screening						LRF logs
Chipping	Chipping							LRF logs
Pelleting	Chipping	Screening	Drying	Grinding	Conditioning	Pelleting	Cooling	LRF logs
Pyrolysis	Drying	Grinding	Fast pyrolysis	Char separation	Liquid recovery			LRF logs
Torrefaction	Chipping	Screening	Drying	Torrefaction	Cooling			LRF logs
Torrefaction + pelleting	Chipping	Screening	Drying	Torrefaction	Pelleting	Cooling		LRF logs
Torrefaction + briquetting	Chipping	Screening	Drying	Torrefaction	Press	Cooling		LRF logs
Steam explosion + pelleting	Screening	Chipping	Drying	Steam Explosion	Pelleting	Cooling		LRF logs
AFEX + pelleting	Screening	Chipping	Drying	AFEX	Pelleting	Cooling		Corn stover bales
Conversion technologies – combustion								
Underfed stoker	Boiler							LRF pellets
Moving bed	Boiler							LRF chips
BFB	Boiler							LRF chips
CFB	Boiler							LRF chips
Dust suspension	Grinding	Boiler						LRF pellets
Conversion technologies – gasification								
Downdraft	Gasifier	Syngas clean-up						LRF chips
Updraft	Gasifier	Syngas clean-up						LRF chips
BFB	Drying	Gasifier	Syngas clean-up					LRF chips
CFB	Drying	Gasifier	Syngas clean-up					LRF chips
Dual fluidised bed	Drying	Gasifier	Syngas clean-up					LRF chips
Entrained flow	Drying	Grinding	Gasifier	Syngas clean-up				LRF pellets

7.2.1 Pre-treatment technologies

The LCOE results for all the pre-treatment technologies are shown in Figure 212, and the Base Case techno-economic data behind these figures are given in Table 37.

Those pre-processing technologies that add the fewest costs to the underlying cost of the biomass are the simplest (chipping and screening); whereas those with the highest LCOE either have relatively low efficiency (e.g. pyrolysis), high capital costs or have a relatively low availability for a new technology based on seasonal feedstocks (e.g. AFEX + pelleting). Chemical washing is more expensive than water washing, mainly due to the additional costs of the chemicals and additional waste water treatment. Drum and belt drying have very similar costs and efficiencies. Briquetting has lower electricity input than pelleting, but higher labour and maintenance costs. Torrefaction to chips is relatively low cost, as the process is energy self-sufficient, but adding pelleting or briquetting increases the capital costs and energy requirements in the plant. Steam explosion pellets currently have significant variable costs from imported natural gas, and even the slightly larger scale (and hence cheaper capex) of the base case steam explosion plant is not enough to offset this.

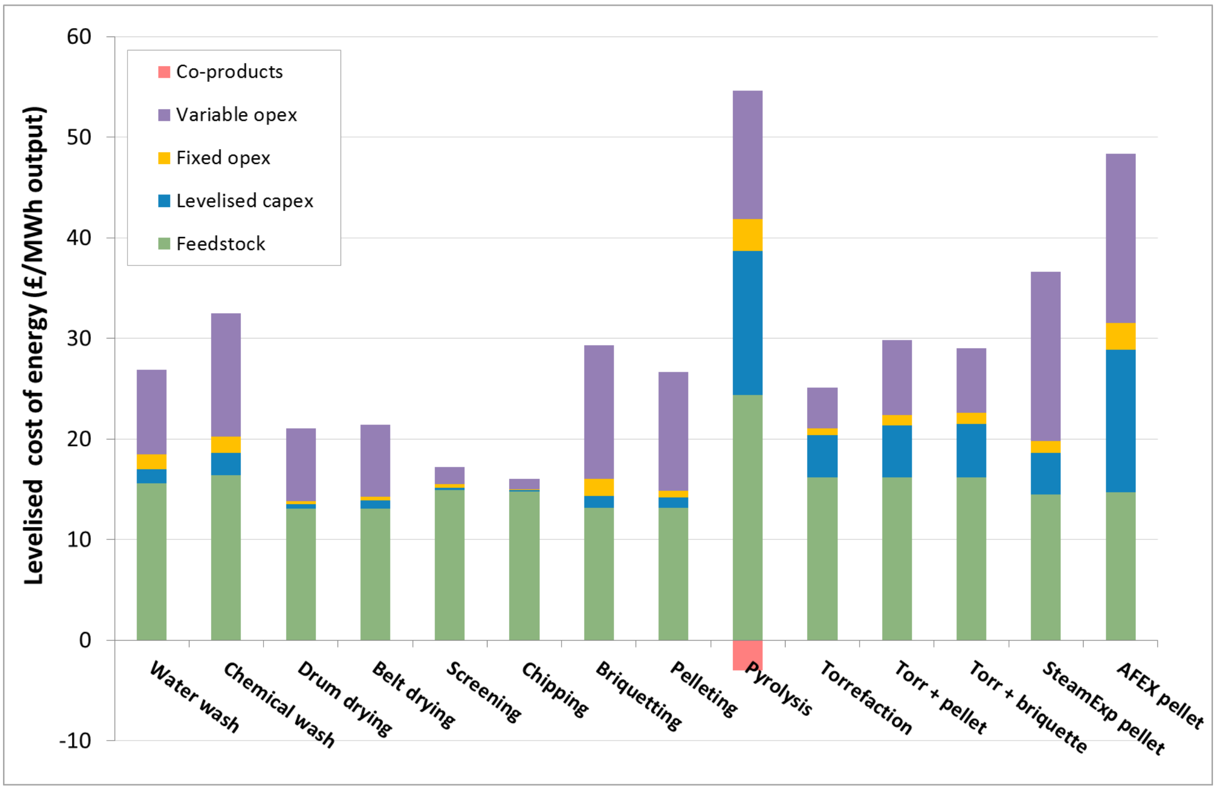

Figure 212: LCOE benchmarking for the TEABPP pre-treatment technologies

Table 37: Base Case data for each pre-treatment technology

Pre-treatment technology	Scale (MW output)	Efficiency (MW out/in)	Total capex (£)	Total fixed opex (£/yr)	Total variable opex (£/yr)	Co-products (£/yr)	Avail-ability (hr/yr)	Lifetime (yr)
Water washing	21.6	0.950	1,910,000	221,000	1,290,000	0	7,085	20
Chemical washing	20.5	0.900	2,670,000	236,000	1,780,000	0	7,085	20
Drum drying	38.7	1.132	1,270,000	93,000	2,260,000	0	8,000	20
Belt drying	38.7	1.132	2,200,000	108,000	2,220,000	0	8,000	20
Screening	33.8	0.990	344,000	100,000	458,000	0	8,000	8
Chipping	68.3	1.000	93,000	15,000	174,000	0	2,500	6
Briquetting	38.3	1.121	3,080,000	472,000	3,830,000	0	7,500	25
Pelleting	40.9	1.121	2,860,000	206,000	3,850,000	0	8,000	20
Pyrolysis	27.6	0.606	26,400,000	703,000	2,770,000	-657,000	7,884	20
Torrefaction	60.6	0.914	17,300,000	300,000	1,870,000	0	7,500	25
Torr + pellet	60.6	0.914	21,200,000	473,000	3,400,000	0	7,500	25
Torr + briquette	60.6	0.914	22,00,000	497,000	2,910,000	0	7,500	25
SteamExp pellet	87.1	1.020	26,300,000	788,000	11,700,000	0	8,000	25
AFEX pellet	13.0	1.008	7,550,000	157,000	982,000	0	4,500	25

7.2.2 Hot water conversion technologies

The LCOE results for all the conversion technologies producing hot water are shown in Figure 213, and the Base Case techno-economic data behind these figures are given in Table 38. No syngas clean-up costs or efficiency losses are assumed, given the direct use of raw syngas in gas boilers.

The difference in scales and operating hours explains why the smaller underfeed stoker (0.4 MW_{th}) levelised Capex is high compared to moving bed boilers (1 MW_{th}). Both the downdraft and updraft gasifiers have significant maintenance cost assumptions, with downdraft costs prominent due to the smaller scale compared to updraft gasifiers (1 vs. 5 MW_{th}). The fluidised beds are all at larger scales, and have very similar costs. The combustion options assume a high-level of automation, hence low variable opex with shift labour not required.

The most efficient heat conversion technologies are moving bed combustion boilers, followed by underfeed stoker combustion boilers, with the range of gasifiers + boilers managing to achieve current heating efficiencies of 60-70% LHV once the efficiencies of gasification are combined with the efficiencies of a syngas boiler.

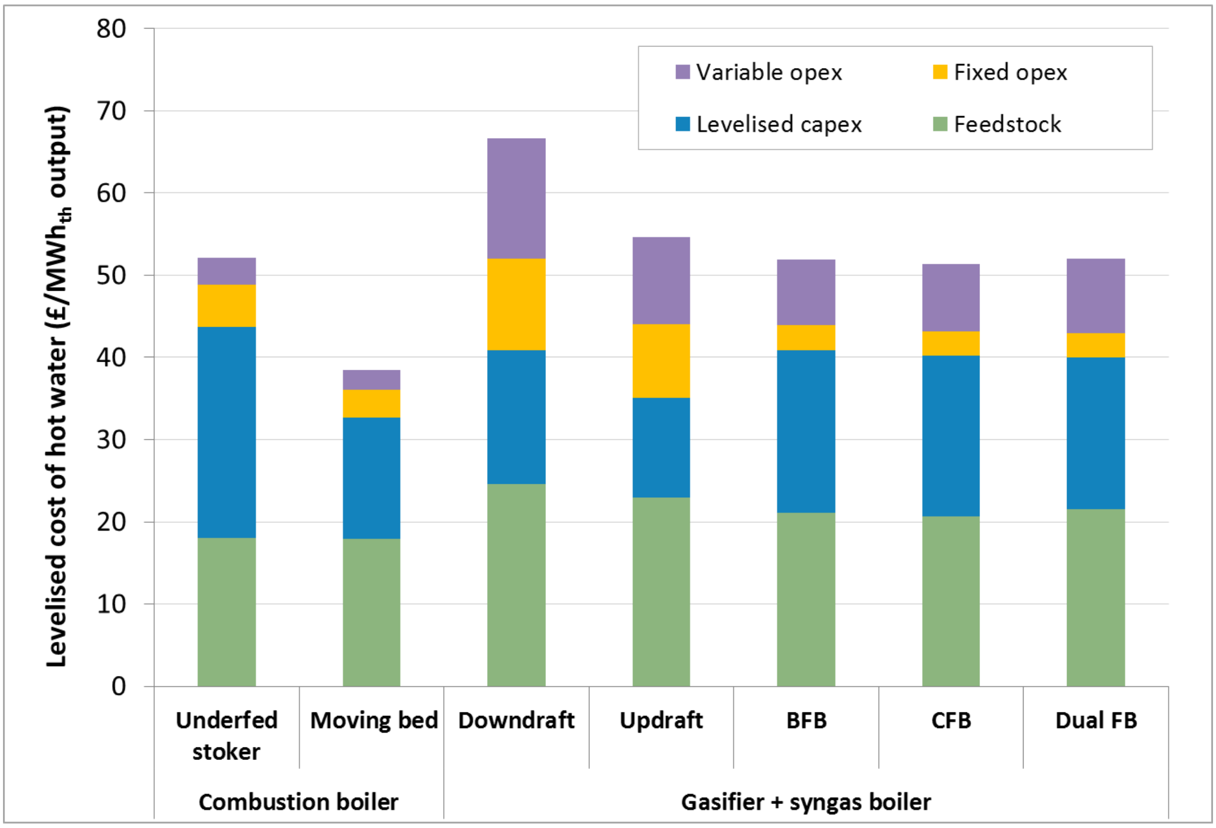

Figure 213: LCOE benchmarking for the TEABPP heating conversion technologies

Table 38: Base Case data for each hot water conversion technology

	Heating technology	Scale (MW output)	Efficiency (MW out/in)	Total capex (£)	Total fixed opex (£/yr)	Total variable opex (£/yr)	Availability (hr/yr)	Lifetime (yr)
Combustion	Underfed stoker	0.4	0.817	372,000	9,800	6,300	4,380	15
	Moving bed	1.0	0.823	679,000	18,000	13,000	5,256	20
Gasifier + syngas boiler	Downdraft	1.0	0.601	1,170,000	87,000	116,000	7,884	25
	Updraft	5.0	0.645	4,340,000	355,000	417,000	7,884	25
	BFB	10.0	0.700	13,400,000	229,000	597,000	7,500	25
	CFB	20.0	0.713	28,000,000	461,000	1,290,000	7,884	25
	Dual FB	20.0	0.688	25,300,000	437,000	1,360,000	7,500	25

7.2.3 CHP conversion technologies

The LCOE results for all the conversion technologies producing combined heat and power are shown in Figure 214, and the Base Case techno-economic data behind these figures are given in Table 39. Note that these costs are given as £ per MWh electricity, so that the heat produced is assigned a co-product credit, reducing the total net LCOE.

The moving bed combustion CHP system is the most expensive due to its low electrical efficiency. The BFB combustion + steam CHP technology is by far the largest plant (50 MW$_e$ output), and hence the cheapest overall LCOE – with all the syngas CHP engines only at 0.5 – 5 MW$_e$, which explains most of the differences seen in the capital costs. Although the gasification + syngas CHP routes have higher feedstock to electricity efficiencies (at 25% to 30% LHV) compared to the combustion options (~20%), the gasifier Opex is also significantly higher, in part due to the syngas clean-up required, plus the smaller scale of the downdraft and updraft gasifiers meaning that fixed labour costs are significant. Dual FB systems have very similar costs to the other BFB or CFB gasifier systems.

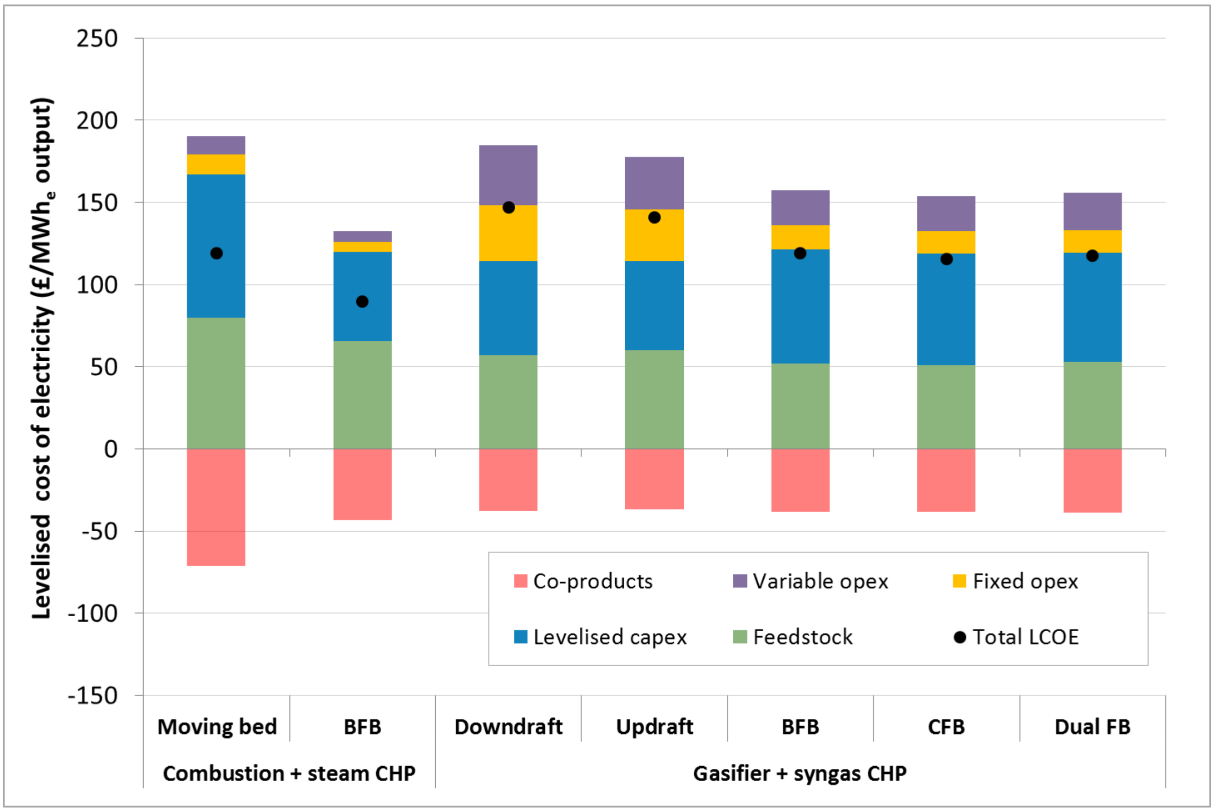

Figure 214: LCOE benchmarking for the TEABPP CHP conversion technologies (dots show net LCOE)

Table 39: Base Case data for each CHP conversion technology

		Scale (MW$_e$ output)	Scale (MW$_{th}$ output)	Efficiency (MW$_e$ out/in)	Total capex (£)	Total fixed opex (£/yr)	Total variable opex (£/yr)	Co-products (£/yr)	Avail-ability (hr/yr)	Life-time (yr)
Combustion + steam CHP	Moving bed	15.9	35.2	0.185	88,200,000	1,430,000	1,330,000	-8,470,000	7,446	20
	BFB	50.0	66.9	0.224	172,000,000	2,180,000	2,510,000	-16,100,000	7,446	20
Gasifier + syngas CHP	Down-draft	0.5	0.5	0.258	1,920,000	124,000	134,000	-139,000	7,884	25
	Updraft	1.9	2.2	0.246	7,470,000	476,000	489,000	-557,000	7,884	25
	BFB	4.6	5.5	0.284	22,000,000	505,000	737,000	-1,320,000	7,500	25
	CFB	9.3	11.0	0.290	45,000,000	993,000	1,570,000	-2,790,000	7,884	25
	Dual FB	9.1	11.0	0.279	41,300,000	956,000	1,570,000	-2,650,000	7,500	25

7.2.4 Power conversion technologies

The LCOE results for all the conversion technologies producing power are shown in Figure 215, and the Base Case techno-economic data behind these figures are given in Table 40.

The smallest (0.5 – 10MW$_e$) and most expensive power generation technology routes are the gasifier + syngas engine cases. Even with the higher electrical efficiency of gas engines compared to steam turbines at small scales, their higher labour costs and levelised capital costs mean that they remain more expensive than combustion + steam turbine options.

The combustion boiler + steam turbine systems have relatively low operating costs compared to gasification systems. The total LCOE costs for moving bed combustion + steam turbine (at 20MW$_e$) are comparable to those of the CFB and Dual FB gasifier + CCGT technologies (at 10MW$_e$). Although the gasification routes have strict/costly syngas clean-up steps with losses in order to meet turbine specifications, the overall electrical efficiency of a BIGCC plant manages to reach 35-40%, compared to ~27% for BFB and CFB combustion + steam turbine.

Syngas clean-up costs are lowest for entrained flow gasifiers, and with the larger 250 MW$_e$ output, the EF gasifier + CCGT plant is able to achieve 40% efficiency and a low levelised Capex. There are similar LCOE values achieved for the largest plant on the chart, dust combustion + steam turbine at 322 MW$_e$ output. This pulverised biomass combustion plant is assuming an ultra-super critical boiler, which with the power requirements for biomass grinding, allows an overall net LHV efficiency of 37%. This explains the low feedstock cost component, along with the large scale leading to the low Capex.

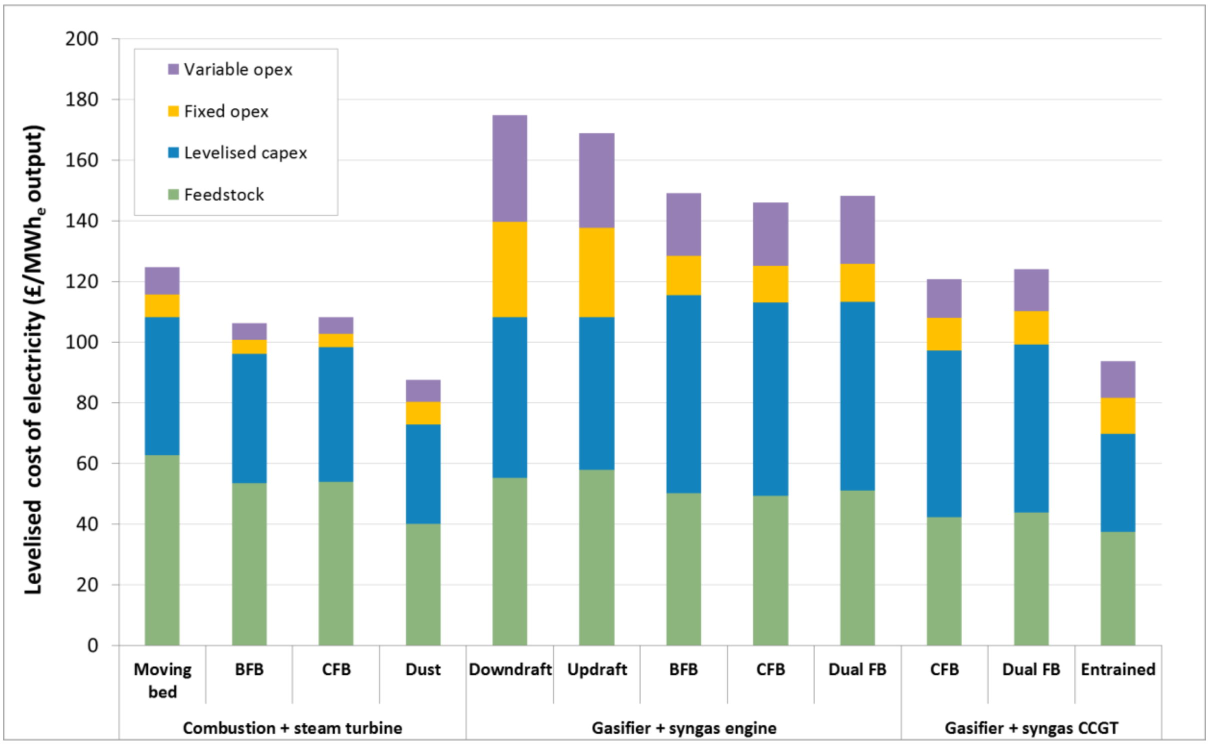

Figure 215: LCOE benchmarking for the TEABPP power conversion technologies

Table 40: Base Case data for each power conversion technology

	Technology	Scale (MW_e output)	Efficiency (MW_e out/in)	Total capex (£)	Total fixed opex (£/yr)	Total variable opex (£/yr)	Avail-ability (hr/yr)	Life-time (yr)
Combustion + steam turbine	Moving bed	20.0	0.236	57,900,000	1,110,000	1,310,000	7,446	20
	BFB	100	0.276	270,000,000	3,310,000	4,120,000	7,446	20
	CFB	100	0.274	280,000,000	3,380,000	4,120,000	7,446	20
	Dust	322	0.369	586,000,000	13,600,000	13,300,000	5,698	40
Gasifier + engine	Downdraft	0.5	0.268	1,780,000	116,000	130,000	7,884	25
	Updraft	1.9	0.255	6,950,000	448,000	475,000	7,884	25
	BFB	4.7	0.295	20,700,000	456,000	719,000	7,500	25
	CFB	9.3	0.300	42,400,000	900,000	1,530,000	7,884	25
	Dual FB	9.2	0.289	38,800,000	863,000	1,530,000	7,500	25
Gasifier + CCGT	CFB	28.2	0.350	113,000,000	2,410,000	2,890,000	8,000	25
	Dual FB	27.9	0.337	105,000,000	2,320,000	2,890,000	7,500	25
	Entrained	251	0.394	580,000,000	23,700,000	23,700,000	7,884	25

7.2.5 Syngas production technologies

The LCOE results for all the gasification technologies producing syngas are shown in Figure 216, and the Base Case techno-economic data behind these figures are given in Table 41.

Note that these output syngas costs are for a clean, polished syngas suitable for use in a gas turbine (or potentially pipeline, although syngas pipeline specifications are yet to be established). These LCOE values are therefore in some cases higher than the £/MWh_th from the heating application charts, as the clean-up steps add Capex and Opex, plus a requirement for power imports to meet parasitic loads (as there is no downstream power generator), and results in an efficiency loss of 5% up to 20% depending on the gasifier type and operational conditions.

Although downdraft and updraft gasifiers are relatively simple, their smaller scale and syngas clean-up costs lead to their levelised Capex being on a par with (or slightly cheaper than) larger fluidised systems. However, the operating and labour costs of these fixed bed systems add significantly to their overall LCOE, leading them to be the most expensive systems for producing clean syngas. The step up in scale from BFB to CFB also explains the drop in Capex, with Dual FB at the same scale as CFB, but with slightly higher costs due to higher power requirements.

Entrained flow gasifiers are able to achieve very large scales and hence the lowest levelised Capex, although still require power import for ASU and feedstock grinding – however, they also have the lowest of the syngas clean-up costs of the gasifiers displayed. Overall system efficiency also improves with scale, with the feedstock cost component decreasing.

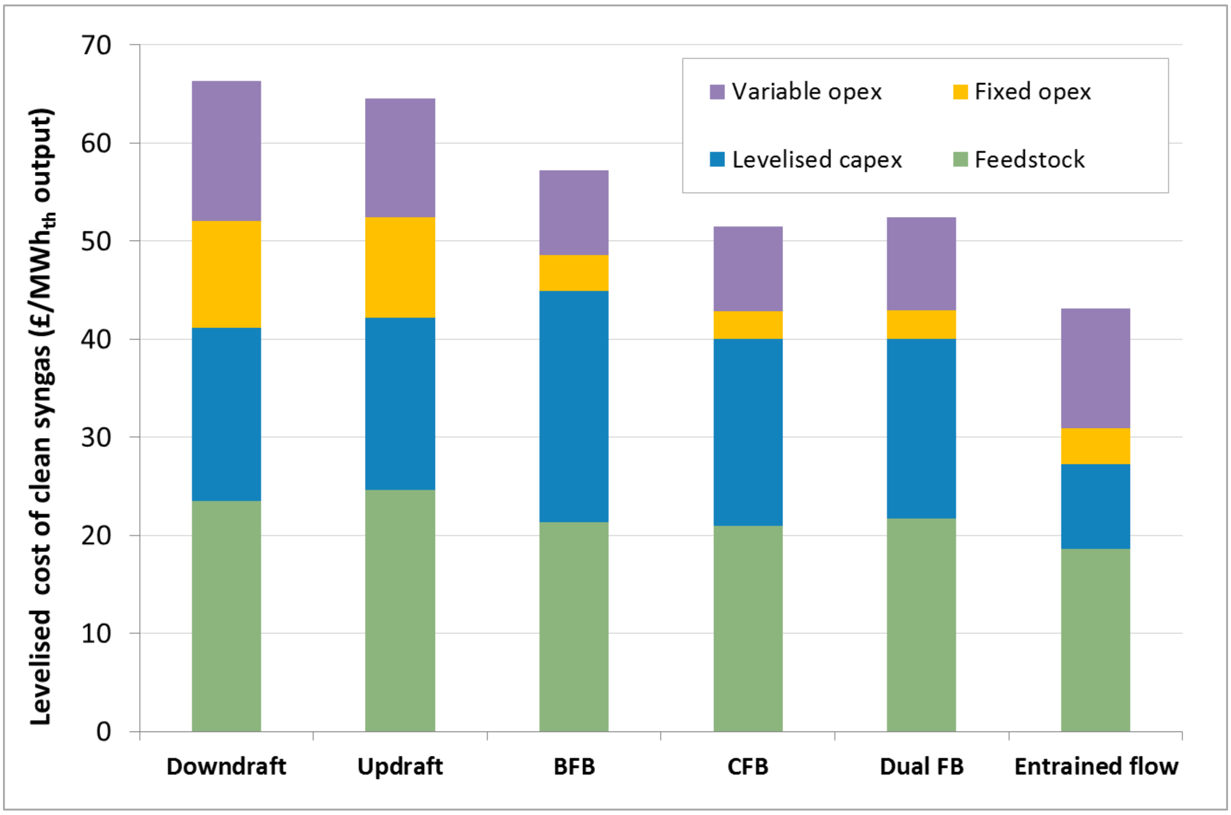

Figure 216: LCOE benchmarking for the TEABPP gasification technologies producing (clean) syngas

Table 41: Base Case data for each gasification technology generating (clean) syngas

	Scale (MW output)	Efficiency (MW out/in)	Total capex (£)	Total fixed opex (£/yr)	Total variable opex (£/yr)	Availability (hr/yr)	Lifetime (yr)
Downdraft	1.0	0.63	1,260,000	85,700	113,000	7,884	25
Updraft	5.0	0.60	6,280,000	406,000	478,000	7,884	25
BFB	10	0.69	16,100,000	274,000	646,000	7,500	25
CFB	50	0.71	69,100,000	1,150,000	3,440,000	8,000	25
Dual FB	50	0.68	62,300,000	1,080,000	3,580,000	7,500	25
Entrained	300	0.79	185,000,000	8,750,000	28,900,000	7,884	25

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[388] Argus (2015) "Biomass prices", 90 day cif spot price of US$ 5.78/GJ

[389] DECC (2015) "2015 DECC non-domestic energy price statistics". Natural gas price, for Large customers Q1 2015. Available at: https://www.gov.uk/government/statistical-data-sets/gas-and-electricity-prices-in-the-non-domestic-sector

7.3 Comparison of key technology features

Table 42 sets out for each technology the current TRL, the expected number of years to reach mass commercial deployment (TRL 9), and the current range of expected plant scales that would be commercially viable/technical feasible, were the technology commercialised. For each conversion technology, a summary is given of the end use applications that have been analysed in WP1. For the pre-processing technologies, a very brief summary is given of some of the key benefits of the intermediate product. The data to fill this table come from the earlier report sections (on current TRL and progression), with the commercial scales being taken from the Excel datasheets for each technology.

Table 42: Conversion and pre-processing technology TRL development, scales (in MW input) and applications

Technology	Current TRL	Years to TRL 9	Commercial MW_{in} scale	Applications
Conversion technologies – combustion				
Underfeed stoker	9	-	0.2-2.9	Heat
Moving bed	9	-	0.5-190	Heat, CHP, power
Bubbling Fluidised bed	9	-	28-417	CHP, power
Circulating Fluidised Bed	9	-	193-1,300	Power
Dust Suspension	8-9	<5	602-1,740	Power
Conversion technologies – gasification				
Updraft	9	-	1-15	Syngas
Downdraft	9	-	0.05-2	Syngas
Bubbling Fluidised Bed	7	10-20	2-20	Syngas
Circulating Fluidised Bed	7	10-20	15-100	Syngas
Dual Fluidised Bed	6-7	15-20	15-100	Syngas
Entrained Flow	6	15-20	100-2,000	Syngas
Syngas Boiler	9	-	0.05-52	Heat
Syngas Engine/CHP	8	<5	0.4-50	Power, CHP
Syngas CCGT	8	5-10	50-800	Power
Pre-processing technologies				
Water Washing	7	10	1-182	↓ alkali, ash
Chemical Washing	4	? 20+	1-182	↓ alkali, ash
Belt Drying	9	-	3-114	↓ moisture
Drum Drying	9	-	3-68	↓ moisture
Briquetting	9	-	1-34	↑ density
Screening	9	-	34-159	↓ rejects
Chipping	9	-	1-150	↓ size
Pelleting	9	-	2-387	↑ density
Pyrolysis	8	5	4-205	→ liquid
Torrefaction	8	5	33-106	↑ LHV
Torrefaction + pelleting	8	5	33-106	↑ density, LHV
Torrefaction + briquetting	5	10	33-106	↑ density, LHV
Steam Explosion + pelleting	8	5	16-301	↑ density, LHV
AFEX + pelleting	5	10-15	4-19	↑ density, LHV

7.4 Data quality and next steps

One of the objectives of WP1 is to collect enough techno-economic data for each pre-processing and conversion technology, of sufficient quality and in a consistent format, in order to input this data into the analysis and down-selection process in WP2 and populate the gPROMS databases in WP3.

Below is a status update on this data gathering, where the main weaknesses in the data and assumptions currently lie, and the next steps to be carried out.

Excel datasheets

Each of the conversion and pre-processing technologies benchmarked in Section 7 has its own underlying Excel datasheet, containing:

- Base Case data for a given plant scale, including input and output flows of each material crossing the plant boundary, the impact of the technology on the biomass parameters, any feedstock limits, and the techno-economics data for the technology. There are complete datasets based on supplier quotes for the most commercial technologies, although at lower TRL, the values are necessarily based on academic literature or estimated proxies from similar equipment. One area that remains more patchy is regarding the less important feedstock limits (i.e. those outside of moisture, size, density, ash content), which are not used in the sensitivities, but could potentially prevent a proportion of a feedstock from being allowable into the plant. These less important rows will be checked and filled if necessary.
- Uncertainty ranges with Min/Max values around the Base Case data, which will be translated along the sensitivity curves when e.g. the plant is resized. These uncertainties reflect the level of comfort with the data provided, as well as the inherent uncertainties from the TRL. In a few cases, there are enough underlying data points to calculate a spread of values, whereas in most cases, +/- % limits have been estimated for each row based on any available evidence from the Base Case data source.
- Sensitivity formulae linking the plant Capex, Opex and efficiency to the scale and the biomass input parameters. In many cases, in order to plot the charts given in Sections 2-4, these sensitivity relationships are provided in a different tab, but linked to the rest of the Base Case data. For each metric, information is given regarding the parameters that influence the metric, and the type of relationship, with the formula either written out or plotted as a chart trend-line. One area that is currently weak in the Excel is how changes in minor biomass characteristics impact availability and efficiency – this process modelling is easier to do in gPROMS than in Excel, given the interaction between 15 or so parameters.
- References and notes, showing sources for each data point (or otherwise not giving the explicit source or project, if it is commercially sensitive).

By completing the benchmarking analysis in WP1, it is known that there is sufficient data to conduct WP2, as D2 will be combining the costs of the different steps (already gathered) together to make full supply chains on demand by the user.

Harmonisation and final improvements

Inputs relating to project development, contingency, insurance, electricity prices, natural gas, ash disposal costs were established and set across the technologies. Similarly, any conversions from older or non-GBP data used a standardised inflate-then-exchange methodology.

All data will be checked in a final review exercise to address cost items that were not able to be harmonised during the benchmarking analysis. This includes looking at the assumptions for the number of staff shifts per plant, and making sure the resulting labour costs are appropriate for the size of the plant modelled. Linked to this is the assumption on operating hours per year; if these are to be changed in the Base Case (say upwards to match other similar technologies), then the input/output flows will be adjusted as well.

The Excel worksheets produced during WP1 will also be standardised, so that every row and column is an identical position between the Excel sheets, so that an automated script can be run to pull the data into gPROMS with the minimum of effort. The sensitivity formulae will need manually input into the WP3 modelling, but this will be carried out once the results of the down-selection in WP2 are clear.

Feedstocks data

Following on from previous meetings, the project team will re-engage with the “Characterisation of Feedstocks” project, in order to formalise the format of the input sheet that will be delivered with the biomass physical and chemical composition data. From the technology data gathered, there is now a clear picture as to which parameters are required (and their units), and which parameters are not currently used. The project team is not expecting a large departure from previous scoping discussions with Forest Research and E.ON, rather a reconfirmation of parameter sets.

7.5 Glossary

AFEX	Ammonia Fibre Expansion
As-received biomass	A reference “as received” feedstock, against which each technology evaluation and quantification is performed. This is not the same as a “green” (i.e. immediately harvested) feedstock, since most feedstocks will be subject to a degree of seasoning before arriving at a plant. Data for each feedstock are given in Section 2.4.
BFB	Bubbling fluidised bed
Capex	Capital costs, generally always referring to the Total Investment Costs unless specifically stated otherwise
CCGT	Combined Cycle Gas Turbine
CFB	Circulating Fluidised Bed
CHP	Combined Heat and Power
CV	Calorific Value
Economic lifetime	The number of years after which money is saved by abandoning the asset. This will always be less or equal to the technical lifetime, and will always hopefully be longer than the period required to pay back any project debt
Equivalence ratio	The actual air to fuel ratio divided by the stoichiometric air to fuel ratio. Affects the quality of the syngas produced.
EPC	Engineering Procurement Construction
EU	European Union
FB	Fluidised bed
FICFB	Fast Internally Circulating Fluidised Bed gasifier technology
FT	Fischer-Tropsch synthesis, a chemical catalytic process which converts syngas into liquid fuels, including diesel, under high temperature and pressure
Gasification efficiency	The performance of a gasifier is defined by the quality and quantity of gas produced. The performance is usually measured by the gasification efficiency. This can be calculated in three ways, by the Cold-gas efficiency, Hot-gas efficiency or Net gas efficiency
Gasification equivalence ratio	The ratio of air present in the gasification step to the amount of fed biomass fuel
gPROMs	General PROcess Modeling System – modelling software edited by Process Systems Enterprise.
Hearth load	The hearth load (also called specific grate gasification rate or space velocity) is the amount of producer gas to be obtained per unit cross sectional area of the throat (or the smallest area of cross section in the reactor). The hearth load of a moving-bed gasifier can be expressed in terms of the fuel gasified, the volume of product gas that is produced, or the amount of energy produced.
H_2	Hydrogen
HHV	Higher Heating Value
IRR	Internal Rate of Return
kW	kilo-Watt
kWh	kilo-Watt hours
kW_{th}	kilo-Watts of heat
LCOE	Levelised cost of energy
LHV	Lower Heating Value
LNG	Liquefied Natural Gas
LTHW	Low temperature hot water

MILENA	ECN’s dual gasifier technology
MTHW	Medium temperature hot water
MW	Mega-Watt
$\mathrm{MW_{in}}$	Mega-Watts input
$\mathrm{MW_{out}}$	Mega-Watts output
MWh	Mega-Watt hours
$\mathrm{MWh_{e}}$	Mega-Watt hours electricity
NPV	Net Present Value
odt	Oven Dried Tonnes – mass of feedstock at 0% moisture content
Opex	Operating costs
ppb	parts per billion, by volume
ppm	parts per million, by mass
ppmv	parts per million, by volume
SNCR	Selective non-catalytic reduction
SOFC	Solid Oxide Fuel Cell
Syngas	Gas composed mainly of hydrogen and carbon monoxide, produced from the gasification of biomass
t	metric tonnes
TEABPP	Techno-Economic Assessment of Biomass Pre-Processing
Technical lifetime	The number of years over which the plant asset is expected to be useful to the average owner, with normal repairs and maintenance, for the purpose it is acquired. This is typically the period over which a project is evaluated when in planning, and over which asset depreciation is charged in company accounting. This is the typical definition of “lifetime” in the TEABPP project, i.e. the service life of a plant
Total Installed Costs (TIC)	Procurement, installation and commissioning of equipment
Total Investment Costs (TinvC)	TIC + Operations & Utilities + Civils & Land + Project development + Contingency. Note all TinvC costs are given as overnight costs (i.e. as if the plant is constructed instantaneously within 2015, without construction costs inflating over years)
tpa	Tonnes per year
tpd	Tonnes per day
TRL	Technology Readiness Level, see section 2.3
Turndown ratio	The ratio of the highest practical gas generation rate to the lowest practical rate.
TUV	Vienna University of Technology, Austria
TWh	Tera-Watt hours (x 10^{12})
US DOE	United States Department of Energy
White pellets	Pellets produced from long rotation forestry biomass, without torrefaction, AFEX or steam explosion
WP	Work Package
w.b.	Wet basis
Yield (with respect to gasifiers)	The yield of a gasifier is used to measure the volume of producer gas generated per kg of feedstock supplied to the system.