A COMPARISON OF INTEGRATED BIOMASS TO ELECTRICITY SYSTEMS

ANDREW JONATHAN TOFT

Doctor of Philosophy

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SUMMARY

This thesis presents a comparison of integrated biomass to electricity systems on the basis of their efficiency, capital cost and electricity production cost. Four systems are evaluated: combustion to raise steam for a steam cycle; atmospheric gasification to produce fuel gas for a dual fuel diesel engine; pressurised gasification to produce fuel gas for a gas turbine combined cycle; and fast pyrolysis to produce pyrolysis liquid for a dual fuel diesel engine. The feedstock in all cases is wood in chipped form. This is the first time that all three thermochemical conversion technologies have been compared in a single, consistent evaluation.

The systems have been modelled from the transportation of the wood chips through pretreatment, thermochemical conversion and electricity generation. Equipment requirements during pretreatment are comprehensively modelled and include reception, storage, drying and comminution. The de-coupling of the fast pyrolysis system is examined, where the fast pyrolysis and engine stages are carried out at separate locations. Relationships are also included to allow learning effects to be studied. The modelling is achieved through the use of multiple spreadsheets where each spreadsheet models part of the system in isolation and the spreadsheets are combined to give the cost and performance of a whole system.

The use of the models has shown that on current costs the combustion system remains the most cost-effective generating route, despite its low efficiency. The novel systems only produce lower cost electricity if learning effects are included, implying that some sort of subsidy will be required during the early development of the gasification and fast pyrolysis systems to make them competitive with the established combustion approach. The use of de-coupling in fast pyrolysis systems is a useful way of reducing system costs if electricity is required at several sites because a single pyrolysis site can be used to supply all the generators, offering economies of scale at the conversion step.

Overall, costs are much higher than conventional electricity generating costs for fossil fuels, due mainly to the small scales used. Biomass to electricity opportunities remain restricted to niche markets where electricity prices are high or feed costs are very low. It is highly recommended that further work examines possibilities for combined heat and power which is suitable for small scale systems and could increase revenues that could reduce electricity prices.

Keywords: biomass, electricity, combustion, fast pyrolysis, gasification

Dedication

This thesis is dedicated to my mother's fortitude, support and sense of humour.

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1. INTRODUCTION

This thesis evaluates the cost and performance of various systems that could be used to generate electricity from biomass at capacities of 1-100 MW_e (MW electric). All systems are fed by wood, delivered to the process in chipped form. This feedstock is converted using combustion, fast pyrolysis or gasification into steam, pyrolysis liquid or fuel gas respectively and these intermediate energy carriers are used in steam turbines, engines or gas turbines for electricity generation. These system options will be described in the next chapter.

This chapter presents an overview of current energy supply and discusses the reasons why biomass to electricity systems should be studied. The chapter begins by discussing the problems associated with the current energy mix and the solutions that are being promoted to alleviate them (Section 1.1). One of the solutions is a diversification of the energy mix to embrace more renewable energies, and biomass is introduced as one such energy source (Section 1.2). The chapter then focuses on the reasons for using biomass in electricity generation (Section 1.3). The problems of electricity generation from biomass using conventional technology are summarised in Section 1.4, leading to an introduction to the concept of integrated biomass to electricity systems. These systems are the focus of this thesis and the objectives of the work are summarised in Section 1.5. The chapter concludes by previewing the rest of this thesis in Section 1.6.

1.1 CURRENT ENERGY SUPPLY

Contemporary energy supply is dominated by fossil fuels, large scale hydroelectric schemes and nuclear power (see Table 1.1 [1]). This situation is not tenable for the following reasons.

1. The combustion of fossil fuels produces carbon dioxide (CO_2) and is widely accepted as a significant contributor to global warming [2]. Over 160 countries are now committed to reducing their CO_2 emissions to 1990 levels by the year 2000 as a result of UN Framework Convention on Climate Change agreed at the UN Earth Summit in Rio de Janeiro in 1992. These countries must now find ways of meeting these targets: solutions include increasing energy efficiency; increased use of natural gas (which produces less

CO_2 per unit of energy); sequestering carbon in trees; and increasing the proportion of non-fossil fuels in the energy mix.

2. Sulphur dioxide (SO_2) and oxides of nitrogen (NO_x) emitted in fossil fuel combustion are blamed for increasing atmospheric acidification (leading to acid rain) and local air quality problems such as smog. The European Commission have imposed strict regulations to control emissions from major polluters such as power stations and road vehicles [3]. The power generation sector has reacted in three ways: the use of 'clean' coal technologies such as the integrated gasification combined cycle (IGCC) that increase efficiency and reduce emissions; further reliance on natural gas, which can produce lower SO_2 and NO_x; and the development of alternative energy resources such as solar, wind, hydropower, and biomass.

3. The supply and demand of crude oil exerts considerable influence on energy markets. National economies are highly sensitive to crises in the politically unstable areas where oil reserves are concentrated (as seen in the Gulf War of 1992) or price controls (the Oil Crises of the early 1970s). There is a political will to diversify energy supplies to embrace indigenous primary energy sources and reduce the influence of any single fuel [4]. This again implies a move away from the current fossil fuel dominated energy mix to one where alternative energy plays a much more substantial role.

4. Fossil fuels are ultimately a finite resource. In the extreme case reserves may actually run out, but the more immediate scenario is that supply will become less reliable and more expensive as the most accessible reserves are depleted [5, 6, 7]. This will again put national economies in jeopardy and is further impetus towards a more diverse and sustainable energy mix.

5. Opportunities for new large-scale hydroelectric schemes are restricted by a combination of environmental issues and financial constraints. Population displacement and the flooding of what is often prime agricultural land for reservoirs are meeting increasing opposition [8]. These problems are compounded by large capital investment requirements, threats of catastrophic dam failures, the diminishing quality of downstream land and damaging impact on the aesthetics of the landscape [9].

6. Nuclear power was once viewed as a viable and environmentally friendly alternative to electricity generation through fossil fuels, but is now less widely accepted. The World

Commission on the Environment and Development has shown that nuclear power is blighted by the environmental problems of waste disposal (both during the lifetime of the plant and at decommissioning) and the risk of major accidents [9].

Table 1.1 - Commercial primary energy supply

	1950 mtoe	%	1970 mtoe	%	1990 mtoe	%
Coal	1000	56.7%	1641	31.8%	2192	27.3%
Crude oil	520	29.5%	2272	44.0%	3101	38.6%
Natural gas	160	9.1%	929	18.0%	1738	21.6%
Hydro	84	4.8%	305	5.9%	541	6.7%
Nuclear	-	-	20	0.4%	461	5.7%
Total	1764		5167		8033	

mtoe million tonnes of oil equivalent; 1 toe = 40.3 GJ

The only prudent reaction to the problems above is a diversification of the energy supply mix to include more renewable energy sources. This is a view supported at international, regional, national and corporate levels. The World Energy Council predicts that energy supplied from renewable sources will double by 2020, while the UN has forecast that 40% of primary energy and 60% of electricity will be generated from renewables within the next century [10]. The EU aims to double its renewable contribution from 4% to 8% of primary energy in the next 10 years [10]. The UK is committed to generating 20% of its 1991 electricity output from renewable sources by 2025, equivalent to 60 TWh/y [10]. At the corporate level, Shell scenarios for future energy supply show a diversification of energy supply and an increase in market share for renewable energies; one such scenario is shown in Figure 1.1 [11].

1.2 BIOMASS FOR ENERGY

Biomass is one of the renewable energies that could play a substantial role in a more diverse and sustainable energy mix. Biomass may be defined as any renewable source of fixed carbon. The term is generally used to describe plant material such as wood, wood residues, agricultural crops and their residues. Industrial and municipal wastes are often also considered as biomass due to their high percentages of food waste and fibre.

Biomass is already a significant source of primary energy. It supplies 14% of global primary energy; 38% of primary energy in developing countries; and 2.5% of primary energy in the EC

[3, 4]. Nearly all of this existing use is for cooking, heating and lighting and is non-commercial (hence its exclusion from Table 1.1).

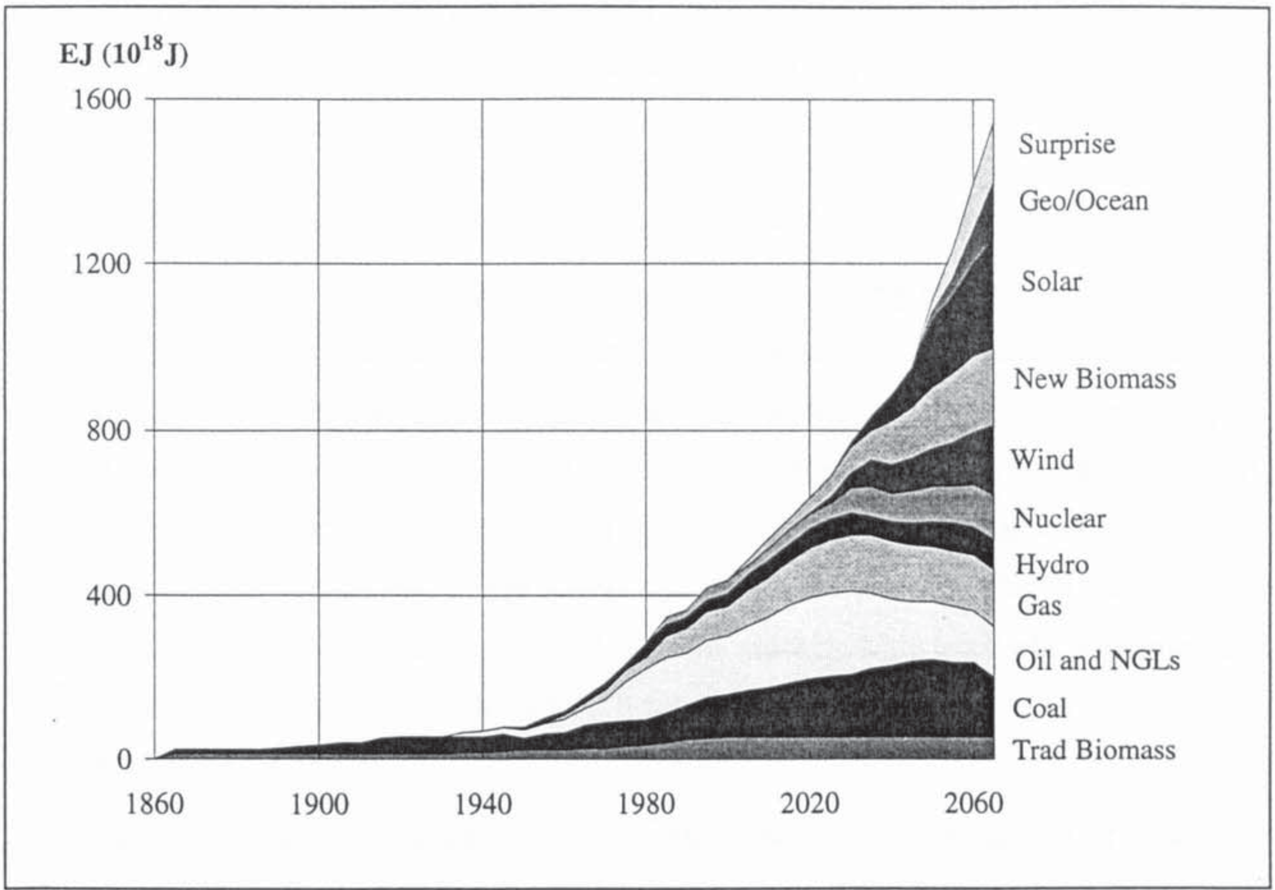


Figure 1.1 - Shell scenario for future energy supply

Biomass use for energy has several environmental benefits:

1. CO_2 released in biomass combustion is equivalent to that consumed during growth of the plant matter. Although there is some CO_2 released as a result of energy consumed by machinery and construction, the CO_2 balance over the life cycle of a biomass to energy project is still better than most alternatives and sustained biomass use offers an excellent opportunity to reduce CO_2 emissions [12, 13].

2. Large scale planting of energy crops (especially wood) would result in a substantial amount of carbon sequestered in plant matter and a reduction in atmospheric CO_2 [2, 14]. It has been shown that a net sequestration of carbon will be achieved for typically the first 100 years after initial planting [15].

3. Most biomass sulphur contents are very low and combustion produces negligible SO_x emissions [12].

4. The cultivation of carefully selected fuel crops can control soil erosion, improve soil fertility, benefit local hydrology, reduce herbicide, pesticide and fertiliser use [16, 17].

5. Ashes from biomass thermal conversion could potentially be used as a fertiliser by biomass producers, creating a closed-loop system and reducing or eliminating the requirement for conventional fertilisers [18].

There are also good social and economic reasons for promoting biomass use:

1. The European Union Common Agricultural Policy currently pays farmers to take 15% of agricultural land out of food production as set-aside land. This land could be used to produce biomass fuel crops, retaining the land in active service and preventing its decline. It has been estimated that 1.5 Mha will be set-aside in the UK by the year 2000, with a European total of 20 Mha [19]. Ultimately the area available for biomass planting could reach 50% of the current agricultural land, or 125 Mha by the year 2070 [20]. Similar policy initiatives in the US are also making large tracts of agricultural land available for alternative uses, with 8-16 Mha of agricultural land available in the short term [21].

2. Biomass production and utilisation is a very labour intensive method of producing energy (3-6 times more labour intensive than power from fossil fuels). While this is often considered an economic disadvantage, high labour requirements are an advantage in this case because the jobs are created in rural communities. Thus biomass production, processing and utilisation offers opportunities for regeneration and diversification of rural economies [19, 22]. A UK estimate has shown that electricity generation from short rotation coppice could create 3.6 jobs/MW_e, compared with 1.5 jobs/MW_e for small-scale hydroelectric schemes and 0.2 jobs/MW_e for wind energy [23].

3. Biomass to energy schemes could be adapted as a means of waste disposal, offering the dual benefits of heat and power generation and a reduction in the amount of material to be sent to landfill [10].

1.3 BIOMASS FOR ELECTRICITY GENERATION

It has already been stated that biomass plays a significant role in non-commercial energy, mainly as a source of heat. Commercial exploitation of biomass energy is more limited. Elliot has summarised the potential for biomass in the energy market, which he divides into the solid boiler fuel sector; the liquid transportation fuel sector and electricity generation [24]. He argues that electricity generation offers the best opportunities on the basis that:

1. Biomass can only compete with other solid fuels where the biomass source and energy plant are local: high transport costs limit more widespread application.

2. Conversion of biomass to liquid transport fuels produces a liquid that is 4-6 times more expensive than conventional fuels. Even with technology advances biomass-derived liquids are unlikely to be competitive without fiscal incentives or a marked increase in crude oil prices (Shell do not expect any dramatic increase in oil prices to be sustainable [25]).

3. Electricity is the most valuable energy commodity, and as such the gap between the cost of biomass and the price of the energy product is greatest. This offers the best opportunity for market penetration.

Biomass is already used for electricity generation, with 8 GW_e of generating capacity in the US and a world-wide generating capacity of 9 GW_e [26]. These plant use various biomass and wastes (mainly wood) in boilers to raise steam that is used to drive a steam turbine.

1.4 INTEGRATED BIOMASS TO ELECTRICITY SYSTEMS

Current biomass-based electricity generation is constrained by low generating efficiencies, high capital costs, poor environmental performance and the limited availability of suitable feedstocks. Integrated biomass to electricity systems are designed to overcome these problems. Integrated biomass to electricity systems comprise (see Figure 1.2):

1. Sustained production of biomass feedstocks;

2. Conversion of solid biomass into liquid and gaseous intermediate energy carriers; and

3. Electricity generation using high efficiency cycles that use the intermediate energy carrier.

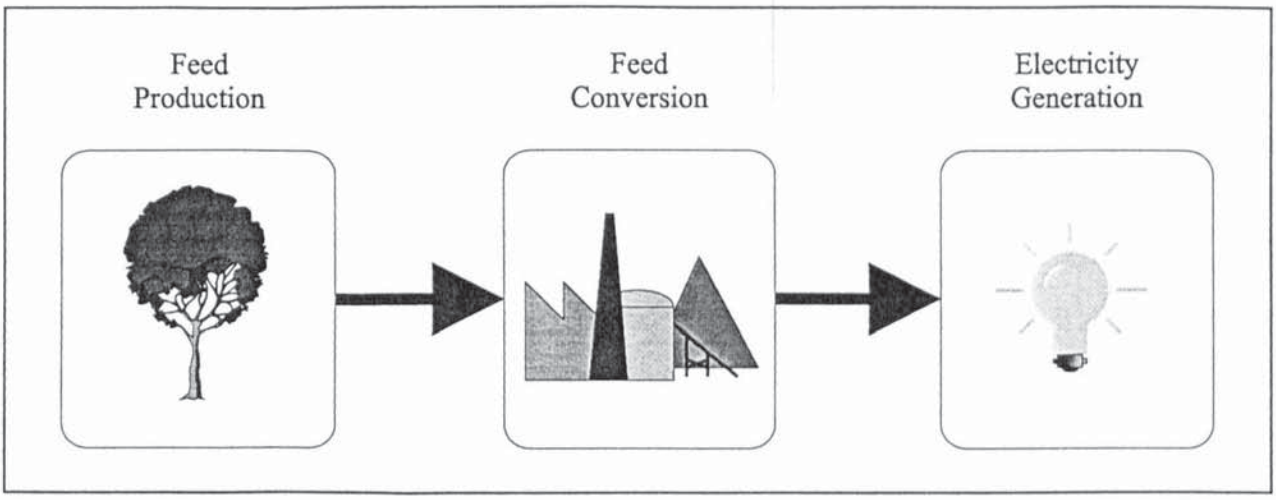


Figure 1.2 - Basic stages in an integrated biomass to electricity system

Feed production is required to substantially increase the biomass resource base. Current biomass resources are predominantly wastes and residues from forestry and agriculture. While such resources offer enough biomass in the short term, they cannot meet the demands that would be imposed if expansion of biomass energy is to reach the levels projected in Section 1.2. For this reason the production of biomass fuel crops is required. Fuel crop development is well-supported by national research programmes [18, 27] and through international initiatives such as International Energy Agency Bioenergy Agreement [28] (now known as IEA Bioenergy). This research is concentrating on increasing yields, reducing production costs and reducing the environmental impact of sustained fuel crop production.

Electricity generation using the steam cycle is established but far from ideal for biomass fuels. Resource limitations and the high costs of biomass transportation limit plant sizes and at low plant sizes steam turbine plant are inefficient generators. In the US the average plant capacity for a biomass-fired generating facility is 17 MW_e with generating efficiencies of around 20% [26, 29]. Also steam cycle capital costs are highly sensitive to scale and at such low capacities the capital investment required is very high [30]. Biomass combustion is further complicated by potentially high particulate emissions and ash fusion problems in the combustor [31, 32].

The integrated biomass to electricity system incorporates a feed conversion step to alleviate the above problems. This step converts solid biomass to an intermediate liquid or gaseous fuel that may then be used in gas turbines or engines for increased generating efficiency at small scale. Investment may be reduced through the use of equipment with less scale-sensitive capital costs. Emissions may be controlled in several ways: primary measures during conversion can reduce the formation of pollutants; secondary measures can remove pollutants by treating either the intermediate energy product or the flue gases from combustion. Lower operating temperatures during conversion help to reduce ash fouling.

Feed conversion techniques may be either thermochemical or biochemical. The latter option uses biological agents to produce the fuel intermediate (e.g. ethanol in fermentation or methanol in anaerobic digestion). Biochemical conversions will not be considered in this work because their application in electricity generation is limited. Anaerobic digestion is only suited to small-scale applications (12 MW_e is considered large [33]), and is usually applied to 'wet' biomass streams or slowly decomposing material at landfill sites [10]. Fermentation can only use the carbohydrate fraction of the biomass for ethanol production and as such part of the biomass is either wasted (although the residues may be used to provide energy) or transformed to carbohydrates using hydrolysis, adding to process cost and complexity. Fermentation is seldom viable without fiscal incentives and is only considered at very large scales [14, 34].

Thus the feed conversion in the integrated biomass to electricity system would be achieved using thermochemical conversion technologies that use heat to produce the fuel intermediate. Thermochemical technologies are combustion, gasification, and pyrolysis (each is described in the next chapter). Gasification produces a fuel gas, pyrolysis produces a liquid fuel, char and a gas, and combustion produces heat although for the purposes of this work it is assumed that the product of combustion is steam, raised in a boiler. Combustion of biomass is an established process [35], although research concerning emissions, co-combustion with conventional uses and ash characteristics is active [36]. Gasification and pyrolysis are less established, and both are well-supported by research programmes internationally, as discussed in Chapter 2.

The generating cycles that could be used in the electricity generating step of the integrated biomass to electricity system are generally established for conventional fossil fuels. There is much less experience of such cycles using the intermediate energy carriers produced by gasification and fast pyrolysis and the problems imposed by the novel fuels are described in the next chapter.

In summary, the integrated biomass to electricity systems aims to combine feed production, feed conversion and electricity generation for efficient, sustainable and viable electricity generation. Although each stage in the system is actively supported by current research and development programmes, much less is known about the integration of the three stages to give an overall system. Given this uncertainty, potential feed producers are reluctant to commit resources to feed production for fear of a lack of biomass conversion facilities while potential generators are discouraged by a lack of biomass resources.

Techno-economic studies are used to predict the cost and performance of bioenergy systems, highlighting development opportunities, increasing awareness of system constraints and helping to reduce uncertainty. Such studies can be used to direct research effort and funding to the most important areas of the system. As such they can accelerate the development of integrated biomass to electricity systems and help to secure a large share of the energy market for this renewable resource.

1.5 OBJECTIVES OF THIS WORK

This work will calculate, evaluate and compare the costs and performance of integrated biomass to electricity systems that use combustion, pyrolysis and gasification in combination with steam turbines, gas turbines and internal combustion engines. The evaluations will include feed transport, feed pretreatment at the conversion facility, feed conversion and electricity generation. Particular features of this work that differentiate it from previous techno-economic studies are:

- the comparison of all three thermochemical conversion technologies on a consistent and comparable basis. Previous studies have compared gasification and combustion only.
- the development of generic models that calculate the cost and performance of systems for a continuous range of capacities from 1-100 MW_e. This allows an examination of the viable operating capacities for each technology and the preferred technologies for each capacity.
- the comprehensive examination of pretreatment, with the development of a model that calculates the costs and performance of feed reception, storage, handling, drying and comminution depending on system capacity, the feedstock and the requirements of the conversion technology.
- an analysis of de-coupled systems, where feed conversion and electricity generation are separated by time or space. This de-coupling option is unique to pyrolysis, and various

de-coupled systems are evaluated to assess whether de-coupling can give pyrolysis a cost advantage over the other technologies.

- an allowance for learning effects. Increased experience of a technology tends to reduce its costs, and the incremental cost savings diminish with each replication. The costs of established technologies are therefore relatively constant while the costs of systems using novel technologies are expected to fall. The impact of this learning effect on electricity production cost is evaluated.

1.6 OUTLINE OF THESIS

Chapter 2 will describe the basic features and status of the options available in feed production, feed conversion and electricity generation. It will also examines the problems that must be overcome in combining the various options to give an overall system. The chapter concludes by selecting the most suitable systems for further evaluation in this study.

Chapter 3 will review previous system studies. The scope, results, strengths and weaknesses of a number of recent comparative studies will be highlighted so that a scope can be developed for this work that supplements previous studies and builds on experience already gained.

Chapter 4 will describe the methodology that will be used to model the biomass to electricity systems. It will identify the performance parameters that will be used to compare each system. Methods for developing investment and production cost relationships will be defined. These definitions will ensure that the models developed in subsequent chapters are consistent and comparable.

Chapters 5-8 will describe the development of the modules (sub-models) that will be used to evaluate the systems. Feedstock and liquid fuel transportation will be discussed in Chapter 5. The feed pretreatment module will be described in Chapter 6. Chapter 7 will present the development of the feed conversion modules and Chapter 8 will describe the electricity generation modules.

Chapter 9 will evaluate and compare biomass to electricity systems using the models reported in Chapters 5-8. Chapter 10 will draw conclusions from the systems evaluations. The thesis will end with recommendations for further work in Chapter 11.

2. THE INTEGRATED BIOMASS TO ELECTRICITY SYSTEM

2.1 INTRODUCTION

The previous chapter explained the background to this work and introduced the concept of a sustainable, integrated biomass to electricity system comprising three stages: biomass production, thermochemical biomass conversion and electricity generation. There are many potential ways of combining feed production, feed conversion and electricity generation using the three thermochemical conversion technologies (combustion, gasification and pyrolysis). This chapter examines these options and determines the actual system combinations that will be evaluated further in the remainder of the thesis.

The system options are evaluated on the basis of their suitability for the generating capacities envisaged for biomass systems, their current status, technological constraints and opportunities and ultimately their near-term potential for biomass to electricity. The chapter does not present details of technology costs and performance beyond that needed to select appropriate systems since more detailed data will be presented in later chapters as the selected systems are modelled.

The chapter is structured as follows.

- A suitable generating capacity range is determined in Section 2.2.
- Wood production methods and fuel characteristics are introduced in Section 2.3.
- The interface between biomass production and biomass conversion is discussed in Section 2.4.
- Systems based on wood combustion are discussed in Section 2.5.
- Potential wood gasification systems are discussed in Section 2.6, including the problems arising at the feed conversion and electricity generation interface.
- Potential wood fast pyrolysis systems are discussed in Section 2.7, with details of the problems that may arise at the feed conversion and electricity generation interface.
- Problems that may arise when integrating the system with its environment are introduced in Section 2.8.
- The chapter concludes by defining the systems that will be evaluated further in Section 2.9.

2.2 GENERATING CAPACITY

It has been suggested that electricity could be generated commercially at scales as low as 100-500 kW_e [37, 38]. However, such systems would suffer from very high capital costs and poor system efficiencies and are unlikely to be economic in all but a few exceptional circumstances such as in very remote locations or on-farm plant where electricity generation is a marginal activity. The lower threshold for commercial electricity generation in an integrated biomass to electricity system is likely to be higher. In the US the minimum generating capacity is around 1.5 MW_e [39]. Asplund suggests a minimum capacity of 3-5 MW_e for cogeneration plant in Europe [40]. The lowest generating capacity that will be evaluated in this study will be 1 MW_e, on the understanding that this is expected to be below viable system capacities and provides a suitable margin to highlight problems at the very small scale.

High electricity generating capacities are constrained by the high cost of biomass transport and the limited availability of biomass resources. Most literature resources suggest that the practical upper limit is between 50 and 100 MW_e [8, 29, 41, 42, 43]. This thesis will consider systems up to a maximum capacity of 100 MW_e. Again, this limit is beyond the level that is likely in the UK, but will allow trends at the higher capacities to be observed. Limited resources at locations in the UK are expected to constrain maximum capacities to far below 100 MW_e.

2.3 FEED PRODUCTION

2.3.1 The selection of wood

There are many biomass resources that are being considered for energy applications including agricultural crops such as miscanthus and switchgrass [44]; agricultural residues such as straw [10]; residues from pulp and paper industries [45] and municipal solid wastes [46]. While all these feedstocks have significant potential as energy sources and are widely supported by energy programmes in Europe and the US, the predominant biomass feedstock for electricity generation is currently wood. Wood accounts for 2/3 of the total US biomass to electricity generating capacity, as shown in Figure 2.1 [47]. In Europe the importance of wood is reflected by the support for wood-based projects under the current THERMIE EC demonstration programme for Biomass and Energy from Wastes. In this programme almost half of the projects in gasification and pyrolysis were based on wood feedstocks, as shown in Table 2.1 [48]. There is far more information available about wood production, handling and processing than any other feedstock. For these reasons this thesis will focus exclusively on wood feedstocks.

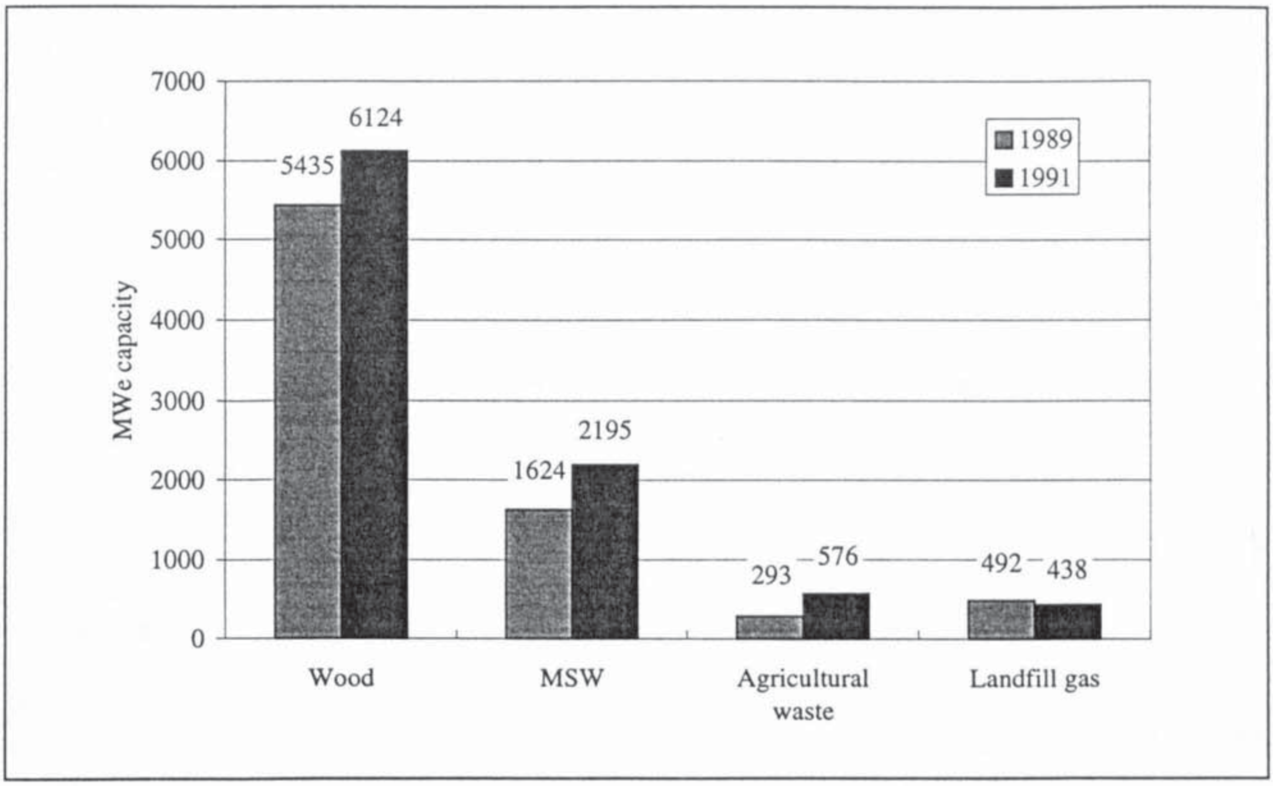


Figure 2.1 - Biomass supply mix for electricity generation in the US

Table 2.1 - Projects supported under the EC THERMIE demonstration programme

	Agri-culture	Forestry	Stock rearing	Aqua-culture	Urban waste	Industrial waste	Total
Biomass harvesting	2	2	-	-	-	-	4
Energy plantations	-	3	-	-	-	-	3
Waste treatment	1	3	2	-	13	4	23
Biogas production/ utilisation	2	-	19	2	33	31	87
RDF production/ utilisation	1	-	-		14	1	16
CHP by combustion	15	**17**	3	-	10	17	**62**
Gasfication and pyrolysis	7	**15**	-	-	4	8	**34**
Compost and fertilisers	-	-	3	-	3	2	8
Chemicals, alcohol biological	1	-	-	-	-	-	1
Chemicals, alcohol, thermo-chemical	8	-	-	-	1	4	13
Production of proteins	-	-	1	-	-	1	2
TOTAL	37	40	28	2	78	68	253

2.3.2 Conventional forestry practices

Wood feedstocks are available as residues from conventional forestry practices or as specially grown energy feedstocks from short rotation forestry (SRF). Conventional forestry wood is the near-term resource, SRF is emerging and is discussed in Section 2.3.3.It has been estimated that annual UK resources of wood from conventional forestry were 3.26 Mt in 1990 and expected to rise to 4.00 Mt in 2010 [51]. Assuming an energy value of 9 GJ/t (lower heating value, green tonne at 50% moisture content), this represents an energy resource of 30 TJ/y (TJ = 10^{15} J).

Wood from conventional forestry can arise in three different ways. The first is the most commonly exploited resource: residues that remain in the forest after the high-value lumber has been removed. This residue has a low value but is expensive to collect and transport [49, 50]. The second option is the use of trees removed in thinning processes (the premature cutting of low quality trees to improve the growth rate of the remaining stock). These trees are of low value as lumber and could be used for energy although the harvesting and collection costs are higher than for residues. The final and most expensive resource is clearfell wood, where all the wood at a site would be cleared for fuel. Such wood has a high value in other markets such as the pulp and paper industry and it is unlikely that the energy market could support such prices. These harvesting techniques and the associated costs in UK practices are reported by Mitchell [51] and typical wood harvesting costs before transport vary widely, as shown in Table 2.2 [52]. (All costs reported in this thesis are presented in US$, 1995 basis and are normalised to this currency using the methodology described in Section 4.5.2; odt are oven dry tonnes, the mass of feed at 0% moisture content).

Table 2.2 - Harvesting costs of conventional forestry wood in UK conditions

Harvesting system	Harvesting cost, \$/odt[a]
Residues - terrain chip	41-47
Residues - landing chip	9-23
Whole tree - terrain chip	12-22
Whole tree - landing chip	25-54
Integrated harvesting - thinning	31-50
Integrated harvesting - clearfell	10-25

2.3.3 Short rotation forestry

Short rotation forestry is a means of producing wood rapidly for energy markets. Trees may be cut back (coppiced) every 3-5 years on plantations with a life of around 30 years, or single stem trees may be cut down after 7-20 years growth. Short rotation forestry is a widely applicable technique, with many different species available to suit the local climate: temperate climates favour willow and poplar, with species such as robinia or eucalyptus preferred for more arid conditions [17].

Short rotation forestry is less established than conventional forestry. Extensive and long term silvicultural trials are underway to develop short rotation forestry methods that increase crop productivity, reduce environmental impacts and, ultimately, reduce production costs. Research topics include site selection; fertilisation and soil nutrient depletion; species selection and development; water requirements; weed control, pest control and disease management, and harvesting techniques [53].

There are 15000 ha of SRC plantations in N. Europe [43] and this figure is expected to rise with the current THERMIE demonstration projects in Europe that specifically support the development of short rotation forestry for energy at three 7-12 MW_e biomass to electricity facilities [54]. Current yields for N. Europe are 10-15 odt/ha/y [20], although these are based on small-scale plantations and larger plantations are likely to be less productive. In the US, the Short Rotation Woody Crops Programme increased yields during the 1980s from 9 to 13 odt/ha/y [55]. A yield of 10 odt/ha/y is realistic. Approximate costs for wood produced by short rotation coppicing in Europe are given in Table 2.3 [17].

Table 2.3 - Production costs of short rotation coppice

Harvesting cycle, years	Yield, odt/ha/y	Cost, \$/odt[a]
5	12	45-52
5	16	42-49
3	12	52-62

2.3.4 Wood fuel characteristics

2.3.4.1 Units

The unit oven dry tonne (odt) is used throughout this thesis to denote the mass of feedstock at zero moisture content; if a feedstock property is measured on a wet tonne basis then the unit "t" is used.

2.3.4.2 Composition

The ultimate analyses of various solid fuels are compared in Table 2.4 [56]. It can be seen that the composition of the wood feedstocks is quite consistent, and much more so than the compositions of the other biomass feedstocks or coal. Secondly, the sulphur content of the biomass feedstocks is negligible, resulting in almost no sulphur emissions in combustion. Nitrogen levels in the wood feedstocks are also low in comparison with fossil fuels, but NO_x emissions are not necessarily reduced because NO_x emissions are influenced by the reactor conditions as well as fuel nitrogen content.

Table 2.4 - Ultimate analyses of various solid feedstocks

Material	Composition, %dry feed						Heating value, GJ/t dry basis	
	C	H	O	N	S	Ash	HHV	LHV
Douglas fir	52.3	6.3	40.5	0.1	0.0	0.8	21.0	19.6
Douglas fir bark	56.2	5.9	36.7	0.0	0.0	1.2	22.0	20.7
Pine bark	52.3	5.8	38.8	0.2	0.0	2.9	20.4	19.1
Western hemlock	50.4	5.8	41.4	0.1	0.1	2.2	20.0	18.7
Redwood	53.5	5.9	40.3	0.1	0.0	0.2	21.0	19.7
Beach	51.6	6.3	41.5	0.0	0.0	0.6	20.3	18.9
Hickory	49.7	6.5	43.1	0.0	0.0	0.7	20.1	18.7
Maple	50.6	3.0	41.7	0.3	0.0	1.4	19.9	19.2
Poplar	51.6	6.3	41.5	0.0	0.0	0.6	20.7	19.3
Rice hulls	38.5	5.7	39.8	0.5	0.0	15.5	15.3	14.0
Rice straw	39.2	5.1	35.8	0.6	0.1	19.2	15.8	14.7
Sawdust pellets	47.2	6.5	45.4	0.0	0.0	1.0	20.5	19.1
Paper	43.4	5.8	44.3	0.3	0.2	6.0	17.6	16.3
Redwood wood waste	53.4	6.0	39.9	0.1	0.1	0.6	21.3	20.0
Alabama oak wood waste	49.5	5.7	41.3	0.2	0.0	3.3	19.2	17.9
Animal waste	42.7	5.5	31.3	2.4	0.3	17.8	17.1	15.9
Municipal solid waste	47.6	6.0	32.9	1.2	0.3	12.0	19.8	18.5
Charcoal	80.3	3.1	11.3	0.2	0.0	3.4	31.0	30.3
Pittsburgh seam coal	75.5	5.0	4.9	1.2	3.1	10.3	31.7	30.6
West Kentucky No.11 coal	74.4	5.1	7.9	1.5	3.8	7.3	31.2	30.1
Utah coal	77.9	6.0	9.9	1.5	0.6	4.1	32.9	31.6
Wyoming Elkol coal	71.5	5.3	16.9	1.2	0.9	4.2	29.5	28.3
Lignite	64.0	4.2	19.2	0.9	1.3	10.4	24.9	24.0

Another useful measure of fuel characteristics is the proximate analysis and typical analyses are presented in Table 2.5 [57]. This analysis shows the high volatile content of biomass feedstocks; this portion of the feedstock is driven off during heating and reacts in the vapour phase. A high volatile content indicates a fuel that is easy to ignite. Moreover, the fixed carbon that remains is far more reactive in biomass than in coal, such that gasification of biomass residual carbon is easier [56]. It can be concluded that the overall reactivity of biomass feedstocks is much higher than for coal.

Biomass ash contents are variable, although ash content of wood feedstocks is generally less than 1%. The ash content of bark is higher, and ranges from 2 to 5%. A wood-bark mixture would therefore be expected to have a lower ash content than a coal feedstock, which is an asset in that the mass of ash residues to be disposed of is lower. However, the composition of biomass ashes is different from coal ashes, notably in its high alkali metals content and this can lead to fouling problems and corrosion of downstream equipment [32].

Table 2.5 - Proximate analysis of various solid feedstocks

		Proximate analysis, dry basis		
		Volatile matter	Fixed carbon	Ash
Bark	Pine	72.9	24.2	2.9
	Oak	76.0	18.7	5.3
	Spruce	69.6	26.6	3.8
	Redwood	72.6	27.0	0.4
Wood	Redwood	82.5	17.3	0.2
	Pine	79.4	20.1	0.5
	Fir/Pine	75.1	24.5	0.4
Coal	Lignite	44.1	44.9	11
	Sub-bituminous	39.7	53.6	6.7
	Bituminous	16.0	79.1	4.9

2.3.4.3 Moisture content

Biomass feedstocks generally contain more moisture than conventional fossil fuels. Freshly cut wood feedstocks will typically contain 45-65% moisture measured on a wet basis [56]. All moisture contents in this report will be given on a wet basis, which is the mass ratio of moisture in the feedstock to the total mass of the feedstock. The alternative (not used here) is to measure the moisture content on a dry basis, using the mass ratio of the moisture in the feedstock to the dry mass of the feedstock.

It will be shown in the descriptions of each technology that the moisture content of a fuel has a major influence on the efficiency of the conversion process and the quality of the conversion products. Moisture in the feedstock must be evaporated in the conversion process, and if this latent heat is not recovered (by condensing flue gases for example) then the energy in the feedstock that was used to dry the feedstock is wasted. Moreover, high moisture contents increase the mass of the material while diluting its energy content: this makes the feedstock more expensive to handle and transport as shown in Section 2.3.4.6 below.

2.3.4.4 Heating value

The heating value of a fuel (sometimes referred to as its calorific value) is that energy released when a unit mass (for solid and liquid fuels) or a unit volume (for a gaseous fuel) of the fuel is burned completely in oxygen at reference conditions. A full definition of these conditions and methods for determining calorific values is given in BS 526 (1961). Two heating values can be quoted: higher heating value and lower heating value. The higher heating value assumes that any water in the combustion products is a liquid at the reference conditions (25°C and 1 bar) and therefore includes the latent heat released as the water vapour condenses. Lower heating values discount the latent heat available from condensing the water vapour since this energy is seldom recovered. The difference between higher and lower heating values is discussed in more detail in Section 4.3.1.

Heating values are given in Table 2.4. There is a wide range of heating values amongst biomass feedstocks, although as seen before the variance is much reduced when only wood is considered. The heating value of all the biomass feedstocks is considerably below that of the coals shown, due mainly to their low carbon content (which is displaced by oxygen). It should be noted that actual energy available per unit mass of wet material is much lower than the heating values presented in Table 2.4 because the moisture dilutes the energy-carrying dry matter. The energy required to evaporate the moisture will also reduce the net energy available during thermochemical conversion.

2.3.4.5 Morphology

Biomass feedstocks are produced in a wide variety of forms such as stems, shoots, chunks, chips, shavings or sawdust that will effect their handling properties, their behaviour in storage and their behaviour in the conversion process. It is usual for the larger feed sizes to be comminuted to chip form before transport because this makes them easier to handle and increases the bulk density and hence decreases transport costs (see Section 2.4.2) [58, 59, 60, 61].

2.3.4.6 Bulk density and energy density

The bulk density of a fuel is the mass of fuel contained in a unit volume and includes the space between the biomass particles. Bulk density is significant in that it changes with feed morphology and higher bulk densities are preferred because transport costs are reduced. A typical bulk density for wood chips on a dry basis is up to 150 kg/m^3 [56], and for a 50% moisture content this gives a bulk density of 300 kg/m^3. If the lower heating value of the wood chips is 19.3 GJ/odt when dry then the lower heating value at 50% moisture content is

8.47 GJ/t (details of this calculation are given in Section 4.3.1). Thus the energy density of the material in transit will be 2.51 GJ/m^3.

For comparison, the bulk density of coal is between 673 and 865 kg/m^3, depending on the type of coal and its morphology [62]. At a bulk density of 700 kg/m^3 and a lower heating value of 30 GJ/t, this gives an energy density of 21 GJ/m^3, or 8.3 times the value for wood chips. The difference between the two is manifested in increased transport costs, storage costs and larger reactor sizes for wood chips.

2.3.5 Summary

This work will assume a wood feedstock due to its near term availability, relatively consistent feed characteristics and the fact that the majority of data for the conversion processes applies to wood applications. It will be assumed that the wood is available for transport to the conversion facility in a chipped form. Further details of the feedstock and its characteristics will be defined in Section 4.3.

2.4 THE FEED PRODUCTION TO FEED CONVERSION INTERFACE

2.4.1 Introduction

The characteristics of wood feedstocks as they are found at harvest or collection are often very different from the feed characteristics demanded by the conversion reactor, and steps may be required to match the feedstock to the process, as indicated by Table 2.6. Every effort should be made to produce a biomass feedstock that minimises the handling and processing required between feed production and feed conversion [63]. However, there is a compromise that must be found since processing during feed production increases the feed cost at delivery. Conversely extra pretreatment equipment at the conversion plant can allow the use of a lower cost feedstock but at the penalty of increased the capital and operating costs [64].

Table 2.6 - Typical interface problems between feed production and feed conversion

Feed Production	Feed Conversion	Interface requirement
Seasonal production (in SRC)	Constant operation	Storage
Feed produced as trees	Small particle sizes required	Comminution
High feed moisture	Low feed moisture	Drying
Dispersed production	Conversion at specific locations	Transport

2.4.2 Transport

Biomass production is dispersed and therefore the feedstock must be collected, loaded and transported to the feed conversion facility. Biomass feedstocks are difficult to handle (a liquid or gaseous fuel would be much easier) and their low energy density (compared in Section 2.3.4.6) makes them expensive to transport. Even when the transport distances are low the fixed costs of loading and unloading can be a significant proportion of the total transport cost [65], but as distances rise transport costs become increasingly important. Ultimately the increased costs of biomass transportation may outweigh the cost savings brought about by scale economies and increased system efficiency as system capacity increases. Therefore the cost of feed transport should be included when evaluating integrated biomass to electricity systems at different scales.

2.4.3 Storage

Feed storage is required to ensure a continuous supply of feed to the conversion reactor despite changes in the feed production rate. If forestry residues are used as the feedstock then the production of feedstock is likely to be continuous [66] and the amount of feed that must be stored is low (enough to allow for disruptions to feed collection due to the weather or other unforeseen problems). However, if the feedstock is produced using short rotation coppicing then harvesting is a seasonal activity restricted to 3-4 winter months of the year [65]. Under these circumstances a significant amount of feedstock must be stored if the feed conversion reactor is to operate throughout the year.

The length of time that the feed is stored is important to the overall system performance and viability. The feedstock will degrade during storage and it may also experience changes in moisture content. The extents of these changes are a function of the morphology of the feedstock, its initial moisture content, the climate and the storage conditions (discussed in Chapter 6). The combined effect of dry matter losses and moisture content changes will be either a gain or a loss in the net energy available in the feedstock, and therefore the amount of feedstock that must be produced to meet the demands of the system.

The location of the feed store is also important. If long term storage is at the forest or plantation, then the biomass producer must bear the cost of dry matter losses. Conversely, the biomass user must bear the costs of feed degradation if there is long-term storage at the feed conversion facility. This aspect of the system was examined in associated work that is attached as Appendix A (introduced in Section 3.4.5), and will not be included here. However any dry matter losses and moisture content changes during short term storage at the conversion facility will be included.

2.4.4 Comminution

At some point in the feed supply chain the biomass must be converted from its original form to a size range that is suitable for the conversion process [63]. It has been noted in Section 2.3.4.5 that it is usual for wood feedstocks to be chipped before transport and this condition is assumed here. The timing of comminution to chips is important since feedstocks with a small particle size lose more dry matter in storage than feedstocks stored in larger pieces. It is possible to harvest and chip in one step and therefore the benefits of reduced dry matter losses must be balanced against the extra handling and processing required by separating harvesting and comminution. This feed supply option was also investigated in the work described in Appendix A, and comminution options outside the feed conversion facility will not be studied here.

Feed comminution at the conversion site will be examined. It is shown in Section 2.5.2 and Section 2.6.2 that the combustion and gasification technologies selected for evaluation will accept a chipped wood feedstock without further processing other than screening to ensure no over-size pieces or contaminants such as metal or rocks enter the reactor (discussed as the processes are modelled in Chapter 7). However, fast pyrolysis in a fluid bed is a more demanding technology and requires a feedstock with a particle size of less than 2 mm if high liquids yields are to be attained (see Section 2.7.2). Thus extra comminution will be required in a fast pyrolysis system. This grinding is extremely energy intensive (see Section 6.8) and it is important that it is included in the evaluation of fast pyrolysis based systems.

2.4.5 Drying

Freshly cut wood has a moisture content of around 50%. Thermochemical conversion of such moist feedstocks is feasible but generally inadvisable because any thermal processing will waste energy in the evaporation of the feed moisture and can reduce the quality of the fuel gas or liquid fuel produced. It is generally more efficient to dry the feedstock using waste heat at the conversion facility. The optimum moisture content for each process is a compromise between the costs of drying and the benefits gained from using a drier feedstock. The preferred moisture contents for each technology are discussed as they are modelled in Chapter 7.

Feed drying can also take place during storage, with moisture contents often approaching 30% after storage for one summer. Research is ongoing to investigate the most appropriate regimes for storage in the field to accelerate the drying process while limiting dry matter losses [67, 68, 69]. Data on moisture content changes during storage vary widely because of differences between the initial feed characteristics, the storage conditions, the harvesting season and the local climate. Drying of the feedstock before delivery will not be evaluated in this study since it has already been examined in the work included in Appendix A.

2.4.6 Summary

The comprehensive evaluation of an integrated biomass to electricity system should include analysis of the interface between feed production and the feed conversion facility. Variations in feed supply have been examined in the project reported in Appendix A and the effects of processing before transport on feed cost will not be featured here. However this thesis will include:

- the costs of wood chip transport since this will vary with system capacity;
- feed handling and storage systems appropriate to the capacity of the system to be evaluated;
- changes in feed characteristics during storage at the conversion facility;
- feed drying to the moisture content specified by the conversion process; and
- the comminution of the feed delivered to fluid bed fast pyrolysis systems to give a <2 mm powdered feedstock.

2.5 COMBUSTION SYSTEMS

2.5.1 Introduction

Combustion in this context refers only to the direct combustion of solid biomass to release heat without the production of an intermediate liquid or gaseous fuel. In combustion the carbon and hydrogen in the fuel are in theory completely oxidised to carbon dioxide and water although in practice some unreacted char and feedstock will remain. As a fuel enters the combustor it is heated and any moisture is evaporated off. Further heating of the dry material causes the volatile components to pyrolyse, or turn to a vapour. These components undergo various reactions to ultimately produce CO_2 and H_2O. Meanwhile the solid carbon that remains after pyrolysis is oxidised to complete combustion.

Combustors have evolved with the changing perception of biomass: initial combustors were nothing more than incinerators, used to eliminate a waste; latest combustors are much more efficient and clean, designed to recover as much energy as possible from a valuable resource while minimising emissions [70]. Combustion technology is fully established and already widely used in biomass applications [35].

2.5.2 Combustion reactor configurations

The most popular combustors for biomass applications are either stoker fired and fluid bed designs. Pile burners and suspension fired combustors are also available but the former type suffers from poor efficiency, high NO_x emissions and control problems while the latter type is highly efficient but generally avoided because it requires extensive feed pretreatment [71].

In stoker fired combustors the feed burns as it moves through the furnace while resting on a grate. Movement through the grate is either achieved by gravity on a stationary sloping grate or mechanically in travelling or vibrating grate designs. The stationary sloping grate is simple but poor feed dispersion on the grate makes the boiler inefficient and difficult to control. Travelling grates are more efficient and offer improved control. The vibrating grate is especially efficient due to even spreading of the feedstock [71].

Fluid bed designs burn the feed in a turbulent bed of inert material that is fluidised by combustion air flowing through it from underneath. The fluid bed acts as a heat sink and promotes heat transfer, mixing and complete combustion at isothermal conditions of around 800-900°C [74]. This avoids the peak temperatures observed in stoker configurations and reduces thermal NO_x emissions. The heat capacity of the bed material allows the combustion of high moisture content fuels and offers extra fuel flexibility [74]. Fluid bed combustors should be capable of accepting a wood chip feedstock with a wide variety of moisture contents. Other benefits include a low feed inventory, high combustion efficiency and reduced maintenance.

Fluid beds combustors feature either a bubbling bed or a circulating bed. Bubbling fluid beds comprise a relatively shallow bed, with fluidising gases moving with enough velocity to suspend the inert material without ejecting substantial amounts of the bed material with the volatiles and combustion gases. In the circulating fluid bed the fluidising gases have sufficient velocity to entrain bed material with the combustion products as they leave the reactor. The bed material is recovered in a cyclone system and returned to the bed along with any other solids such as char and ash. Virtually complete combustion of the feedstock is assured in circulating fluid beds by virtue of char recycling [72]; carbon conversion is not as high in bubbling fluid beds since some fine char material will inevitably leave the reactor with the combustion gases.

Capital costs of fluid bed systems are around 10% higher than pile burners and they incur higher operating costs by virtue of the power requirements for the fans needed for the fluidising gas [71]. However higher capital and operating costs may be offset by improved environmental performance, efficiency and feed flexibility.

Although grate-fired combustors are the norm for older biomass-fired plant [73], fluid bed combustors are rapidly becoming the preferred technology for biomass combustion because of their low NO_x emissions [74, 75, 76]. For this reason only fluid bed combustors will be modelled during the systems evaluations. This is a simplification since in reality a combustor would be selected to suit the particular system and grate-fired boilers may well be more suited to some cases, especially at the smaller scale.

Fluid bed boilers have been commercially available for over 20 years [74] at capacities ranging from 15 to 715 MW_{th} input. Over 110 units are operating or are planned for operation in the US [41], all with performance guarantees from the vendor. La Nauze [77] lists over 50 commercial installations that operate on biomass with capacities of 2.5-94 MW_{th}. Bubbling fluid beds tend to be limited to the lower size range, while circulating fluid beds are reported over the entire capacity range.

2.5.3 Applications

There are a number of ways of generating electricity using the heat produced in combustion, including the steam turbine, the reciprocating steam engine, Stirling engines, indirect fired gas turbines and direct fired gas turbines. These options have been reviewed in an IEA evaluation [36] that showed that the steam turbine is the only established generating technology. The other options had efficiency advantages but were not available commercially and most were confined to small scale applications. This work will only consider combustion with steam turbines. Direct combustion of solid biomass to raise steam for a steam turbine is the established technology for electricity generation. There is already 6 GW_e of wood-fired combustion and steam cycle generating capacity in the US [78]. There are substantial data in the literature on such plant and combustion may be considered here as a benchmark for the more novel systems.

Steam turbines are the established prime mover for power generation. The basic steam cycle is the Rankine cycle whereby boiler feed water is heated in a boiler to raise steam; the steam is expanded in a steam turbine to convert its thermal energy into mechanical energy that drives a generator; the expanded steam is condensed in a condenser; and finally the condensate is pumped to the boiler to be turned to steam again. Full expansion of the steam in the steam turbine (termed condensing turbines) produces the maximum amount of power. A common alternative is the non-condensing turbine where the steam is not fully expanded but extracted from the turbine at an intermediate pressure to provide process steam or heat. A diagram of the Rankine cycle with superheat is shown in Figure 2.2 [79]. In this instance a condensing turbine is used and the steam is superheated to increase the average temperature at which heat is supplied and hence the cycle efficiency is increased. There are many variations of the basic Rankine Cycle such as regenerative cycles and reheat cycles that may be examined in most textbooks on steam power generation [e.g. 79, 80].

Steam cycles are bound by thermodynamic and materials limitations to modest efficiencies of around 35% [81]. Such cycles are optimised through the use of high pressure, highly superheated steam combined with complex steam generation, reheat and regeneration options. This extra complexity and the materials demands imposed by high pressure steam increase capital costs dramatically at small scale, with only minor increases in system efficiency. Thus such enhancements are only viable in the larger steam cycles where scale economies can be realised [71]. Steam turbine efficiencies are also low at small scale with the biomass combustion systems can only attain overall generating efficiencies to around 30% [41].

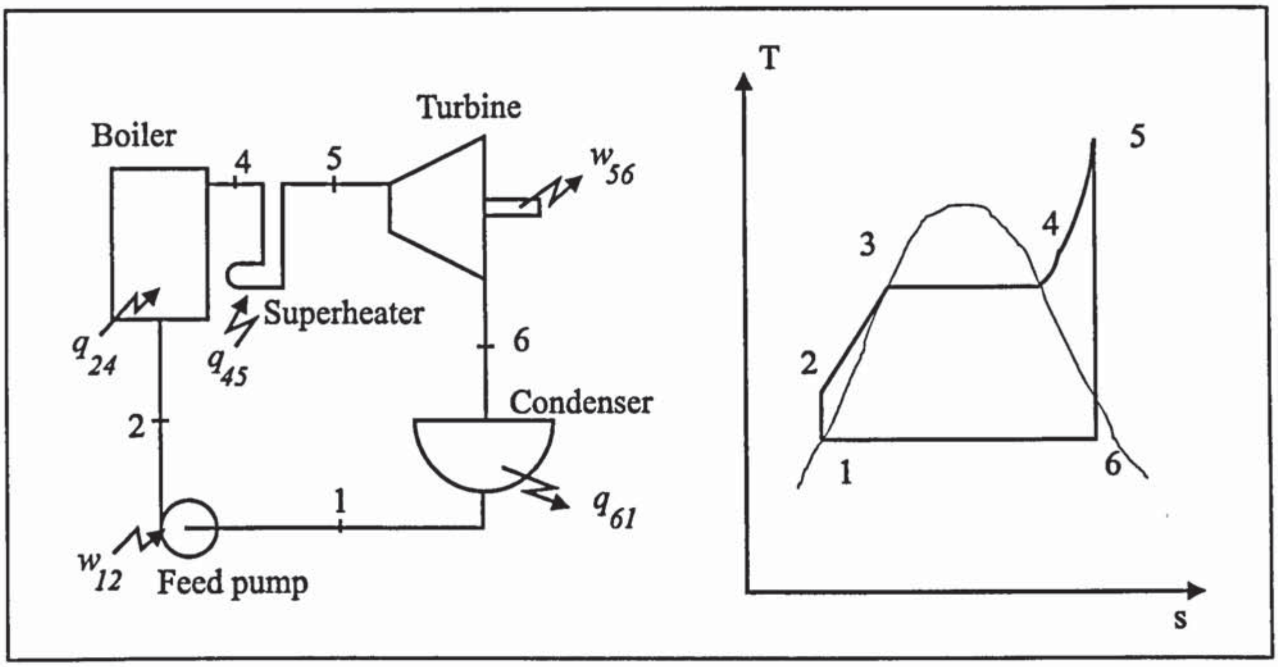


Figure 2.2 - The Rankine steam cycle

2.5.4 Constraints

Unlike the fast pyrolysis and gasification systems considered below, there are no major technology issues that must be resolved when using steam from a biomass-fired boiler in a steam turbine other than the omnipresent drive to reduce emissions.

One issue that has arisen recently is the problem of ash disposal. Proponents of biomass energy suggested that ash recycling to the forest is one advantage of biomass energy and yet the increasing production of biomass ashes is starting to cause concern in some countries [82]. The IEA is supporting the compilation of an international database of biomass ash characteristics in an effort to clarify this issue [36]. Another approach has been to concentrate the harmful constituents in a small portion of the total ash by control of the reactor temperature, flue gas cleaning temperatures and flue gas condensation [83]. This work will not consider the disposal costs of ash, but it would be a useful to assess the process costs or revenues generated in later analyses.

2.5.5 Summary

Combustion of biomass to raise steam for use in a steam turbine is the only established method of generating electricity from biomass and there is substantial data available on the costs and performance of such systems. It will therefore be used as a benchmark for the comparison with the more novel technologies described below. Although there are many combustor configurations, fluid bed combustion is gaining popularity due to its improved environmental performance, its wide capacity range and its fuel flexibility. For these reasons, this work will assume a fluid bed or circulating fluid bed combustor. Such combustors are able to process a wood chip feedstock with a wide range of moisture contents.

2.6 GASIFICATION SYSTEMS

2.6.1 Introduction

Development of gasification technology dates back to the end of the 18th century when hot gases from coal and coke furnaces were used in boiler and lighting applications. Gasification for power was first demonstrated in 1878 [84]. In the 1920s and 1930s there were some 12000 stationary gas producers in the US and 150 companies manufacturing gasifiers in Europe. The technology rapidly expanded during World War 2 in transport applications, to the extent that 900000 gas powered vehicles were operational in 1942 [85]. Gasification of coal is well-established and biomass gasification has benefited from activity in this sector. However the two technologies are not directly comparable due to differences in char reactivity, proximate composition, ash composition, moisture content and density.

Thermochemical gasification is the conversion by partial oxidation at elevated temperature of a carbonaceous feedstock into a gaseous energy carrier consisting of permanent, non-condensible gases. The process comprises three stages [86]:

1. Drying at 100-250°C, releasing water vapour and CO_2 at the higher temperatures.
2. Pyrolysis, where the biomass is thermally degraded in endothermic reactions at 300-500°C to give a gas and condensable tars, and leaving behind a char.
3. Gasification, which is the partial oxidation of the char, tars and gases at temperatures above 700-800°C in a variety of endothermic and exothermic reactions.

The heat for the endothermic pyrolysis stage is provided by exothermic reactions during gasification. This heat may be supplied directly by partial oxidation of the pyrolysis products in the gasification reactor or indirectly by char combustion in a separate combustor.

2.6.2 Reactor configurations

Gasifiers have been designed in various configurations, with the main options shown in Figure 2.3. There are other less established designs in development such as the twin fluid bed and the entrained bed that are in development but are not considered here due to a lack of information for modelling. Each type has been fully reviewed elsewhere [87, 88] and the characteristics of each type are summarised in Table 2.7 [89]. More details of the specific gasifiers featured in this study will be given when the process is modelled in Chapter 7.

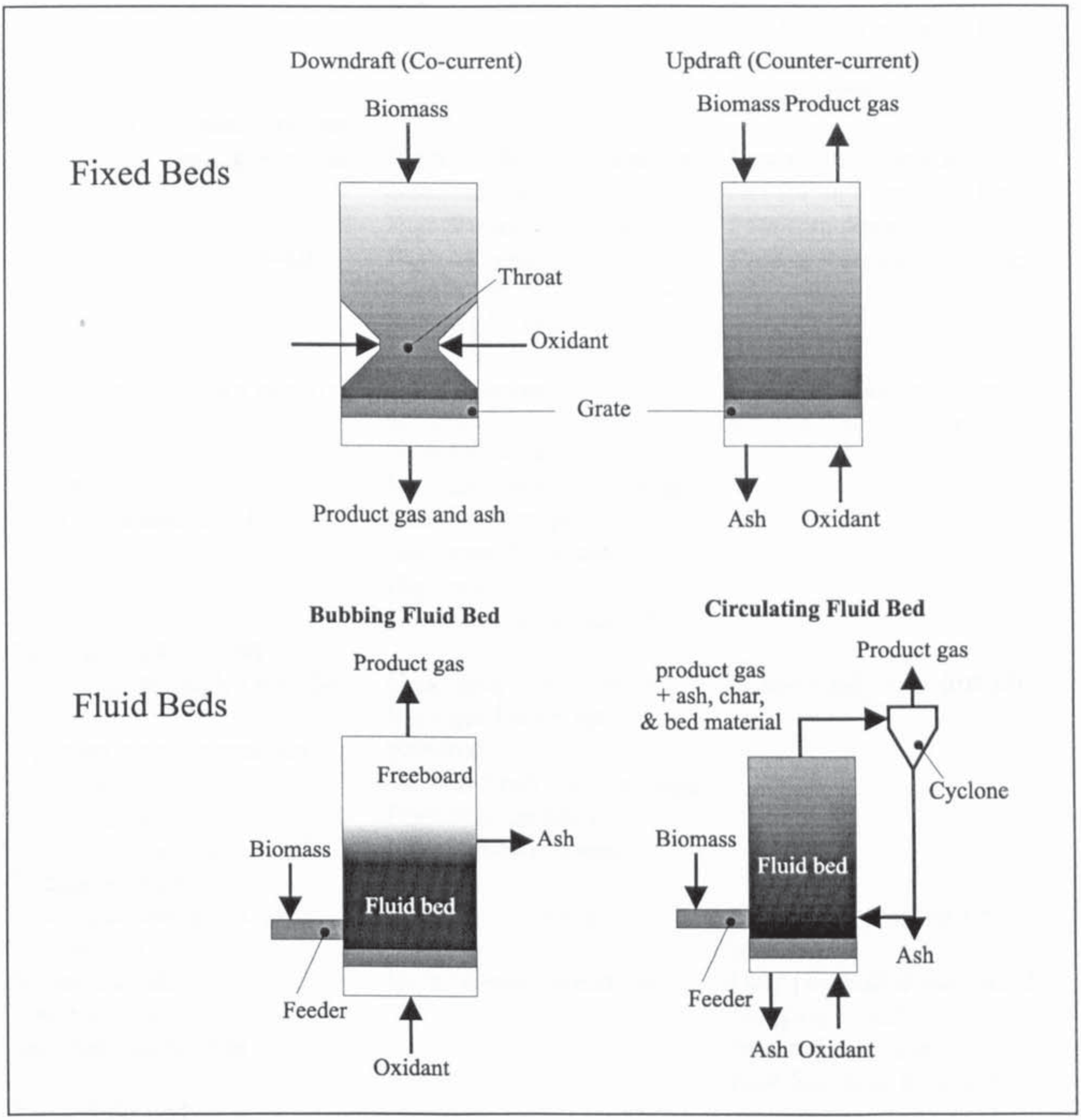


Figure 2.3 - The main gasifier configurations

Table 2.7 - Comparison of Gasifier Reactor Configurations

Mode of Contact & Gas Quality	Features	Limitations
Downdraft or Co-current		
Solid moves down, gas moves down Very low tar levels Moderate particulate levels	Simple, robust construction Proven for certain fuels High carbon conversion High residence time of solids Low ash carry over	Close specification on feed Very limited scale-up Low specific capacity Not for high moisture feeds Poor turn down capability Clinker formation on grate
Updraft or Counter-current		
Solid moves down, gas moves up Very high tar levels Moderate particulate levels	Simple, robust construction Good scale up potential High thermal efficiency High carbon conversion High solids residence time Suitable for direct firing	Low specific capacity Not for high moisture feeds Poor turn down capability Clinker formation on grate
Fluid bed		
Gas moves through a bubbling bed Inert solid stays in the reactor Fair tar levels High particulate levels	Good temperature control Good scale-up potential In-bed catalysts possible Increased particle size range Good scale up potential High specific capacity High reaction rates Good turn down capability	Low feedstock inventory Carbon loss with ash
Circulating fluid bed		
Gas moves through a bubbling bed Particles are elutriated and recycled Low tar levels Very high particulate levels	Good temperature control Very good scale-up potential Increased particle size range High reaction rates High carbon conversion	In bed catalysis is difficult
Entrained flow		
Fine feed carried by high velocity gas No inert solids Low tar levels Very high particulate levels	Very good scale-up potential High carbon conversion	Costly feed preparation required Only practical above ~10 t/h Slagging of ash Materials of construction Low feedstock inventory
Twin fluid bed		
Pyrolysis occurs in the 1st reactor Char combustion in 2nd reactor heats the bed material for the first reactor High tar levels High particulate levels	MHV gas using air only In-bed catalysts possible	Complex, costly design Only practical above ~5 t/h Scale-up possible but complex

A variety of biomass gasification processes have been developed commercially [144]. Of these processes, only fluid bed technologies are capable of operating within a 1-100 MW_e generating system and they are the predominant large-scale commercial technology [90]. Fluid bed gasifiers are the only type that are currently considered in electricity generating plant of greater than 1 MW_e [24, 91] and this work will focus entirely on this type of gasifier.

Fluid bed gasifiers are available from a number of manufacturers in thermal capacities ranging from 2.5MW_{th} to 150MW_{th} for operation at atmospheric or elevated pressures. Atmospheric bubbling bed gasifier manufacturers include JWP Energy Products, PRM Energy, Pyropower, and SEI [41, 92]. Energy Products of Idaho (EPI) have sold more than 40 units at capacities up to 73 MW_{th} [93]. Pressurised bubbling bed systems are being developed by IGT and Tampella. Tampella estimate that pressurised fluid beds are only likely to be economic at scales over 110 MW_e [93]. Atmospheric circulating fluid bed suppliers include Ahlstrom, Batelle, Gotaverken Energy, and Lurgi; while Bioflow in Finland is developing a pressurised system.

Fluid bed gasifiers are capable of handling a wide variety of feed particle sizes including chipped feedstocks [86]. Thus no further comminution of a wood chips feed would be required during pretreatment, although removal of larger particles would be necessary to prevent damage to the reactor feeding mechanism and to maintain an efficient, controllable process.

Fuel gases may be produced using various oxidants: oxygen, air, steam or a combination. The oxidant is usually air in biomass gasification since this avoids the expense and risks associated with oxygen separation. Another process option is the operating pressure, which may be near-atmospheric or pressurised. Atmospheric gasifiers are less expensive but the efficiency of the overall system can benefit from pressurised gasification, as discussed in Section 2.6.6.2.

2.6.3 Fuel gas characteristics

Ideally, the process produces only a non-condensible gas and an ash residue. In reality, incomplete gasification of char and the pyrolysis tars will produce a gas containing varying levels of the contaminants shown in Table 2.8 [89] and an ash residue containing some char. The composition of the gas and the level of contamination varies with the feedstock, reactor type and operating parameters, and typical gas characteristics are shown in Table 2.9 [86].

Treatment of the gas to remove the contaminants is a major barrier to the implementation of gasifiers in electricity generation, as will be discussed in Section 2.6.6.3.

Table 2.8 - Gasifier product gas contaminants

Contaminant and examples	Problems	Clean-up method
Particulates • Ash, char, fluidised bed material	Erosion	Cyclones, barrier filtration, wet gas scrubbing
Alkali metals • Sodium, potassium compounds	Hot corrosion	Cooling, condensation, filtration, adsorption
Fuel-bound nitrogen • Mainly ammonia and HCN	NO_x formation	Scrubbing, Flue gas treatments such as SCR
Tars • Refractive aromatics	Clogs filters, difficult to burn, deposits internally	Tar cracking and tar removal
Sulphur, chlorine • HCl, H_2S	Corrosion Emissions	Lime or dolomite, scrubbing, absorption

Table 2.9 - Gasifier product gas characteristics

	Gas composition, %v/v dry					HHV, MJ/Nm^3	Gas quality	
	H_2	CO	CO_2	CH_4	N_2		Tars	Dust
Fluid bed air-blown	9	14	20	7	50	5.4	Fair	Poor
Updraft, air-blown	11	24	9	3	53	5.5	Poor	Good
Downdraft, air-blown	17	21	13	1	48	5.7	Good	Fair
Downdraft, oxygen-blown	32	48	15	2	3	10.4	Good	Good
Multi-solid fluid bed	15	47	15	23	0	16.1	Fair	Poor
Twin fluidised bed gasification	31	48	0	21	0	17.4	Fair	Poor

2.6.4 Gasification applications

Biomass gasification can be used to produce heat, steam, bulk chemicals or electricity, as shown in Figure 2.4. Of the options shown for electricity generation, only gas turbines and diesel engines will be considered further on the grounds of applicability to scale and their potential for exploitation in the near term.

One alternative that should be considered in further work is the application of fuel cells. Fuel cells are electrochemical devices similar in many ways to batteries. In a battery the chemical store of ions is used up so that the charge eventually drains away; in a fuel cell, charge is constantly replenished by a fuel gas and oxidant. Fuel cells are promoted on the grounds of high efficiency and low emissions [94, 95] and they are emerging as a commercial technology for natural gas [96]. Gas specifications are extremely stringent to prevent electrolyte fouling, and is expected that this factor will limit applications to natural gas in the near-term [132].

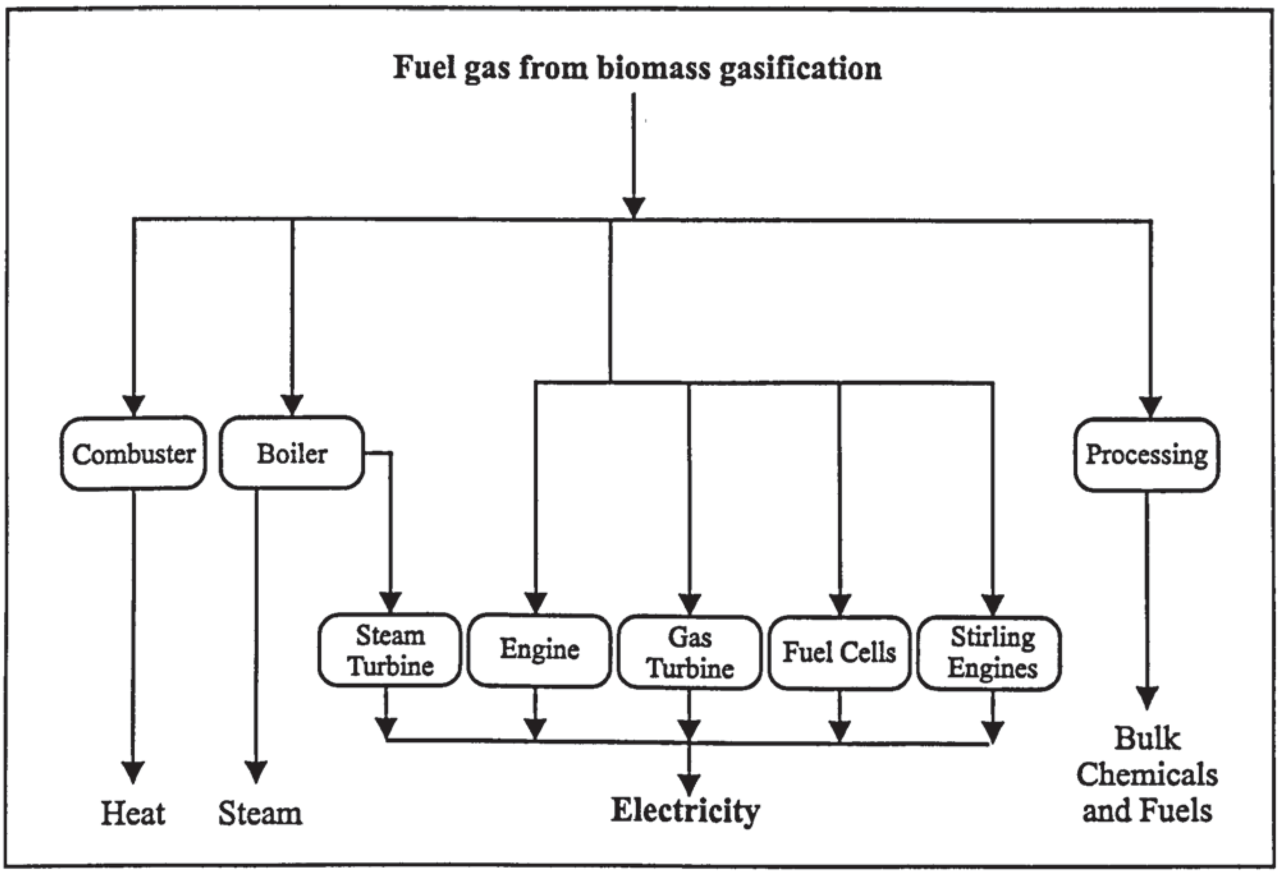


Figure 2.4 - Gasification applications

2.6.5 Integrating gasifiers and gas turbines - Activities

Gas turbine cycles have many advantages over steam cycles, notably higher efficiency; lower specific capital cost, especially at small scale; short lead times by virtue of modular construction; lower emissions; higher reliability and simpler operation [8, 97, 98]. The integration of gasification with gas turbines was first demonstrated at the Cool Water demonstration coal integrated gasifier/combined cycle (CIG/CC) 100 MWe facility in the United States [99, 100]. Gasification of coal for combined cycles is now so established that the HMIP in the UK are expected to declare the gasification combined cycle the best available technology for generating electricity from low quality high-sulphur fuels [101]. The status of biomass applications is not as advanced but there are many demonstration projects active [87, 102], as shown in Table 2.10 [103] and gas turbine tests on biomass fuel gases are underway by a number of organisations to support these projects [104, 105, 106].

Table 2.10 - Activities in gasification to electricity

Organisation	Gasifier	Technology	Generator	Status	MWe
Aerimpianti	Studsvik	CFB	Steam turbine	Operational	6.7
Battelle Columbus	Battelle	MSFB	Gas turbine	Operational	0.2
Battelle Columbus	Battelle	MSFB	Gas turbine	Design	12
Bioflow	Ahlström	Pressure CFB	Gas turbine	Operational	6
Border Biofuels	Wellman	Updraft	Engines	Abandoned	5
Elkraft/Elsam	Tampella	Fluid bed	Gas turbine	Design	7
ENEL	Lurgi	CFB	Gas turbine	Design	12
General Electric	GE	Updraft	Gas turbine	On hold	-
GEF	TPS	CFB	Gas turbine	Design	27
PICHTR	IGT	Pressure O_2 FB	not specified	Operational	2-3
North Powder	JWP (EPI)	FB	Steam turbine	Shut down	9
MTCI	MTCI	Fluid bed	Gas turbine	Design	4
Vattenfall	Tampella	Pressure FB	Gas turbine	Abandoned	60
VUB	VUB	Fluid bed	Gas turbine (Brayton)	Design	0.6
Yorkshire Water	TPS	CFB	Gas turbine	Design	8

CFB - Circulating fluid bed; MSFB - Multi-solid fluid bed; FB - Fluid bed

Further details of four projects in Europe are given in Table 2.11 [91, 107]. The Värnamo plant in Sweden is the only currently operating integration of a gas turbine with a gasifier. This process operates on a wood waste or wood chip fuel that is dried to 10-20% before gasification in a rotary dryer using hot gases generated by burning approximately 10% of the incoming feed. The reactor is a pressurised circulating fluidised bed whereby a cyclone is used to recycle the solids entrained in the fuel gas, including ash, char, sand and a dolomite or limestone catalyst. The catalyst is used to enhance cracking of tars in the reactor, and tars are also cracked thermally by virtue of the high reactor temperature of 950-1000°C. The fuel gases are cooled after the cyclone to approximately 350°C before filtration in a hot gas filter (see Section 2.6.6.3) to clean the gas to the required standards imposed by the gas turbine. The gas turbine is an aero-derivative EGT Typhoon (see Section 2.6.6.1) which is used in a combined cycle (see Section 2.6.6.4) to generate a total of 6 MW_e. The gas turbine combustion chambers have been specially designed for the low heating value gas. There is currently no flue gas treatment for NO_x but space is available at the gas turbine flue gas train for selective catalytic reduction is required.

Table 2.11 - Details of current European gasification projects for electricity

	Värnamo	ARBRE	Biocycle	Energy farm
Location	Sweden	UK	Denmark	Italy
Status	Operational	Design	Design	Design
Output	6 MW_e 9 MW_e	8 MW_e	7.2 MW_e 6.8 MW_{th}	11.9 MW_e
Feedstock	Waste wood	SRF wood	SRF wood	SRF wood
Gasifier	Pressurised CFB (Bioflow)	Atmospheric CFB (TPS)	Pressurised fluid bed (RENUGAS)	Atmospheric CFB (Lurgi)
Tar removal	Catalytic dolomite cracking	Catalytic dolomite cracking	Thermal cracking and dolomite	Water scrubbing
Gas cleaning	Hot gas filter	Water scrubbing	Hot gas filter	Bag filter
Gas turbine	EGT Typhoon	EGT Typhoon	EGT Typhoon	EGT Typhoon
Cycle	Combined cycle	Combined cycle	Combined cycle	Combined cycle
Electrical efficiency	Not known	30.6%	39.8%	33%
Plantation area required	Not applicable	2800 ha	1325 ha	3680 ha
SRF Productivity	Not known	12	9	10-15
Electricity price	Not known	13.68 ¢/kWh	8.46 ¢/kWh	14.92 ¢/kWh
Heat price	Not known	Not applicable	2,92 ¢/kWh	Not applicable

The other three systems shown in Table 2.11 are part-funded by the EC THERMIE Demonstration programme [107]. The UK venture is particularly relevant to this work. This project is aided by a Non-Fossil Fuels Obligation (NFFO) contract that guarantees a price for electricity generated by biomass gasification that is higher than the normal pool price for the first 15 years' operation. This is funded by a levy on all UK electricity sales which is used to finance the generation of electricity from non-fossil fuel sources. The majority of the revenue raised is used to subsidise nuclear power, with the remainder used to promote the development of renewable energy. Contracts are awarded by competitive tender after the announcement of NFFO orders that specify the target amounts of new generating capacity required in each renewable sector. Bids for contracts must include the price that the developer would require for the electricity generated and the price for each sector is set after evaluating the bids. The aim of the scheme is to provide the extra revenue required to support initial projects so that learning effects (see Section 4.7) can bring prices down with experience to levels where a subsidy is not required to compete with conventional generating technologies. The contract for project ARBRE was granted under the 3rd NFFO order and set a price for the electricity of 13.7 ¢/kWh (8.65 p/kWh). This figure will be used later in Chapter 9 (Results) for comparison with the production costs that are calculated in the model.

2.6.6 Integrating gasifier and gas turbines - issues

There are several issues that must be resolved in the integration of gas turbines with biomass gasification, notably:

1. The firing of the gas turbine on low heating value gas;
2. The most suitable gasifier operating pressure;
3. Fuel gas cleaning and cooling; and
4. The selection of the gas turbine cycle.

2.6.6.1 Selection of the gas turbine

There are two types of gas turbine: industrial and aero-derivative, compared in Table 2.12. The high efficiency of aero-derivative gas turbines has tended to favour their development over industrial turbine-based systems. This is surprising since industrial gas turbines are proven in low heating value gas applications such as the firing of blast furnace gases [108]. The compact combustion chambers used with aero-derivative gas are not ideal for low heating value gases and the gas turbine may need to be re-designed to account for [109, 110, 111, 112]:

- difficulties in achieving complete combustion while maintaining low NO_x emissions;
- difficulties in maintaining combustion stability;
- the increased pressure drop across the fuel injection system; and
- the increased mass flow in the gas turbine.

Table 2.12 - Comparison of Aero-Derivative and Industrial Gas Turbines

Aero-derivative	Industrial
High efficiency in single cycle	Lower efficiency in single cycle
Low specific size	Large size, featuring large, external combustion chambers
Parts more sensitive to damage	High reliability
Can be replaced if repairs are required	Repaired in-situ
Capacities from 0.3-42 MW_e	Capacities from 0.2-200MW_e

2.6.6.2 Gasifier pressure

If the gasifier is considered in isolation, atmospheric gasification is generally preferred over pressurised gasification because reactor feeding is less complex and all of the safety issues associated with pressurised systems are avoided. Blackadder [134] reports that the more compact pressurised system is only cost effective at fuel inputs above 150-200 MW_{th} if the gasifier is considered in isolation. The integration of a gasifier with a gas turbine makes selection of the gasification pressure more complicated, due to the issues shown in Table 2.13. Blackadder has estimated that the difference in efficiency between the two options may only be 1-2 percentage points although other studies have calculated differences of up to 6 percentage points at around 30 MW_e [113]. System developers have been unable to agree on the optimum system, although there is general support for the pressurised gasification route at higher system capacities (above 50 MW_e) [134].

Table 2.13 - Comparison of atmospheric and pressurised gasification for gas turbines

Atmospheric Gasification	Pressurised Gasification
Easier reactor feeding	Complicated reactor feeding. High operating costs due to pressurisation with inert gas.
Low-cost reactor	High-cost reactor
Fuel gas must be compressed: • compressor cost • compressor power consumption • gas must be cooled to near ambient • sensible heat lost in cooling	Fuel gas already compressed: • only booster compressor required • cooling only to gas turbine injector limit (400-500°C) • sensible heat retained
Gasifier air flow is independent of the gas turbine	Gasifier air flow is bled from the gas turbine compressor, with potential control problems

2.6.6.3 Fuel gas treatment

Gas turbines are highly sensitive to fuel gas quality, and since the fuel gas is expected to be contaminated by the impurities presented in Table 2.8 it must be treated to remove the contaminants. Typical gas quality requirements are shown in Table 2.14 [114]. Fuel gas purity is particularly important in low heating value gases since more fuel (and therefore contaminants) are injected into the combustion chamber per unit of power output. It has also been noted in Table 2.7 that the gas must be cooled after leaving the gasifier to ambient temperatures if gas compression is required or to the gas turbine fuel valve temperature limit (around 450degC) if the gas is already pressurised. Two basic gas treatment methods have been proposed [114, 115]:

- hot gas filtration, where the gases are partially cooled to around 500°C to condense alkali metal vapours onto particulate in the gas. Gas cooling is followed by a hot gas filter that removes both the particulate and the condensed alkali metals. The gas is delivered to the gas turbine at relatively high temperatures of around 450°C that allow tars in the gas to be retained as vapours.
- wet gas scrubbing, where the gases are cooled to under 150°C and then passed through a wet gas scrubber. This removes particulate, alkali metals, tars and soluble nitrogen compounds such as ammonia.

Wet gas scrubbers are considered to be the more established gas cleaning technology although there is little experience of their application with biomass gasification gases. Hot gas filters are currently the subject of a great deal of research and development activity [116, 117], and are perceived to be the better solution if their technical problems can be overcome because the tars and sensible heat in the product gas are retained and the effluent stream that would be produced in wet gas scrubbing is avoided. Wet gas scrubbing is more suitable to atmospheric gasifiers where the gas must be cooled anyway before compression to the gas turbine combustion chamber pressure.

If wet scrubbing is used then the removal of tars will reduce the heating value of the gas. In this situation then thermal or catalytic cracking of the tars [118] can be adopted to produce non-condensible hydrocarbon gases from the tars and so retain the chemical energy of the fuel gas.

Gas cleaning techniques are considered further during the modelling of the gasification conversion steps (Section 7.4 and 7.5).

Table 2.14 - Notional Gas Turbine Fuel Gas Specification

Contaminant	Tolerance
Minimum gas heating value (LHV)	4-6 MJ/Nm3
Minimum gas hydrogen content	10-20%
Maximum alkali concentration	20-1000 ppb
Maximum delivery temperature (set by fuel valve materials)	450-600°C
Tars at delivery temperature	All in vapour form or none
Maximum particulate (ash, char etc) level at turbine inlet	Concentration, ppm_w
Particle size, µ	0.1
>20	1.0
10-20	10.0
4-10	
NH_3	No limit
HCl	0.5 ppm
S (H_2S+SO_2 etc.)	1 ppm
N_2	No limit
Combinations	
Total metals	< 1 ppm
Alkali metals + sulphur	< 0.1 ppm

2.6.6.4 Gas turbine cycle

Combustion in a gas turbine typically converts less than 40% of the available energy in the fuel to electricity, most of the remaining energy is lost in the hot gas turbine exhaust gases. The gas turbine may be used in a simple cycle without recovering the heat from the exhaust gases. Such cycles may be viable in peak shaving applications where operation is infrequent and low investment costs are paramount. Base load applications often include heat recovery where the efficiency gains and consequent lower fuel costs can justify the increased capital cost. Two cycle options are:

1. The combined cycle; and

2. The steam injected cycle.

Both cycles raise steam by passing the gas turbine exhaust gases through a heat recovery steam generator. In the combined cycle this steam is expanded in a steam turbine in a bottoming steam cycle; extra power is produced in the steam turbine with only minor reductions in gas turbine power output due to higher temperature exhausts and an increased back pressure. In the steam injected gas turbine (STIG) cycle the steam is injected with the fuel at the combustion chamber. This has two effects: mass flow through the turbine increases with consequently more power output and NO_x emissions may be reduced by controlling flame temperatures in the combustor. STIG is now established in conventional fuels applications [119]. The relative efficiencies of simple cycles, combined cycles and STIG cycles are shown in Figure 2.5 [120, 121, 122, 123, 124].

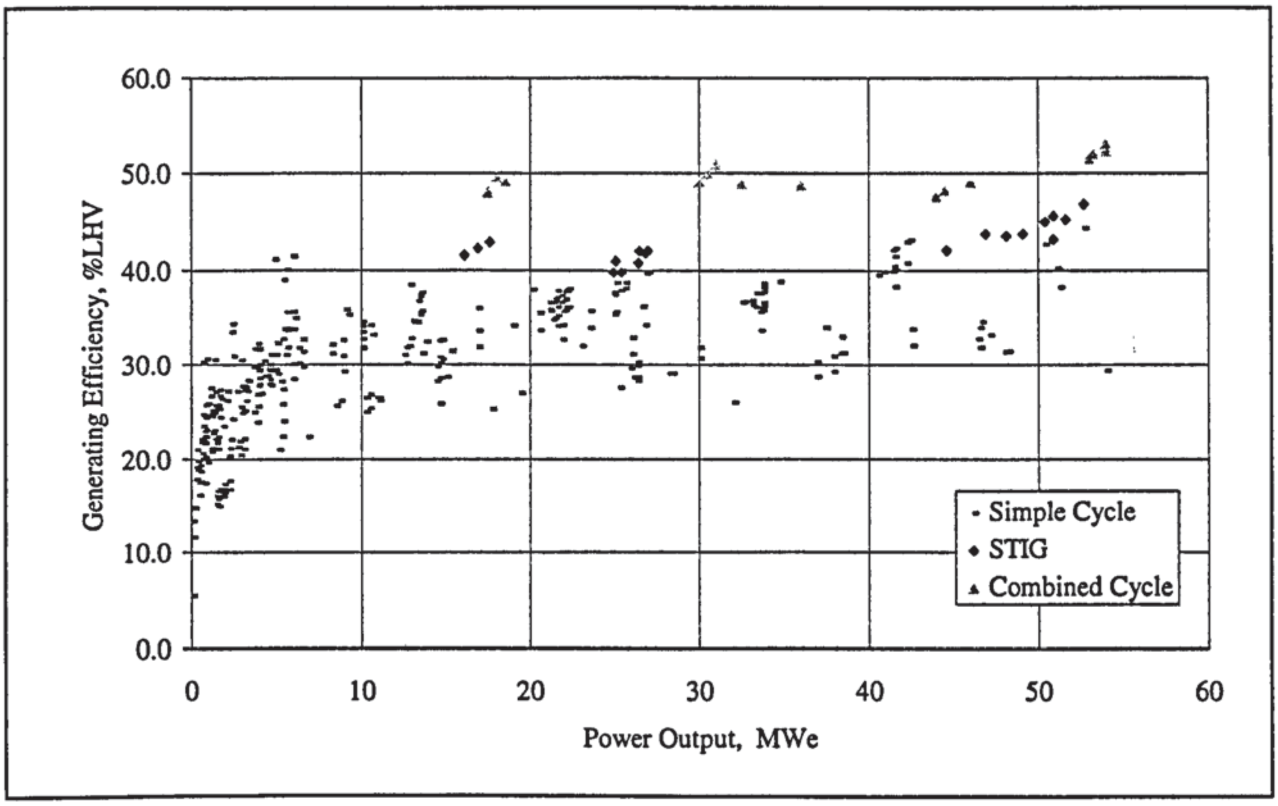


Figure 2.5 - Gas turbine combined cycle efficiencies, natural gas fired

All of current large scale gasification and gas turbine cycles are based on combined cycles and on this basis only gas turbine combined cycles will be considered in this work. Although STIG has been promoted as a suitable option for biomass-based cycles [125] some doubts are now emerging about the suitability of STIG to biomass gasification applications. Studies have shown that steam injection is technically feasible with such fuels but that the amount of steam injection possible is limited, given the extra mass flows imposed by using a low heating value gas [111]. The system efficiency in STIG is lower than combined cycles, so feed costs are increased. Studies by EPRI have shown that overall operating costs are higher due to factors such as a need for large quantities of demineralised feed water [41], even though the high costs of the scale-sensitive rankine cycle are eliminated.

2.6.7 Integrating gasifiers and internal combustion engines

The operation of diesel engines using a variety of low heating value gases is an established practice [33, 126, 127, 128, 129, 130]. Internal combustion engines may be compression-ignited (diesel cycle) or spark-ignited (Otto cycle). Diesel engines are available in larger sizes than spark-ignited engines and are generally favoured above 1 MW_e. They can be adapted for use with many liquid and gaseous fuels. They may also be run in dual-fuel mode, where a diesel pilot fuel is used to ignite a less conventional fuel. Such engines can switch between dual fuel and diesel-only operation without interrupting generation, a significant benefit in unconventional systems where the supply of a biomass-derived fuel may not be reliable. Diesel engines therefore offer robustness; easier maintenance; multi-fuel capability; and ability to use poor quality fuels [131, 132].

In contrast, spark-ignited engines tend to be smaller, high speed engines. They may only be used with high quality liquid fuels, although they can be adapted for use with low heating value gases. They are more sensitive to fuel contaminants than the diesel engines. Spark-ignited engines are not as tolerant of changes in fuel quality as diesel engines and more sensitive to fuel contaminants [93]. They cannot switch from one fuel to another easily: as such a spark-ignited engine would be totally dependent on the feed conversion technology, a factor that reduces generating reliability. Given these disadvantages, spark ignited engines will not be considered further in this study.

Dual fuel diesel engines for operation using low heating value gases may be regarded as fully developed [93], although integration of a biomass gasifier and engine is not established.

Since the gas is required at near atmospheric pressures in engine applications there is no advantage to pressurised gasification. The main issue that must be resolved is treatment of the fuel gas to cool and clean it to the specifications demanded by the engine [133]. The fuel gas must be cool at injection to the engine and therefore wet scrubbing is the preferred gas treatment method.

Diesel engine efficiencies are significantly higher than both steam cycles and gas turbine cycles at small scale (see Chapter 8). Their capital costs are also relatively insensitive to scale when compared with steam cycles and gas turbine cycles. As a result they are a low-cost, high efficiency option in small scale systems. Blackadder [134] recommends the gasifier and diesel engine option for capacities of up to 5 MW_e. At higher capacities, the combination of scale economies and increasing cycle efficiencies in the alternative cycles can erode the advantages that diesel engines enjoy at the lower scale. One of the aims of this work will be to examine the upper limit where engine generators should be used.

2.6.8 Summary

Gasification technologies are commercial and there is substantial information available to allow their costs and performance to be modelled. The only gasifiers that have been developed to the capacities required for operation of a 1-100 MW_e power plant are fluid bed or circulating fluid bed designs. This work will only consider these reactor configurations. The fluid bed offers high fuel flexibility and it is assumed that the reactor will be capable of processing wood chips.

The integration of a gas turbine with a prime mover for electricity generation is less established. The main issue to be resolved is the clean-up of the fuel gas to the specifications demanded by the prime mover. The integration of gasifiers and gas turbine is the subject of a great deal of research and development and should be evaluated further in this work on the grounds of its status and applicability to the range of capacities required.

There is wide experience of dual fuel engines using low heating value gases but more limited experience of commercial electricity generation using gasifiers and engines. Dual fuel engines are more tolerant of fuel gas contamination than gas turbines and as such they are a near-term opportunity for electricity generation since the problems of gas clean-up are to some extent alleviated. The integration of gasifiers and dual fuel engines will be considered further in this work.

The gas turbine option could be applied to both atmospheric and pressurised gasification. Given the limited time available for this study, only one of these options could be selected for further evaluation. The pressurised gasification system was chosen on the basis that it provided the most interesting alternative to the other systems: it is expected to be a high efficiency, high capital cost system in contrast to the engine-based technology and the steam cycle.

2.7 PYROLYSIS SYSTEMS

2.7.1 Introduction

Pyrolysis is an ancient process first developed over 10000 years ago for the production of charcoal and tars [135]. Wood pyrolysis was a major source of chemicals prior to the advent of the petrochemical industry in the first half of this century [136] although the development of pyrolysis for fuels production began later, prompted by the energy crises of the 1970s [137]. Various pyrolysis processes have been investigated, with each using different operating parameters to produce a particular mix of solid, liquid and gaseous products (see Table 2.15) [138]. Of the processes shown in Table 2.15, fast pyrolysis for liquids is currently attracting the most support because it can produce high yields of a liquid product with a diverse range of potential applications as shown in Section 2.7.5 [139, 140, 141, 142].

Table 2.15 - Comparison of Pyrolysis Processes

Process	Carbonisation	Slow Pyrolysis	Fast Pyrolysis for Liquids	Fast Pyrolysis for Gases
Temperature, °C	300-500	400-600	450-600	700-900
Pressure, bar	1	0.1-1	1	1
Gaseous Product				
Yield, %dry feed	Up to 150	Up to 60	Up to 30	Up to 80
LHV, MJ/Nm^3	2.5-5.5	4.5-9	9-18	13.5-18
Liquid Product				
Yield, %dry feed	Up to 25	Up to 30	Up to 70	Up to 20
LHV, MJ/kg (dry product)	18	18	22	20
Solid Product				
Yield, %dry feed	Up to 40	Up to 30	Up to 15	Up to 15
LHV, MJ/kg	29	29	29	29

Pyrolysis is the thermal degradation of biomass in the absence of an oxidising agent whereby the volatile components of a solid carbonaceous feedstock are vaporised in primary reactions by heating, leaving a residue consisting of char and ash. Pyrolysis vapours and gases from the primary reactions pass out of the reactor and a liquids recovery system condenses the condensible fraction, leaving non-condensable gases. Before the vapours are condensed they are subjected to secondary reactions that crack the vapours to produce a higher proportion of gas and char. These secondary reactions may be reduced by limiting the residence time of the gas/vapour mixture, its temperature and any contact with char (which acts as a catalyst).

2.7.2 Reactor configurations

Fast pyrolysis research and development is active at a number of institutions using a variety of reactor configurations, as listed in Table 2.16 [143]. Each system has been designed to meet the requirements of a fast pyrolysis reactor: rapid heating of the feedstock to moderate temperatures, prompt quenching of the pyrolysis vapours to minimise secondary reactions and the removal of char as it forms. Actively researched reactor configurations fall into three main categories: fluid bed pyrolysis and variations thereof, ablative pyrolysis and vacuum pyrolysis. These processes are reviewed fully in work by Bridgwater and Bridge [138, 144].

Table 2.16 - Reactor Configurations for Fast Pyrolysis

Reactor Type	Organisation
Ablative plate	U. Aston
Ablative vortex	NREL
Other ablative	BBC/Castle Capital
Circulating fluid bed	CRES, ENEL/Pasquali
Entrained flow	Egemin
Fluid bed	CPERI, IWC, NREL, Union Fenosa, U. Aston, U. Hamburg, U. Leeds, U. Sassari, U. Stuttgart, U. Waterloo
Horizontal vacuum moving bed	Pyrovac/U. Laval
Rotating cone (transported bed)	U. Twente, Schelde/BTG
Transport bed with solids recirculation	Ensyn

Fluid bed pyrolysis reactors use the excellent heat transfer characteristics of fluid beds to rapidly heat the feedstock to approximately 500°C mainly by conduction from the heated bed material. Heating the bed is achieved by either heating the reactor in bubbling bed designs or by heating the bed material in an external reactor if circulating fluid beds or transported beds are used. The thermal energy would usually be provided by combustion of the char and off-gas by-products. In all reactors the requirement for rapid heating demands rapid heat fluxes and this has proved to be a major design challenge.

In ablative pyrolysis the feed particles are heated by sliding them at high contact pressures against a heated surface. This has the advantage of continually exposing fresh biomass for pyrolysis as the primary char is abraded off and the pyrolysis gas and vapours are carried away. Since exposure to primary chars is reduced, catalysis of secondary reactions is also reduced. Char is not abraded from reacting particles in the fluid bed to the same extent (there is some slight abrasion due to the turbulent bed material) and therefore primary vapour products will immediately be exposed to char in fluid bed designs. The result of this is that ablative reactors can tolerate larger particle sizes than fluid beds. The former type can process chips [145]; the latter type is limited to particle sizes of less than 2 mm if the process is to maintain high liquids yields [146, 147, 148].

In vacuum pyrolysis the heat is supplied by molten salts flowing through tubes that are welded to form a plate [149]. Feedstock passes over the plates in an agitated moving bed and very high heat fluxes are achieved. The vacuum conditions encourage vaporisation with low entrainment of solids. The molten salts are heated using the pyrolysis off-gas, with an induction heater providing reserve thermal energy when required.

2.7.3 Products

There are three products of pyrolysis: char, a liquid and a medium heating value gas. The char may be sold or used internally to provide heat for the process. The gas has a medium heating value and could be exported but it is usually more economic to use it internally [150]. The organic liquid is a homogenous mixture of organic compounds and water in a single phase and is the product of most interest for fuels applications (see Section 2.7.5). Its main fuel properties are presented in Table 2.17 [151, 152]. The relative yields of each product is a function of a variety of process parameters that are described in Chapter 7.

2.7.4 Status of Fast Pyrolysis

Fast pyrolysis has been in development in North America since 1980 and experience in Europe has accumulated over the last 8 years, with a wide variety of reactor configurations at the experimental stage and two demonstration plants in operation [143]. Small-scale research is well-supported and the main issues that remain unresolved in fast pyrolysis for fuels are product quality, system efficiency and technology scale-up.

Only fluid bed configurations have been scaled up to commercial capacities. Ensyn have the most developed system, based on a transported bed. The company has been commercial since 1989 and has three operational plants with several at the design or construction stage [153, 154, 155]. The largest plant is at Red Arrow Products in Wisconsin, producing a liquid for chemicals extraction with the residue used to fire a 6 MW_{th} input boiler. The system has a feed capacity of 25odt/d, operating 5 days per week, 24h/d (operation is expected to reach 6 days per week in 1995) [156]. A 650 kg/h plant has recently been commissioned in Italy for ENEL for the production of liquid fuels [157].

Considerable development work on small-scale fluid bed pyrolysis has been carried out at the University of Waterloo in Canada [146, 158], and a demonstration plant based on this technology is operated by Union Fenosa in Spain that has achieved a feed capacity of 160 kg/h (dry feed basis) [159, 160].

Ablative pyrolysis has been developed at the bench scale by the Aston University (UK) [145, 161] and at pilot scale by the National Renewable Energy Laboratory (NREL, US) [162, 163, 164], but has not been demonstrated at large scale. One attempt to develop the NREL system to 32.7 odt/d feed input by Interchem [165] failed due to lack of funds for full development.

A single commercial vacuum pyrolysis plant has been built in Belgium for the disposal of waste tyres [149]. The process is expensive and produces low liquids yields. The main interest in vacuum pyrolysis is in waste disposal and the extraction of chemicals.

Table 2.17 - Comparison of pyrolysis liquid and conventional fuel oil characteristics

		Pyrolysis liquid	Diesel	Heavy fuel oil
Density	kg/m3 at 15degC	1220	854	963
Composition, as produced	%C	48.5	86.3	86.1
	%H	6.4	12.8	11.8
	%O	42.5	-	-
	%S	-	0.9	2.1
Viscosity	cSt at 50degC	13	2.5	351
Flash point	°C	566	70	100
Pour point	°C	-27	-20	21
Conradson carbon residue	%wt	17.8	<.1	9
Ash	%wt	0.13	<.01	0.03
Vanadium	ppm	0.5	<1.	100
Sodium	ppm	38	<1.	20
Calcium	ppm	100	<1.	1
Potassium	ppm	330	<1.	1
Chloride	ppm	80	<1.	3
Sulphur	%wt	0	0.15	2.5
Water	%wt	20.5	0.1	0.1
LHV	MJ/kg	17.5	42.9	40.7
Acidity	pH	3	-	-

This work will only consider fluid bed fast pyrolysis since it is the only configuration with sufficient published data available for modelling and its scale-up has been demonstrated. However, the extra tolerance of the ablative system could give it a commercial advantage because the need to grind a chipped feedstock is eliminated. Since no commercial ablative processes exist the relative costs of fluid bed pyrolysis and ablative pyrolysis systems cannot be examined at this time but it is recommended that the two systems are compared in future work.

2.7.5 Fast pyrolysis applications

There are a wealth of potential opportunities for fast pyrolysis in heat, chemicals, fuels and electricity production, as indicated by Figure 2.6. This discussion will be confined to the various routes to electricity generation. Fast pyrolysis is one of the technologies supported under the latest NFFO order (NFFO-4), which is an indication that it is near-term commercial opportunity for electricity generation [166].

The extraction of chemicals followed by use of the liquid residues for fuels has been demonstrated commercially by the Red Arrow plant in Wisconsin that produces food flavourings and boiler fuel [156]. A system that produces chemicals and electricity could not be modelled unless the chemicals to be extracted and their market opportunities are known, and this information is generally proprietary. Therefore this option will not be evaluated but the economic potential of the option could merit future investigation on a case study basis. Generation of electricity via upgrading processes such as zeolite catalysis and hydrotreating are also unlikely to be viable due to the high cost of upgrading [139]. Pyrolysis liquids are used commercially as a boiler fuel but this is unlikely to be economic for dedicated electricity generating systems unless feed costs are negligible because of the low efficiency and high capital cost of the steam cycle [167]. Most development for electricity generation is focused on the use of raw pyrolysis liquids in gas turbine or diesel engine applications, described below.

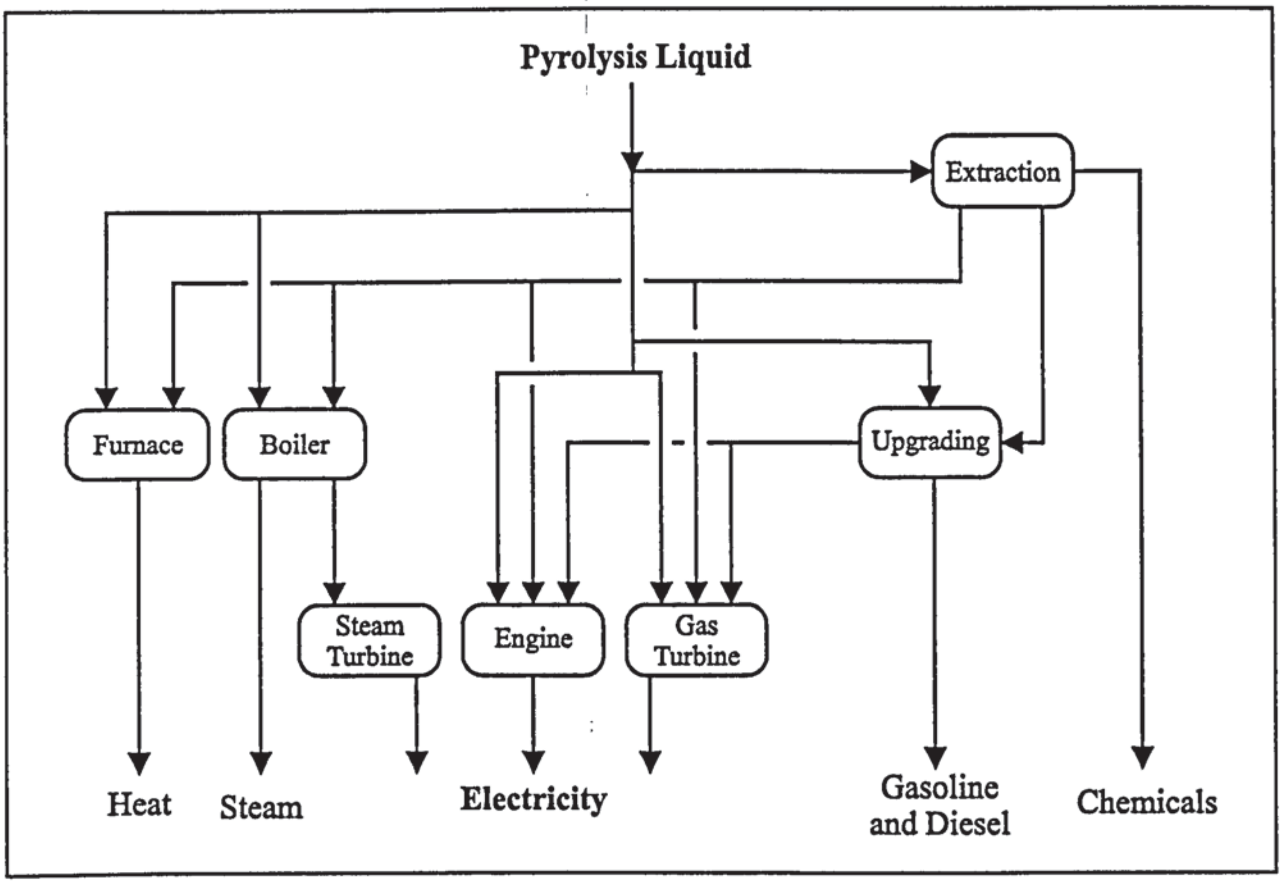


Figure 2.6 - Fast Pyrolysis Applications

2.7.5.1 Integration with Internal Combustion Engines

Diesel engine firing by pyrolysis liquids has been successfully carried out in limited tests. VTT in Finland have carried out engine tests using a small (4.8kW) single cylinder Petter test diesel engine [168, 169]. Larger scale development of diesel engine systems is ongoing through the work of Ormrod Diesels in the UK [170] and Wärtsilä Diesels in Finland [152]. Pyrolysis liquids are very different to conventional diesel fuels but it has been shown that crude pyrolysis oil burns well but is difficult to ignite [168]. Pilot-ignition engines, which use a small amount of an auxiliary fuel to ignite the main fuel are expected to solve this problem. Table 2.18 introduces the fuel characteristics that are likely to be important when integrating fast pyrolysis with a diesel engine [143]. All of the problems are expected to be soluble and diesel engine applications are the most likely opportunity for fast pyrolysis systems in the near term. Therefore this application will be evaluated further here.

2.7.5.2 Integration with Gas Turbines

Minimal development work has been carried out on the use of gas turbines with fast pyrolysis. Early tests on a combustor rig designed to simulate a slurry-fed gas turbine showed the potential for pyrolysis liquid combustion in a gas turbine [171]. The tests showed that entrained char in the bio-oil tends to block fuel injection systems; that ash fouling occurred downstream of the gas turbine; that bio-oil cause corrosion to turbine components; and that smoke emissions increased. More recent tests at Orenda in Canada [172] are evaluating the firing of a 2.5 MW_e industrial gas turbine. The turbine has been run successfully for 1 hour (as reported in May 1996) on 100% pyrolysis liquids [173] while flame tunnel tests are examining the long-term resistance of turbine parts to corrosive attack from alkali metals in the ashes entrained in pyrolysis liquids.

There is generally less experience of gas turbines using fast pyrolysis liquids and this option will not be evaluated further at this time. It is expected that gas turbine applications will be developed in the longer term and future work should examine this option as experience grows.

2.7.6 De-coupling

De-coupling is the separation in time or space of the conversion and generation stages of the biomass to electricity system. De-coupling is only available for fast pyrolysis systems where it is viable to store and transport the intermediate energy carrier since it is a liquid. Conversely the steam produced in a combustion system must be used immediately in the steam turbine and low heating fuel gas cannot be stored or transported for long distances economically. Combustion and gasification systems must therefore be used in close-coupled configurations where the conversion and generating stages occur concurrently and at the same site.

De-coupling offers several potential system configurations, with the four main options shown in Figure 2.7. In each case there is an interaction between transport costs and capital costs that could result in a lower production cost for the electricity. Since these four options are not available in combustion or gasification based-systems, de-coupled fast pyrolysis systems may be more cost-effective than the alternative technologies in particular circumstances. This work will evaluate fast pyrolysis in the four configurations shown below to highlight where de-coupling could be used to advantage.

Table 2.18 - System requirements for fast pyrolysis liquids

Characteristic and Effects	Solution
<u>Suspended char</u>	
• Erosion, • Equipment blockage, • Combustion problems from slower rates of combustion. • Deposits and high CO emissions.	• Hot vapour filtration; • Liquid filtration; • Modification of the char for example by size reduction so that its effect is reduced; • Modification of the application.
<u>Alkali metals</u>	
• Solids deposition in combustion applications including boilers, engines and turbines. • Corrosion of turbine blades in high performance gas turbines.	• Hot vapour filtration; • Processing or upgrading of oil; • Modification of application; • Pretreat feedstock to remove ash
<u>Low pH</u>	
• Corrosion of vessels and pipework.	• Careful materials selection; • Stainless steel and some olefin polymers are acceptable
<u>Incompatibility with polymers</u>	
• Swelling or destruction of sealing rings and gaskets.	• Careful materials selection.
<u>Temperature sensitivity</u>	
• Liquid decomposition on hot surfaces leading to decomposition and blockage; • Adhesion of droplets on surfaces below 400°C.	• Recognition of problem and appropriate cooling facilities; • Avoid contact with hot surfaces > 500°C.
<u>High viscosity</u>	
• High pressure drops in pipelines leading to higher cost equipment and/or possibilities of leakage or even pipe rupture.	• Careful low temperature heating, • and/or addition of water, • and/or addition of co-solvents such as methanol or ethanol.
<u>Water content</u>	
• Complex effect on viscosity, heating value, density, stability, pH, homogeneity etc.	• Recognition of problem; • Optimisation with respect to application.
<u>In-homogeneity</u>	
• Layering or partial separation of phases; • Filtration problems.	• Modify or change process or parameters; • Change feedstock to low lignin; • Additives; • Control water content.

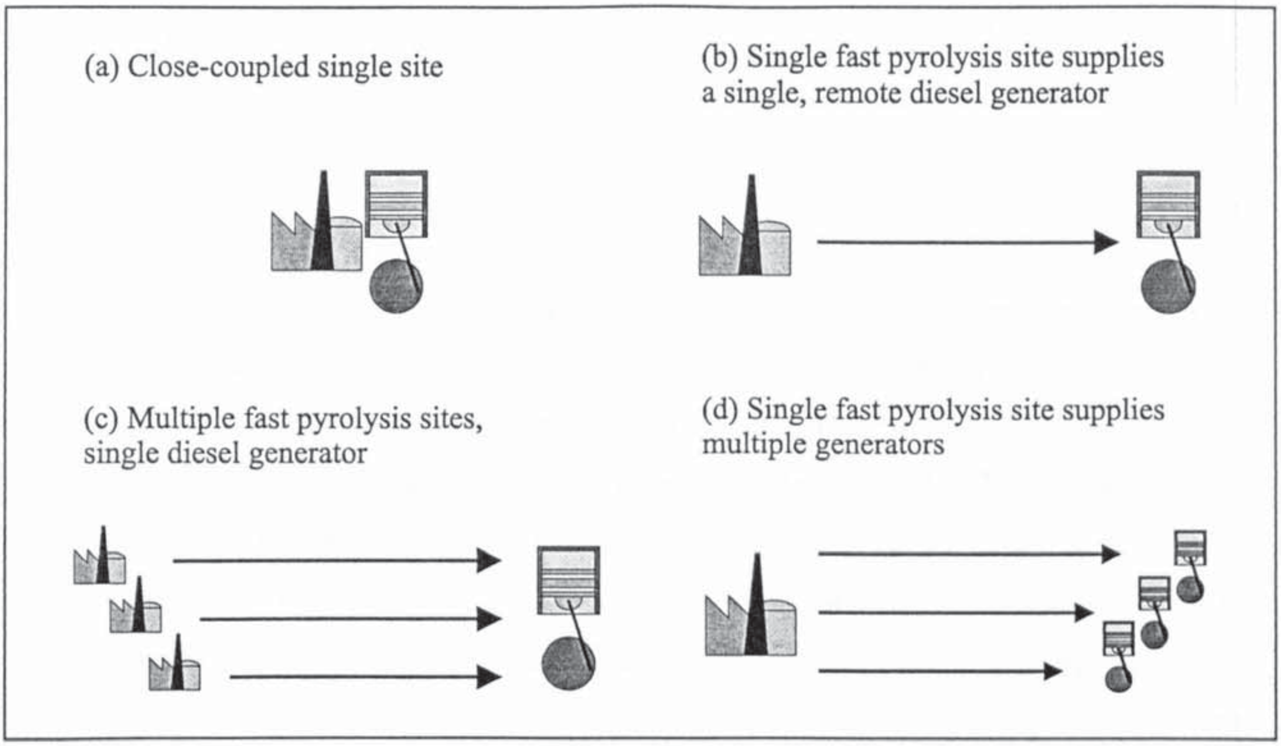


Figure 2.7 - System de-coupling options for fast pyrolysis systems

2.7.7 Summary and selection of systems

Fast pyrolysis is the least developed of the thermochemical conversion technologies. It has now achieved commercial status but information regarding the costs and performance of large scale reactors is scarce. There is, however, a lot of performance data for bench and pilot scale reactors, especially in fluid bed configurations.

The use of fast pyrolysis liquids has been demonstrated at the small scale and there are active projects in the UK and Sweden that are firing large diesel engines in dual fuel mode using fast pyrolysis liquids. This application of fast pyrolysis will be examined further. Gas turbine applications are not as developed and are dismissed at this stage.

Fast pyrolysis allows the de-coupling of feed conversion and electricity generation. This could be an advantage and it would be useful to examine the options that de-coupling presents in this work.

2.8 THE INTERFACE WITH THE ENVIRONMENT

Biomass may be regarded on a global scale as a renewable, environmentally beneficial source of renewable energy. On a more local level, the use of biomass for power production will produce solid residues, liquid effluents and gaseous emissions just as any other industrial process. There are also health and safety issues to be addressed that may affect the performance of the plant and the well-being of those who work on it. As the widespread use of biomass for electricity becomes a more likely near-term prospect, the issue of environmental impacts is increasingly debated [13].

The original brief for this work had included an analysis of the environmental impacts and health issues involved. To this end, a methodology was developed for the initial evaluation of a bio-energy system that was both rigorous and yet simple. This methodology, the Bioenergy Environmental Evaluation Scheme (BEES) is attached as Appendix B. The methodology was applied to the four systems selected for evaluation in this chapter (summarised in Section 2.9) but no satisfactory way could be found to integrate the environmental analysis with the technical and economic evaluations that follow in this thesis. For this reason the work is not included here but the analysis, results and conclusions of the environmental evaluations may be found in Appendix B.

2.9 SUMMARY

Four biomass to electricity systems have been selected for further evaluation. All systems will be fed by a wood chip feedstock. The transport of the feedstock will be included in the process evaluations since this could have an important impact on the feed cost at delivery to the conversion facility in large scale systems. Pretreatment of the feedstock at the conversion facility prior to its use should also be included in the evaluations.

Four combinations of feed conversion and electricity generation will be examined:

1. Combustion and steam cycle;
2. Atmospheric gasification and gas-fired dual fuel engine;
3. Pressurised gasification and gas turbine combined cycle; and
4. Fast pyrolysis and liquid-fired dual fuel engine.

The reactors will all be based on either fluid bed or circulating fluid bed configurations.

The combustion and steam cycle is the only established system evaluated. It will therefore be used as a reference case.

The fast pyrolysis system may be de-coupled, in that the feed conversion stage can be separated from the electricity stage. None of the other systems have this option, and must be close-coupled. The extra flexibility that this offers the fast pyrolysis system will be evaluated in this work.

3. PREVIOUS SYSTEMS STUDIES

3.1 INTRODUCTION

This chapter examines previous techno-economic studies of bioenergy systems. It serves two purposes:

- it provides comparable results that may be used for reference in the results presented in Chapter 9; and
- it shows the novel aspects of the current work by placing it in the context of existing studies.

Previous system studies vary widely, not only in the core technologies used but in the boundaries of the systems that are modelled and the financial scope of the capital and operating costs that are used. This variation between studies means that comparisons between their results can only be made with caution and if the methodology of the study is known. Therefore the focus of this chapter is on system studies that have documented methodologies and detailed published results. These system studies will be examined in terms of the:

- system boundaries used;
- technologies that are studied;
- the range of capacities evaluated;
- the depth of the technical analysis; and
- the scope of the financial data used.

Key results will be presented for comparison later. A brief summary of other system studies where the methodologies and results are not as well documented is also given. The chapter concludes by highlighting the features of this work that will make it unique.

3.2 ASTON UNIVERSITY, UK

Aston University has been active in techno-economic analyses of biomass conversion systems for many years. The most recent work (1991-1994) examined fast pyrolysis and pyrolysis upgrading through the development of a dedicated techno-economic model, BLUNT (Biomass Liquefaction and Upgrading by Novel Technologies) [139, 150]. A previous model, AMBLE

(Aston Model of Biomass to Liquid Energy) was developed between 1988 and 1991 to examine processes for producing liquid fuels from biomass via gasification [174, 175].

In both cases the upstream limit to the system is the delivery of the feedstock to the conversion facility. In AMBLE the downstream limit to the system is the production of a liquid energy product. In BLUNT the downstream limit is the production of a liquid energy product or electricity generated using the pyrolysis liquid or the upgraded products. In reality the electricity generation models are very weak and to all intents and purposes the downstream limit of the BLUNT model should be regarded as the production of a liquid energy product. A summary of the key system criteria for both models is given in Table 3.1.

Both systems use a step model approach to build up the system costs. The system is split into a sequence of stages, with mass balances, energy balances, capital cost calculations and operating cost calculations performed for each step and summed to give overall system results. The system capacity in all cases is fixed by specifying a feed input that may be anywhere between the limits shown in Table 3.1. The capital costs have been calculated by multiplying the total costs of the major equipment items by a factor to give "battery limit costs"; a rigorous definition of battery limit costs is not given and it is assumed to be installed plant costs (see Section 4.5).

The core of the BLUNT model is a generic fast pyrolysis step model that calculates the mass and energy balance, utility requirements and economic performance for fast pyrolysis. The generic pyrolysis model has been adapted to simulate the performance of three specific processes (Ensyn, Fenosa and NREL processes) although capital costs for the generic system are used in all cases. This can be misleading since the NREL process is ablative pyrolysis and is likely have different capital and operating costs than the other technologies which are both fluid bed based (even these may differ in cost). In principle specific performance data should not be applied to generic costs and vice-versa. The model includes processes for the upgrading of pyrolysis liquids to produce gasoline or diesel oil (zeolite catalysis or hydrotreating respectively). A very basic electricity generation step is included, using diesel engines or gas turbines depending on scale.

Table 3.1 - System criteria evaluated by AMBLE and BLUNT

	AMBLE	BLUNT
Feedstocks	Wood Refuse/MSW Wheat Sugar beet	Wood Straw Sorghum bagasse
Products	Gasoline Diesel Methanol Fuel alcohol Ethanol Gasoline blending stock	Pyrolysis oil Crude hydrotreated oil Crude aromatics Diesel (refined hydrotreated oil) Gasoline (refined aromatics) Electricity
Processes	Gasification (4 options) Pyrolysis (2 options) Liquefaction (1 option) Fermentation (3 options)	Pyrolysis (generic and 3 specific processes) Hydrotreating Zeolite cracking (present and potential technology) Diesel engines or gas turbines
Capacity limits	100 - 2000 daf[a] t/d	25 - 1000 daf t/d
System scope includes:		
Feed production	No	No
Feed transport	No	No
Feed pretreatment	Yes	Yes
Feed conversion		
Combustion	No	No
Gasification	Yes	No
Pyrolysis	Yes - but poorly modelled	Yes
Electricity generation	No	Yes - but poorly modelled
Multiple sites/de-coupling?	No	No

a daf t = dry ash free tonne

Pretreatment has been included in both cases with step models for reception, storage, screening, comminution and drying of the various feedstocks. The pretreatment models in BLUNT are based to a large extent on those in AMBLE. The concept of using process steps in pretreatment is a useful one, although in this earlier work there is very little variation in the equipment used as scale increases. Particular problems with the BLUNT model are a very low power consumption for grinding of the feedstock and problems in fixing the drying capacity (based on the ratio of input and output moisture contents rather than the actual moisture removed).

The balance between technical and economic modelling in this work is very good, and while the detail on either side is not exhaustive, it is appropriate to the data that was available and the associated uncertainties. Uncertainties in the model were tested by sensitivity analyses (the variation of input parameters in a base case by a fixed percentage to see the variation in results). The sensitivity analyses revealed the overwhelming importance of the feedstock cost and system capacity to the system production costs. An example of results from the BLUNT model is given in Figure 3.1.

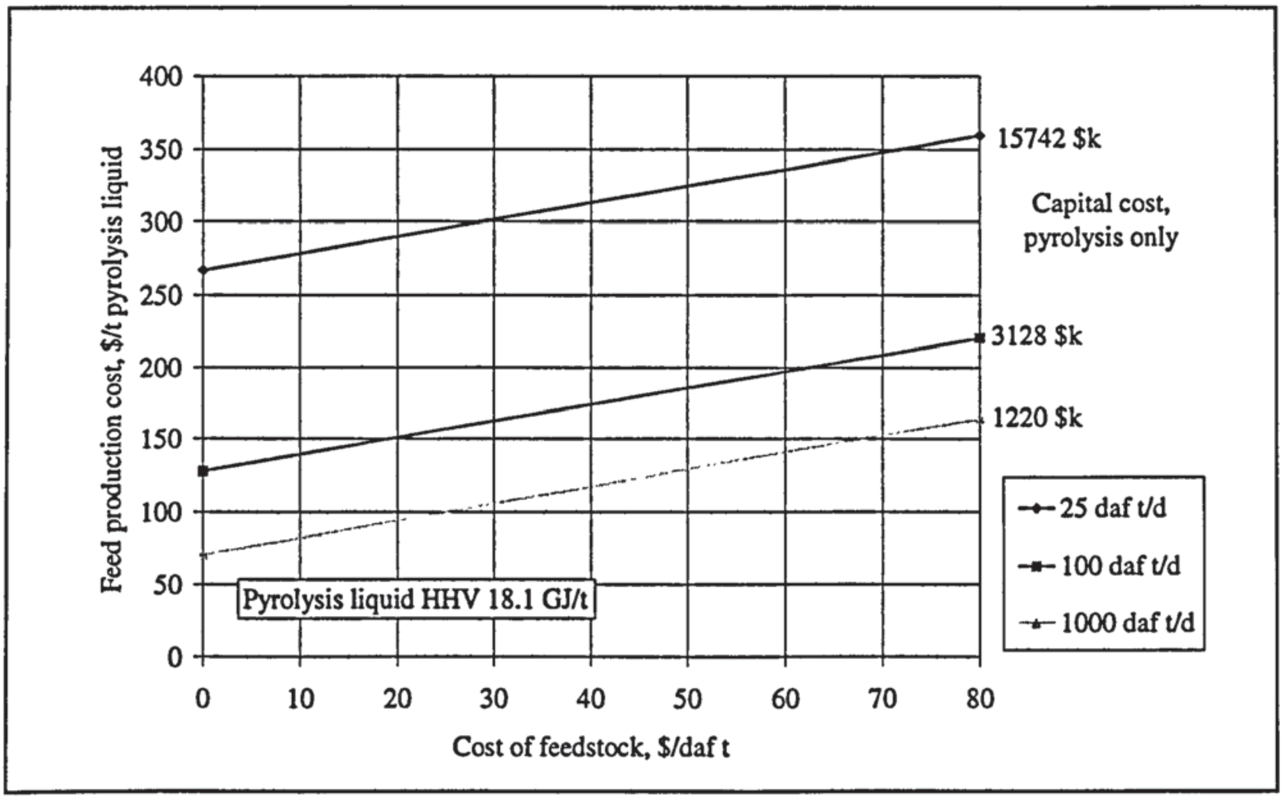


Figure 3.1 - Pyrolysis liquid production costs, BLUNT model

3.3 BTG, NETHERLANDS

The Biomass Technology Group is based in Enschede in the Netherlands and has been responsible for many reports comparing feed production techniques, pretreatment options and conversion systems to highlight potential biomass to energy systems for the Netherlands. A key focus of this work is the CO_2 that can be avoided using the various biomass to energy systems. This work is largely co-ordinated by NOVEM, the Netherlands' agency for energy and the environment. This agency has commissioned a number of reports that examine the costs and performance of various aspects of the biomass to electricity system under the National Programme on Global Air Pollution and Climate Change.

This work is worthy of note for the number of biomass to energy systems studied and the wide scope of each study, as shown in Table 3.2. The system boundaries are far reaching and encompass all aspects of the biomass to electricity system from the production of the feedstock to the delivery of the energy product to the consumer. Even with such a variety of processes, fast pyrolysis is not featured in any way. Summaries of this work are given by Van den Heuval [176, 177]. The overall results cannot be compared with any results in this work because the methodology for the study is not clear. However, a summary of the relative costs of several systems that were analysed is presented in

Table 3.3 [176]. The main conclusions of the work are that none of the technologies are viable under current conditions and that the thermochemical processing routes for heat and power provided a more effective means of CO_2 reduction that the liquid fuels production routes.

Table 3.2 - System criteria evaluated by BTG

Feedstocks	Wheat
	Straw
	Sugar beet
	Rapeseed
	Maize
	Poplar
	Miscanthus
Products	Electricity and/or heat
	Transport fuels (ethanol, rape methyl ester, methanol)
Processes	Combustion
	Combustion and steam cycle
	Gasification and gas turbine combined cycle
	Fermentation
	RME production
	Gasification and methanol synthesis
Capacity limits	5-50 MW_e, 500 MW_e in co-combustion
System scope includes:	
Feed production	Yes
Feed transport	Yes
Feed pretreatment	Yes
Feed conversion	
Combustion	Yes
Gasification	Yes
Pyrolysis	No
Electricity generation	Yes
Multiple sites/de-coupling?	No

Table 3.3 - Results from BTG analyses

System	Electricity cost, ¢/kWh
5 MW_e CHP by combustion	26
50 MW_e combustion	16.5
500 MW_e co-combustion of miscanthus (10%) and coal	15
5 MW_e gasification and gas turbine CHP	20
50 MW_e gasification and co-firing with coal	16
50 MW_e gasification and co-firing with natural gas	15

As well as the overall studies, several reports have been produced that focus on specific parts of the system [e.g.178, 179]. One particularly useful report reviews the costs and performance of biomass transport and pretreatment systems [180] and the data presented there has been used in Chapters 5 and 6 as these aspects of the system are modelled.

3.4 INTERNATIONAL ENERGY AGENCY

The International Energy Agency Bioenergy Agreement (now IEA Bioenergy) has 15 member countries who co-operate in various activities aimed at promoting the development of biomass for energy [28]. The IEA operates in three year cycles, and in each triennium the objectives are split into Tasks, encompassing biomass production, harvesting and utilisation as well as the use of municipal solid wastes. Within each Task there are activities which examine certain areas using funds provided by the countries that join the activity. These activities offer an opportunity for international collaboration, and have resulted in several techno-economic studies. These are discussed below.

3.4.1 IEA Techno-economic assessment of biomass liquefaction processes, 1990

This activity examined the techno-economics of 4 systems in current and potential cases. The systems are shown in Table 3.4. The report from the study [140] gives a detailed account of the technical and economic parameters and methodologies that were used. The evaluation of the potential cases was performed by assuming improvements to the technologies on a case by case basis. In the pyrolysis case the future reactor design used (circulating fluid beds rather than fluid beds) allowed the use of fewer reactors with a larger capacities and consequent savings in capital and labour costs. The pyrolysis systems include drying from 50% to 7% moisture content in flash dryers and grinding of the feedstock to below 1 mm in a two stage grinding process. Results for the current and future wood pyrolysis systems are presented in Table 3.5.

Table 3.4 - System criteria evaluated by the IEA TEA of biomass liquefaction processes

Feedstocks	Wood
	Peat
Products	Liquid fuels
Processes	Liquefaction, current and future
	Fast pyrolysis, current and future
Capacity limits	1000 odt/d
System scope includes:	
Feed production	No
Feed transport	No
Feed pretreatment	Yes
Feed conversion	
Combustion	No
Gasification	No
Pyrolysis	Yes
Electricity generation	No
Multiple sites/de-coupling?	No

Table 3.5 - Results for wood pyrolysis, IEA TEA of biomass liquefaction processes

	Current	Potential
Wood input, odt/h	41.7	41.7
Pyrolysis liquid produced, t/h wet	28.06	25.83[b]
Feed conversion efficiency, %LHV	62%	68%
Internal power consumption, MW_e	6.3	3.5
Total plant cost, $\$k_{1995}$[a]	55655	33928
Labour requirement, persons	31.5	26.5

a *converted from 1990 basis using $CI_{1990} = 105$, $CI_{1995} = 122$*

b *The potential case produced less liquid even though its efficiency was deemed to be greater. An analysis of the mass and energy flow spreadsheets showed that the assumed organics yield in both cases was 66% odt basis. In the current case this yield was reached after losses in the condenser, in the potential case this yield was before losses. It is not known which case is correct.*

3.4.2 IEA Pyrolysis/Liquefaction Activity (1989-1991)

The Pyrolysis/Liquefaction Activity (1989-1991) assessed the technical and economic potential of selected systems for the conversion of biomass to transportation fuels. Two processes were evaluated in three detailed case studies under this brief [181]. The processes were wood pyrolysis and upgrading to give gasoline based on NREL process data; and straw liquefaction at high pressure to give a fuel oil based on the MANOIL (University of

Manchester) data. The systems included all steps from the acceptance of a delivered feedstock through conversion to the production of a liquid product.

Each case study required the collection and evaluation of performance data; the development of a process flowsheet; calculation of mass and energy balances; capital cost calculations based on individual equipment items (including feed reception, storage, handling and pretreatment); and estimation of production costs. A sensitivity analysis was included to assess the influence of system parameters on the economics. The published report is very detailed and is a useful source of data for equipment costs. This level of detail is only possible if the resources available are very substantial or the number of analyses are few.

Table 3.6 - System criteria evaluated by the Pyrolysis/Liquefaction Activity (1989-1991)

Feedstocks	Wood
	Straw
Products	Gasoline
	Fuel oil
Processes	Liquefaction
	Fast pyrolysis and upgrading
Capacity limits	1000 daf t/d
System scope includes:	
Feed production	No
Feed transport	No
Feed pretreatment	Yes
Feed conversion	
Combustion	No
Gasification	No
Pyrolysis	Yes
Electricity generation	No
Multiple sites/de-coupling?	No

3.4.3 IEA Pyrolysis Activity (1992-1994)

This activity examined the cost and performance of wood-based electricity generating systems, using ASPEN PLUS™ to model the performance of a number of case studies and applying economic relationships to the results. The methodology and results are reported in [113]. The ASPEN modelling was carried out at VTT (see Section 3.8.6). The financial analysis is based on Aston University data: it will be summarised here but is not used in any validation of the results of this work since a large part of the basic financial data is common to both studies. A summary of scope of the study is given in Table 3.7.

Table 3.7 - Key system criteria evaluated by the Pyrolysis Activity (1992-1994)

Feedstocks	Wood
Products	Electricity
	Electricity and heat
Processes	Pressurised integrated gasification combined cycle
	Atmospheric integrated gasification combined cycle
	Pressurised gasification and steam-injected gas turbine
	Atmospheric gasification and diesel engine
	Fast pyrolysis and diesel engine
	Fast pyrolysis and gas turbine combined cycle
Capacity limits	Sized to give approximately 5, 30 or 60 MW_e
System scope includes:	
Feed production	No
Feed transport	No
Feed pretreatment	Yes - using a VTT drying model and Aston financial data
Feed conversion	
Combustion	No
Gasification	Yes
Pyrolysis	Yes - but only as cost for pyrolysis liquid production
Electricity generation	Yes - but poorly modelled
Multiple sites/de-coupling?	No

A total of 21 systems were evaluated by varying the processes, scales, products and drying technology used. The calculated cost and performance of the gasification systems are given in Table 3.8. Although this study attempted to compare gasification and pyrolysis systems, modelling of pyrolysis has actually been avoided. Instead a cost of pyrolysis liquid is used and the performance of gas turbine combined cycles and diesel engines fired by pyrolysis liquid has been modelled. The costs of the fast pyrolysis liquid were calculated using the BLUNT model (described above). Although this gives some idea of the relative costs and performance of pyrolysis and gasification the results cannot be reliably compared until the scope of all systems is the same.

Key results of this work were:

- IGCC systems are only viable using current costs at scales above 30-40 MW_e.
- There is no clear advantage to atmospheric or pressurised gasification.
- STIG is not competitive since the amount of steam that can be injected is limited.
- The investment costs of small scale diesel engine systems are too high, due mainly to the costs of drying.

- Steam drying gives lower production costs than flue gas drying, but this was only tested in two medium scale cogeneration systems where steam was available.
- Pyrolysis-based system efficiencies are low.
- Pyrolysis costs are high but their ability to be used in de-coupled systems may compensate for this (this hypothesis was not tested).

ASPEN is a complex and detailed process simulation tool and as such a lot of effort has been put into the technical modelling of the systems. The economic analysis is much less detailed in comparison, and the authors strongly emphasise the potential errors in this area. It must however be realised that there are no examples of large scale gasification systems at the 30 MW_e and 60 MW_e scales modelled. As such there is just as much uncertainty inherent in the technical modelling. It can be concluded that if there is a high level of uncertainty in both the technical and economic performance then there is little point in modelling the technical performance in great detail with only a cursory examination of the costs - a more appropriate balance should be found.

Table 3.8 - Results from the IEA Pyrolysis Activity for gasification systems

Gasification pressure	Dryer type	Power generation			Capital cost	Electricity cost[a]
		Net	Losses	%LHV	$/kWe	¢/kWh
Pressure gasification						
IGCC, power only	Flue gas	31.9	0.8	47.1	2700	7.9
IGCC, cogeneration	Flue gas	30.0	0.8	44.3	3000	7.0
IGCC, cogeneration	Steam	29.1	1.0	42.9	3100	6.6
IGCC, power only	Flue gas	59.8	1.3	45.1	2200	6.9
Atmospheric gasification						
IGCC, power only	Flue gas	6.6	1.4	37.4	4800	12.9
IGCC, power only	Flue gas	33.2	6.9	40.9	2700	8.1
IGCC, power only	Flue gas	62.3	11.6	39.5	2200	7.2
IGCC, cogeneration	Flue gas	29.1	6.7	37.2	3100	7.1
Pressure gasification						
STIG, power only	Flue gas	4.9	0.2	28.9	7100	18.4
STIG, power only	Flue gas	23.6	0.9	34.9	3500	10.2
STIG, cogeneration	Steam	23.5	0.9	34.9	3700	8.0
STIG, cogeneration	Flue gas	23.6	0.9	34.9	3600	8.6
Atmospheric gasification						
Diesel engine	Flue gas	4.9	0.1	33.9	4600	13.6
Diesel engine	Flue gas	24.6	0.4	33.9	2800	9.4

a Based on 25 $/t feed cost, 50% moisture (dried at the plant to 15%); 20 year life; 5% interest rate; 5000 h/y operation

3.4.4 IEA Combustion Activity (1992-1994)

A study was undertaken under the IEA Task X Combustion Activity of options for small-scale combustion CHP systems (2-3 MW_e). This was co-ordinated by dk-Teknik in Denmark and the results are reported by Jakobsen [182] and summarised by Hustad [36]. Five technologies were investigated in terms of their status, investment costs, efficiency, and near-term potential. The technologies were: steam turbines, steam engines, Stirling engines, indirectly fired gas turbines and directly fired gas turbines (using a pressurised downdraft combustor). The qualitative results of the study are quite useful, in that the current status, opportunities and constraints for each technology are reported (these results are used in Chapter 2).

The quantitative comparisons are not as useful since varying system definitions have been used for each technology: some of the analyses include the combustor and some concentrate solely on the prime mover. Nor has any attempt been made to put all capital costs on the same financial basis. A simple analysis of costs for a 2 MW_{th} heat output CHP system is given for each of the 5 systems, where the electricity output is variable and depends on each technology. The revenue from electricity sales is used to offset the heat price. In the comparison only the base investment costs and efficiencies are varied between the systems. As such the analysis is rather simplistic: all capital costs are scaled using a scale factor of 0.8 despite probable changes in sensitivity to scale; all maintenance costs are constant (3% investment); labour is constant at 1.5 persons. A summary of the findings of the report are given in Table 3.9 but they should be used with caution, as the authors themselves are keen to point out.

Table 3.9 - IEA Combustion Activity comparison of 2 MW_{th} CHP systems

	Steam turbine	Steam engine	Stirling engine	Direct-fired gas turbine	Indirect-fired gas turbine
Thermal output, MW_{th}	2.00	2.00	2.00	2.00	2.00
Power output, MW_e	0.44	0.48	1.05	1.16	0.33
Efficiency, %	14.6	16.7	27.5	28.6	10.0
Investment, $\$k_{1994}$ [a]	1682	1097	779	2164	1808
Heat sales price, \$/MJ [b]	11.12	9.11	6.92	10.12	10.12

a *Year has been assumed. The investment costs do not include items common to all systems such as buildings, the heat distribution network*

b *Electricity sale price fixed to 5 ¢/kWh; wood chip cost (45% moisture) 5.5 \$/GJ*

3.4.5 IEA Interfacing (1992-1994)

The IEA Interfacing Activity was a collaboration between two IEA Bioenergy Agreement Activities (Task IX - Technoeconomic assessment of wood fuel; Task X - Interfacing). This work featured early versions of the systems models developed in this thesis and a summary of the work and its findings is given in the paper attached as Appendix A.

The aim of this work was to examine the interactions at the system interfaces described in Chapter 2. To accomplish this a spreadsheet based model was constructed that calculated the costs and performance of every stage in the system from the establishment of a feed crop to the delivery of ethanol or the supply of electricity to the grid. The feed conversion and electricity generating options modelled were exactly those studied in this thesis. Each stage was modelled on a spreadsheet and the spreadsheets were combined in an overall systems model called BEAM (BioEnergy Assessment Model). The reader is referred to Appendix A for the results and conclusions of the analysis.

The feed production spreadsheets that were used modelled wood production from short rotation forestry and conventional forestry under UK conditions. These models were provided under licence to the IEA activity by the University of Aberdeen Wood Supply Research Group. The BEAM model used earlier versions of the techno-economic models developed for this thesis for pretreatment, conversion and generation. These earlier versions were much simpler and less rigorous than the models described in later chapters.

3.5 PRINCETON UNIVERSITY, US

Larson, Williams et al [30, 125, 183] have been evaluating various biomass to electricity systems since the mid-1980s at Princeton University, US. The studies have largely concentrated on bagasse as the feedstock and the use of gasifiers and gas turbines in various cycles. This work used a modelling package called STEAM that was developed specially for simulation of complex gas and steam turbine cycles and uses Turbo-Pascal with limited application of FORTRAN. The package demands the definition of a process from a flowsheet of the components required such as pumps, compressors, combustors and heat exchangers. The user needs a considerable amount of information before attempting to model a particular system. After the process has been defined, the package will iterate and find the stable steady-state operating conditions.

Capital costs are calculated by scaling base capital costs for the entire system. Published economic analyses were confined to a basic review of capital costs for the total systems, assumed feedstock costs and O&M (operating and maintenance). The justification for this is that the processes are not yet established and more accurate data is not yet available. Just as in the ASPEN modelling, there is an imbalance in the work: it is assumed that performance data is reliable enough to model the technical features of a system with great detail, but uncertainties are used to justify a relatively cursory examination of costs.

Lately work by Consonni and Larson [112, 184] has focused on three types of gasifiers: pressure circulating fluid bed, atmospheric fluid bed and atmospheric indirectly heated. Systems using these gasifiers and the General Electric LM2500 gas turbine have been examined in detail, focusing on near term solutions to utilising low heating value gases that involve slight de-rating of the turbine because of a necessity to lower turbine inlet temperatures. Further case studies have been analysed that consider future gas turbine concepts.

Work by Marrison at Princeton [185] has extended the original systems studied to include the transport of the feedstock to the processing site. The aim was to analyse the opposing effects of increasing feed transportation costs and decreasing capital costs as system capacity increases. The results were calculated using a set of relationships derived from previous data on feed production costs, transport costs, capital costs, efficiencies and operating costs. Relationships for transport costs were developed separately and in some detail using a geographical information system (GIS) to produce yields that vary with local soil type and transport distances that account for the actual road system in the region. The results from the GIS analysis were then reduced down to a few general relationships that were applied to give the results. The advantage of this approach is that a few simple continuous relationships are used to illustrate a principle that could be later examined by more detailed modelling or case studies. The drawback of such an approach is that many simplifications have been introduced. For example, generating efficiencies are independent of scale for all but the steam plant, even though scale effects on efficiencies will have a major role in determining feed requirements, feed transport costs and hence the optimum plant size.

Marrison showed that the impact of economies of scale on capital costs are much more significant than the impact of increasing transport costs with scale. Depending on the combustion or IGCC system analysed, the interaction between the two scale effects produced minimum electricity costs at scales between 114 MWe and 519 MWe, outside the range that is likely due to limited feed availability. Marrison also demonstrated that the scale effects are very important at small scale and virtually insignificant above a certain threshold: most electricity production costs reached a level within 1% of the minimum level at capacities around 50% of the required capacity for minimum costs.

3.6 UNIVERSITY OF ULSTER, N. IRELAND

A variety of wood-based generating systems have been studied at the University of Ulster, all based on combustion or gasification. Prime movers have included steam turbines, gas turbines and engines, and the latest work also considers fuel cells [186, 187]. The processes are modelled using ECLIPSE, a modelling package for the PC designed for the analysis of chemical processes and adapted for use in power systems analysis by the University of Ulster. Each system analysis is very detailed and requires the preparation of a flow diagram, stream conditions, mass and energy balances, utilities requirements, capital costs and operating costs. McIlveen-Wright has used ECLIPSE to analyse over 56 wood-based generating systems, with the scope of the study shown in Table 3.10. Key results for combustion and IGCC systems are given in Table 3.11 and Table 3.12 [186].

The analyses are wide ranging and detailed. The study features a thorough examination of the effects of transport costs, which were found to have little impact in all but the largest plant sizes and even then the effects were marginal unless the feed was very dispersed. Another feature of the work is an examination of the effects of drying on the combustion systems. This was uneconomic in plant sizes up to $100MW_e$, where the extra efficiency obtained did not justify the increased capital cost. Even in the very large plant the production cost improvement was marginal. However, the study considered drying down to 13% moisture content (wet basis), which is rather extreme when further improvements in combustor performance are small when drying below around 40% (see Section 7.2.3). Thus a study with less severe drying might have given improved system efficiencies without the high capital cost penalties seen here.

The analyses of the different IGCC options summarised in Table 3.12 allows a comparison of both pressurised and atmospheric gasifiers as well as two gas cleaning techniques (hot gas filtration and wet gas scrubbing). One unusual result was that the internal power consumption of the pressurised gasification cases was consistently higher than the respective atmospheric case. One would expect the fuel gas compression required in the latter case to make the power consumption of the atmospheric case higher (as seen in Table 3.8). The consistency between specific capital costs in atmospheric and pressurised gasifiers is also striking, where the lower costs of the atmospheric gasifier are cancelled out by their lower power output.

Table 3.10 - Key system criteria evaluated by McIlveen-Wright

Feedstocks	Wood
Products	Electricity
Processes	Combustion without feed drying Combustion with feed drying Gasification and gas turbine combined cycles Gasification and gas turbine simple cycles Gasification and gas turbine steam injected cycles Gasification and SI engines
Capacity limits	From 10 to 10000 odt/d
System scope includes:	
Feed production	No
Feed transport	Yes
Feed pretreatment	Yes - but fixed equipment throughout (except for the dryer)
Feed conversion	
Combustion	Yes
Gasification	Yes
Pyrolysis	No
Electricity generation	Yes
Multiple sites/de-coupling?	N

Table 3.11 - Combustion system performance and cost data, McIlveen-Wright

Plant Size	odt/d	10	10	100	100	500	500	1000	1000
Dryer[a]		No	Yes	No	Yes	No	Yes	No	Yes
Heat input	MWth LHV	2	2	20.1	20.1	100.5	100.5	201.1	201.1
Gross power	MWe	0.43	0.43	5.3	5.5	28.1	29.6	56.2	59.1
Internal consumption	MWe	0.08	0.1	0.6	0.7	2.6	3.1	5	5.8
Net power	MWe	0.35	0.33	4.7	4.8	25.5	26.5	51.2	53.3
Efficiency	%LHV	17.5	16.6	23.1	23.9	25.4	26.4	25.4	26.5
Specific investment[b]	$/kWe	11230	13197	3110	3547	1900	2303	1479	1748

a dryer dries feed from 50% to 13% moisture content

b Original cost data in £$_{1992}$ was updated using CI_{1992} = 160, CI_{1995} = 170, Exchange rate in 1995 1 US$ = 0.632 £

Table 3.12 - IGCC systems performance and cost data, McIlveen-Wright

General Electric Gas Turbine No.	LM 1600	LM 1600	LM 1600	LM 5000	LM 5000	LM 5000	MS 6001	MS 6001	MS 6001
Gasifier	Press.	Press.	Atmos.	Press.	Press.	Atmos.	Press.	Press.	Atmos.
Gas cleaning	Filter	Scrub	Scrub	Filter	Scrub	Scrub	Filter	Scrub	Scrub
Heat input, MW_{th} LHV	60	76.2	59.2	147.5	185.4	143.4	183.9	227.3	185.2
Gross power output, MWe	29.1	35.8	21.2	70.6	86.2	49.8	86.3	97.3	59.9
Internal power, MWe	4.4	5.6	0.7	10.9	13.6	1.5	12.1	14.4	1.9
Net power output, MWe	24.7	30.2	20.5	59.7	72.6	48.3	74.2	82.9	58
Efficiency, %LHV	41.1	39.7	34.6	40.5	39.1	33.7	40.4	36.5	31.3
Specific investment, $/kWe[a]	4010	3783	4215	3120	2972	3263	2882	2992	3166

a *Original cost data in $£_{1992}$ was updated using $CI_{1992} = 160$, $CI_{1995} = 170$, Exchange rate in 1995 1 US$ = 0.632 £*

The variety of systems analysed, coupled with the complexity of the ECLIPSE approach, have inevitably led to a few compromises in the analysis of the pretreatment system where the reception and handling equipment is the same for all system capacities, ranging from 0.3 MW_e to over 600 MW_e (a very high upper limit). In reality the front end of the conversion system would change significantly with such a wide range of capacities, from basic manual handling to highly automated systems that exchange increased capital cost for lower operating costs.

3.7 UTRECHT UNIVERSITY, NETHERLANDS

This institution is contributing to an EC JOULE II project entitled "Energy from biomass: an assessment of two promising systems for energy production". The project compares combustion and gasification technologies for the ultimate aim of building a 30 MW_e IGCC system in the Netherlands and a biomass combustion plant in Ireland. The capacity of the 30 MW_e IGCC plant has been fixed by the capacity of the gas turbine, the General Electric LM2500. The combustion option will be implemented either by building a new unit or retrofitting a peat-fired boiler.

Van den Broek [71] has published a review of biomass combustion systems that includes a qualitative assessment of the available technology and an quantitative assessment of modern combustion systems. The quantitative assessment examines 6 existing plants, 2 that are in construction and 2 concepts in terms of their efficiency, investment costs and emissions. Some of the plants studied are CHP units and the system performance has been adapted using a fully described methodology to predict their performance when producing power only. The results of the survey are given in Table 3.13.

The study offers a variety of cost data for the systems surveyed. This data has been updated to the same currency and time base but no attempt has been made to analyse the cost data further to bring all the cost data to the same financial scope (variation in financial scope is discussed in Section 4.5).

In addition to the data in Table 3.13 the study drew the following conclusions:

- all boiler types are still constructed, no design has a clear advantage;
- the majority of boilers use a wood feedstock;
- the maximum expected capacity for a biomass combustion plant in Europe is 50 MW_e; and
- the steam cycle efficiencies are very scale dependent.

The gasification system options have also been reviewed and are reported by Faaij [188]. The two studies are independent, which is surprising given that the intention of the project is to compare the gasification and combustion options. Faaij initially examined three IGCC options, using three different gasification technologies, each using circulating fluid bed technologies. One option, the twin bed gasifier was eliminated on the grounds of its non-commercial status. A further option was the pressurised circulating fluid bed gasifier, but this was rejected on the grounds of technical uncertainty in the gas clean-up system and system control. This confined the study to an analysis of the cost and performance of IGCC using atmospheric gasification and wet gas scrubbing. The elimination of the pressurised system is dubious: hot gas filtration is currently the subject of much research and its performance is no more uncertain than wet gas scrubbing in biomass gasification applications; the only biomass IGCC system in existence is a pressurised system and the results from this demonstration will help to clarify both gas clean-up and control issues. Costs were not considered in the decision on the grounds of uncertainty but it could be argued that the uncertainty in both atmospheric and pressurised processes is considerable.

Table 3.13 - Combustion plant survey results, Utrecht University

Plant	Fuel (%wet)	Power MW_e[a]	Efficiencies, %LHV Boiler	Turbine	Overall	Capital cost $\$_{1992}/kW_e$
Zurn Travelling grate	Wood (50%)	25	-	-	28	1200-1600
Delano I BFB	Ag, waste (24%)	27	86	35	29	-
McNeil Travelling grate	Wood (47%)	50	83	39	30	1800
Måbjergkærket CHP Vibrating grate	Straw, wood, MSW	34	89	36	30	2900
Händleöverket CHP CFB	Wood (50%)	46	89	38	32	1100[b]
Enköping CHP Vibrating grate	Wood (45%)	28	96	37	33	1900
Grenaa CHP CFB	Coal, straw	27	100[c]	37	35	2500
EPON co-fire Pulverised coal	Demolition wood	20	-	-	37	800[d]
WTE Pile grate	Whole trees (44%)	100	90	41	38	1500
ELSAM CFB	Coal, straw, wood	250	-	-	44	-

a *CHP capacities and efficiencies have been converted to give the expected performance in power only production*
b *Costs for CFB boiler and pretreatment only.*
c *Efficiencies are probably about 5-10% lower because of inaccurate data, this would lead to a electrical efficiency of 32-33%.*
d *Additional costs for additional investments for wood co-firing (pretreatment and burner)*

The atmospheric system was simulated using ASPEN-PLUS. A logistics study for the area was used to define the feed transport costs for five potential feedstocks, including forestry thinnings (as chips) and waste wood. Drying of the feedstocks in a flue gas dryer was included. The analysis for the clean wood feedstock at 50% moisture content produced a net generating efficiency of 40.3% based on LHV, a net output of 29.04 MW_e and an internal power consumption of 7.51 MW_e. 6.53 MW_e of the total internal power consumption was due to the fuel gas compressor required to raise the fuel gas pressure to the 26 bar gas turbine inlet pressure.

Some useful information on gas turbine fuel tolerances are presented and an analysis of potential emissions. Ash residues and the scrubber effluent were acknowledged but no effort was made to assess their disposal costs.

Quite a detailed assessment of costs for the system has been undertaken. This includes an interesting analysis of first and future plant cost. Each major plant item has been assigned a range of expected costs based on manufacturers' data and literature values. The low estimates are added together to give what is defined as the potential future cost; the highest costs for each equipment item are added to give the 1st plant cost. After factors are added in to give total plant costs the investment required is 76 M$ for the 1st plant, 57 M$ for future plant. Given a 29MW$_e$ output, this equates to a range of 1954-2615 \$/kW$_e$. The financial analysis extends to an evaluation of the operating costs including labour, utilities, ash disposal and capital charges. The wood based system gives a final electricity cost of 7.87-10.8 ¢/kWh. A sensitivity analysis is also given that highlights the importance of process efficiency, capital costs and the operating hours per year to overall production costs.

In conclusion this analysis is a thorough examination of a single system using a variety of feedstocks. The background to the results is well-documented and while any comparison between this and other studies should be approached with caution, these details make such comparisons more informed. Ultimately though it is a shame that resources have prevented the application of the analysis to pressurised gasification or other conversion technologies so that a comparison could be made.

3.8 OTHER WORK

3.8.1 Electric Power Research Institute, US

The US Electric Power Research Institute (EPRI) has developed a model for evaluating the cost and performance of various electricity generating systems based on wood, waste or co-fired feedstocks [189]. This model, BIOPOWER, was made available under licence to Aston University under the terms of the IEA Interfacing Activity described above so that it could be used to support and validate the combustion model reported in Chapter 7. The electricity generating options include several combustion technologies and a gasification gas turbine combined cycle option. The model has been constructed using spreadsheets and depends on user input for key parameters such as boiler efficiency and generator efficiency. These values do not change automatically as the system changes and the model must be used with great care to produce meaningful results.

3.8.2 National Renewable Energy Laboratory, US

NREL are conducting an evaluation of the costs and performance of a biomass IGCC system incorporating the Battelle Columbus twin-bed gasifier system. The system has been modelled using ASPEN and the technical and economic results have been published by Mann [190]. The study will continue with a life cycle analysis (LCA) for the process that aims to establish all the human and ecological impacts of the system for the immediate and global environment. The net capacity of the system is 122 MW_e with an internal consumption of 15.2 MW_e (11% of gross output) and the reported capital costs are 1108 $/kW$_e$ (1990 basis). This cost is significantly below those suggested by other authors, although the financial scope of the costs has not been given. The net plant efficiency quoted is 35.4% on an HHV basis. Assuming a 6% hydrogen content and a 50% moisture content for the wood fuel this equates to a net efficiency 43.4% on an LHV basis.

NREL have also examined the economics of fast pyrolysis in a separate study, based on the ablative vortex reactor that is under development at the laboratory. Gregoire [191] has reported on the costs of a 900 t/d system, simulated using ASPEN PLUS. The total plant costs for the ablative pyrolysis system was 17.3 M$, although the financial results are of little relevance now since they are based on 1983 costs (updated to 1992).

3.8.3 Energy Technology Support Unit, UK

RECAP (Renewable Energy Crop Analysis Program) is a computer application developed at the Energy Technology Support Unit (ETSU) in the UK, and launched in January 1995 [192]. RECAP is used to evaluate the costs of producing a short rotation coppice feedstock to supply an electricity generator or combustor and calculates the overall profitability of the system.

The model claims to be capable of evaluating systems from the production of biomass to the generation of electricity or heat. In reality only feed production may be calculated with any reliability. A feed transport model has been included. While this is very detailed in some respects (fuel usage per vehicle is calculated for example) there are fundamental flaws in calculating transport distances. The pretreatment system analysis assumes air drying only with no specific drying machinery. As such the feedstock is kept in store while it dries to the moisture content specified by the conversion process and an auxiliary fuel is used in the interim, an odd solution. Combustion and gasification are modelled using very simple models that clearly have little relation to actual practice (gasification requires a feed moisture content that is fixed at 30% for example) [193].

The model combines these flaws with a polished user interface and some rigorous analysis tools such as Monte Carlo simulations to evaluate process risk. In short the scope of RECAP has been extended far beyond the capabilities of the data within it. The feed production work is very detailed and reliable: it would have been far better to limit the scope of the first version of RECAP to this stage in the biomass to electricity system and extend this scope in the later versions that are planned.

3.8.4 Elliott and Booth, Shell International

Elliot and Booth have published several papers in support of the 27 MW_e GEF Brazilian IGCC project [14]. This work gives a projected capital cost for the project of 2700 $/$kW_e$, and uses a learning curve to show that costs could fall to 1300 $/$kW_e$ after 10 replications of the plant. This learning curve is discussed in Section 4.7. The technology to be used in the GEF project has not been finalised - it may feature either pressurised or atmospheric gasification. Since the capital costs of each process are known to be very different [134], it would appear premature to discuss capital costs, even though the 2700 $/$kW_e$ figure is now widely quoted. The basic capital costs should therefore be viewed with some uncertainty.

3.8.5 Ensyn Technologies Inc., Canada

Papers on the Ensyn RTP fast pyrolysis process often include cursory summaries of the process economics. There is however, a paper published in partnership with Beckman that goes into a little more detail [194]. This study is based on a 25 t/d operation, with the costs scaled up using a 0.6 scale exponent to give costs at 100 and 250 t/d. The best capital cost for the 25 t/d plant were 917 $k, with an estimated range up to 1511 $k (1992 figures have been updated by a factor of 122/113). Pretreatment costs are not included in the analyses and a zero feed cost is assumed. The authors acknowledge that this is very optimistic and that feed pretreatment costs may add 10-20 $/odt to the feed cost.

3.8.6 VTT, Finland

The Technical Research Centre of Finland (VTT) have studied many bioenergy systems based mainly on wood and peat. These studies generally apply ASPEN in the investigation of gasification and gas turbine cycles [195]. Gasification and combustion cycles are being evaluated as part of a European JOULE project co-ordinated by ENEL, an Italian Utility company. This work has highlighted the difficulties of achieving a viable system at very small scale and the study reports that systems under 0.5 MW_e cannot be supported financially under any of the conditions used in the work. IGCC is promoted at scales above 30-50 MW_e.

Costs for current and future cases have been examined by increasing the number of available operating hours for future cases and reducing capital costs, although the basis for reducing the capital costs is not given. Published results are presented in Table 3.14 [195]

Table 3.14 - VTT analyses of combustion and IGCC cycles

	Gas turbine	Capacity, MW_e	Efficiency, %LHV	Capital cost, $k [a]	Production cost, c/kWh [b]
Steam cycle	-	5.7	21.1	2158	4.95
IGCC, now	Frame 6	60.3	45.3	1778	5.71
IGCC, future	Frame 6	60.3	45.3	1397	4.06
IGCC, now	PGT 10	16.4	44.4	2920	8.88
IGCC future	PGT 10	16.4	44.4	2285	6.34

a costs are updated to from ECU_{1994} using a CI ratio of 160/155 and an exchange rate of 0.813 ECU/$
b current costs are calculated using 4440 operating hours/y; future costs use 4750 operating hours/y

Current work is also evaluating fuel cells [196] and has shown the efficiency improvement that could be gained by integrating a solid oxide fuel cell into a pressurised gasification integrated gas turbine combined cycle (an increase of 10% from 49.9% to 59.1%). However, the authors are keen to stress that there is a lot of development required before such systems are demonstrated.

3.9 SUMMARY

The system boundaries of each study are different. The start point or upstream limit to the system may be the production of the feedstock, the transport of the feedstock, the delivery of a raw feedstock to the facility or the input of a prepared feedstock to the conversion reactor. The end point or downstream limit to the systems can may be the production of a fuel or the generation of electricity. Similarly the economic scope can vary: capital costs usually extend beyond basic equipment costs but they can incorporate or exclude various features such as engineering costs, interest and contingency. For these reasons it is very difficult to compare results from different analyses unless it is clear what has been included and excluded from the results.

No single study has evaluated all three thermochemical conversion technologies. Combustion and gasification have been extensively examined as individual cases and in comparative studies. Pyrolysis has been less thoroughly explored, reflecting its less established status. There have been isolated examples of costs in the literature for pyrolysis liquid production and the technology was covered by the BLUNT and IEA Pyrolysis Activity work. The BLUNT model, however, lacked integrity in the electricity generation calculations while the IEA work merely assumed that liquid was purchased to fire the generating equipment. It can be concluded that a comparison of all three thermochemical conversion routes using a consistent methodology would be a useful addition to current knowledge.

De-coupled systems have not been examined in any studies and in all of the system studies operation at a single site only is examined. It would be novel to assess how fast pyrolysis systems may be manipulated through de-coupling and how this affects the competitiveness of this technology in relation to combustion and gasification.

The starting point of the current study will be the loading of feedstock for transport to the conversion sites(s). The effect of transport costs with changing system capacity has already been examined in work by Marrison and McIlveen-Wright. It has also been included in other studies such as the work by Faaij. Thus this element of the current work is unlikely to yield any new information about the effects of feed transport in close-coupled systems, but the analysis of transport will be unusual in this case where multi-site and de-coupled systems are examined.

This work will include the costs and performance of the pretreatment system required for the reception, storage, handling, drying and comminution of the feedstock at the conversion facility prior to the entry to the reactor. Pretreatment has been considered in most of the recent system studies, but the quality of the analysis is variable. In the majority of cases the focus of the pretreatment work is on the dryer, because of the impact of feed moisture content on the quality of the product in the case of gasification and also because of the overall effect of drying on system performance and economics. The studies rarely pay attention to the changes that would occur in the reception, storage and handling parts of the pretreatment chain as the system capacity changes. This work will examine this aspect of the system in detail for the first time.

Various studies have examined the potential impact of learning on process economics and recognised that the gasification and fast pyrolysis technologies are inherently more expensive than combustion because the latter process is more established. Several approaches to this problem have been used, with future systems given either better performance, increased reliability or lower capital costs. Improvements in performance have only been used in case study work where the limited number of cases allows potential performance improvements to be evaluated in detail. Reliability is often used by VTT and can be a good indicator of learning because of the total electricity production increases. The most common approach is to reduce capital costs through either the application of learning factors or by varying contingencies. Learning factors have been promoted by Elliot and Booth and are often cited as a justification for investment in the first-of-a-kind plant despite high costs. The use of learning factors will be adapted for this work.

Finally, many studies use sensitivity analyses to analyse the effect of uncertainties on the results. There are other ways of examining uncertainty such as Monte-Carlo analyses that offer a more detailed study of changing input parameters but these are complex and less transparent. This work will also adopt sensitivity analyses as a means of testing the effect of changes to assumed values in the models.

4. METHODOLOGY

4.1 INTRODUCTION

This chapter establishes the procedures that will be used to model the systems that were specified at the end of Chapter 2. The structure of the model is described in Section 4.2. The characteristics of the feed that will be used in all systems are presented in Section 4.3. Parameters for the measurement and comparison of system performance are defined in Section 4.4. The scope of the capital costs used is defined in Section 4.5, and the methods used to convert all capital costs to this basis are described. Procedures for calculating production costs are defined in Section 4.6. Finally a discussion of learning effects and their application in this work is presented is Section 4.7.

4.2 MODEL STRUCTURE

4.2.1 Definition of approach

In Chapter 3 it was seen that the techno-economic evaluations were carried out on two basic levels: general studies that evaluated a large number of systems, usually by varying capacity, and case study analyses that evaluated a more limited number of systems in greater detail. In the former case the work was usually supported by one or more techno-economic models that calculated the details of process performance and capital costs were estimated once the mass and energy balances were known. Such models are suitable for work that must evaluate a large number of systems. Their disadvantage is that the development of the model requires considerable resources to ensure robust and accurate results. It is important that the underlying data, the relationships derived thereof, and the interdependent algorithms between system parameters are comprehensively researched. Where data is not available, the assumptions used are valid and justifiable. The case study analyses examined each case in isolation and the performance and cost criteria were calculated manually. Case studies tend to produce a more detailed examination of a limited number of systems. Their advantage is that they can include many system features that would be difficult to model accurately such as the local topography, multiple feedstock availability and cost, and specific environmental control equipment to comply with local regulations.

This thesis will evaluate biomass to electricity systems through the use of a specially developed set of techno-economic models. This approach is more suitable in this work because:

- the number of system configurations is large;
- some of the technologies to be modelled are insufficiently advanced to warrant their detailed modelling given a lack of suitable data (e.g. fast pyrolysis at commercial scales);
- uncertainties can be examined more easily using sensitivity analyses; and
- the model can be updated and expanded to allow further systems evaluations in the future.

4.2.2 The modelling platform

The model will be developed using spreadsheets. An alternative would have been to use a programming language to create a dedicated application and the two options are compared in Table 4.1. The reasons for using spreadsheets were:

- Calculation speed is no longer an issue with modern computers unless the model is very complex.
- Graphical capabilities are not vital; model integrity and ease of development is far more important. The ability to clearly present input and output data is the main concern. Spreadsheets can present input data adequately and are excellent at producing results that can be manipulated easily for presentation.
- Interaction with the IEA Interfacing Activity (see Section 3.4.5) would be enhanced.
- Previous experience at Aston (the AMBLE and BLUNT models) has shown that modelling using a programming language produces a model that is difficult to update or expand. It was considered important that this work could be developed further if required at the end of the current project.

4.2.3 Modelling the System

The systems have been broken down into parts that are modelled separately using a step model approach similar to that used in the AMBLE and BLUNT models described in Section 3.2. Each step model is referred to as a module and the modules that must be developed are shown in Figure 4.1. A definition of the limits within the overall system for each module are given in Table 4.2.

Table 4.1 - Comparison of modelling platforms

Criteria	Programming Language	Spreadsheet
Example	Visual Basic, Toolbook, Borland Delphi, Visual C++	Microsoft Excel, Lotus 123
Calculation Speed	Fast	Can be slow.
Graphical Capabilities	Good with the current range of windows development tools	Poor.
Distribution	Model may be distributed as a stand-alone package.	The user must have the required version of the spreadsheet application.
Transparency and continuity	Underlying calculations are hard to follow unless the code is accessible and the user is versed in the language.	Underlying calculations may be easily tracked by anyone familiar with spreadsheets (providing the spreadsheet is clearly laid out).

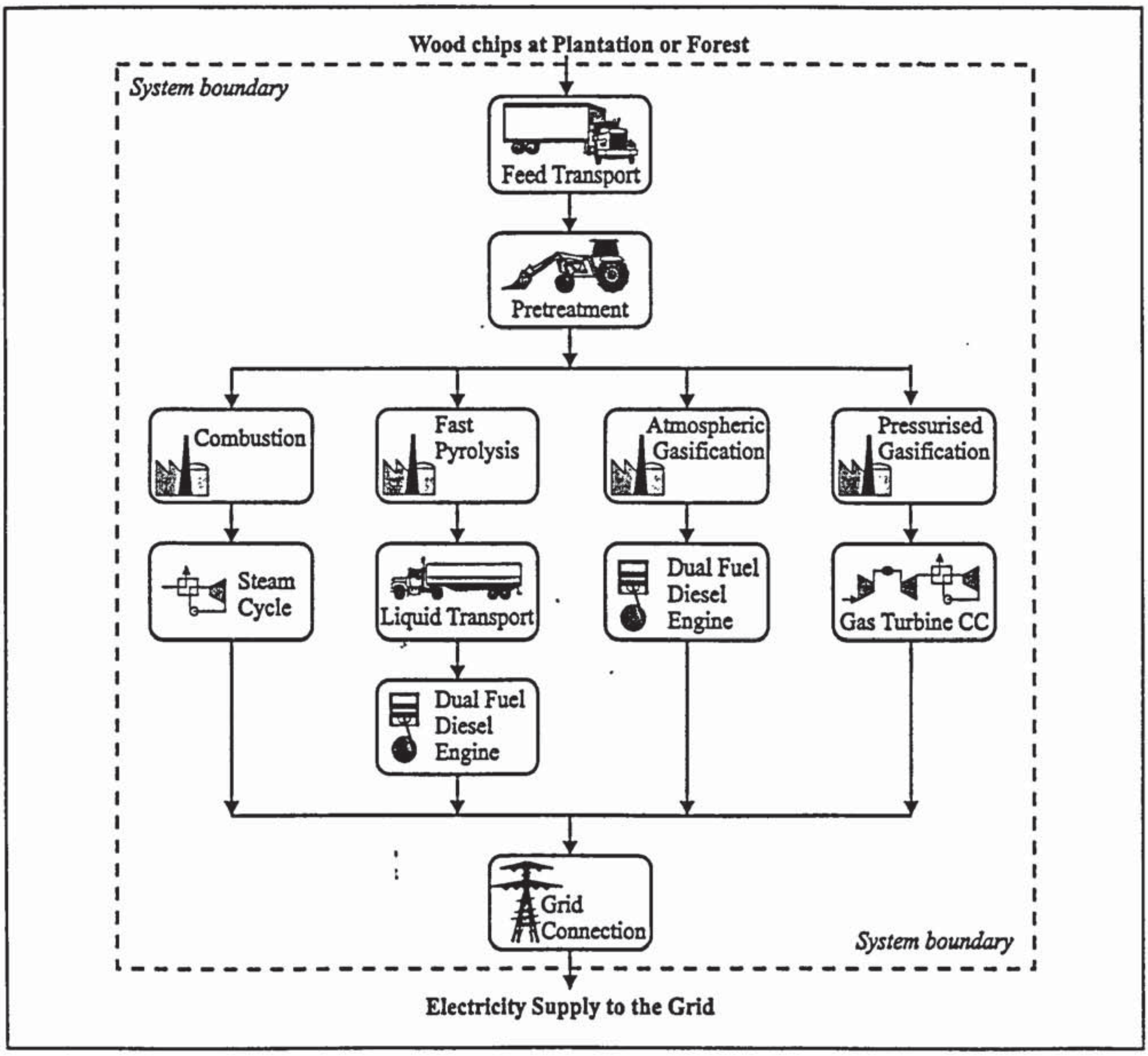


Figure 4.1 - Modules required in systems modelling

Table 4.2 - Module limits

Module	Upstream limit	Downstream limit	Described in
Feed transport	Wood chips ready for loading	Wood chips at the entrance to the conversion facility	Chapter 5
Feed pretreatment	Wood chips at the entrance to the conversion facility	Prepared feedstock at entry to the conversion reactor feeding system	Chapter 6
Combustion	Wood chip entry to the combustor	Steam supply to the steam turbine	Section 7.2
Fast pyrolysis	Wood powder entry to the reactor	Pyrolysis liquid supply to storage	Section 7.3
Atmospheric gasification	Wood chip entry to the reactor	Clean, cool fuel gas at engine injection system	Section 7.4
Pressurised gasification	Wood chip entry to the reactor	Clean, hot fuel gas at gas turbine combustion chamber	Section 7.5
Pyrolysis liquid transportation	Pyrolysis liquid from storage at the pyrolysis site	Pyrolysis liquid in storage at the generation site	Chapter 5
Steam cycle	Steam supply to the steam turbine	Gross electricity at generator terminals	Section 8.2
Liquid dual fuel engine	Pyrolysis liquid from storage	Gross electricity at generator terminals	Section 8.3
Gas dual fuel engine	Clean, cool fuel gas at engine injection system	Gross electricity at generator terminals	Section 8.4
Gas turbine combined cycle	Clean, hot fuel gas at gas turbine combustion chamber	Gross electricity at generator terminals	Section 0
Grid connection	Gross electricity at generator terminals	Net electricity to grid	Section 8.6

4.3 FEEDSTOCK CHARACTERISTICS

The feedstock at entry to all systems is wood, available for transport in chipped form and with the physical characteristics shown in Table 4.3. It is expected that the feedstock will contain a significant proportion of bark since the removal of bark would make the feed too expensive for the energy market (if the bark were removed the clean wood feedstock could be sold more profitably as a pulping feedstock). The characteristics in Table 4.3 are derived from the data given in Section 2.3.4 and assume that the wood is mixed with bark in the ratio 15% bark:85% wood.

Table 4.3 - Feedstock properties as delivered to the conversion site

Elemental composition		
C	%odt	51.8
H	%odt	5.7
O	%odt	40.9
N	%odt	0.1
S	%odt	0.0
Ash content	%odt	1.1
Particle size	mm	<50mm
Bulk density	odt/m^3	0.15
Moisture content	%wet basis	Up to 50%
HHV	GJ/odt	20.5
LHV	GJ/odt	19.3

Feed moisture content and feed cost before delivery will be variables in the model. The default moisture content is 50%. The default cost of feed before transport is $40/odt, based on the data presented in Section 2.3.2 and Section 2.3.3 (all costs are in $US, 1995 basis, see Section 4.5.2). The costs of feedstock will have an important influence on systems with different efficiencies and a variation of feed costs will be explored in the systems evaluations.

4.3.1 Feed heating value

The lower heating value of the feedstock is set in Table 4.3. All efficiencies in this work will be related to the lower heating value of the feedstock on the basis that this is the norm in power generation studies; that most prime mover specifications are defined for LHV; and because lower heating values are more relevant in comparative studies since they allow for differences in feed or fuel composition [71].

The lower heating value given in Table 4.3 is the amount of energy that would be recovered during the combustion of 1 odt of material if the material was dry. The amount of energy that will be recovered from the combustion of 1 odt of wet material is lower since energy is used to evaporate the moisture associated with each dry tonne. Therefore the lower heating value given is adjusted to give the lower heating value of the feedstock at its current moisture content at the specific point in the system. This is achieved using Equation 4.1 and Equation 4.2 where 2.454 is the latent heat of evaporation in GJ/t of water at the reference conditions for measuring heating values (20°C and 1 bar).

$$LHV_{x\%\ moisture}, GJ/dry\ t = LHV_{0\%\ moisture} - \frac{x}{(1-x)} \times 2.454 \tag{4.1}$$

$$LHV_{x\%\ moisture}, GJ/t = LHV_{0\%\ moisture} - \left(LHV_{0\%\ moisture} + 2.454\right) \times x\% \tag{4.2}$$

4.4 PERFORMANCE PARAMETERS

4.4.1 System efficiencies

System performance is measured in terms of its ability to convert the energy in the delivered feedstock into power supplied to the grid. Since the two dual fuel engines use an auxiliary fuel in electricity generation then this is included as an energy input to the system so that the four system efficiencies can be compared on a consistent basis. The various energy fluxes in the system that are considered are shown in Figure 4.2. All thermal energy flows are measured on an LHV basis. Net system efficiency will be calculated using Equation 4.3.

$$\text{Net system efficiency, } \eta_{e,net} = \frac{E_{e,net}}{E_{th,del} + E_{th,aux}} \tag{4.3}$$

where $E_{e,net}$ = net annual power output to the grid, GJ/y

$E_{th,conv}$ = annual energy value of the conversion technology product, GJ/y

$E_{th,aux}$ = annual energy value of the auxiliary diesel fuel if used, GJ/y

Other useful efficiencies that can be used to compare system performance are the gross system efficiency (Equation 4.4), the conversion efficiency (Equation 4.5) and the generation efficiency (Equation 4.6). This gross system efficiency is useful in highlighting the effects on net efficiency of internal power consumption, which is expected to be particularly high in fast pyrolysis where the feedstock must be ground. It should be noted that the conversion and generation efficiencies are gross, in that they are based purely on the energy in the feed or fuel and ignore power consumption.

$$\text{Gross system efficiency, } \eta_{e,gross} = \frac{E_{e,gross}}{E_{th,del} + E_{th,aux}} \tag{4.4}$$

$$\text{Conversion efficiency, } \eta_{conv} = \frac{E_{th,conv}}{E_{th,pret}} \tag{4.5}$$

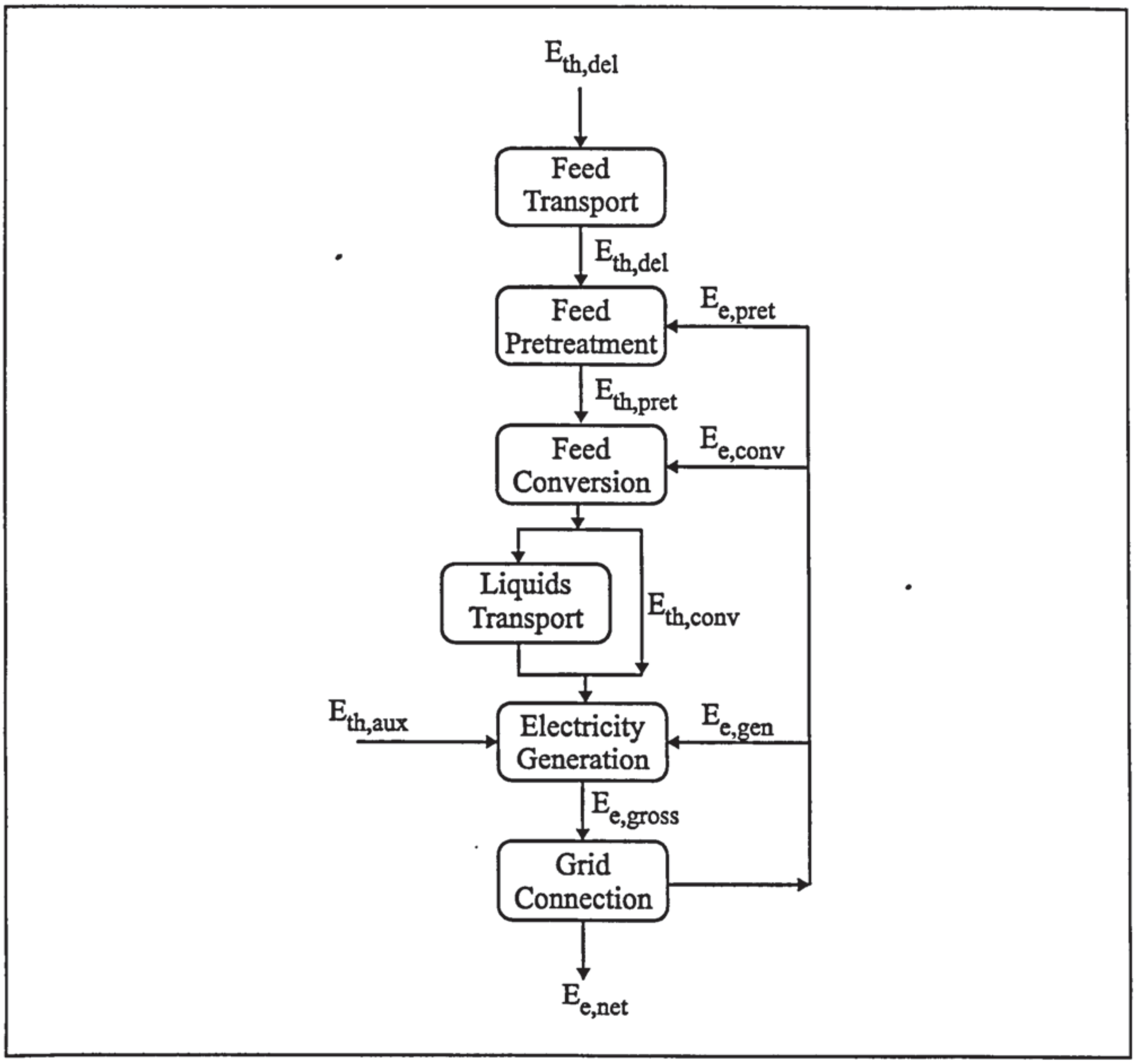


Figure 4.2 - Energy flow through the system

$$\text{Net generating efficiency, } \eta_{gen} = \frac{E_{e,gross}}{E_{th,conv} + E_{th,aux}} \tag{4.6}$$

where $E_{e,gross}$ = the gross annual power output before internal consumption, GJ/y

$E_{th,conv}$ = the energy in the conversion product, GJ/y

$E_{th,pret}$ = the energy in the prepared feedstock at the conversion reactor, GJ/y

4.4.2 Availability

One aspect of the system performance that is set in the system definition is the availability of the conversion technology and the generating cycle. This is the percentage of the year that the two stages will operate. In close-coupled systems the conversion and generating cycles are intimately linked and the two stages must operate at the same time. Therefore only one availability is specified. In fast pyrolysis systems the availability of the fast pyrolysis conversion stage and the dual fuel engine generating stage may be defined independently because the system can be de-coupled. Availability includes stoppages for planned and unplanned maintenance. The availability in all systems will be set to 90% by default. However, availability will be considered in the systems evaluations in Section 9.5.

4.5 CALCULATING CAPITAL COST

4.5.1 Introduction

Capital costs are required for two reasons:

- they show the total amount of money that is invested in plant construction;
- they contribute to production cost calculations either directly as interest on loan capital or indirectly a way of estimating components of the production cost.

Capital cost is one indicator of the amount of risk involved in funding a project since the capital is tied up in a venture that may ultimately fail. A high capital cost system that produces a low product cost by virtue of a high efficiency may therefore not be as attractive as a low capital cost system with a higher product cost if the risk inherent in the former system is higher. This aspect of the systems is discussed in Section 9.5.

Capital costs may be reported with varying financial scopes, as presented in Table 4.4. It is this variation in capital cost scope that makes it difficult to compare the results of different systems studies. The aim here is to calculate the total investment that would be required to finance the installation of a system to the point where it is ready to operate. This is the total plant cost. Total capital employed is not used since the working capital can be recovered at the end of the project. Methods used to convert capital costs to total plant costs are described in Section 4.5.3 and Section 4.5.4.

Table 4.4 - Scope of capital costs

Name	Scope
Equipment Cost, EC	The purchase costs of the major equipment items including: • Processing equipment; • Raw materials handling and storage; and • Product handling and storage.
Direct Plant Cost, DPC	The equipment cost plus the cost of: • Installation of major equipment items; • Instrumentation and its installation; • Piping and its installation; • Electrical equipment and its installation; • Buildings; • Yard improvements; • Service facilities; and • Land.
Installed Plant Cost, IPC	The direct plant cost plus: • Engineering and supervision; and • Construction.
Total Plant Cost, TPC	The installed plant cost plus: • Contractor's fee; and • Contingency
Total Capital Employed	Total plant cost + working capital.

4.5.2 Cost Year and Location

All costs are updated and relocated as necessary to give costs in US$, 1995 basis for a plant located in the UK.

Updating costs requires a cost index that is used to adjust for inflation in the industry. Various cost indexes are compiled relating to different industries as reported by the IChemE [197] and Peters [198]. International cost indices from the Chemical Engineer are used here to allow the use of international cost data [199]. Conversion to US$ is applied after the costs are updated to 1995 in their original currency. Average international exchange rates for 1995 are taken from OECD data [200]. The most recent cost data is used wherever possible to avoid errors in updating costs to 1995 since updating capital costs using cost indices does not allow for changes to the technology and general industry trends are only partially applicable to particular equipment or processes.

There are different costs associated with equipment purchased in different countries that can lead to errors when converting costs between countries. To account for this, location factors should be used to correct for the relative manufacturing costs in each country. Location factors are specific to a particular time and the most recent location factors that could be found were based on 1988 figures [201]. It was decided that to convert costs to a 1988 basis before applying location factors and updating to 1995 would probably introduce more errors that it would resolve. Therefore location factors have not been used in this work.

4.5.3 Development of total plant costs using ratios

This procedure is used where costs are available for whole processes or substantial parts of a process. Calculating total plant costs from individual equipment costs is described in the next section. It is based on established cost estimation procedures reported by Peters whereby the total plant costs are calculated from equipment, direct or installed plant costs by using ratios based on cost breakdowns for general processing plant as shown in Table 4.5 [198]. The ratios that are used are presented in Table 4.6.

4.5.4 Development of total plant cost using factors

This method produces total plant costs from the costs of individual major equipment items. The method is based on factors published by the IChemE [197] and adapted by Bridgwater to give the factors shown in Table 4.7 [113]. The direct plant cost is the equipment cost plus a fraction of equipment cost that corresponds to each factor in Table 4.7, calculated using Equation 4.7. Once the direct plant cost is known then the factors shown in Table 4.8 are used to convert to total plant costs.

$$\text{Factor} = a\left(\text{Equipment cost in mild steel, US\$}_{1991}\right)^{b} \qquad (4.7)$$

Table 4.5 - Breakdown of Total Plant Costs based on Modular Equipment Costs

Material handled	Solids		Solids/Fluids		Fluids	
	%TPC	EC=100	%TPC	EC=100	%TPC	EC=100
Equipment Cost	25.8%	100	24.2%	100	20.7%	100
Purchased Equipment Installation	11.6%	45	9.4%	39	9.7%	47
Instrumentation and Controls (installed)	2.3%	9	3.1%	13	3.7%	18
Piping (installed)	4.1%	16	7.5%	31	13.7%	66
Electrical (installed)	2.6%	10	2.4%	10	2.3%	11
Buildings (including services)	6.5%	25	7.0%	29	3.7%	18
Yard improvements	3.4%	13	2.4%	10	2.1%	10
Service facilities	10.3%	40	13.3%	55	14.5%	70
Land	1.6%	6	1.5%	6	1.2%	6
Direct Plant Cost	68.2%	264	70.9%	293	71.6%	346
Engineering and supervision	8.5%	33	7.7%	32	6.8%	33
Construction expense	10.1%	39	8.2%	34	8.5%	41
Installed Plant Cost	86.8%	336	86.9%	359	87.0%	420
Contractor's fee	4.4%	17	4.4%	18	4.3%	21
Contingency	8.8%	34	8.7%	36	8.7%	42
Total Plant Cost	100.0%	387	100.0%	413	100.0%	483
Working capital	17.6%	68	17.9%	74	17.8%	86
Total Capital Employed	117.6%	455	117.9%	487	117.8%	569

Table 4.6 - Conversion Factors for Module Investment Costs

Material handled	Solids	Solid/Fluids	Fluids
TPC/IPC	1.15	1.15	1.15
TPC/DPC	1.47	1.41	1.40
TPC/EC	3.87	4.13	4.83

Table 4.7 - Equipment Cost Conversion Factors

Factor	Average values for equipment in mild steel, $US\$_{1991}$ basis a (constant)	b (power)	Adjustments
Erection	1.924	-0.261	* 0.56 if low e.g. erection included as in large tanks * 1.32 if high e.g. some site fabrication * 4.26 if very high e.g. much site fabrication
Piping[a]	34.347 (liquids) 31.953 (gases)	-0.380 (liquids) -0.358 (gases)	* 0.30 if very low e.g. ducting only * 0.71 if low e.g. small diameter or ducting only * 1.42 if high e.g. large diameter or complex
Instruments	13.942	-0.330	* 0.46 if very low e.g locate only * 0.80 if low * 1.28 if high
Electrical	4.2112	-0.231	* 0.23 if very low e.g. lighting only * 0.83 if low e.g. for ancillary drives only * 1.46 if high e.g. transformers and switchgear
Civil	1.997	-0.231	* 2.25 if high * 2.90 if very high
Structures and Buildings	4.990	-0.244	* 0.83 if very low e.g. negligible * 0.35 if low e.g. open air or ground level * 1.18 if high e.g. covered building * 1.89 if very high e.g. elaborate under cover
Lagging	10.338	-0.419	* 0.61 if low e.g. service only * 1.16 if high * 1.84 if very high e.g. cold lagging

a only one value should be used, either liquids or gases

Table 4.8 - Conversion of direct plant costs to total plant costs

Item	Range	Factor Used
Engineering, design and supervision	10-20% DPC	0.15 DPC
Management overheads	5-20% DPC	0.10 DPC
Installed Plant Cost (IPC)		1.25 DPC
Commissioning	1-10% IPC	0.05 IPC
Contingency	0-50% IPC	0.10 IPC
Contractors fee	5-15% IPC	0.10 IPC
Interest during construction	7-15% IPC	0.10 IPC
Total Plant Cost (TPC)		1.45 IPC = 1.81 DPC

4.5.5 Contingency

Both the ratios and factorial methods for producing total plant costs include a factor for contingency of 10% of the installed plant cost. This contingency is included in capital cost estimations to allow for:

- changes in the project scope;
- inaccuracies in the cost data;
- inaccuracies in the process design;
- unknown cost factors such as the costs of transportation;
- changes in the value of money during construction;
- unexpected site problems such as local regulations or labour problems;
- adverse weather;
- subcontractors' delays; and
- organisational complexities.

Some of the factors above are expected to be particularly critical to less established processes where inaccuracies and unexpected costs are more likely to effect the costs. In such circumstances a higher contingency could have been added but since the capital costs used for these processes are derived from 1st plant costs (see Section 4.6.7), there is already likely to be a substantial contingency included and so a 10% contingency is used throughout.

4.6 CALCULATING PRODUCTION COSTS

Detailed production costs must account for all the direct, indirect and fixed costs incurred by the process as listed in Table 4.9 [198]. Such detail can only be achieved on the basis of historical costs for similar plant and even then differences in operation conditions, location, scale or many other factors can reduce the value of such data. Given this limitation and the lack of historical data a more basic method of calculating production costs is required. Several methods have been published for the rapid estimation of production costs [198, 202, 203, 204] by adding factors for materials, labour, utilities, capital charges and the selling price. This work will calculate the production costs by summing:

- Capital amortisation (see Section 4.6.2);
- Materials costs (see Section 4.6.3);
- Labour (see Section 4.6.4);
- Utilities (see Section 4.6.5);

- Plant overheads (see Section 4.6.6); and
- Plant maintenance (see Section 4.6.6).

This allocation of production costs has already been used in work by Cottam [150] and Double [175] to produce previous techno-economic models at Aston University.

Table 4.9 - Components of production costs

Manufacturing cost	Direct production costs	Raw materials Operating labour Operating supervision Utilities Maintenance and repairs Operating supplies Laboratory charges Royalties (if not on lump sum basis) Catalysts and solvents
	Fixed charges	Depreciation Taxes on property Insurance Rent
	Plant overhead costs	Medical Safety and protection General plant overhead Payroll overhead Packaging Staff facilities Control laboratories Plant superintendence Storage facilities
General expenses	Administrative expenses	Executive salaries Clerical wages Engineering and legal costs Office maintenance Communications
	Distribution and marketing expenses	Sales offices Salesmen expenses Shipping Advertising Technical sales service
	Research and development	

4.6.1 Time value of money

All production costs are expressed in real terms in \$US, 1995 basis, and are assumed constant throughout the life of the project. In reality, the actual production costs each year will increase in nominal terms because of inflation. However, revenues would also be expected to rise in nominal terms because of inflation and balance out the increase in costs. Therefore it is reasonable to use constant costs provided that this is applied consistently.

4.6.2 Capital amortisation

Capital amortisation is the money required to pay back the loan on capital required to set up the plant. It is calculated by the using Equation 4.8.

$$\text{Fixed charge, \$k / y} = TPC \times i \times \frac{(1+i)^{l}}{(1+i)^{l}-1} \tag{4.8}$$

where TPC = Total plant cost, \$k

i = annual nominal interest rate, %

l = length of project, years (assumed to be the same as the loan period)

This fixed charge is constant in nominal terms and must therefore be adjusted to real terms for consistency with all other production costs. The cost in real terms of capital amortisation can be calculated for each year of the project by applying Equation 4.9. An average of the annual charges is used to give the approximate cost of capital amortisation in real terms.

$$\text{Annual charge, \$k / y} = \frac{1}{(1+f)^{n}} \tag{4.9}$$

where n_x = project year

f = annual rate of inflation, %

4.6.3 Materials

The costs of wood chips before transport are defined in the input parameters on an oven dry tonne basis, and a default value of \$40/odt before transport is used as defined in Section 4.3. The cost of all other materials such as auxiliary fuels are calculated as required and are reported in the module descriptions in the following chapters.

4.6.4 Labour costs

The labour requirement per shift is estimated for each part of system based on known labour levels in existing or similar plant, and these estimations are noted as each module is reported. The labour cost is calculated by multiplying the labour requirement by the number of shifts and a cost per person per year. A 5 shift system is assumed. A labour cost of $30000/person/y is used which is a variable in the model.

4.6.5 Utilities

Utility requirements are reported in each module description. The utilities used are boiler feed water, cooling water, and electricity. The cost of boiler feed water used is assumed to be the cost of town water, taken as $1.00/t [205]. It is assumed that cooling water is abstracted from an unsupported source at a cost of $0.017/t [206].

Electricity requirements are taken from the gross generating output. However, in de-coupled fast pyrolysis plant the fast pyrolysis and pretreatment power requirements must be met by buying power or through the addition of generating capacity at the fast pyrolysis site. It is assumed that the power is supplied by the grid at a cost of $0.10/kWh (a default value that is variable in the model). It would be interesting in further work to examine how the electricity requirements of a remote fast pyrolysis plant could be met by use of part of the pyrolysis liquid since it is likely that such systems would be located in areas where electricity would not be available.

4.6.6 Maintenance and Overheads

Maintenance is charged as a percentage of total plant costs in the pretreatment and conversion modules. The default value of 4% is used which is a variable in the model. The maintenance costs for the generation modules were calculated per unit of electricity produced (in kWh), since typical costs were available. These are also variable and will be discussed in each module description.

Other fixed costs include insurance, taxes, rent and general overheads. These are typically taken as 2-10% of the total plant costs [198] and a value of 4% will be used as a default in this work, which is variable in the model. The effects of varying this percentage will be tested in the sensitivity analyses presented in Chapter 9.

4.6.7 Profit

The aim of this work is to compare the costs of electricity production. As such no profit element is included. In reality a return on the investment would be expected and profit is discussed during the systems evaluations in Section 9.4.

4.7 LEARNING EFFECTS

It is a normal feature of new processes that their initial costs reduce with experience. This phenomenon was first recognised in aircraft manufacture [204] and has also been proven in electricity generation by Lloyd [207]. The learning effect can be observed by plotting unit production costs in real terms against cumulative production on a log-log scale. The points tend to fall on a straight line with a negative slope of between 20-30%. This percentage reduction in cost per doubling of cumulative production is referred to as the learning factor [208]. Strictly speaking the line defined above is an experience curve because it combines all the elements of cost reduction including increased productivity through learning, innovations and scale economies. Learning curves are limited to cost reductions through increased productivity but the two terms tend to be synonymous in the biomass community.

Learning may be manifested in two ways: an improvement in equipment performance and a reduction in capital costs. Performance improvements are likely in new processes but their impact is difficult to predict and no attempt is made in this work to adjust performance to account for future plant operation. Instead, it is assumed that their will be some reduction in capital costs as the more novel technologies become established. Elliott and Booth [24] support the view that that plant replication will cut capital costs in the GEF Brazilian IGCC plant (see Section 3.8.4) from $2500-3000/kW_e$ to $1300-1500/kW_e$ after five to ten plant replications. These reductions will be achieved by virtue of value engineering (adjusting process design, operating parameters and process design to minimise production costs), replication of standard components and the use of factory built modules. A halving of capital costs by the 10th plant is equivalent to a learning factor of 20%, and is in line with orthodox process economics for novel processes. A learning factor on capital of 20% will also be assumed in this work.

While this learning factor may be applied to all capital costs, the status of the technologies in each system should also be considered. The combustion plant cost data is based on established technology since many combustion plant have been built. Combustion plant costs in this work are defined as 100th plant costs, and as such it would require the construction of 100 new plant to achieve a capital cost reduction of 20%. Thus the total plant costs are tending towards an asymptote as seen on Figure 4.3.

The fast pyrolysis and gasification module cost data is, in contrast, based on first or near-first plant cost data and the costs produced from this data are regarded in this work as first plant costs. As such the costs will rapidly reduce in successive systems as the number of plant in operation doubles. This rapid reduction in plant costs during the early development of a process is shown in Figure 4.3. It is debatable, however, whether such a dramatic reduction in plant costs would occur if the whole system is considered: parts of the system such as feed pretreatment have already been established, and could be considered 100th plant cost if considered in isolation.

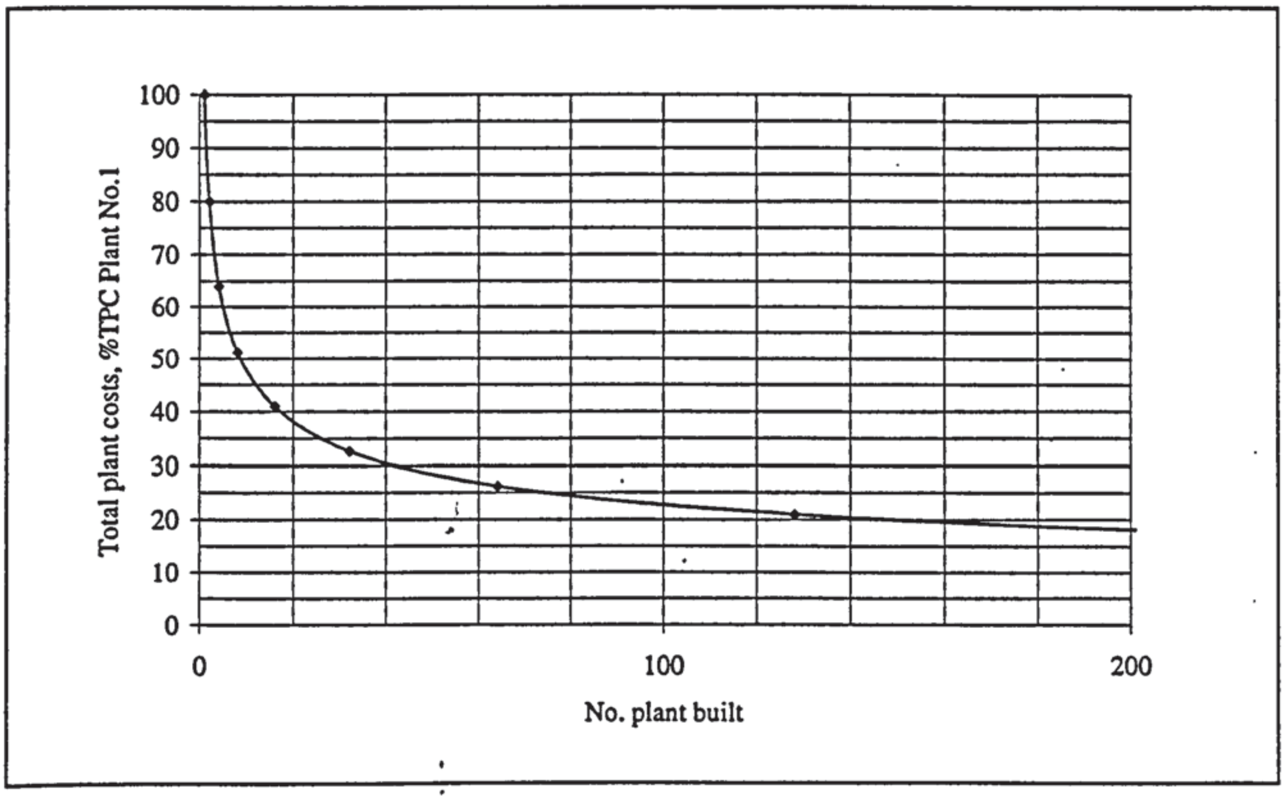


Figure 4.3 - Capital cost reductions with plant replication

Most of the results in Chapter 9 will calculate the costs for a plant to be built now using current costs and technology. Under these conditions the combustion system will be considered 100th plant and all other systems will be 1st plant. The costs of future plant will then be evaluated in two ways. In both cases the costs of the combustion plant is kept constant and the costs of the 10th system using each of the novel technologies is calculated by applying a 20% learning factor and either:

- assuming all parts of the systems are 1st plant cost; or
- assuming only the conversion module costs are 1st plant and that the rest of the system costs (pretreatment, generating cycle and grid connection) are for 100th plant.

The plant cost for the 10th plant (after 9 replications) is calculated using Equation 4.10 and Equation 4.11. Equation 4.10 is used to regress the costs of the current plant to 1st plant costs if the data is considered 100th plant. Equation 4.11 then calculates the capital cost expected for the future system.

$$TPC_1 = \frac{TPC_{now}}{(1-LF)^{\frac{\ln(n)}{\ln(2)}}} \tag{4.10}$$

$$TPC_{future} = TPC_1 \times (1-LF)^{\frac{\ln(n+r)}{\ln(2)}} \tag{4.11}$$

where TPC_1 = total plant cost of the first plant, \$k

TPC_{now} = total plant cost based on current costs used in the module, \$k

TPC_{future} = total plant cost of future plant after replications, \$k

LF = learning factor, %

n = plant number of current plant costs

r = number of replications

4.8 NOMENCLATURE

a,b... = Constants

DPC = Direct plant cost

$E_{th,del}$ = Energy available in the feed delivered to the conversion facility, GJ/y LHV basis

$E_{th,pret}$ = Energy available in the feedstock as fed to the conversion reactor after pretreatment, GJ/y LHV basis

$E_{th,conv}$ = Energy available in the energy carrier produced in conversion, GJ/y LHV basis

$E_{th,aux}$ = Energy available in the auxiliary fuel used in power generation, GJ/y LHV basis

$E_{e,gross}$ = Gross electricity output before internal consumption, GJ/y

$E_{e,net}$ = Net electricity output after internal consumption, GJ/y

$E_{e,\,pret}$ = Internal electricity consumption in the pretreatment module, GJ/y

$E_{e,\,conv}$ = Internal electricity consumption in the conversion module, GJ/y

$E_{e,\,gen}$ = Internal electricity consumption in the generation module, GJ/y

EC = Equipment cost

HHV = Higher heating value, GJ/odt

i = nominal interest rate, %/y

IPC = Installed plant cost

LF = Learning factor

LHV = Lower heating value, GJ/odt

N = Plant number of a plant using current costs

n = number of years

n_x = project year

odt = oven dry tonne, the mass at 0% moisture content

r = The number of plant replications

TPC = Total plant cost

5. TRANSPORT MODULES

5.1 INTRODUCTION

This chapter describes the development of the transport modules that calculate the transport distances and costs incurred during the transport of wood chips from source to the conversion facility. Relationships for pyrolysis liquid transport distances and costs in de-coupled systems are also developed.

Road transportation is assumed in all cases for the reasons described in Section 5.2. The relationships required to calculate the transport distances are developed in Section 5.3. Feed transport costs are discussed in Section 5.4 and the costs of transporting pyrolysis liquid are discussed in Section 5.5.

5.2 SELECTION OF TRANSPORT SYSTEM

Bulk transport of biomass can be achieved by road, rail or water. Road transport is the normal mode of transport in bioenergy systems [209] since it offers flexibility and is particularly suited to systems where the material is transported over distances of less than 100 km [210, 211]. Biomass to electricity facilities require a low cost feedstock and this tends to limit transport distances to this range. The cost of long distance haulage is only recoverable if the biomass has a high value, for example as feedstock for the pulp and paper industries. The US Department of Energy have predicted that maximum haulage distances will be 65-97 km, with a mean distance of 40 km [29] whereas rail and waterway transport are suited to distances over 150 km [57, 211]. Therefore road transport is assumed for all cases.

5.3 TRANSPORT DISTANCES

Locating the feed conversion facility and the electricity generating site (one and the same in a close-coupled system) is recognised as one of the most important aspects of a viable system since both feed transport costs and grid connection costs are affected [212]. A full analysis of transport distances can only be performed on a case by case basis, taking account of actual feed production areas, local topography, the road network and other case-specific features. Such an approach would not offer the consistency and flexibility required for this work and instead a simplified network of feed production sites, feed conversion sites and (where appropriate in de-coupled sites) electricity generating sites is developed here. The assumptions that are used simplify a complex issue but ensure that every system is treated equally and objectively in such a way that all system configurations can be evaluated consistently.

5.3.1 Key assumptions

The following assumptions have been made:

1. Feed production is evenly distributed over a circular feed supply area (see Section 5.3.2).

2. Where there is more than one feed conversion facility, each conversion facility is identical. The total feed supply area is split into identical sectors and each sector supplies feed to a single site (see Section 5.3.3).

3. Each feed conversion facility is located at that point in its feed supply area that minimises the total direct distance from all of the feed sources to the conversion facility (see Section 5.3.4).

4. The road network is regular and symmetrical such that a single "winding factor" can be used to convert the direct distance between source and conversion facility into an actual road distance (see Section 5.3.5).

5. In de-coupled fast pyrolysis systems, either conversion or generation must take place at a single site located at the centre of the feed supply area (see Section 5.3.6)

5.3.2 Feed production within the feed supply area

It has been assumed that feed production is evenly spread throughout the supply area. An alternative and probably more likely scenario (used by McIlveen-Wright [186]) is that feed production would be more concentrated in the immediate vicinity of the conversion facility. This has not been used here because it would complicate the analysis of multiple sites. The assumption of a constant feed production density is more likely to be true for small scale systems and there may be an error introduced as capacities near 100 MW_e.

The size of the feed supply area is calculated from Equation 5.1. In this equation the actual land used only in feed production is adjusted to give the total area required for all uses by application of a land area limitation. The land area limitation is the percentage of the total area that is used in feed production.

$$A = \frac{Q_{del}}{Y} \times \frac{1}{LAL} \qquad (5.1)$$

where Q_{del} = the amount of feedstock required by all conversion facilities, odt/y
Y = the wood production yield, odt/ha/y
LAL = the land area limitation, %

A default of 10 odt/ha/y is used for the land yield. A default of 5% is used for the land area limitation in accordance with the previous work presented in Appendix A. Both are variables in the model and their impact will be investigated by sensitivity analysis.

From this the radius of the feed supply area is calculated using Equation 5.2.

$$\text{Feed Supply Radius (R), km} = \frac{1}{\pi} \times \sqrt{A, \text{ha}\left[\frac{10^{-2}\,\text{km}^2}{\text{ha}}\right]} \qquad (5.2)$$

5.3.3 The feed supply area for multiple conversion facilities

Two alternative scenarios were considered as ways of supplying feed to multiple conversion facilities: the circular feed supply area could be split into sectors or the total feed supply area required could be made up of individual smaller circles. The two options are shown in Figure 5.1. Two differences emerge between the options:

1. As the number of feed conversion facilities increases, if diminishing circles are used (or any other shape) then a decision must be made about how they must be positioned. This subjective analysis is avoided by using sectors.

2. If de-coupled multiple fast pyrolysis facilities are used to supply fast pyrolysis liquid to a central diesel generator then Figure 5.1(f) and Figure 5.1(g) would give different pyrolysis liquid transport distances. One option would have to be selected and this may not be appropriate to all cases. Again, splitting the total area into sectors avoids this issue.

It was decided in the light of the above that dividing the total feed supply area into sectors provided the more consistent approach.

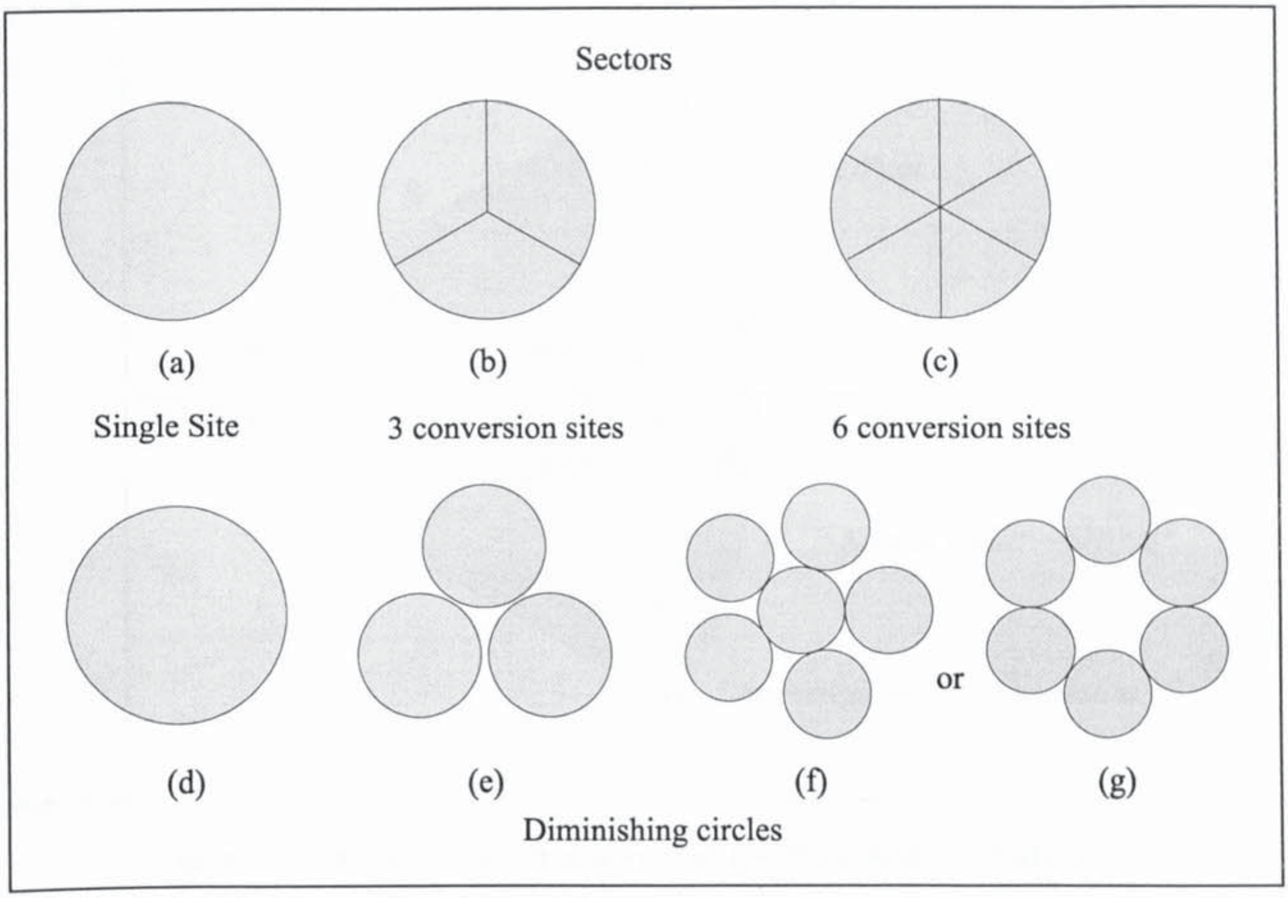


Figure 5.1 - The conceptual feed supply scenarios

5.3.4 The average direct distance from feed source to conversion facility

It has been assumed that all conversion facilities will be located at that point in its feed supply area that minimises the total distance between feed sources in the feed supply area and the conversion facility. This is the ideal case that is unlikely to be achieved in reality but that does ensure all systems are treated equally. The total distance between the feed sources and the facility that they serve can be calculated for any number of conversion sites with reference to Figure 5.2. An analogy with 1st moments of area has been used to locate the feed conversion facility. It this analysis each elemental area exerts a "moment" on the conversion facility equal to the product of the elemental area and the distance to the conversion facility. Continuing this analogy, the location of the point in a given area that minimises the total moment (and therefore the total distance to the elemental areas) is the centroid of the area [213]. Thus the conversion facility will be located at the centroid of the circle (for a single conversion facility) or the centroid of a sector (for multiple cases).

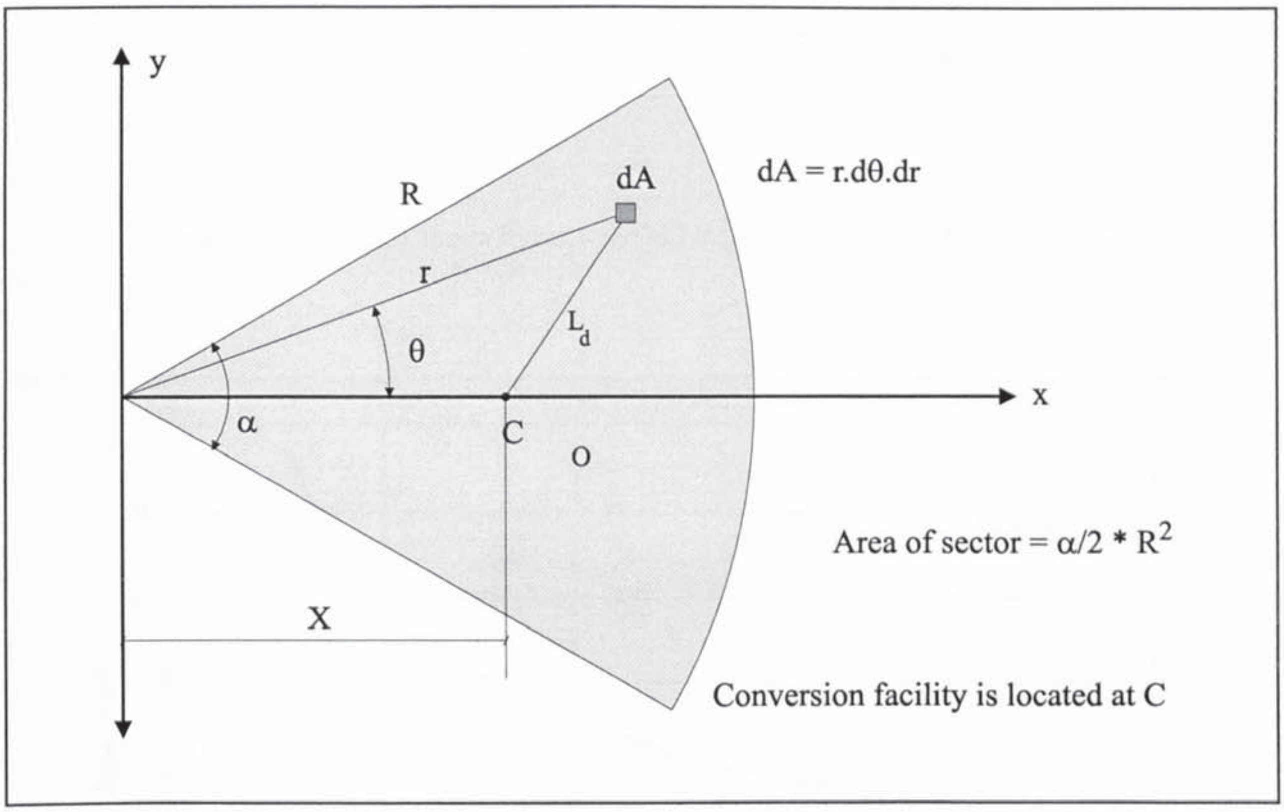


Figure 5.2 - Analysis used to derive average direct transport distances

The location of the centroid is a known property of a sector and the distance X shown in Figure 5.2 can be calculated using Equation 5.3. If a single conversion facility is used then X = 0, confirming that the feed conversion facility would be located at the centre of the feed supply area.

$$\text{Distance to centroid (X), km} = \frac{2}{3} \times \frac{\sin(\alpha/2)}{\alpha/2} \times \text{R} \qquad (5.3)$$

where α = the sector angle, rad = 2π/(the number of conversion facilities)
R = the radius of the total feed supply area, km

The average distance from any element in the sector is given by calculating the total moment for all elemental areas and dividing by the sector area. Equation 5.4 has been derived from Figure 5.2 and can be used to calculate the average direct distance to the conversion site. The double integral has been evaluated numerically to give the results shown in Figure 5.3. Thus the average direct distance between a feed source and the feed conversion facility that it serves is calculated using Equation 5.5.

$$\text{Direct distance } (L_D),\ \text{km} = \frac{2}{\alpha}\int_0^R\int_0^\alpha r.\sqrt{(r^2 + X^2 - 2.r.X.\cos\theta)}\,d\theta\,dr \tag{5.4}$$

$$\text{Direct distance } (L_D), \text{km} = R, \text{km} \times 0.6747 \times (\text{No. conversion facilities})^{-0.4647} \tag{5.5}$$

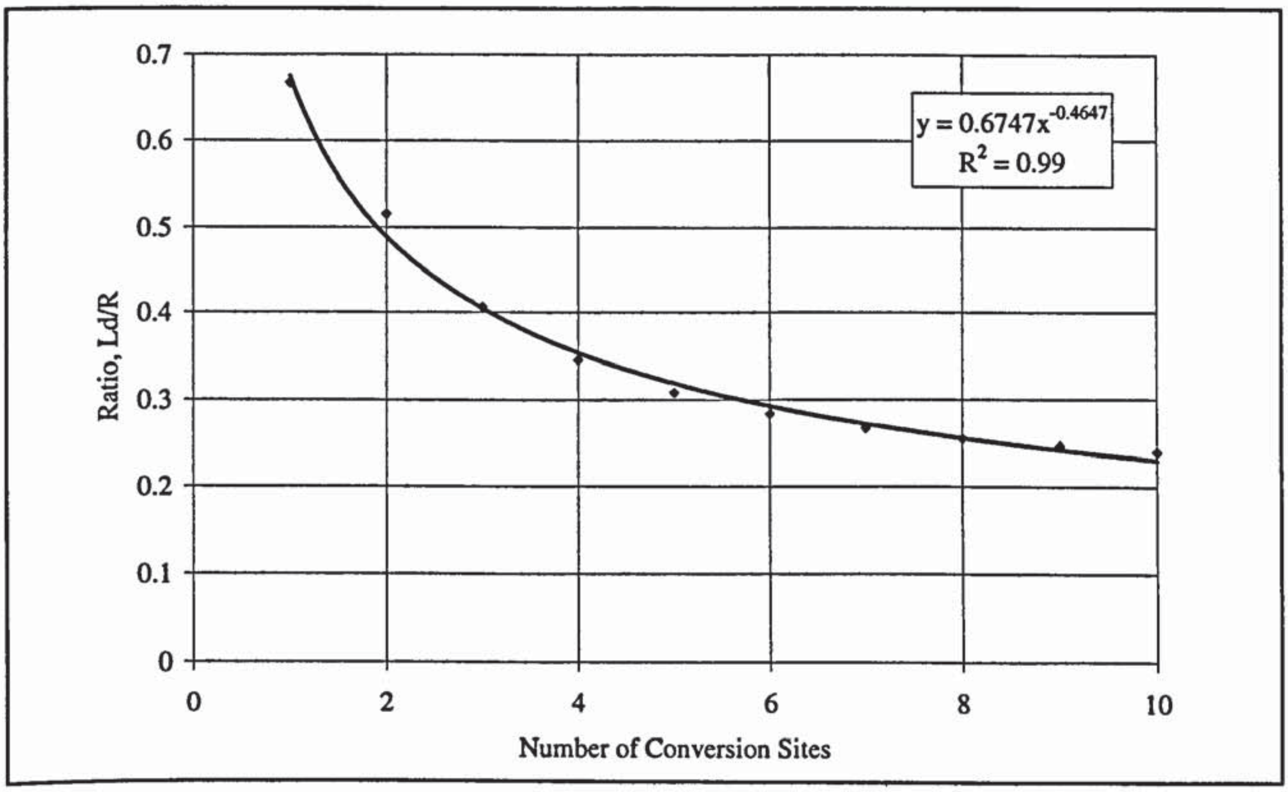


Figure 5.3 - Relationship between the number of sites and the average direct transport distance

5.3.5 The average road distance from feed source to conversion facility

The actual distance travelled between the feed source at the conversion facility will be higher than the direct distance because the transport vehicles must follow the existing road network. Thus the direct distances are modified by using the general case shown in Figure 5.4. This assumes such that the actual route taken between a feed source and the feed conversion facility runs along two sides of a right-angled isosceles triangle. With reference to Figure 5.4:

$$AB^2 = AC^2 + CB^2 = 2AC^2 \qquad \therefore AB = \sqrt{2}AC$$

$$\text{Winding factor} = \frac{L_r}{L_d} = \frac{2AC}{AB} = \frac{2AB}{\sqrt{2}} \times \frac{1}{AB} = \sqrt{2}$$

Thus the mean direct distance given in Equation 5.5 is multiplied by a winding factor 1.41 to give the actual road distance travelled. The determination of an actual winding factor for each system would require much more rigorous analysis using geographical information systems and would be particular to a specific area. A recent study by Marrison [185] has performed such an analysis and produced a winding factor of 1.4. While this figure is only pertinent to the particular case studied, it does confirm that the value used in this study is reasonable.

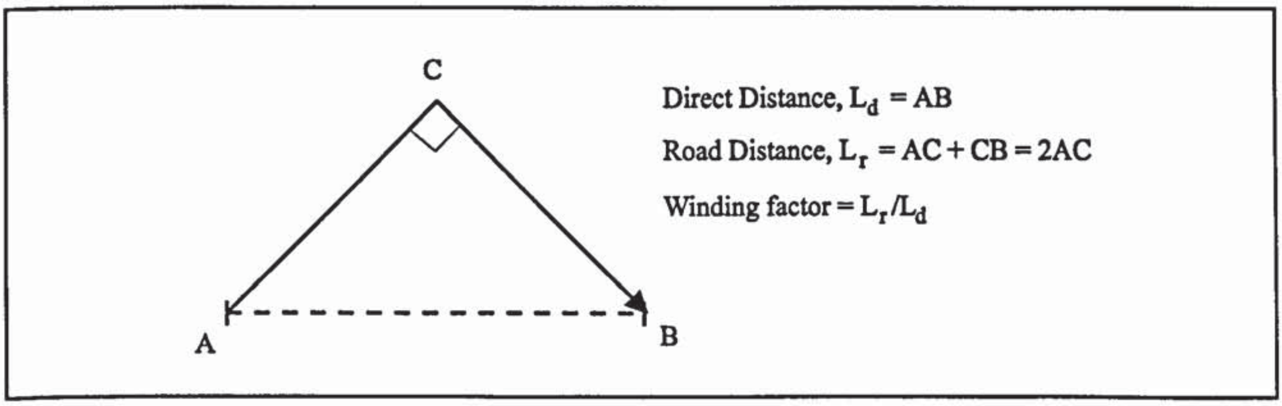


Figure 5.4 - The assumed road scheme used to give a winding factor

5.3.6 Transport distances used in de-coupled fast pyrolysis systems

The system configurations that may be introduced if the conversion technology is fast pyrolysis was discussed in Section 2.7.6. The transport modules must be able to calculate the transport distances involved in systems that use either multiple fast pyrolysis sites to supply a single generator or multiple generators supplied by pyrolysis liquid from a single fast pyrolysis facility. To simplify the analysis and maintain consistency the layout of the system in these two cases is as shown in Figure 5.5. In each case the multiple sites are assumed to be located at the centroid of a sector and the distance from the centre of the feed supply area to each site is calculated using Equation 5.3. The single site, whether the fast pyrolysis site or the generating site, is located at the centre of the feed supply area. The direct liquid transport distance is given by Equation 5.3 in both cases and the winding factor defined in Section 5.3.5 is applied to give the road transport distance.

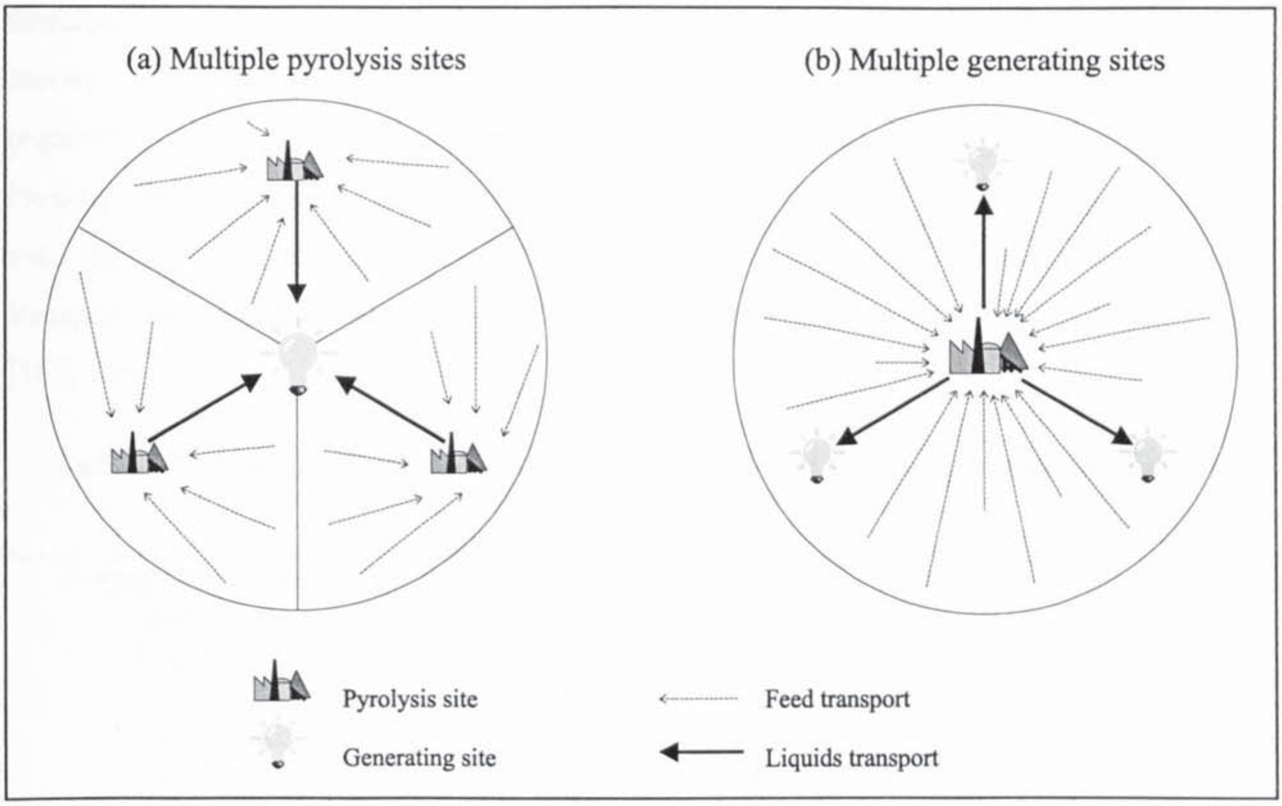


Figure 5.5 - Layout of Systems using Multiple Pyrolysis or Generating Sites

5.4 FEED TRANSPORT COSTS

Transport costs are influenced by the carrying capacity of the vehicles, since this dictates the number of trips that are required to supply the feed. Vehicles are regulated in terms of maximum vehicle dimensions and their maximum payload, such that the carrying capacity of a vehicle will be limited either by volume or mass depending on the bulk density of the feedstock as exemplified by Table 5.1 [211]. If the volumetric limit is reached first then transport costs must be calculated on a unit volume basis; conversely, achieving maximum payload would mean that transport costs must be calculated on a mass basis.

Hauliers aim to maximise the load carrying capacity of their vehicles by reaching the payload limit requiring the bulk densities shown in Table 5.1 [209, 211]. Large tree sections tend to have lower bulk densities because of the unfilled space between pieces, but comminuted wood can reach the required limit. This is one of the reasons why feed is usually transported as chips, supported by improved handling characteristics and increased vehicle options [59, 60, 214]. In most cases the bulk density of wood chips would be sufficient to reach maximum payload. McDonald has examined wood chip bulk densities for various species and moisture contents and recorded values of 225-479 kg/m^3 [211]. In a few cases closer packing would therefore be required but McDonald and Hankin have shown that vibration, compaction or simply the force of blowing the chips from a chipper into the truck are sufficient to give the required bulk densities. There are insufficient data available to assess the extra costs of such loading methods. The current module will assume that the feedstock is transported with a bulk density sufficient to reach the payload of the vehicle in all cases. Thus the costs of transport are calculated on a mass basis, an assumption that has been used in other studies [185, 186, 215].

Table 5.1 - Limiting bulk densities in relation to truck size and payload in Finland

Overall vehicle length, m	12	16	18	22
Load length, m	10.2	12.5	7.5 + 7.5[a]	8.0 + 10.5[a]
Load width, m	2.3	2.3	2.3	2.3
Load height, m	2.4	2.4	5.4	2.4
Load volume, m^3	55.4	67.9	85.0	100.4
Total weight, t	22.0	36.0	42.0	42.0
Payload, t	11.8	21.2	26.5	25.8
Limiting bulk density, kg/m^3	213	312	312	257

a Includes trailer

Total transport costs are a sum of fixed charges for loading and unloading the feedstock and charges to cover transport that are variable with distance. Unfortunately most of the data available in the literature simplifies this relationship to a single variable charge. Only data that separated fixed and variable costs is shown in Table 5.2. The costs for transport of a wet tonne of feedstock were calculated for distances between 0 and 100 km for each of the values shown in Table 5.2. The data produced was used in Figure 5.6.

The feed transport module uses the mean transport cost shown in Figure 5.6, resulting in Equation 5.6.

$$\text{Wood chip transport cost, } \$/t = 2.60\$/t + \left(0.090\$/t/km \times L_r,\ km\right) \qquad (5.6)$$

It can be seen from Figure 5.6 that the costs of feed transport can vary widely, reflecting differences in vehicle, road network, fuel costs and feed characteristics. There is a lot of uncertainty associated with the fixed and variable charges used, especially since no data could be found in the literature for UK conditions (some data containing just variable costs was available by Mitchell [51]). The effect of changing transport costs will be evaluated by sensitivity analysis.

Table 5.2 - Feed transport costs

Location	Year	Handling charge $/t	Transport charge $/t/km	Source
US	1991	0.22	0.04	[216]
US	1991	2.43	0.04	[216]
Brazil	1992	0.67	0.08	[185]
Netherlands	1990	1.42	0.07	[42]
Netherlands	1990	1.24	0.12	[42]
Canada	1992	0.96	0.01	[217]
Netherlands	1995	0.39	0.05	[218]
Netherlands	1995	0.38	0.09	[188]
Netherlands	1995	0.82	0.06	[188]
New Zealand	1995	4.89	0.12	[50]
New Zealand	1995	8.92	0.13	[50]
New Zealand	1995	3.31	0.11	[50]
US	1990	1.15	0.18	[219]

a *handling charge includes loading and unloading*

b *transport charge is based on the one way distance between feed source and feed user. The return trip is included.*

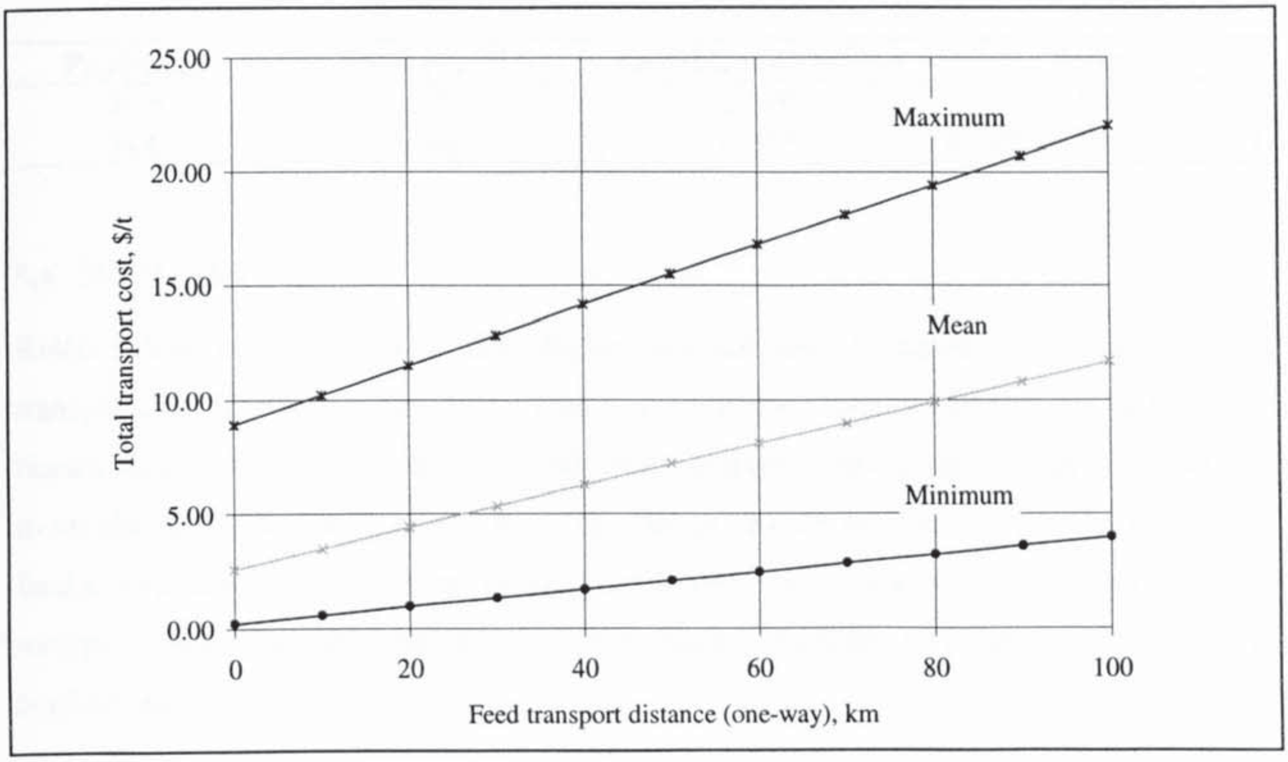


Figure 5.6 - Transport costs reported in the literature

5.5 PYROLYSIS LIQUID TRANSPORT COSTS

No data was available in the literature for pyrolysis liquid transportation. Instead, tanker haulage transport charges for fuel oil distribution in the UK are used. Two costs were obtained, as shown in Table 5.3. The variation in fixed charge is notable: the Shell data is composite data for the whole oil industry and includes the fixed cost burden of the distribution centres; Linkman on the other hand reported that the fixed cost would be negligible since it is simply the cost of pumping the liquid. The latter case has been assumed here. The validity of this assumption is questionable and it is recommended that future work examines this aspect of the system costs in more detail especially since no account has been made for the special properties of the pyrolysis liquid. For example the acidity of the liquid will dictate the use of specialist tankers and the viscosity of the pyrolysis liquid may demand special pumping equipment and heated lines at the pyrolysis and generating sites. The default transport cost for liquids is calculated using the fixed and variable charges shown in Equation 5.

$$\text{Pyrolysis liquid transport cost, } \$ / t = 0.00\$ / t + \left(0.50\$ / t / km \times L_r, km\right) \quad (5.6)$$

Table 5.3 - Fuel oil transport costs

Payload, t	Fixed cost, $/t	Variable cost, $/t/km	Source
30.5	5.27	0.048	Shell UK [220]
24.0	none	0.053	Linkman Tankers [221]

5.6 SUMMARY

Relationships are described in this chapter that are used to calculate the cost of feed transportation from the source of the feed to a conversion facility and the cost of pyrolysis liquids transportation in a de-coupled fast pyrolysis-based system from a fast pyrolysis facility to an electricity generating plant. A set of assumptions are used to simplify the analysis of feed availability and the locations of the various conversion and generating sites to allow the transport costs to be calculated on a consistent basis irrespective of plant capacity or system configuration.

Feed transport costs are important on the premise that as plant capacity increases, the transport costs element of a delivered feed cost will increase. This negative scale economy counters the cost reductions that can be achieved through the construction of larger conversion and generation plant. In de-coupled fast pyrolysis systems it is important to calculate the costs of liquid transport so that the relative costs of feed and liquid transport can be assessed and used in the decision on where the fast pyrolysis plant would be best positioned: at the feed source or with the generator.

The assumptions that are made to produce a consistent analysis have simplified the issue enormously, but a more exhaustive analysis would have required a case-by-case examination of the transportation variables such as appropriate vehicles, the road network and the actual locations of the feed sources and conversion facilities. While this would be important when implementing a particular system, this type of analysis cannot be applied at the generic level.

There are some areas where the model could be improved with further work.

- Specific costs for wood chip transport under UK conditions are required.
- The option of calculating transport costs would be useful to improve accuracy when transporting low bulk density feedstocks.
- The extra costs associated with the fast pyrolysis liquid characteristics such as high viscosity and low pH should be evaluated.

Ultimately, significant resources should not be invested in applying transport costs to a generic model. The costs are too-case specific and work by Marrison [185] and McIlveen-Wright [186] has already demonstrated that capital costs impose a much greater impact with changes in scale.

5.7 NOMENCLATURE

A = The total land area that supports the system, ha

Q_{del} = The total feed required by the system, odt/y

Y = the wood production yield, odt/ha/y

LAL = the land area limitation, the percentage of the total land available that is used in feed production, %

R = the radius of the total land area A, km

r = the direct distance from the centre of A to a feed source, km

L_d = the direct distance from a feed source to the conversion facility that it supplies, km

α = the sector angle of a feed supply sector that supplies a conversion facility, rad

X = the distance from the centre of a circle to the centroid of one of its sectors, km

L_r = the road distance travelled in transporting either feed or liquid, km

6. THE FEED PRETREATMENT MODULE

6.1 INTRODUCTION

This chapter describes the development of the feed pretreatment module. The aim of the pretreatment module is to model the cost and performance of all the equipment necessary to ensure a reliable supply of a wood feedstock with characteristics that match the requirements of the feed conversion process. In all systems the feed will be delivered as chips with the characteristics described in Section 4.3. The need for feed pretreatment was introduced in Section 2.4, where it was shown that some handling, storage and processing of the delivered feedstock is usually required at the feed conversion facility.

The development of a flexible techno-economic model for the reception, handling, storage and pretreatment (hereafter collectively referred to as "pretreatment") of wood chips at the feed conversion facility prior to feed conversion is complicated by the number of operations required, the range of feed characteristics involved and the wide capacity range. For this reason, pretreatment has been broken down into a number of steps. These steps are modelled in isolation and the overall pretreatment requirements are met by a sequence of steps, shown schematically in Figure 6.1. Transfer between steps is always included in the upstream step of the two steps.

The equipment requirements, performance, capital cost and operating costs for each step are discussed in the following sections, as shown in the list below.

1. Reception - Section 6.2
2. Screening (large scale systems only) - Section 6.3
3. Bulk storage - Section 6.4
4. Screening - Section 6.5
5. Re-chipping - Section 6.6
6. Drying - Section 6.7
7. Grinding (fast pyrolysis systems only) - Section 6.8

8. Buffer storage - Section 6.8.2

Capital costs and power consumption for each step and options within each step have been calculated from the main equipment items required. The data sheets that were used are presented in Appendix C. Labour and power requirements are considered for each step as the steps are described. Operating costs for the entire module are summarised and discussed in Section 6.10.

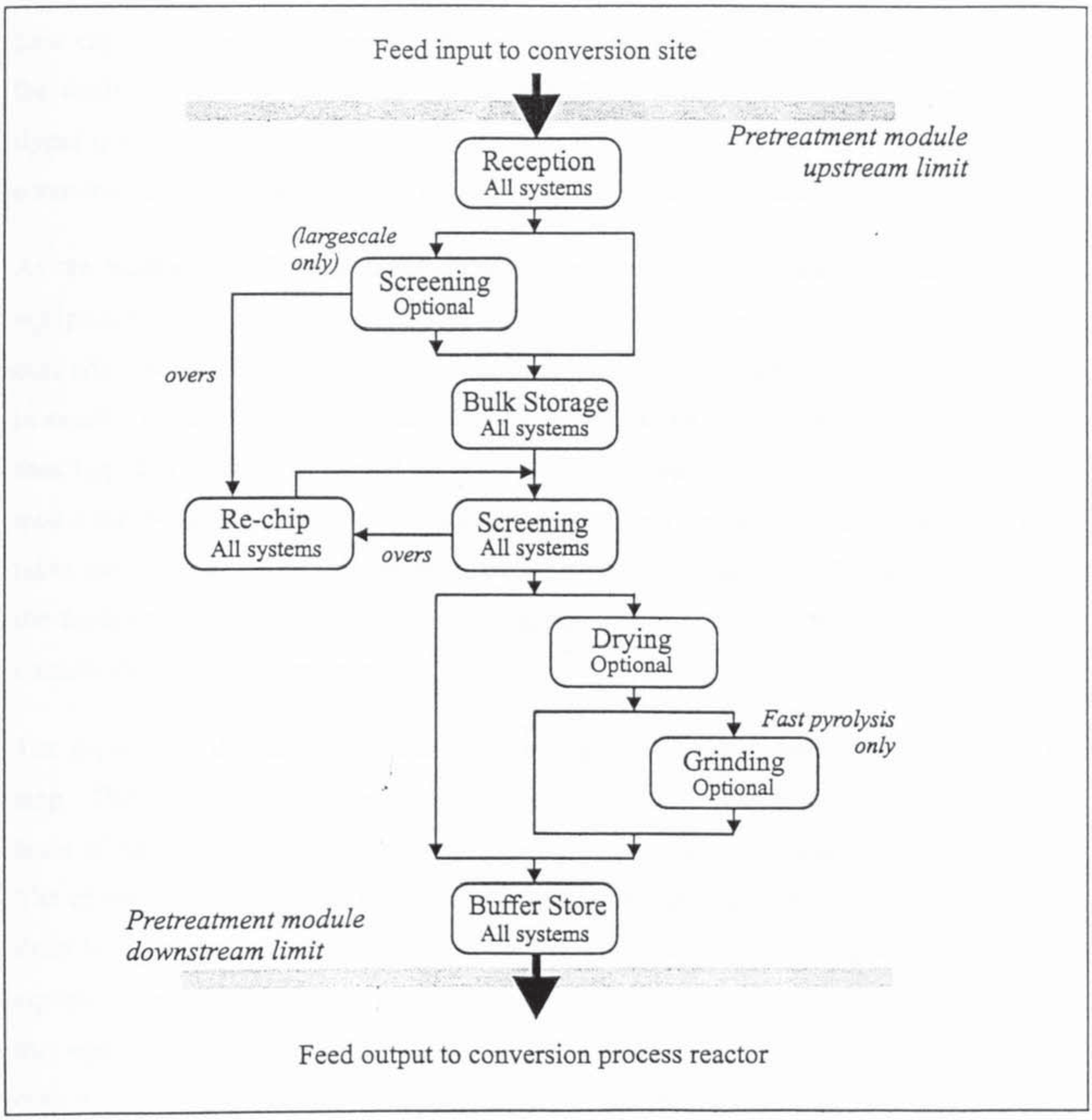


Figure 6.1 - Sub-modules in the pretreatment module

6.2 FEED RECEPTION (ALL SYSTEMS)

The reception step includes the equipment necessary to unload the incoming vehicles and record the deliveries for stock control and payment of suppliers. The key part of the reception step is the equipment used to unload the trucks since laden trucks should not be held up at the facility before unloading. This can incur high costs [222] and cause a bottleneck to the rest of the system. Thus as the frequency of truck deliveries increases equipment must be installed that can rapidly unload their cargo.

Low capacity systems neither need nor can afford unloading equipment [223]. In such cases the facility would rely on self-unloading vans (fitted with walking floors for example) or tipper trucks. The extra equipment required on the vehicles reduces their payload and their economic range but this should not impact on a small capacity system.

As the number of deliveries increases, it becomes necessary to install specialist unloading equipment called truck dumpers. These allow the use of larger capacity vehicles that are more cost effective over long hauls and more rapid unloading. The truck and trailer (or trailer only in smaller truck dumpers) are attached to a platform on the truck dumper and the platform is then tipped to an angle of around 60° whereupon the feed falls under gravity from the trailer into a receiving pit. Trailer-only truck dumpers offer a capital cost saving but each trailer takes longer to unload because the truck and trailer must be separated. Truck dumpers unload the feedstock into a live-bottomed receiving bin which in turn discharges to a conveyor to transfer the feed to the next step.

The capacity of the various unloading options has been used to define three options in this step. The limits for each system have been developed in Table 6.1 [222, 223, 224] on the basis of delivery on weekdays and during daylight hours only (10 hours/day for 5 days/week). The unloading time required for each delivery and its capacity gives a theoretical maximum daily feed input. Since the frequency of deliveries is unlikely to be regular, feed reception equipment must be adequately sized to cope with high volume periods. One rule of thumb is that equipment should be able to accommodate delivery of half the expected daily capacity over a third of the working day [222]. This give the actual capacities shown in Table 6.1. Thus three equipment options are used:

- Small scale (<293 t/d) - no unloading equipment;
- Medium scale (293-720 t/d) - a small truck dumper is used;

- Large scale (>720 t/d) - multiple truck dumpers are used, each with a maximum daily capacity of 1350 t/d.

Table 6.1 - Maximum unloading capacities

Unloading equipment	Unloading time, min	Truck capacity, t[a]	Maximum capacity, t/d[b]	Actual capacity, t/d	Approximate output, MW_e [c]
None	30	22	440	293	6.6
Small truck dumper	15	27	1080	720	16.3
Large truck dumper	8	27	2025	1350	30.6

a *t in this thesis is always used to refer to wet tonnes, odt are used to define mass on a dry basis*
b *Delivery 10 h/d, 5 d/week*
c *Feed LHV 19.2 GJ/odt, moisture content 50%, generating efficiency 32%, generation 8000h/y*

The complications that can arise in designing feed pretreatment systems are immediately apparent at this stage: what combination of truck dumpers should be used to meet a particular capacity? According to the data in Table 6.1 a daily delivery rate of 1400 t/d could be met by either 2 small truck dumpers, two large truck dumpers or one of each. On a capital cost basis 2 small truck dumpers would be the best solution while having 2 large truck dumpers would allow the system to continue operating at almost full capacity if one of them were to break down. The combination of large and small truck dumper would probably offer the best compromise. Such questions arise whenever multiple trains (i.e. replicated equipment) are used to meet the system capacity. In this case it is assumed that capacities over 1350 t/d are met by multiple large truck dumpers. Cost reductions could be achieved through other unloading equipment combinations but any other solution would have been unduly complicated.

The reception step should also record the feed as it enters the site to aid stock control and the payment of suppliers. At the small scale specific equipment for weighing the delivery trucks to determine their loads is too expensive. Instead, stock control is carried out either on a volume basis (by the truck load), or through the use of roadside truck scales [223]. Weighing the trucks on entry to and exit from the site is a better system that is cost effective in the medium and large scale options. Thus both will include a truck scale, with the medium scale system using a smaller scale that only weighs the trailer. Just as in the case of the small scale truck dumper, this offers a capital cost saving but reduces the maximum frequency of trailers that can be accommodated.

Each option must include the transfer of material to the next step. In the small scale reception option it is assumed that transfer to the next step (feed storage) is achieved using a front end loader. In the medium scale reception system the size of the feed storage pile would be too large for a front-end loader and so an open belt conveyor is used to transfer feed to storage. Ideally, the feed should be screened before transfer by conveyor to protect the conveyor and prevent blockages at transfer points [225]. However, all feed will be screened immediately after storage anyway to ensure the feed delivered to the reactor is in the correct size range. In view of this, and because the transfer conveyor will not be complex at the medium scale, feed screening is not included in systems at this scale to save costs.

At large scale the distance between the feed reception area and feed storage can increase to several hundred metres and the conveying system would become very susceptible to blockages. Therefore the large scale reception option includes a short open belt conveyor to pre-storage screening (see Section 6.3).

6.2.1 Capital costs

Total plant costs have been calculated from equipment costs for the main equipment items using the factors method described in Section 4.5.4. These costs were compiled from various sources [181, 224, 226, 227] into a data sheet presented in Appendix C. The data sheet was used to calculate the capital costs of each option for various feed input capacities, with the results shown in Figure 6.2. At the medium and large scale the basic cost is due to the truck dumper(s), with a slight variation in costs due to the changing capacity of the main output conveyor. An average of the costs for each step, corresponding to an additional truck dumper, was used in the model. Thus the capital costs for each system capacity are as presented in Table 6.2.

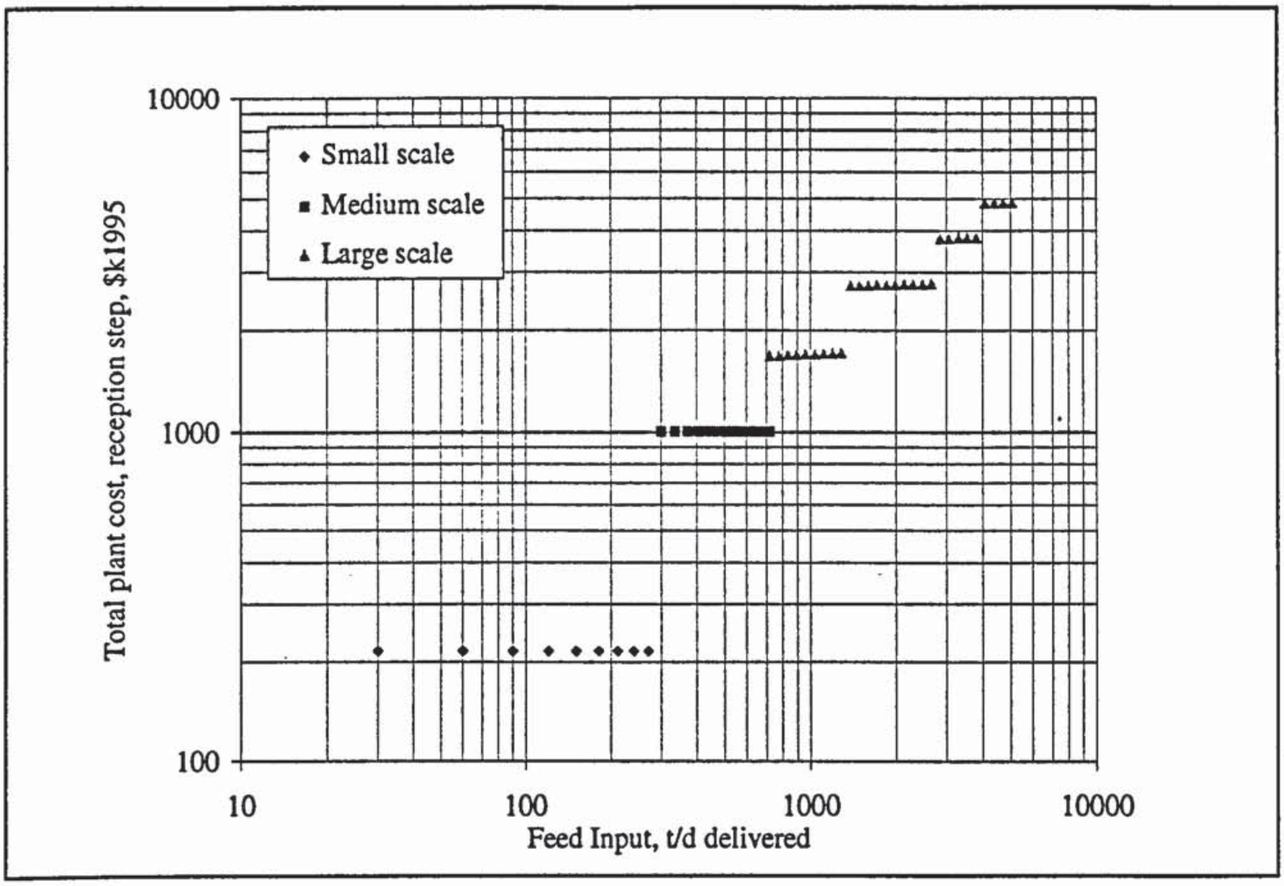


Figure 6.2 - Total plant costs, reception step

Table 6.2 - Total plant cost relationships, reception step

Defining capacity = $Q_{d,del}$ wet tonnes per day delivered		
Option	**Constraints**	**TPC relationship**
Reception, small	$Q_{d,del} < 293$	$TPC_{rec} = 206$
Reception, medium	$293 < Q_{d,del} < 720$	$TPC_{rec} = 1004$
Reception, large	$720 < Q_{d,del} < 1350$	$TPC_{rec} = 1687$
	$1350 < Q_{d,del} < 2700$	$TPC_{rec} = 2746$
	$2700 < Q_{d,del} < 4050$	$TPC_{rec} = 3802$
	$4050 < Q_{d,del} < 5400$	$TPC_{rec} = 4845$

6.2.2 Power consumption

The installed power requirement for the truck scales and conveyors was taken from [181, 224] and is included in the data sheets in Appendix C. The power consumption for this step is calculated using the relationships in Table 6.3. The power consumption for all steps is discussed in Section 6.10.2.

Table 6.3 - Installed power requirements, reception steps

Defining capacity = Q_d, wet tonnes per day delivered		
Option	**Constraints**	**Installed power required, kWe**
Reception, small	$Q_{d,del} < 293$	None
Reception, medium	$293 < Q_{d,del} < 720$	$P_{rec} = 79$
Reception, large	$720 < Q_{d,del} < 1350$	$P_{rec} = 123$
	$1350 < Q_{d,del} < 2700$	$P_{rec} = 241$
	$2700 < Q_{d,del} < 4050$	$P_{rec} = 363$
	$4050 < Q_{d,del} < 5400$	$P_{rec} = 484$

6.2.3 Labour requirements

Throughout the pretreatment module labour requirements are estimated for each step by adding labour requirements for each main equipment item, based on tables of operator requirements published by Peters [198] and Ulrich [228]. The following operator requirements are assigned for the equipment required in the various reception options:

- Front end loaders: 1 operator
- Truck dumpers: 1 operator
- Drag chain conveyor: 0.2 operator
- Belt conveyor: 0.2 operator

Receiving bins are static items that should not require attention. Labour requirement for the truck scale is assumed to be part of the truck dumper labour.

After the total labour for pretreatment had been calculated using this method (see Section 6.10.3), it was found that the total labour was too high (more than total plant labour for existing plant in many cases). One of the reasons for this was the amount of labour that had been assigned to the various belt conveyors in the system. It was therefore decided to ignore all labour for belt conveyors (treating the conveyors as part of the equipment that feeds them instead).

Since labour requirements for the system are calculated on a five shift basis and the labour for the reception step would only be required during daylight hours, 5 days/week, the number of operators required has been divided by 5. Thus the labour requirement for the reception step is as shown in Table 6.4. Overall labour costs are discussed collectively in Section 6.10.3.

Table 6.4 - Labour requirements, reception steps

Option	Constraints	Labour requirement/shift
	Defining capacity $= Q_d$, wet tonnes per day delivered	
Reception, small	$Q_{d,del} < 293$	0.20
Reception, medium	$293 < Q_{d,del} < 720$	0.24
Reception, large	$720 < Q_{d,del} < 1350$	0.24
	$1350 < Q_{d,del} < 2700$	0.48
	$2700 < Q_{d,del} < 4050$	0.72
	$4050 < Q_d < 5400$	0.96

6.3 PRE-STORAGE SCREENING (LARGE SCALE SYSTEMS ONLY)

Screening is required in all systems to ensure that the feed supplied to the conversion reactor is in the required size range and to remove contaminants such as ferrous metal and grit or rocks. Screening is applied in all systems after storage (see Section 6.5). At this stage screening is used to produce a feedstock that is easier to handle and less likely to damage or cause blockages in the conveyor system. Pre-storage screening is used in existing large-scale combustion plant in the US [229]. This step is only used in the large scale systems where damage to the feed transfer conveyors to storage could be a significant problem and where the plant is large enough to support the extra expense of two screening steps.

The screening step includes equipment for the removal of:

- large rocks and grit;
- ferrous metal; and
- over-size feed pieces.

The pre-storage screening step is supplied by conveyors that are included in the large-scale reception option. These conveyors transfer their load to a second conveyor (the start of the screening equipment) across a short gap. The momentum of the feedstock carries it across the gap but dense particles such as rocks fall into a trap under the belts due to their higher density [224].

An overhead magnet is positioned over the belt conveyor for the removal of ferrous metal [224].

The standard screening options are disc screens and vibrating screens [222, 230]. Vibrating screens, essentially large sieves, are less popular due to limited capacity, higher power consumption and a tendency to clog. Disc screens are more usually chosen because of their higher capacity and lower power consumption (1130 m^3/h may be processed using only a 15kW motor [231]). Disc screens comprise an array of rotating shafts with interwoven discs. The material moves across the discs and the disc spacing is arranged so that material of a suitable size drops between the disks and is passed. Disk screens are self-cleaning with little risk of clogging and they are adjustable to cope with different incoming feeds or screening requirements. They are quieter than other screens and can be easily inserted into the conveying system. This step will use a single screen that will be capable of sorting the feedstock required for all capacities up to 100 MW_e.

The screen removes over-size pieces that could damage or block the conveyors. This oversized material is dumped to storage on a concrete pad and later transferred to the main re-chip operation (after the post-storage screening operation, see Section 6.6) by front end loader. This is considered a house-keeping operation and a front-end loader specified in the storage step (see Section 6.3.1) will carry out this task.

Transfer of passed feedstock to the next step (Storage) is achieved by an open belt conveyor.

6.3.1 Capital costs

Total plant costs have been calculated from equipment costs for the main equipment items using the factors method described in Section 4.5.3. These costs were compiled from various sources [181, 224, 227] into a data sheet presented in Appendix C. The data sheet was used to calculate the capital costs of each option for various feed input capacities, with the results shown in Figure 6.3. Thus the capital costs for the step are calculated using Equation 6.1. The defining capacity used is $Q_{h,del}$, the wet tonnes delivered per hour.

$$TPC_{snl} = 75.85\left(Q_{h,del}\right)^{0.4498} \tag{6.1}$$

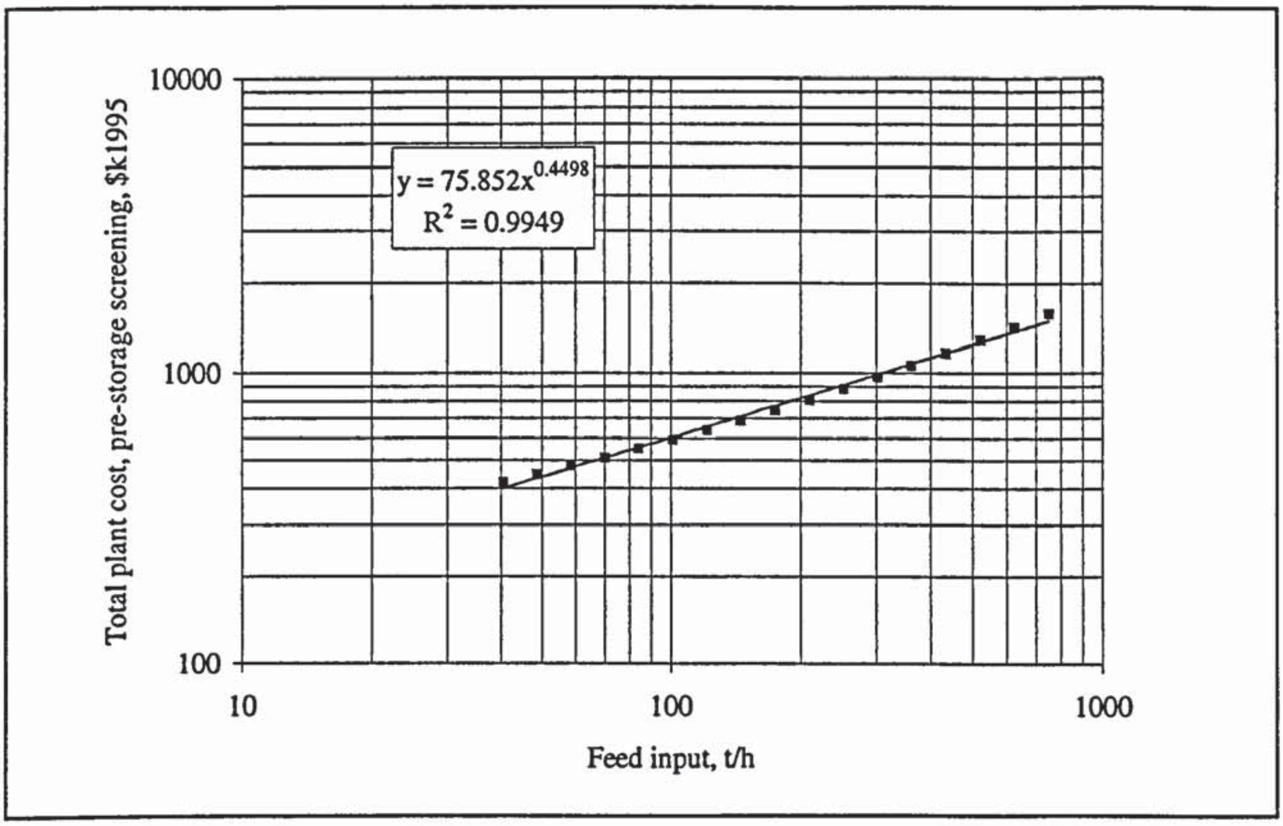


Figure 6.3 - Total plant costs, pre-storage screening step

6.3.2 Power requirements

The installed power requirement for disk screens has been taken from product literature from Rader [231]. Other power requirements were taken from an Enfor report into wood pretreatment systems [224]. This data was included in the data sheets in Appendix C and the power consumption for this step is calculated as the capital costs were derived, producing the relationship given as Equation 6.2. The defining capacity, $Q_{h,del}$, is the feed delivered per hour.

$$P_{sn1} = 0.574\left(Q_{h,del}\right)^{0.739} \tag{6.2}$$

6.3.3 Labour requirements

The following labour requirement is taken from a table by Ulrich [228].

- Screen: 0.05 operator

No data was available for the magnet but the labour requirements for this are expected to be minimal. The labour requirement suggested for the screen is rather low, perhaps reflecting the reliability of disk screens. It was decided to add 0.2 operators for the screen rather than use the value given above, thus assuming that the screen will require as much attention as a belt conveyor.

Since labour requirements for the system are calculated on a five shift basis and the labour for this step would only be required during daylight hours, 5 days/week, the number of operators required has been divided by 5. This makes the labour requirement for this step is 0.04 operators/shift.

6.4 STORAGE (ALL SYSTEMS)

6.4.1 Feedstock changes in storage

Feedstock changes during storage were introduced in Section 2.4.3. This section determines the values that will be used in the model to calculate dry matter losses and moisture content changes during storage. There are a number of factors that influence these changes, notably the storage facility, the quantity of feed stored, local climate, storage period, time of year, and initial feed characteristics [232]. These variations make it difficult to quantify the changes in the general case, as Mitchell reported during a study of UK wood handling practices [233]. Data for changes in moisture content and dry matter losses are given in Table 6.5. The figures for a single month are average changes based on the total absolute change. In fact the incremental monthly changes will change through the storage period (as the data reported by Gjølsjø [234] has shown) but the convention is to average the changes as practised here.

The reported data on moisture content change shows considerable variation. This variation is likely to increase in short term storage where the behaviour of the pile will be highly influenced by the prevailing weather conditions. Indeed, some reported data has shown increases in the moisture content of material kept in uncovered storage piles [233, 235] (the storage periods were not reported and hence the data was not included in Table 6.5). In view of this variation, the default used in the model will be 0% moisture content change during storage.

Some loss of dry matter will occur due to fungal and bacterial attack, but the data in Table 6.5 shows the extent of the losses are also difficult to quantify. It is known that the amount lost increases in feedstocks with a high moisture content such as forestry wastes [180] and on this basis a default dry matter loss of 2.5% per month is assumed, being at the higher end of the data presented.

Table 6.5 - Reported changes in feedstock during storage

No. months in storage	Starting moisture content, %[a]	Moisture reduction, %	Total dry matter losses, %	Moisture reduction/ month, %	Dry matter loss/month , %	Ref.
1.0[b]	-	-	0.7	-	0.7	[180]
1.0[b]	-	-	3.7	-	3.7	[180]
1.8	41.5	8.9	2.3	4.9	1.3	[69]
1.8	39.9	4.3	2.6	2.4	1.4	[69]
2.5	59.0	14.0	5.0	5.6	2.0	[236]
3.0	57.0	11.0	19.1	3.7	6.4	[237]
4.1	41.5	10.8	7.5	2.6	1.8	[69]
4.1	39.9	3.1	8.5	0.8	2.1	[69]
6.1	41.5	10.4	7.5	1.7	1.2	[69]
6.1	39.9	10.6	8.7	1.7	1.4	[69]
7	40.0	10.0	7.5	1.4	1.1	[238]
11	~50	negligible	-	-	2.9	[235]
13	-	-	-	-	2.5	[235]

a *all percentages are absolute differences (i.e. percentage points)*
b *based on a relationship by Nellist of 0.001 to 0.005%/h, the upper limit is for high moisture feedstocks*

6.4.2 Storage period

Low feed inventories on site are preferred because the feed will lose dry matter while in storage, storage takes up a lot of space, and because of the risk of fires. However, such factors must be weighed against the need to ensure a continuous supply of feedstock through interruptions to the delivery schedule. The amount of feed kept in storage is therefore a compromise. Although storage of up to 60 days has been reported in large facilities [225], stockpiles of 10-30 days are more usual [47, 222, 224]. This work will assume feed storage for three weeks in all cases.

6.4.3 Storage equipment

Storage facilities and material reclaim equipment vary considerably with the amount of feed that must be handled, the seasonal availability of feedstock and the constraints on space at the facility. Three storage options are defined here, selected for convenience to match the capacity of the three reception options. Although the equipment selected is suitable for their capacity ranges they represent just three of many arrangements that could have been used.

The maximum capacity of the small-scale reception option (293 t/d, 5 d/week) equates to a storage requirement of 4395 t or 17580 m^3 if feed is stored for three weeks and has an approximate bulk density of 0.25 t/m^3 (all the following calculations use a bulk density of 0.25 t/m^3 since the actual bulk density will vary with moisture content). At this scale only the most basic storage and reclamation equipment is used [223]. Feed is stored in a flat-topped pile with a maximum height of 6 m. This pile is set on a drained concrete base to reduce contamination of the feed with the underlying soil and contamination of the soil with the leachate from the pile. The feed is stacked manually by a front end loader (included in the reception option). Feed is also blended and reclaimed by the same front end loader and transferred to a live bottom day bin sized to hold the feed required for 24 h operation. Continuous extraction from the day bin is by means of a drag chain conveyor that can tolerate the inconsistencies likely to arise in the feed before it has been screened. The drag chain conveyor deposits the feedstock onto a short open belt conveyor that transfers the feed to the screening step.

The medium scale reception option limits are 293-720 t/d, equating to 17580-43200 m^3 in storage for the medium scale storage option. This feed is transferred from the previous step by open-belt conveyor (included in the medium scale reception option). This conveyor meets a second belt conveyor that is used to elevate the material and dump to the top of the storage pile. Feed blending and general house-keeping in the storage pile is managed by a front end loader. Feedstock is reclaimed from under the pile by a reciprocating stoker reclaimer that is ideal for material retrieval from large, deep, unrestricted piles and the maximum 23 t/h reclaim rate that will be encountered [224]. The stoker reclaimer discharges to an open-belt conveyor that transfers the feedstock 50 m to the screening step.

For storage capacities in excess of 43200 m^3, storage is divided between a main stockpile and a reclaim pile [225]. The reclaim pile is located near the rest of the pretreatment equipment and offers a buffer in the case of any failure in the main transfer conveyors from the bulk storage pile. Material is deposited on the main stockpile by means of an open belt conveyor with various trippers along its length to spread feed on the pile as required. The pile is managed by means of tracked bulldozers (each moves a maximum of 150 t/h). These bulldozers shift feed as required to reclaim pits at the edge of the pile that are fitted with under-pile chain conveyors [224]. Each pit can reclaim a maximum of 50 t/h. Feed reclaimed from the main stockpile is transferred 100 m to the reclaim pile. The reclaim pile is stacked using a radial belt-conveyor with extraction from under the pile by means of a screw reclaim system [224]. The reclaim pile holds sufficient feed for 24 h operation. An additional front end loader is included to manage the reclaim pile and for general house-keeping duties. The screw reclaim system supplies an open belt conveyor to post-storage screening.

6.4.4 Capital costs

Total plant costs have been calculated from equipment costs for the main equipment items using the factors method described in Section 4.5.3. These costs were compiled from various sources [181, 224, 227] into a data sheet presented in Appendix C. These costs have been used to produce the cost data shown in Figure 6.4, plotted against the rate of reclaim, $Q_{h,rlm}$. The rate of reclaim is the wet tonnes per hour that must be reclaimed to meet the conversion reactor capacity. The very low scale exponent in the medium scale storage option is caused by the fixed cost of the front end loader. Capital costs for the pretreatment step are calculated using Equation 6.1.

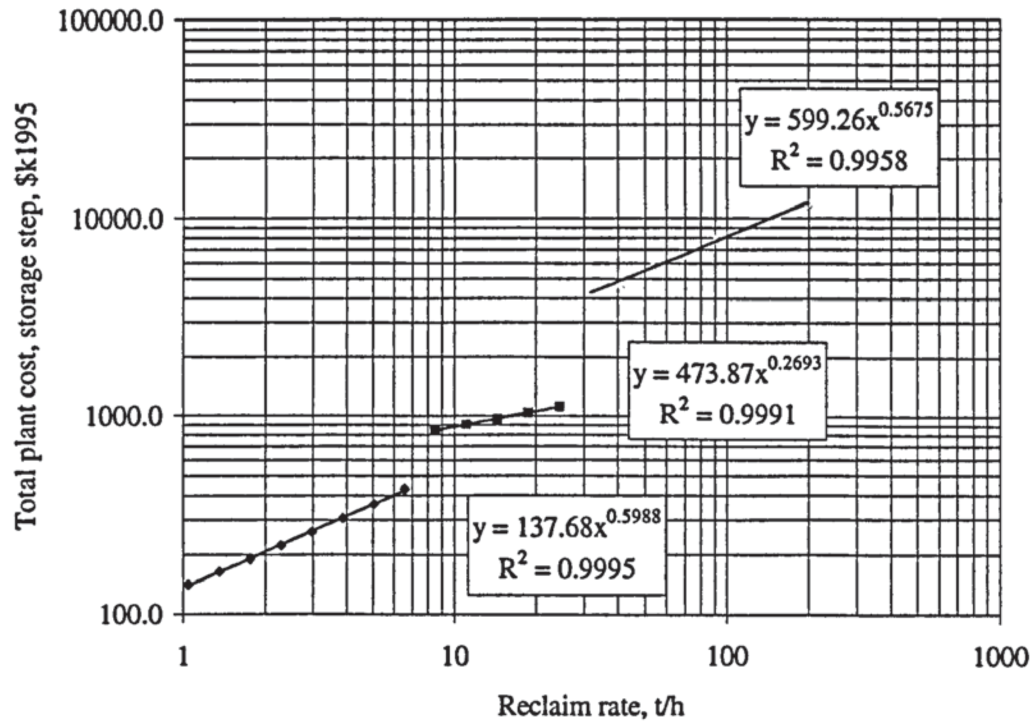


Figure 6.4 - Total plant costs, storage step

Table 6.6 - Total plant cost relationships, storage step

Defining capacity $= Q_{h,rlm}$, wet tonnes per hour reclaimed from storage		
Option	**Constraints**	**TPC relationship**
Storage, small	$Q_{h,rlm} < 8.7$	$TPC_{sto} = 137.68(Q_{h,rlm})^{0.5988}$
Storage, medium	$8.7 < Q_{h,rlm} < 21$	$TPC_{sto} = 473.87(Q_{h,rlm})^{0.2693}$
Storage, large	$21 < Q_{h,rlm}$	$TPC_{sto} = 599.26(Q_{h,rlm})^{0.56751}$

6.4.5 Power consumption

The installed power requirement for the equipment required is taken from the literature [181, 224] and is included in the data sheets in Appendix C. The power consumption for this step was calculated from the data sheets for various capacities and the results were used to produce the relationships in Table 6.7. Utilities for all steps are discussed in Section 6.10.2.

Table 6.7 - Installed power requirements, storage step

Defining capacity $= Q_{h,rlm}$, wet tonnes per hour reclaimed from storage		
Option	**Constraints**	**Installed power relationship**
Storage, small	$Q_{h,rlm} < 8.7$	$P_{sto} = 2.103(Q_{h,rlm})^{0.703}$
Storage, medium	$8.7 < Q_{h,rlm} < 21$	$P_{sto} = 20.52(Q_{h,rlm})^{0.504}$
Storage, large	$21 < Q_{h,rlm}$	$P_{sto} = 5.753(Q_{h,rlm})^{0.794}$

6.4.6 Labour requirements

The labour requirements for this step were assigned as follows [228]:

- Front end loaders: 1 operator
- Bulldozer: 1 operator
- Truck dumpers: 1 operator
- Drag chain conveyor: 0.2 operator
- Stoker reclaimer: 0.2 operator
- Radial stacker and screw reclaimer: 0.2 operator

No data was available for the radial stacker and screw reclaimer system, and although this system is supposed to operate without labour it is likely to require some minimal attention. The labour requirement for equipment used to put feed into the storage pile is divided by 5, since it is not used continuously, as are front end loader labour requirement since it assumed that they will run during daylight hours only. Thus the labour requirement for the reception step is as shown in Table 6.8. Overall labour costs are discussed collectively in Section 6.10.3.

Table 6.8 - Labour requirements, storage steps

Defining capacity = $Q_{h,rlm}$, wet tonnes per hour reclaimed from storage		
Option	**Constraints**	**Labour requirement**
Storage, small	$Q_{h,rlm} < 8.7$	0.2
Storage, medium	$8.7 < Q_{h,rlm} < 21$	0.4
Storage, large	$21 < Q_{h,rlm}$	0.2+ 0.2*bulldozers + 0.2*front end loaders + 0.2*chain conveyors

6.5 POST-STORAGE SCREENING (ALL SYSTEMS)

The main purpose of screening at this stage is to ensure that the feedstock particle size range is suitable for the conversion technology and to remove items that would impede the operation of the conversion reactor such as rocks, grit and tramp ferrous metal. Screening is also used to make the feed easier to handle and convey. Screening is still required in all systems, even with pre-storage screening to ensure material that has compacted and lumped together in storage is broken up.

The screening equipment used is exactly that used in the pre-storage screening step, except that the step includes an extra conveyor to transfer the over-size material to the re-chip step. Another notable difference is that the equipment used in post storage screening will be smaller because the reclaim rate from storage is less than the delivery rate to storage (due to different operating periods).

Screening could be used to produce three streams by using two screens: oversize, undersize and passed material. Only screening for oversize material is considered here since it is assumed that all the conversion processes can tolerate a small fraction of fines. Over-size material may be dumped but it is more usual for it to be comminuted and added to the passed material. This comminution is carried out in the re-chip step (see Section 6.6).

6.5.1 Capital costs

Total plant costs are based on the same data used in Section 6.3.1, with the addition of a conveyor to the re-chip step. A data sheet (presented in Appendix C) was used to calculate the capital costs of each option for various feed input capacities, with the results shown in Figure 6.5. Thus the capital costs for the step are calculated using Equation 6.2. The defining capacity used is $Q_{h,rlm}$, the wet tonnes reclaimed per hour.

$$TPC_{sn2} = 112.45\left(Q_{h,rlm}\right)^{0.3435} \tag{6.2}$$

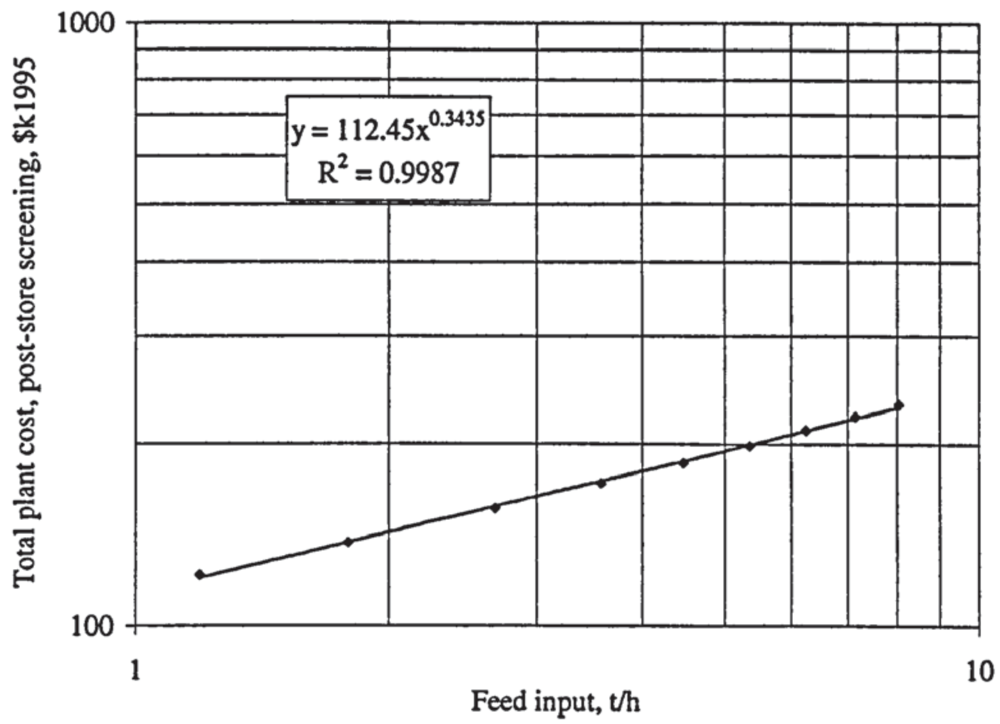


Figure 6.5 - Total plant costs, post-storage screening step

6.5.2 Power requirements

The installed power requirement for disk screens has been taken from product literature from Rader [231]. Other power requirements were taken from an Enfor report into wood pretreatment systems [224]. This data was included in the data sheets in Appendix C and the power requirement for this step is calculated as the capital costs were derived, producing the relationship given as Equation 6.4. The defining capacity, $Q_{h,rlm}$, is the feed reclaimed per hour in wet tonnes.

$$P_{sc2} = 1.718\left(Q_{h,rlm}\right)^{0.427} \tag{6.4}$$

6.5.3 Labour requirements

The labour requirements for this step are the same as those for the pre-storage screening step (see Section 6.3.3). Since this screening step operates continuously then the labour requirement is 0.2 operators/shift.

6.6 RE-CHIPPING (ALL SYSTEMS)

The re-chipping step chips any over-size material rejected during screening and returns the chipped output by open belt conveyor to the main feedstock stream from the post-storage screening step. In large-scale systems the re-chipping of material could have been linked with the pre-storage screening step, as seen at the 60 MW_e Burlington wood combustion plant [225], but since all systems must incorporate post-storage screening all re-chipping is associated with post-storage screening to simplify the model. This also reduces the cost of the equipment, since it is assumed that the step operates continuously. By default, 5% of the total feed is over-size, although this value is a variable.

Reviews of comminution equipment are given by Pierik [218] and Belli [239]. There are many equipment options that should be matched to each case based on throughput, the characteristics of the feed, the size reduction ratio (dictated by the size of the feed at input and the size required at output by the conversion technology), capital costs and operating costs. This study considered the two most popular machines, as stated by Bloomfield [222]: hammermills and knife hogs. Hammermills can tolerate tramp metal and other feed contaminants but their power consumption is very high [240]. Knife hogs are more sensitive to contaminants but have the advantages of lower power consumption and they are more tolerant of bark in the feedstock. A knife hog has been selected here since most contaminants should have already been removed during screening and the feedstock is expected to contain some bark.

6.6.1 Re-chipping

Equipment costs for the major equipment items required by this step have been taken from data by Carr [241], Garrett [227] and Simons [224]. A data sheet (presented in Appendix C) was used to calculate the capital costs of each option for feed input capacities between 1 and 10 t/h, with the results shown in Figure 6.6. Thus the capital costs for the step are calculated using Equation 6.4. The defining capacity used is $Q_{h,ovr}$, the wet tonnes of oversized feed that are chipped per hour.

$$TPC_{rcp} = 194.18\left(Q_{h,ovr}\right)^{0.3398} \tag{6.4}$$

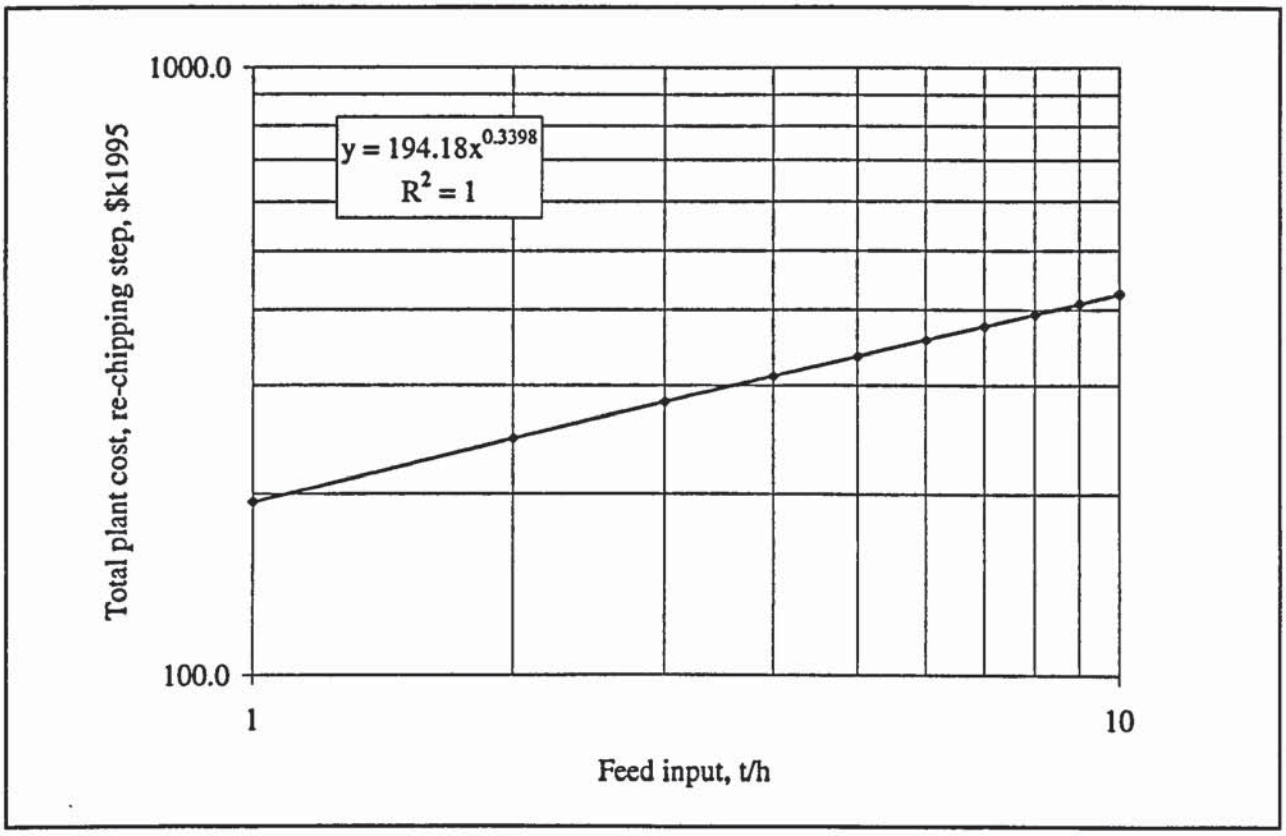


Figure 6.6 - Total plant costs, re-chipping step

6.6.2 Power requirements

The installed power requirements for knife hogs were provided by Carr [241]. Other power requirements were taken from an Enfor report into wood pretreatment systems [224]. This data is included in the data sheets in Appendix C and the power consumption for this step was calculated as the capital costs were derived, producing the relationship given as Equation 6.5. The defining capacity used is $Q_{h,ovr}$, the wet tonnes of oversized feed that are chipped per hour.

$$P_{rcp} = 23.01\left(Q_{h,ovr}\right)^{0.456} \tag{6.5}$$

6.6.3 Labour requirements

Labour requirements are estimated from a table of operator requirements published by Ulrich [228]. This table gave a minimum labour requirement of 0.5 operators for crushers, mills and grinders. It was felt unlikely that an operator would spend half his time at the knife hog, and so a value of 0.25 was used instead.

6.7 DRYING (ALL SYSTEMS)

This step dries the feedstock from its moisture content as it leaves the store to the moisture content required by conversion process. Drying of high moisture feedstocks improves the performance of the conversion processes, as reported in Section 7.2 (combustion), Section 7.3 (fast pyrolysis), and Section 7.4 (gasification). These sections will define the moisture content required by each process, summarised below:

- 35% for combustion;
- 15% for gasification; and
- 10% for fast pyrolysis.

There are many types of dryer available and their relative merits are discussed by Pierik and others [218, 224]. The three types of dryer suitable for high volume, low final moisture content applications are:

- rotary dryers;
- fluidised bed dryers; and
- steam dryers.

This work uses rotary dryers only, since they are the most established dryer type and enough data is available to model their cost and performance. Fluidised bed dryers and steam dryers are less established although initial experiences have indicated better drying efficiencies than for rotary dryers [113, 218] and they would be worth investigating in further work as more data becomes available.

Rotary dryers are rotating inclined drums through which hot gases and the material to be dried are passed. The rotating of the drum agitates the feed, enhancing drying and encouraging the passage of the feed as it dries. Large particles take longer to pass through the dryer, allowing them the opportunity to dry to the same moisture content as fine particles and in this way the rotary dryer can handle a moderately heterogeneous feedstock if necessary [231]. Operation is simple, the dryer is easy to control and electricity consumption is low. Rotary dryers are generally open systems which vent moisture vapour and drying gases to atmosphere. One of the problems of the open system is that the hot gases also drive off volatiles from the feedstock which are emitted causing fumes known as "blue haze" [82]. Careful control of the drying temperature can alleviate this problem.

Additional components of the drying system are the feeder and discharge, the hot gas supply, and the outlet gas emission controls. The feeding system consists of a hopper and a metering screw for controlled feed into the dryer. Where multiple dryers are required each has its own feeding system. Ideally, hot gas would be supplied by heat recovery from elsewhere in the system. Alternatively hot gases may be supplied by heated air or flue gases from a dedicated combustor, offering improved process control but at the expense of increased capital costs and fuel costs. This work will assume hot gas is provided by waste gases elsewhere in the system (this hypothesis is tested as the conversion and generation modules are developed in Chapters 7 and 8). At discharge large particles pass directly onto a conveyor and smaller particles entrained in the drying gas are usually removed in a cyclone to join the outlet conveyor. Very fine particles are captured in a baghouse filter.

A closed belt conveyor is used for transfer to the next step. This protects the dried feedstock from the elements and reduce dust as the feed is transferred.

6.7.1 Dryer energy requirement

This section determines the energy required by the dryer to evaporate feed moisture from an evaluation of 20 systems reported in reviews and manufacturers' literature [218, 224, 231, 242] and presented in Appendix C. This data is analysed to establish the performance of a generic rotary dryer. This information is not used directly in the model but is used in Chapters

7 and 8 to determine whether enough energy would be available elsewhere in the system for drying.

Dryer performance will be considered in terms of drying efficiency, defined here as the ratio of the theoretical evaporation energy requirement to the actual energy used. The theoretical energy requirement has been calculated based on the energy required to raise the moisture in the wood from 25°C to 100°C and the latent heat of water at 100°C. This equates to 2572 GJ/twe (tonne of water evaporated). The actual energy required will be higher due to the energy required to heat the feedstock, heat losses to the environment and any heating of the water vapour to temperatures above 100°C. For example, Nellist has calculated that a dryer would in reality consume a minimum of 3300 GJ/twe [237].

Figure 6.7 plots dryer efficiency against the dryer evaporation load (twe/h). It can be seen that there is a slight increase in efficiency as the amount of water to be evaporated (i.e. the dryer capacity) increases. This is reasonable given that a larger dryer would incur lower heat losses. However, the increase is very slight and the scatter of the data is considerable. Therefore it was decided to simplify the situation and assume that dryer efficiency was independent of dryer evaporation load.

Since the dryer will be required to meet a variety of feed moisture conditions, both at dryer inlet and exit, it is important to assess whether feed moisture effects dryer efficiency. Figure 6.8 compares the dryer efficiency with the feed moisture conditions. No correlation can be seen. It may be concluded that in generic dryers the efficiency of the dryer is independent of feed moisture conditions and thus no adjustment will be required in this respect when examining different systems.

Another issue is the temperature of flue gas used for drying. Hot gas sources will be available from various parts of the system at temperatures ranging from 100 to 800°C, and one use of such heat would be in drying. Figure 6.9 compares the dryer efficiency with the hot gas temperature conditions. In both plots the data is widely scattered and no correlation can be seen between the dryer efficiency and any of the operating conditions.

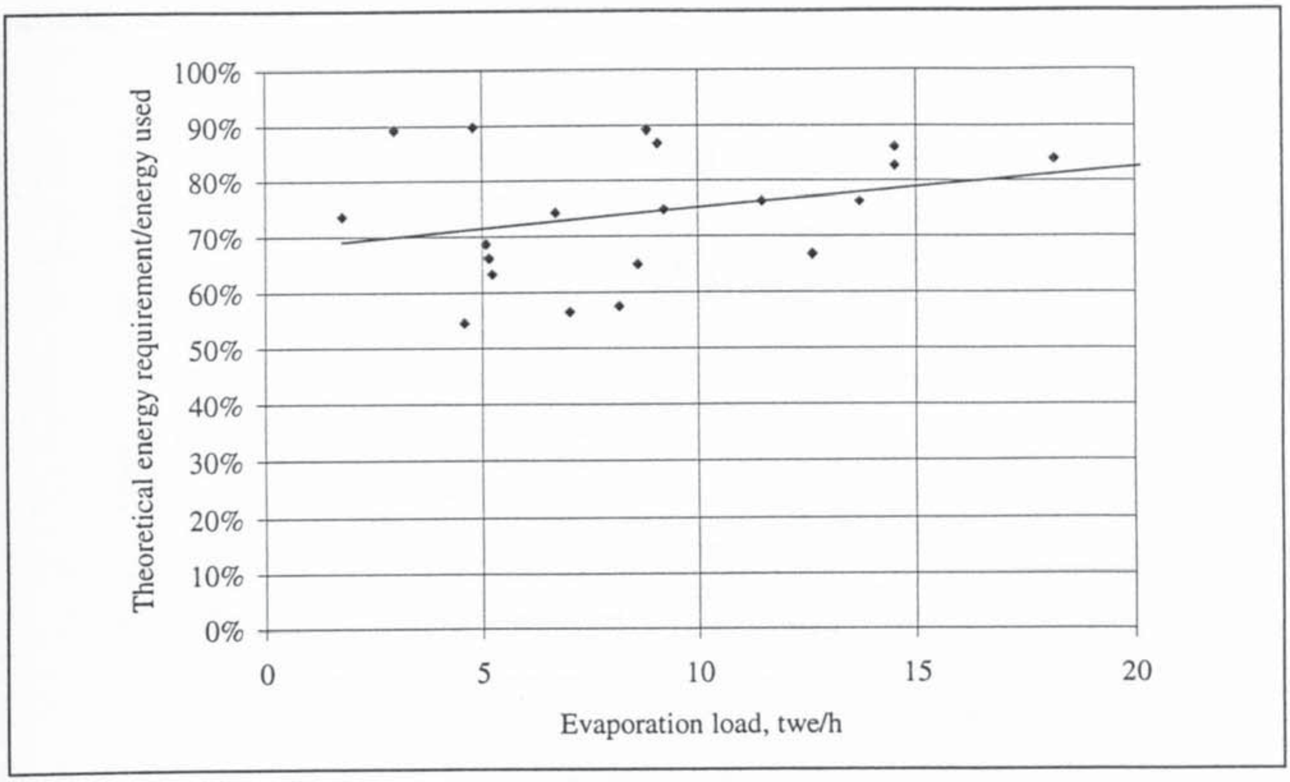


Figure 6.7 - Dryer efficiency as a function of evaporation load

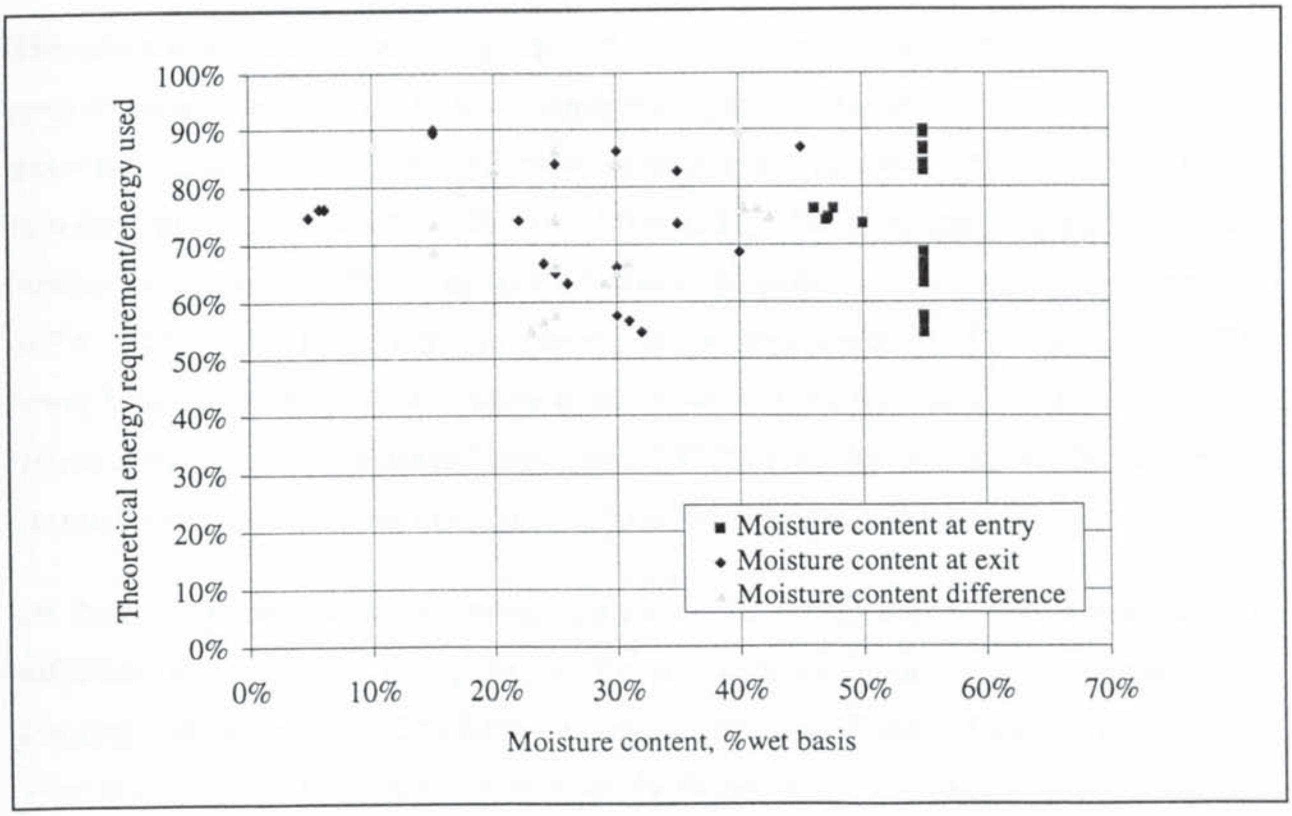


Figure 6.8 - Dryer efficiency as a function of feed moisture conditions

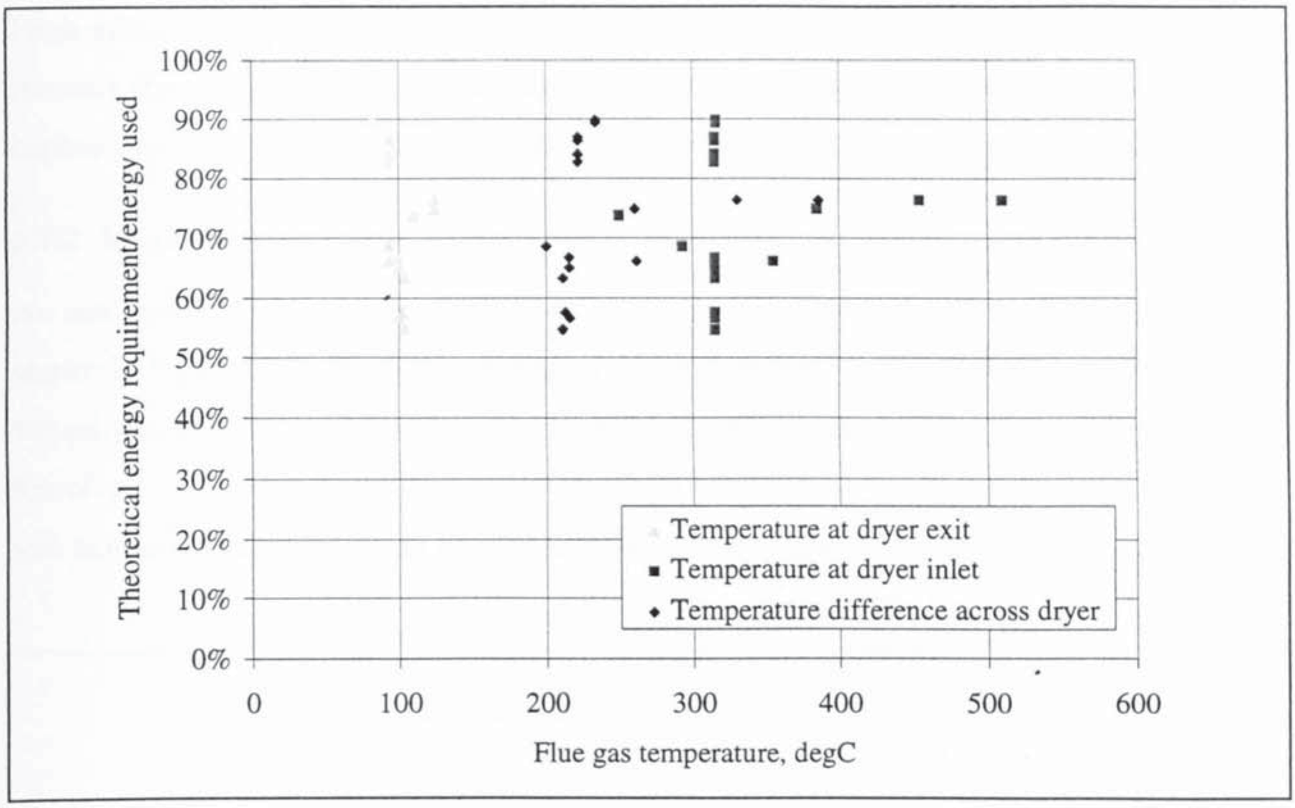


Figure 6.9 - Dryer efficiency vs hot gas temperature

The inlet temperatures are generally higher than current practice would dictate, because of the need to reduce blue haze [82]. Inlet temperatures can be reduced by recirculating dryer exit gases which also increases the humidity at the dryer inlet. This slows the drying rate (helping to reduce blue haze) and reduces the risk of fires [231]. The dryer exit temperatures are very similar and average 102°C. One general review suggested a flue gas exit temperature of 107°C [222]. Other references give a lower limit at the dryer exit of 80°C [218, 242]. This lower limit would allow more energy to be extracted from the flue gas but could cause a plume at the stack. The average temperature of 102°C given by the data will be used in later chapters when examining the availability of heat from elsewhere in the system.

On the basis of the data shown above, a single average energy requirement is assumed to be sufficient for a generic drying model for all dryer conditions. The average energy consumption for drying is 3.553 GJ/twe (a drying efficiency of 72%). This figure will be used when assessing the availability of heat to dry the feedstock from elsewhere in the system. It is important to note that this is an average energy requirement, with the range of energy requirements in the literature being 2861 to 4703 GJ/twe (89% to 54% efficiency). Therefore where there is insufficient hot gas for drying it may be possible to use a more efficient dryer.

High efficiency dryers would probably incur extra cost due to more complex features that increase the agitation of the feed and the contact between the feed and the hot gas. Since the capital costs used are average costs then average performance must also be assumed.

6.7.2 Maximum dryer size

An analysis of dryer size was carried out on all the data available, and the results of this are shown in Figure 6.10. From this analysis it can be seen that the range of evaporation loads is virtually continuous up to 20 twe/h, with two exceptional cases above this capacity. It was therefore decided that a maximum capacity of 20 twe/h would be imposed for a single dryer, and that multiple dryers would be used to meet capacities above that amount.

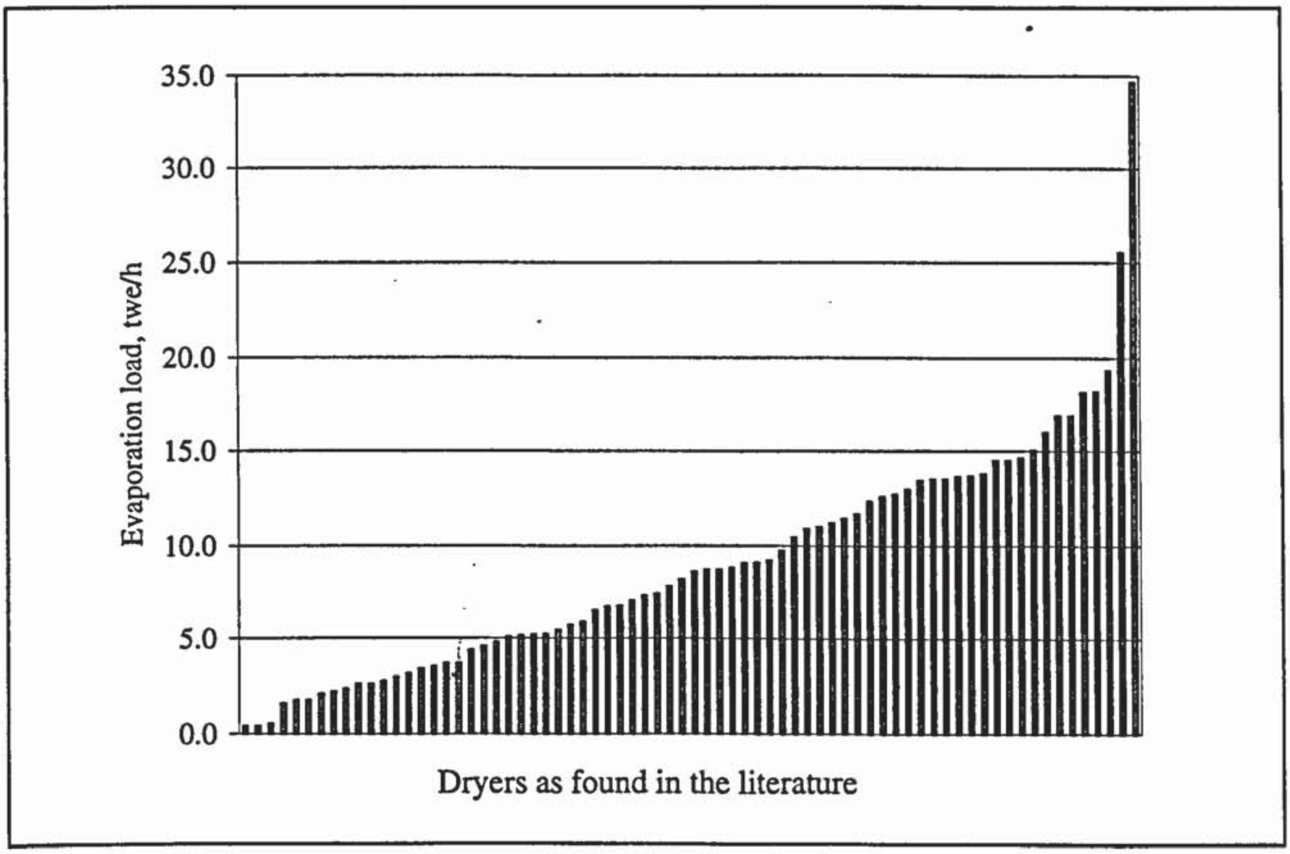


Figure 6.10 - Dryer evaporation loads

6.7.3 Capital costs

The capital costs of the drying step have been calculated using published data for whole drying units [218, 224, 231, 242, 243, 244, 245]. Since this data referred to a collection of equipment items the ratios method (see Section 4.5.3) was used to give the total plant cost of each dryer. The use of the evaporation load to size the dryer gave the best correlation between the cost data but the variation in the costs (shown in Figure 6.11) is still significant, reflecting the differences in dryer configuration that are possible. The costs given in the literature did not include the costs of transfer of the dried feedstock to the next step. This is calculated separately using equipment costs in Garrett [227] and converting to total plant costs using the factors method (see Section 4.5.4). Where multiple dryers are used it is assumed that there would be some common components such as the exhaust gas treatment system. For this reason an exponent of 0.9 has been applied to the number of dryers used. The total plant cost relationship is given by Equation 6.6. Two defining capacities are required to calculate the total plant cost of the dryer step: the tonnes of water evaporated per hour per dryer, $Q_{h,dry}$, and the total volume flow rate of the dried feed to the next step, $V_{h,dry}$.

$$TPC_{dry} = \left(n_{dry}\right)^{0.9} \times \left[226.94\left(Q_{h,dry}\right)^{0.8714}\right] + 13.97\left(V_{h,dry}\right)^{0.3} \qquad (6.6)$$

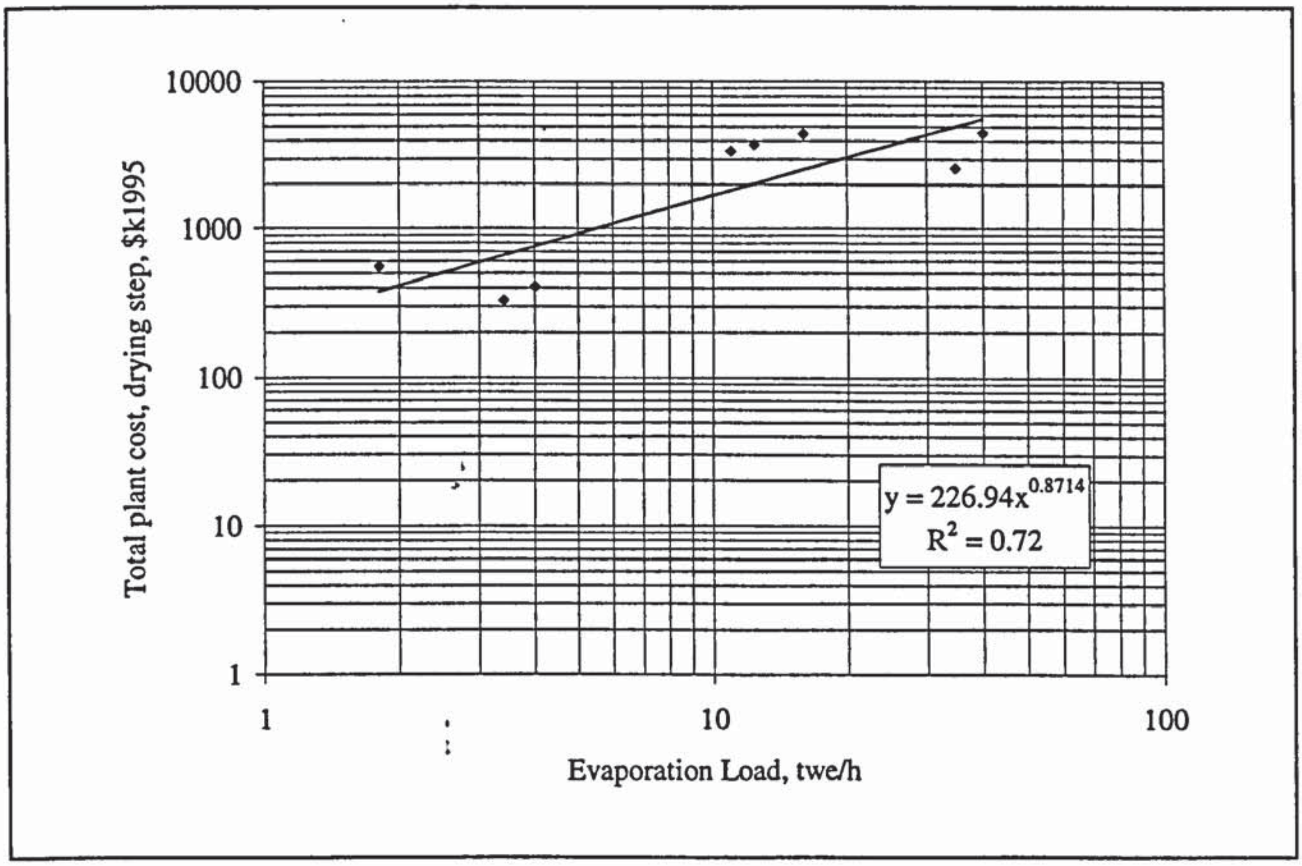


Figure 6.11 - Total plant costs, drying step (not including transfer)

6.7.4 Power requirements

Power requirements for some of the drying systems shown in Appendix C were plotted against evaporation load to give the chart shown in Figure 6.12. A relationship for dryer power was also suggested by the Enfor report [224], and is shown by the line in Figure 6.12. It can be seen that the relationship fits reasonably to the data points (given their scatter) and so the relationship was used to calculate the dryer energy requirement, as given in Equation 6.7. The defining capacities required are the evaporation load per dryer in twe/h, $Q_{h,dry}$, and the total volumetric flow of dried feed from the drying stage in m^3/h, $V_{h,dry}$.

$$P_{dry} = n_{dry}\left[\frac{250}{30}\left(Q_{h,dry}\right)\right] + 0.06\left(V_{h,dry}\right)^{0.7} \qquad (6.7)$$

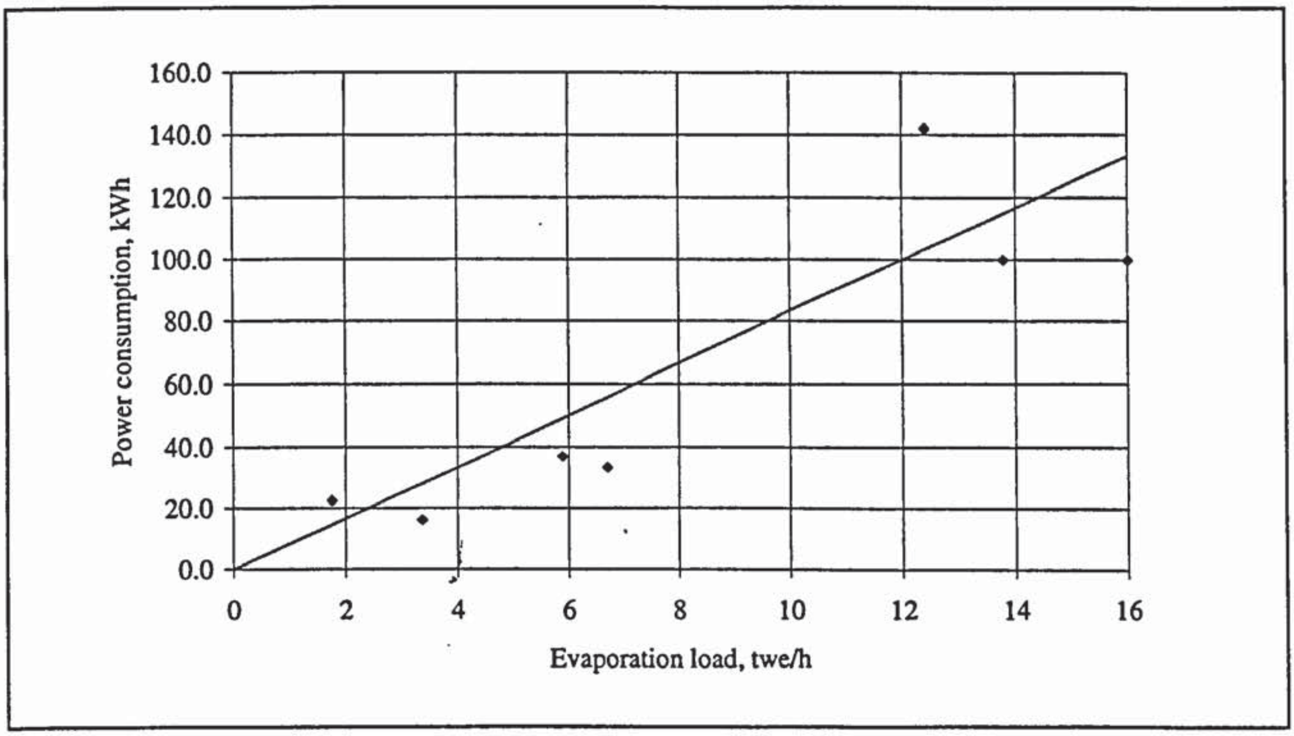


Figure 6.12 - Rotary dryer power requirements

6.7.5 Labour requirements

Peters [198] has estimated that the labour requirement for a rotary dryer is 0.5 workers per shift. Multiple dryers could be supervised using less labour since their operation would be similar, they would be located together and they would share components such as the exhaust air treatment equipment. Therefore an exponent of 0.75 was applied to give a labour requirement as shown in Equation 6.8.

$$\text{Operators}_{dry} = 0.5 \times \left(n_{dry}\right)^{0.75} \tag{6.8}$$

6.8 GRINDING (FAST PYROLYSIS SYSTEMS ONLY)

The grinding step is required by the fast pyrolysis systems only and turns the screened, dried wood chip feedstock into a powder with a particle size of less than 2 mm, as discussed in Section 2.7.2.

The grinding operation is accomplished in multiple pulverizers [224, 246]. These crush the incoming feed between rows of rotating hammers and an impact plate. The feed then travels with the rotating hammers within a screen until it has been sufficiently broken up to pass through the screen. An induced draft fan is used to encourage the passage of comminuted material through the screen. A 1.6 mm feedstock requires two pulverizers in series, the first to pulverise the chips to 3.2 mm and the second to pulverise to 1.6 mm. The second machine will be 150% bigger and consume 150% more power to achieve the same throughput. The feed should be dropped from a height of around 1.5 m to ensure that it passes between the hammers and impact plate. Therefore an elevated surge bin and metering feeder are required for each pulverizer. Bucket conveyors are used to elevate the feed to the surge bin.

6.8.1 Capital costs

Pulverizers are available only in limited sizes and therefore multiple trains are required to meet larger feed capacities. Based on the equipment sizes given in [224], two basic options are used. The first option has a maximum capacity per train of 4 t/h and is used for small-scale systems up with a total capacity of 12 t/h (i.e. a maximum of three trains). The second option uses trains with a maximum capacity of 8 t/h and multiple trains are used to meet any capacity above 12 t/h.

The only costs that were available for comminution to less than 2 mm particles in bulk quantities were given in $Canadian$_{1982}$ [224]. Other work has discussed the cost of wood comminution, but the analysis focuses on the power consumed [180, 247]. It can be seen in Appendix C that the cost of the pulverizers is high and as such their influence on the overall capital costs is significant. It is advisable that future work examines the cost of this comminution step in more detail given the age of the data used here. The data sheet in Appendix C was used to plot the capital costs for various grinding rates as shown in Figure 6.13. The relationships used in the model to calculate grinding capital costs are given in Table 6.9. The high scale exponents are notable, indicating that the there is little economy of scale to be gained by using bigger machines.

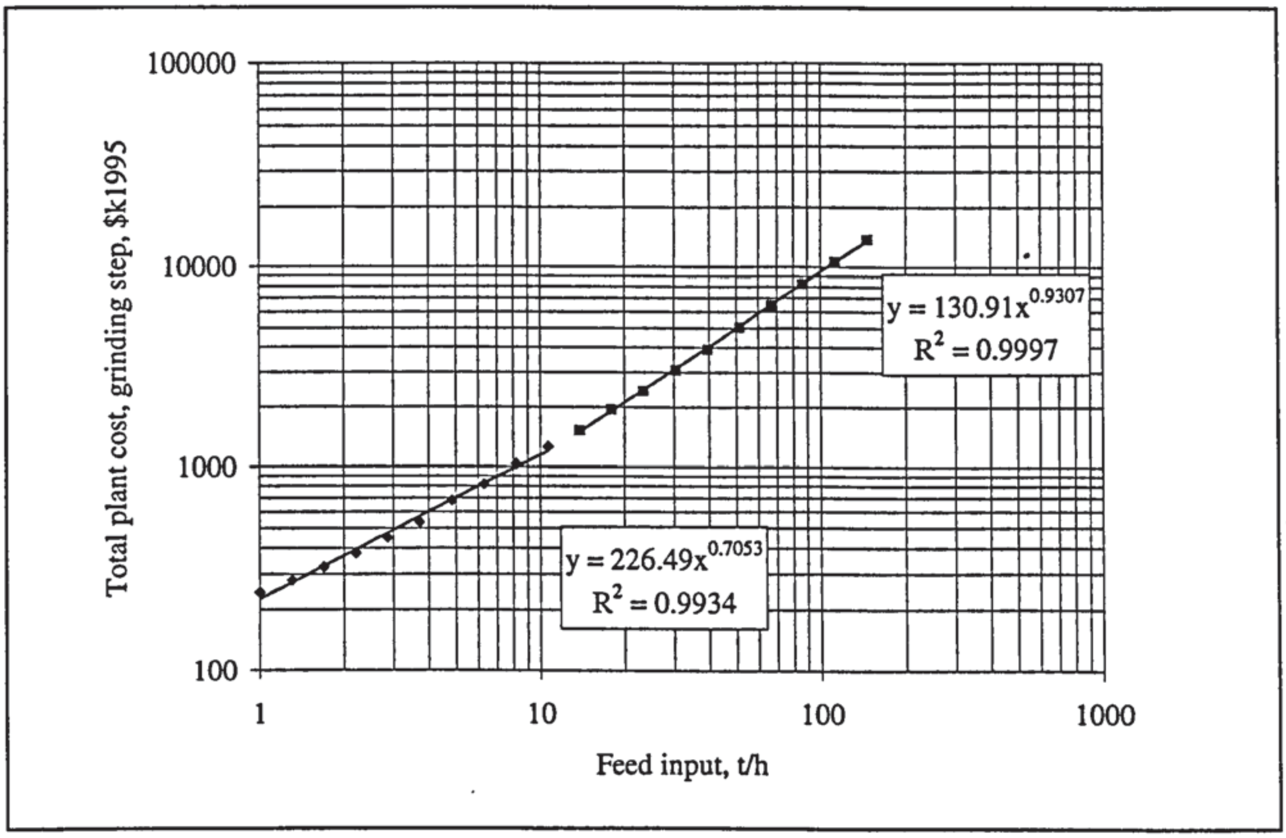


Figure 6.13 - Total plant costs, grinding step

Table 6.9 - Total plant cost relationships, grinding step

Defining capacity = $Q_{h,grd}$ wet tonnes per hour into the grinders from drying		
Option	**Constraints**	**TPC relationship**
Grinding, small scale	$Q_{h,grd} < 12$	$TPC_{grd} = 226.5(Q_{h,grd})^{0.7053}$
Grinding, large scale	$12 < Q_{h,grd}$	$TPC_{grd} = 130.9(Q_{grd})^{0.9307}$

6.8.2 Power requirements

The power consumed in grinding wood to very small particle sizes is extremely high, and this relationship is expected to have an important impact on the economics of fast pyrolysis, which is the only system that requires this step. Examples of power consumption requirements are given in Table 6.10. The huge power requirements at the very small scale are clear. The power consumption relationships presented in Table 6.11 are based on Enfor data [224]. It is interesting to note that for the feedstock defined in Section 4.3, 1 t of feed with 10% moisture has a lower heating value of 17.12 GJ/t. The grinding step would require approximately 110 kWh/t to grind the material, which is 0.396 GJ or 2.3% of the energy in the material. If the electricity is generated at an efficiency of 30%, the energy used in grinding increases to 7.7% of the feed energy. This energy consumption is therefore highly significant in the efficiency of the overall system.

Table 6.10 - Comminution power requirements for small particle sizes

Machine	Particle size in, mm	Particle size out, mm	Power required, kWh/t	Source
Pulveriser	25	3.2	43.4	[224]
Knife mill	Not known	3.2	53	[247]
Hammer mill	Not known	3.2	128	[247]
Not known	Not known	3	40-150	[180]
Pulveriser	3.2	1.6	65.4	[224]
Knife mill	Not known	1.6	132	[247]
Hammer mill	Not known	1.6	137	[247]
Not known	Not known	<2	40-200	[180]
Pulveriser	10	0.8	212	[246]
Pulveriser	10	0.5	532	[246]

Table 6.11 - Power requirements, grinding step

Defining capacity = $Q_{h,grd}$ wet tonnes per hour into the grinders from drying		
Option	**Constraints**	**TPC relationship**
Grinding, small scale	$Q_{h,grd} < 12$	$P_{grd} = 111.4(Q_{h,grd})^{0.9917}$
Grinding, large scale	$12 < Q_{h,grd}$	$P_{grd} = 109.4(Q_{h,grd})$

6.8.3 Labour requirements

The same labour requirements are assumed as for the chipper (Section 6.6.3), with a basic labour requirement per grinder of 0.25 operators per grinder. Where multiple grinders are used the labour for several grinders has been adjusted as shown in Equation 6.9 to account for common components and operating characteristics.

$$Operators_{grd} = 0.25 \times \left(n_{grd}\right)^{0.75} \tag{6.9}$$

6.9 BUFFER STORAGE (ALL SYSTEMS)

The buffer storage step includes the equipment required to provide a small store of prepared feedstock immediately before transfer to the feeding mechanism of the conversion process. A bucket conveyor is used to elevate the screened wood chips or powder from the transfer conveyor to the top of the buffer storage silo. Multiple silos are used as necessary to give a total storage capacity equivalent to 4 h operation, with each silo having a maximum capacity of 500 m^3.

6.9.1 Capital costs

The costs of buffer storage are simply the cost of a live-bottomed silo, with multiple silos used as required to comply with the maximum holding capacity of 500 m^3. The costs used are calculated in the data sheet given in Appendix C and the capital costs for a range of volumetric flow rates are given in Figure 6.14. Thus the capital costs for the step are calculated using Equation 6.10. The defining capacity used is $V_{h,bfr}$, the volume of material required by the conversion reactor per hour.

$$\mathrm{TPC_{bfr}} = \mathrm{n_{silos}}\left[98.82\left(\mathrm{V_{h,bfr}}\right)^{0.3531}\right] \tag{6.10}$$

6.9.2 Power requirements

The power requirements of the live bottomed silos were given by the Enfor report [224] and the relationship used in the model was as given by Equation 6.11.

$$P_{bfr} = n_{silos}\left[0.717\left(V_{h,bfr}\right)^{0.7}\right] \tag{6.11}$$

6.9.3 Labour requirements

The only labour required is assumed to be associated with the screw conveyor at the base of each silo. A labour requirement of 0.2 operators is assigned to this item on the basis of the labour table given by Ulrich [228]. Where multiple silos are used the labour for several silos has been adjusted as shown in Equation 6.12 to account for common components and operating characteristics.

$$Operators_{bfr} = 0.25 \times \left(n_{grd}\right)^{0.75} \tag{6.12}$$

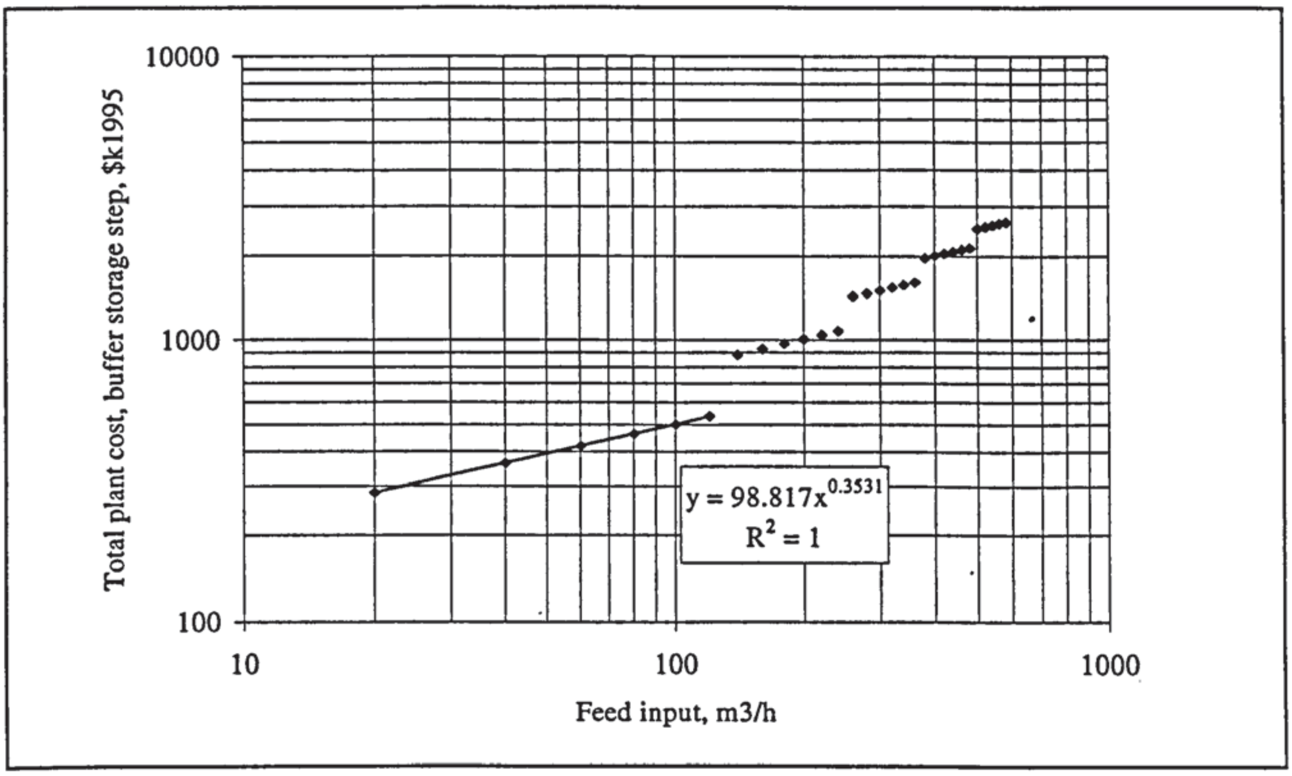


Figure 6.14 - Total plant costs, buffer storage step

6.10 OPERATING COST CALCULATIONS

6.10.1 Capital Amortisation

Capital is amortised using the standard relationship given in Chapter 4. This is a simplification since the equipment used is likely to have different working lives and some items may need replacing during the life of the project.

6.10.2 Utilities

The front end loaders and bulldozers consume 30 l/h and 35 l/h of diesel respectively [224] and are assumed to operate every day during daylight hours only (10 h/d) since when they are not in use in feed reception they may be required for general housekeeping duties. A diesel cost of 0.51 $/l is used [248].

Where power is consumed by equipment, the operating hours per year are required so that the annual power consumption can be calculated. It is assumed that power required by the reception, pre-storage screening and storage equipment is consumed during delivery hours only, whereas the other equipment operates over the same hours as the conversion technology.

Power consumed through the year is subtracted from the gross power output as shown by Figure 4.2. This (with the internal power consumption of the conversion and generation modules) gives the net electricity output per year that is used to calculate the electricity production cost. In the case of de-coupled fast pyrolysis systems, the electricity is charged to the pyrolysis facility as discussed in Section 4.6.5.

6.10.3 Labour

Labour requirements have been specified for each step based on a 5 shift system (see Section 4.6.4). It is useful at this stage to examine how the labour requirements add up over the expected range of feed capacities that are expected. Assuming the feed characteristics given in Section 4.3 and a constant system efficiency of 25%, the labour requirements for a range of capacities are shown in Figure 6.15 for a combustion system. A comparison of the same data with known total labour requirements for US combustion facilities is given in Figure 6.16 (the sources of the combustion plant data are not given as combustion plant labour is discussed in Section 7.2.4.1).

The breakdown of plant labour in Figure 6.15 highlights some inflexibility in the model. At the very small scale it is likely that the labour requirement would be less than that shown, since an operator would not need to operate the front end loader continuously (for example). This over-specification of labour at the small scale could be overcome by adding functions to each labour requirement that accounted for capacity as well as the equipment items but this would be unduly complicated.

The comparison of pretreatment labour with full plant labour in Figure 6.16 shows that the figures are approximately correct. One would expect the proportion of labour in smaller plant to be higher given that there is more manual handling of the material.

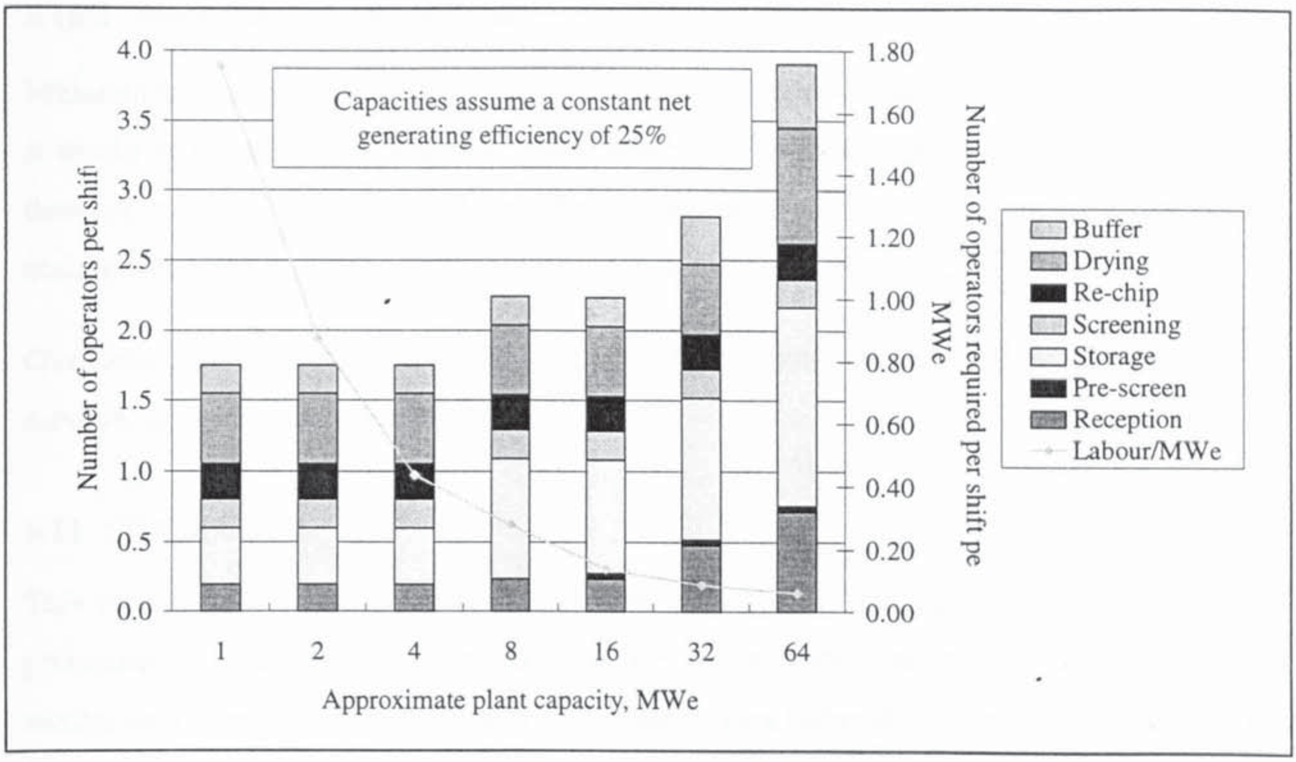


Figure 6.15 - Consolidated labour requirements for pretreatment

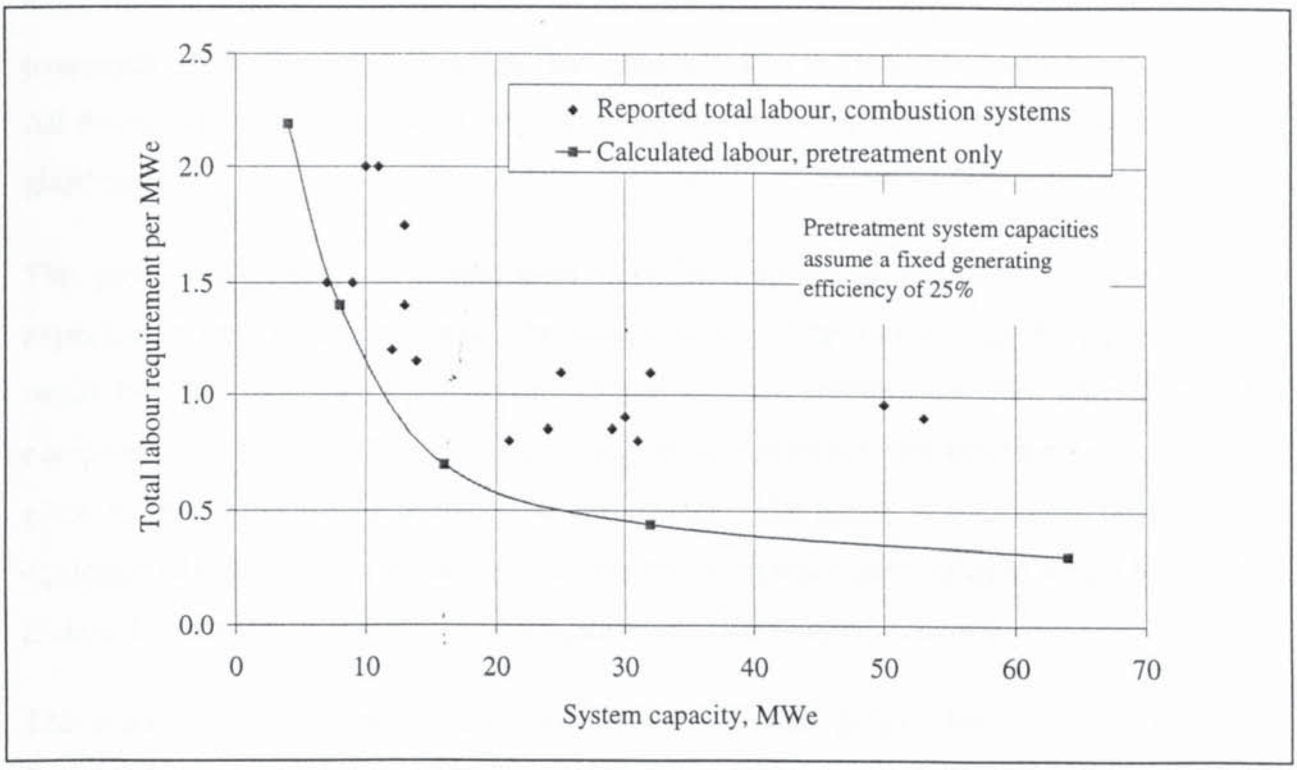


Figure 6.16 - Labour requirements for pretreatment compared with plant data

6.10.4 Maintenance and overheads

Maintenance costs for the different equipment used in pretreatment are highly variable. While it would be feasible to calculate the maintenance cost on an equipment item basis, it has been decided to simplify the model and for consistency with the other modules an overall maintenance cost will be used in accordance with the methodology in Section 4.6.6.

Overheads are also calculated using an overall percentage of the total plant costs for the module, as described in Section 4.6.6.

6.11 SUMMARY

This module models the processes required to carry out the reception, storage, handling and pretreatment of a wood chip feedstock to ensure its continuous supply to the conversion reactor in a form that suits the needs of the conversion technology. The module models the processes required in a sequence of process steps, with each process step model containing relationships for the total plant cost, power requirement and labour requirement of that step. Capital, power consumption and labour are summed for all step models that are required to meet the feed specification imposed by the conversion technology. Capital amortisation, overheads and maintenance costs are calculated based on the total plant cost for all the steps. All the capital costs are based on the costs of established equipment and are regarded as 100th plant costs.

The processes required in pretreatment have been modelled explicitly because they are expected to vary with capacity and between systems. The variation in the equipment that could be used in each step is extensive and this has complicated the definition of the equipment that has actually been used. In particular, the model uses strict limits on where one piece of equipment would be used and not another. The reality is a lot more flexible. The equipment defined here is suitable for the job and is representative of what would be used but it should always be noted that other equipment could have also been used.

The module reported here fulfils the objectives of this project but could be improved. Possible further work is listed below.

- Improve the algorithms for selecting equipment when using multiple trains to try and reduce the extra capital costs imposed by over-sized equipment.

- Examine the drying step in more detail, looking at the relationship between dryer cost and efficiency and evaluating alternative types of dryers.
- Examine the grinding step in more detail, given its high cost and high power consumption. In particular, more up to date costs for the pulverizer should be obtained.
- Examine the effect over the project life of replacing equipment, since many items will not last for the full term of the project.
- Assign maintenance costs by equipment or step.

6.12 NOMENCLATURE

Any mass values in tonnes are wet tonnes unless the units odt (oven dry tonnes) are used.

All costs are US$, 1995 basis unless specifically noted otherwise.

$Q_{d,del}$ = the daily feed delivery rate, t/d

$Q_{h,del}$ = the hourly feed delivery rate, t/h

$Q_{h,dry}$ = the tonnes of water evaporated from a single dryer, twe/h

$Q_{h,grd}$ = the hourly feed rate through the grinding step, t/h

$Q_{h,ovr}$ = the hourly overs rate through the re-chip step, t/h

$Q_{h,rlm}$ = the hourly feed rate from storage, t/h

P_{bfr} = Installed power required by the buffer storage step, kW_e

P_{dry} = Installed power required by the drying step, kW_e

P_{grd} = Installed power required by the grinding step, kW_e

P_{rcp} = Installed power required by the re-chip step, kW_e

P_{rec} = Installed power required by the reception step, kW_e

P_{sc1} = Installed power required by the pre-storage screening step, kW_e

P_{sc2} = Installed power required by the post-storage screening step, kW_e

P_{sto} = Installed power required by the storage step, kW_e

TPC_{bfr} = Total plant costs, buffer storage step, $\$k_{1995}$

TPC_{dry} = Total plant costs, drying step, $\$k_{1995}$

TPC_{grd} = Total plant costs, grinding step, $\$k_{1995}$

TPC_{rec} = Total plant costs, reception step, $\$k_{1995}$

TPC_{rcp} = Total plant costs, re-chip step, $\$k_{1995}$

TPC_{sn1} = Total plant costs, pre-storage screening step, $\$k_{1995}$

TPC_{sn2} = Total plant costs, post-storage storage step, $\$k_{1995}$

TPC_{sto} = Total plant costs, storage step, $\$k_{1995}$

$V_{h,dry}$ = Total volume flow rate from dryers, m^3/h

$V_{h,bfr}$ = Total volume flow rate into buffer storage, m^3/h

n_{dry} = No. of dryers

n_{grd} = No. of grinders

n_{silo} = No. of silos in buffer storage

7. FEED CONVERSION MODULES

7.1 INTRODUCTION

This chapter describes four modules (sub-models) that calculate the cost and performance of converting the prepared wood feedstock into an intermediate energy carrier. The four feed conversion modules are summarised in Table 7.1, and their position in the overall model can be seen in Figure 4.1. These conversion processes were selected in Chapter 2 for evaluation and their principles and status were described in that earlier discussion.

Table 7.1 - The feed conversion modules

	Combustion	**Fast Pyrolysis**	**Atmospheric gasification**	**Pressurised gasification**
See	Section 7.2	Section 7.3	Section 7.4	Section 7.5
Previous module	Pretreatment, Chapter 5			
Feed conditions at reactor entry	Wood chips, 35% moisture	Wood powder, 10% moisture	Wood chips, 15% moisture	Wood chips, 15% moisture
Energy carrier produced	Superheated steam	Pyrolysis liquid	Cool, clean fuel gas	Hot, clean fuel gas
Next module	Steam cycle, Section 8.2	Liquid fired dual fuel engine, Section 8.3	Gas fired dual fuel engine, Section 8.4	Gas turbine combined cycle, Section 8.5

The methodology for calculating the cost and performance criteria has been described in Chapter 4. This chapter applies this methodology to produce conversion modules and each module description includes:

- The constraints imposed by the conversion technology on the wood feedstock;
- The performance of the conversion step, expressed as a ratio of the energy in the intermediate energy carrier to the energy in the prepared feedstock (defined in Section 4.4.1);
- The total plant costs for the equipment covered by the conversion module; and
- The operating costs incurred.

7.2 THE COMBUSTION MODULE

The combustor and boiler that are modelled here are referred to collectively as a combustor. The relationships derived in the rest of this section model apply to a generic fluid bed or circulating fluid bed combustor. While there are significant design differences between the two configurations, at this level their cost and performance parameters are very similar [41].

7.2.1 Feed requirements

Fluid bed combustors will accept a wide range of particle sizes up to 50 mm [77]. When feedstocks contain a high proportion of fines these may be blown through the combustor before complete combustion but the module will assume that fines are not significant in the wood chips used. Therefore special feed preparation for size is not required other than screening to remove over-size material.

Low moisture content feedstocks are preferred because they increase combustion efficiency; they reduce flue gas volumes and hence the capital cost of flue gas equipment; they reduce carbon carry-over and particulate emissions; they raise the dew point in the stack and allow more control over the combustion process. Although drying the feedstock before combustion can benefit the combustion process, the system as a whole can suffer by virtue of the additional capital cost for the dryer and changes to the system to make heat available for drying [186, 242].

The viability of drying the feedstock in combustion systems is a complex issue and a thorough examination of this topic would require a detailed examination of the waste heat available in the system and the feed characteristics. In this generic combustion model, it has been assumed that the default wood feedstock (at 50% moisture content, see Section 4.3), will be dried to a moisture content of 35%, which is considered a reasonable compromise between enhanced system performance and the increased capital costs [57]. It will be seen in Section 7.2.2 that combustor efficiency is only marginally improved by drying below this level and hence the return on capital invested in more extensive drying equipment is reduced.

7.2.2 Performance

In this work the defining criterion of the conversion technology is its ability to convert the chemical energy in the feed into usable energy in the intermediate energy carrier produced (Section 4.4.1). For this module, this definition gives Equation 7.1.

$$\eta_{conv,comb},\% = \frac{\text{Energy added to the feed water and steam, GJ / h}}{\text{Lower heating value of the feed input, GJ / h}} \tag{7.1}$$

where $\eta_{conv,comb}$ = the conversion efficiency of the combustion module, %

The energy efficiency is calculated by considering the mass and energy flows into and out of the reactor. The reactor is considered as a sub-system: the processes within the sub-system are not important, all that is required are the mass and energy inputs and outputs. Assuming the bed material inventory is constant, the mass and energy inputs for the combustor are as shown in Figure 7.1. The energy flows have been simplified by including the air preheater within the limits of the sub-system and assuming that all the energy required to preheat the air comes from the flue gases.

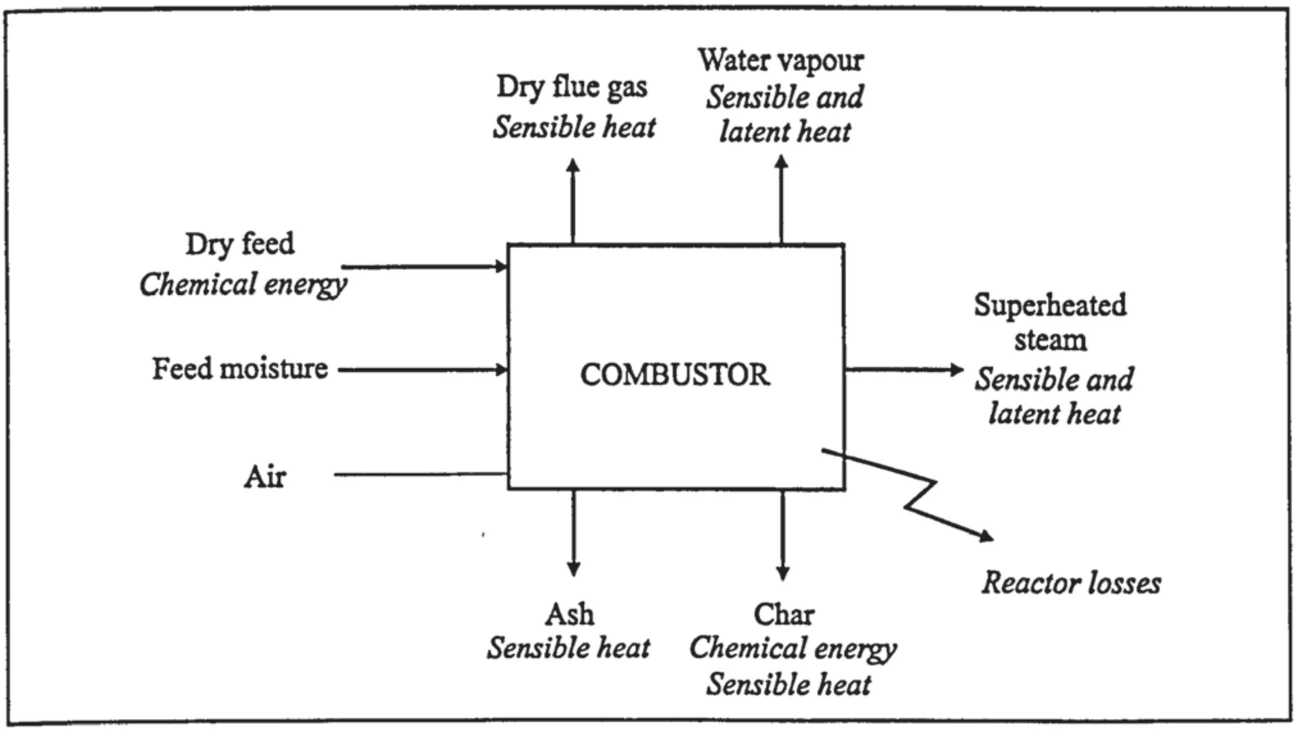


Figure 7.1 - The combustor mass and energy flows

The energy added to the feed water to produce superheated steam is found by difference, after accounting for all the other energy fluxes shown in Figure 7.1. This is a standard approach to calculating the efficiency of a combustor [249]. The mass and energy balance over the combustor is performed on a spreadsheet that allows the efficiency to be calculated for any feed moisture content and combustor input capacity. This data sheet is included in Appendix D.

The mass balance requires a stoichiometric air requirement. This is calculated on the data sheet in Appendix D and uses the ultimate analysis of the feedstock given in Section 4.3. Excess air will be required above the stoichiometric air flow to ensure complete combustion. The excess air required to give complete combustion in fluid beds has been reported to be as low as 15-25% [41, 45] due to the turbulence in the bed that encourages mixing of the air and fuel. However, performance data on 5 fluid bed plants in California gives a range of 36-62% excess air with an average of 46.6% [31] and this mean value is used in the module such that the actual air is 147% of the stoichiometric air requirement. This excess air is still low compared with grate-fired combustors where the excess air is commonly 100% above the stoichiometric requirement [250].

The mass balance assumes a burn out rate of 99%, typical for fluid bed and circulating fluid bed combustors [41, 71, 72, 74]. The 99% of the feedstock that burns is assumed to burn completely, the 1% of feedstock that remains is assumed to be completely unreacted. These two assumptions simplify the stoichiometry required for the mass balance. In reality a small quantity of incomplete combustion products will be found in the flue gas [31, 182], and the feed that leaves with the ash will be partially reacted and be in the form of char. However, these factors can be ignored since the carbon conversion efficiency is so high that the simplifications will have a negligible impact. The rest of the mass balance calculations are trivial and the mass balance can be seen in Appendix D.

The energy balance is based on the higher heating value of the feedstock, although the efficiency that is calculated will be based on the lower heating value to follow the conventions set down in Chapter 4. The higher heating value is used since this represents the total chemical energy in the feedstock. The higher heating value of the feedstock is calculated from the lower heating value using Equation 7.2, where the hydrogen content of the feedstock is as specified in Section 4.3.

$$\mathrm{HHV_0, GJ/odt} = \mathrm{LHV_0, GJ/odt} + \left(\mathrm{H, \%odt} \times \frac{18}{2} \times 2.454\,\mathrm{GJ/t}\right) \tag{7.2}$$

where HHV_0 = Higher heating value of feed at 0% moisture content, GJ/odt

LHV_0 = Lower heating value of feed at 0% moisture content, GJ/odt

H = Hydrogen in feedstock, %odt

PAGE

NUMBERING

AS ORIGINAL

$Q_{h,pret,dry}$ = Dry feed input to the combustor, odt/h

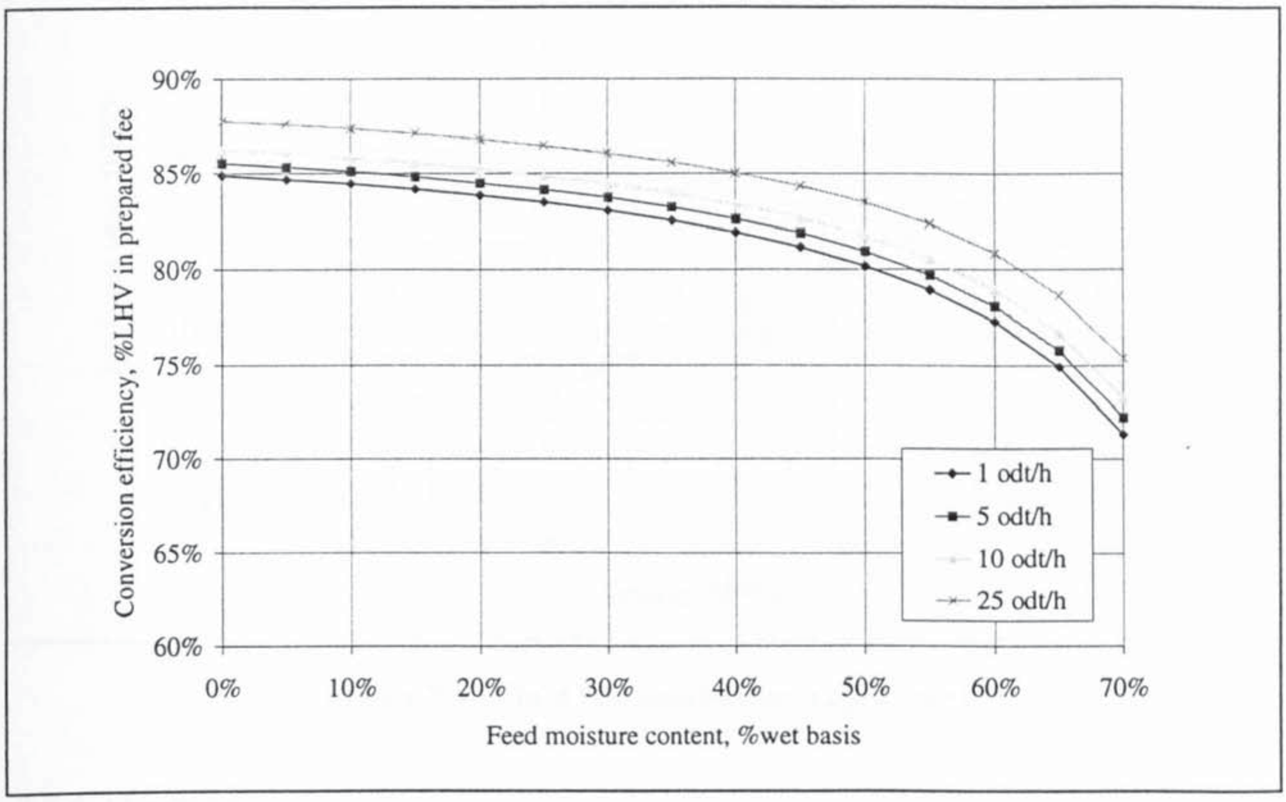


Figure 7.2 - Combustion efficiency as a function of feed moisture content

7.2.3 Capital costs

Reported capital costs of bubbling and circulating fluid bed combustors and boilers have been normalised to a total plant cost basis using the ratios method described in Section 4.5.3. The results are shown in Figure 7.3 [29, 41, 71, 251]. The best regression curve is shown on the graph and total plant costs for the combustion module are calculated using Equation 7.8. The scale exponent shown may appear to be higher than expected, given that steam cycle plant are acknowledged to be very sensitive to scale, but the capital costs have been based on the thermal input. Given that the efficiency of the combustor increases with scale, the actual cost of plant based on their output will decrease more rapidly than Equation 7.5 would suggest.

$$TPC_{conv,comb},\ US\$_{1995} = 697\left(E_{th,pret}\right)^{0.77} \qquad (7.5)$$

where $E_{th,pret}$ = the energy in the prepared feedstock, MW_{th} LHV

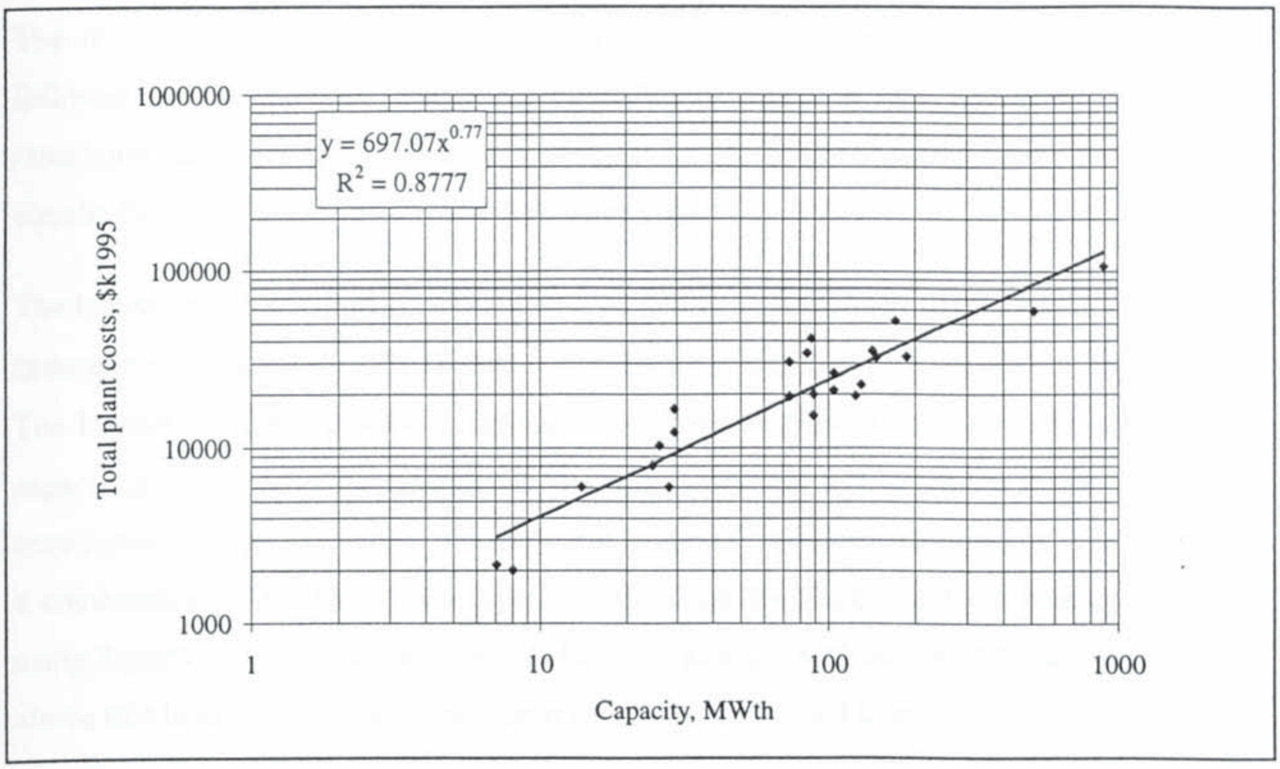


Figure 7.3 - Fluid bed combustion capital cost

7.2.4 Operating costs

7.2.4.1 Labour requirement

No labour requirements could be found for just the combustor in a generating system (i.e. excluding pretreatment and the steam cycle) and so the labour requirement for the combustor must be derived from reported labour requirements for whole systems, reported by Easterly [47] and others [251, 240, 71].

The first stage in this procedure is to extract a component of the labour for pretreatment, since this was already accounted for in the pretreatment module. Ideally, the calculated labour for pretreatment could have been removed directly, but the stepped nature of the pretreatment costs and their inaccuracy at small scale distorted the remaining labour, preventing any further meaningful analysis. Instead an arbitrary function was used to account for pretreatment labour. This function assumes that 66% of the labour in a small scale facility (1 MW_e) is for pretreatment of the wood chips, reducing logarithmically to 33% at 100 MW_e. This allows for the high manual handling during reception, storage and handling in small scale systems and the more automated processing in larger facilities.

The results of this procedure are given in Table 7.2. The reported labour is given first, followed by the assumed labour for pretreatment which is then subtracted to give the remaining labour for the combustion and steam cycle. This remaining labour has been split equally between the conversion and generating step.

The labour for the combustion module was plotted against thermal input, assuming a constant generating efficiency of 25% (Figure 7.4). Two regression curves are shown on this chart. The logarithmic curve has the advantage that it does not predict a steep rise in labour at low capacities, but it gives poor results at the high scales and would predict a very low labour requirement at high capacities. The converse is true of the power relationship. For this reason a combination of the two curves is used with labour for the combustion module calculated using Equation 7.6 at thermal inputs below 140 MW_{th}, and Equation 7.7 used at capacities above that limit. In both cases the labour calculated is the total labour for 5 shifts.

$$\text{Labour}_{\text{combustion}} = \left(-0.0488\ln\left(E_{th,pret}\right)+0.3001\right)\times E_{th,pret} \tag{7.6}$$

$$\text{Labour}_{\text{combustion}} = \left(0.9098\left(E_{th,pret}\right)^{-0.5539}\right)\times E_{th,pret} \tag{7.7}$$

where $E_{pret,MWth}$ = Energy available in the prepared feedstock, MW_{th}

These equations will give labour requirements of between 0.23 and 0.03 operators/year/MW_{th}. Evald suggests that a typical combustion plant requires 0.55 operators/year/MW_{th} following a survey of 30 Danish CHP facilities [252]. The labour requirement varied widely, with values from 0.2 to 1 operators/year/MW_{th} between capacities of 1-10 MW_{th} input. The labour requirements given by Equation 7.7 shows a reasonable agreement in this capacity range, producing values of 0.12-0.23 operators/year/MW_{th}, given that the pretreatment labour requirement is approximately 1-0.6 operators/year/MW_{th} over the same range.

Table 7.2 - Allocation of labour in combustion systems

Capacity		Total labour		Pretreatment labour	Remaining labour	Combustion labour		Source
MWe	MWth[a]	Total	/MWe	/MWe[b]	/MWe	/MWe[c]	/MWth[a]	
7	28	11	1.50	0.95	0.55	0.47	0.12	[47]
9	36	14	1.50	0.94	0.56	0.47	0.12	[47]
10	40	20	2.00	1.24	0.76	0.62	0.15	[47]
11	44	22	2.00	1.23	0.77	0.62	0.15	[47]
12	48	14	1.20	0.73	0.47	0.37	0.09	[47]
13	52	18	1.40	0.85	0.55	0.42	0.11	[47]
13	52	23	1.75	1.06	0.69	0.53	0.13	[47]
14	56	16	1.15	0.69	0.46	0.35	0.09	[251]
21	84	17	0.80	0.46	0.34	0.23	0.06	[47]
24	96	20	0.85	0.48	0.37	0.24	0.06	[47]
25	100	28	1.10	0.61	0.49	0.31	0.08	[47]
29	116	25	0.85	0.46	0.39	0.23	0.06	[47]
30	120	27	0.90	0.48	0.42	0.24	0.06	[71]
31	124	25	0.80	0.43	0.37	0.21	0.05	[47]
32	128	35	1.10	0.58	0.52	0.29	0.07	[47]
50	200	48	0.96	0.45	0.51	0.22	0.06	[240]
53	212	48	0.90	0.41	0.49	0.21	0.05	[47]

a *Assuming a constant generating efficiency of 25%*
b *Pretreatment labour calculated using a logarithmic relationship between 1 and 100 MW_e; 66% of labour is pretreatment at 1 MW_e, 33% of labour is pretreatment at 100 MW_e.*
c *Half of the remaining labour.*

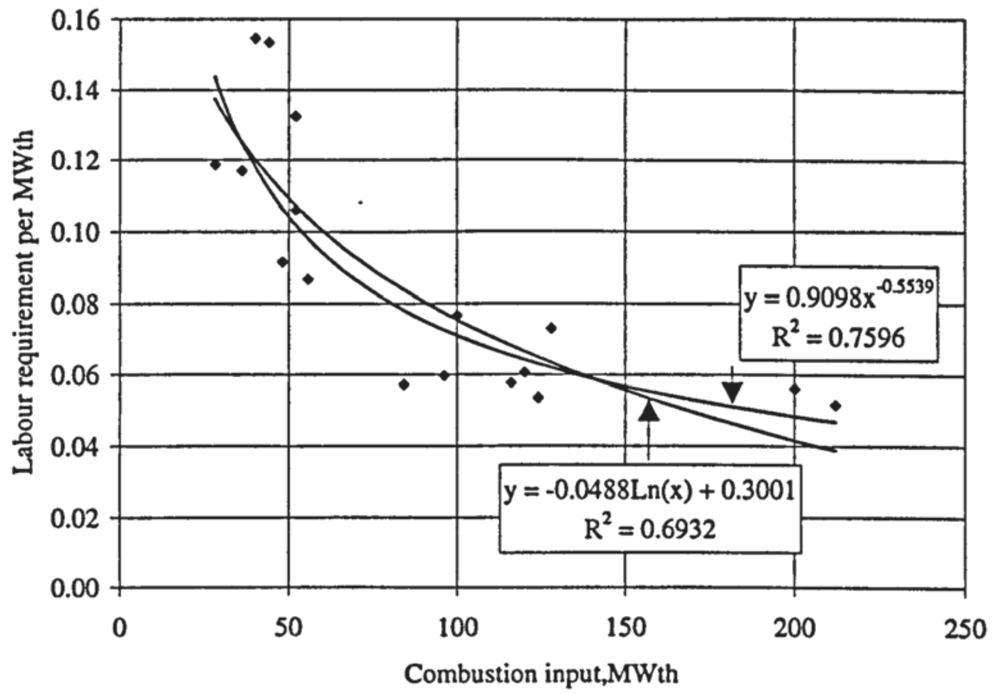


Figure 7.4 - Specific labour requirements, combustion module

7.2.4.2 Utilities

The only utility that is considered here is the internal power consumption. Boiler feed water and cooling feed water requirements of the steam cycle are calculated in the steam cycle module (see Section 8.2.4.1).

Reports of the auxiliary power consumption for fluid bed combustion plant rarely consider the combustor alone and it has been difficult to estimate the power consumption for the module. A survey of 5 Californian fluid bed and circulating fluidised bed combustion plant [31] showed power consumption for the whole combustion and generating system of between 10.6-17.7% of the gross power output for net capacities of between 18.6 and 31.9 MW_e. The average internal power consumption was 13.6% of the gross power output, which was approximately 4% of the energy input. Evald [252] showed that the power consumption (excluding the heating system) in 30 Danish wood combustion plant was 1.6-2.4% of the thermal input in MW_{th} (the average value was 1.9%). On the basis of this limited data, a value of 2% of the thermal input is used as a default in the module. The impact of internal consumption on the overall system will be tested by sensitivity analysis in Chapter 9.

7.2.5 Waste heat available

This section briefly examines the amount of energy available in the flue gases that could be used to dry the incoming feedstock to 35%. The combustion performance data sheet in Appendix D shows the values that are used given an undried feed moisture content of 50%. The calculations shown assume that the flue gases enter the dryer at 175°C and leave the dryer at 102°C, as specified in Section 6.7.1. The dryer evaporation load is given by Equation 7.8. The energy available is the sum of the enthalpy available in the dry flue gas and the enthalpy available in the water vapour. This is calculated using Equation 7.9.

$$\text{Dryer load, twe / h} = Q_{h,dry} \times \left(\frac{x_{in}}{1 - x_{in}} - \frac{x_{out}}{1 - x_{out}} \right) \tag{7.8}$$

$$\text{Energy available, GJ / h} = \left(E_{flue} \times \frac{(T_{in} - T_{out})}{(T_{in} - T_o)} \right) + \left(Q_{H_2O} \times (h_{in} - h_{out}) \right) \tag{7.9}$$

where x_{in}, x_{out} = the moisture content at entry and exit of the dryer

E_{flue}	= the enthalpy of the flue gases at entry to the dryer
T_{in}, T_{out}	= the temperature of the flue gas at entry and exit of the dryer
T_o	= the reference temperature, 20°C
Q_{H2O}	= the mass flow rate of water vapour, t/h
h_{in}, h_{out}	= the enthalpy of the water vapour at entry and exit of the dryer
twe	= tonne of water evaporated

Given the dryer energy requirement of 3.553 GJ/twe specified in Section 6.7.1, there is a shortfall of around 20% between the energy available in the flue gas and the energy required by the dryer. This does not necessarily mean that there is insufficient energy to support drying, but is does means that the system should be examined in more detail for sources of waste heat. The shortfall could be met by a combination of a more efficient dryer, a lower dryer exit temperature or a higher flue gas temperature. A more efficient dryer would not change the system performance but could cost more and this extra cost has not been calculated. The lower dryer exit temperature may produce a plume at the dryer exit that could be against local regulations. Frea [242] reports that temperatures as low as 80°C could be used at the dryer exit, which reduces the energy deficit to 2% without changing the efficiency of the system. A higher flue gas temperature would reduce the energy recoverable for steam raising and air preheating and would therefore impact on the system. It is assumed that a combination of the above can be used to fulfil the drying requirement without significantly changing the system performance or cost and no adjustment will be made at this stage. However, this energy deficit does highlight an uncertainty in the model that should be examined in more detail in further work.

7.3 THE FAST PYROLYSIS MODULE

7.3.1 Feed requirements

The feedstock delivered to the fast pyrolysis facility is wood chips of an unspecified species mixed with 15% bark and with the characteristics given in Section 4.3.

It has already been noted in the original discussion on fast pyrolysis in Section 2.7.2 that fast pyrolysis in a fluid bed requires a feed particle size of less than 2 mm for high liquids yields [146, 147, 148, 151, 253]. For this reason the delivered wood chips must be ground to a powder before entering the reactor. This step is carried out by equipment modelled in the pretreatment module (6.8) and is only necessary for this conversion technology.

A feed moisture content of 10% is specified, which will be achieved by drying the delivered feedstock before grinding. This process is also modelled in the feed pretreatment module (Section 6.7). Most published data on fast pyrolysis is based on dried feedstocks with moisture contents of 4-7%. Higher moisture contents have two effects: firstly the moisture in the feed reduces heating rates and increases char production; this char catalyses secondary reactions of the primary tars [145] and reduces the organics yield. Secondly this extra moisture in the feed dilutes the liquid product so that its energy value is reduced. It is likely that the cost of feed drying in commercial processes would prohibit such low moisture contents as 4-7%, and hence the specification of 10%. Another reason to use a slightly higher feed moisture content is that moisture in the liquid reduces its viscosity and make it easier to use in an engine [254]. It would certainly be worth investigating the trade off between the costs of drying and the effects on the properties of the liquid product in further work.

7.3.2 Performance

In this work the defining criterion of conversion technology is its ability to convert the chemical energy in the feed into usable energy in the intermediate energy carrier produced (Section 4.4.1). For this module, this definition gives Equation 7.9.

$$\eta_{conv,pyr},\% = \frac{\text{Lower heating value of the pyrolysis liquid produced, GJ / h}}{\text{Lower heating value of the feed input, GJ / h}} \tag{7.9}$$

where $\eta_{conv,pyr}$ = the conversion efficiency of the fast pyrolysis module, %

Just as in the combustion module, the fast pyrolysis module may be considered as a sub-system that includes the pyrolysis reactor; its feeder and the liquids recovery system. The mass and energy flows as shown in Figure 7.5. In this work the reactions and pathways within the sub-system are not important: all that is required is a relationship to give the usable energy in the energy product. The energy in the conversion product can be established by calculating the energy output in the pyrolysis liquid as the product of:

- the pyrolysis liquid yield; and
- the lower heating value of the pyrolysis liquid.

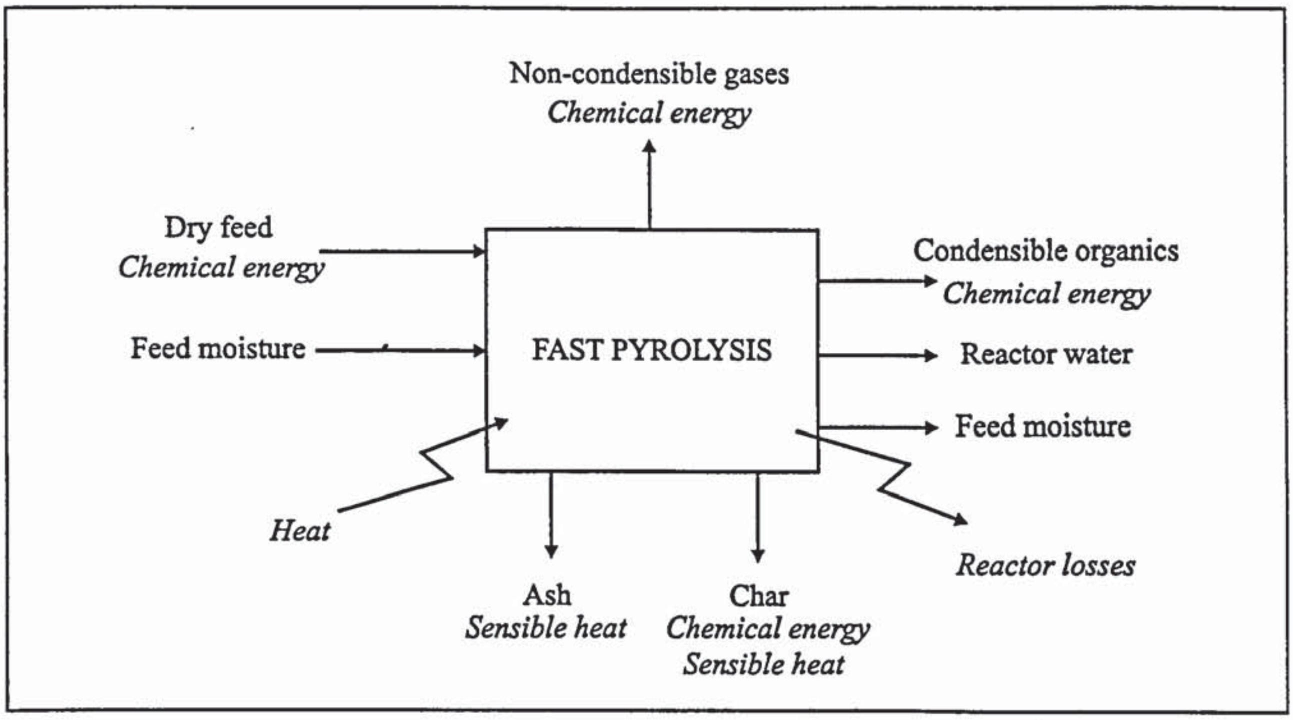


Figure 7.5 - Mass and energy flows in fast pyrolysis

7.3.2.1 The pyrolysis liquid yield

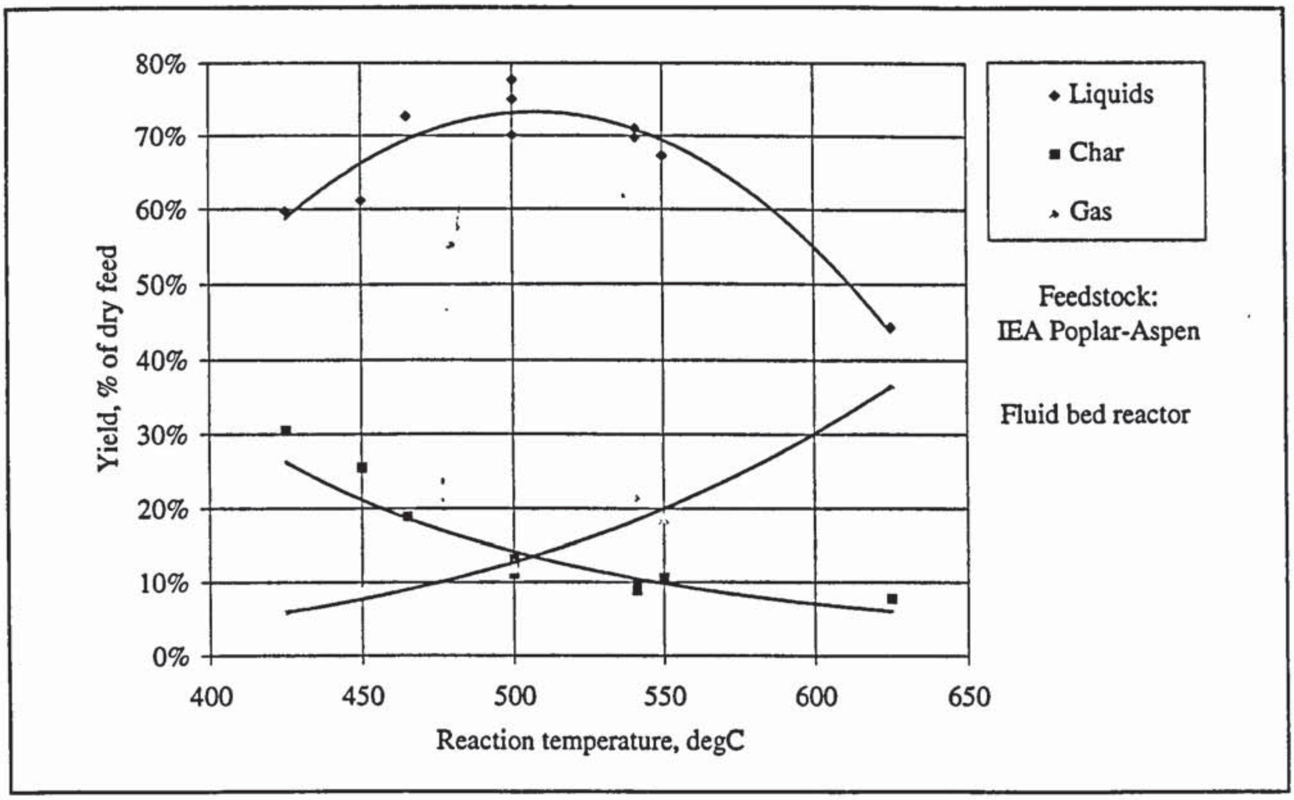
The current approach to fast pyrolysis process design for fuels is maximisation of liquid product yield, even though this liquid may not be the optimum quality. This will be the basic premise of this section: to establish the maximum liquids yield that could be attained in a commercial fast pyrolysis system. More specifically, the aim is to maximise the organic fraction of the liquid product since it is this fraction that carries the energy (see Figure 7.5). Fast pyrolysis yields and product characteristics are a function of a number of interdependent process parameters that have been investigated in bench or pilot-scale equipment. This section will use experimental data from small-scale equipment to predict yields in commercial facilities with the wood/bark feedstock. Experiments have shown that maximum organics yields in fluid bed fast pyrolysis are achieved using the following conditions [145]:

- a reactor temperature of 450-625°C;
- a gas/vapour product residence time of <2s;
- a gas/vapour product temperature of 400-500°C;
- atmospheric reactor pressure;
- a particle size of less than 2 mm; and
- a biomass moisture content of less than 10%.

Other issues that could impact on yields are the liquids recovery equipment and any treatment of the vapours or liquid to improve its properties.

The reaction temperature is the single most significant influence on the liquids yield, and yield data with respect to temperature can be used to establish what the theoretical maximum yield would be in a system operated for liquid fuels production as a basis for predicting commercial yields. The effect of reaction temperature on yields is exemplified in Figure 7.6 [147], showing that maximum liquids are achieved at temperatures around 500°C. In this case the organics and reaction water have been considered together; all further graphs will separate the organic and water yields so that the analysis can focus on the production of condensible organics.

The optimum reaction temperature varies with the feedstock. Figure 7.7 illustrates the effect of feed material on organics yield for small-scale fluid bed fast pyrolysis at different reaction temperatures and vapour residence times of less than 1s [147, 151, 255, 256, 257]. It can be seen that the variation in yield between the wood feedstocks is much less than the variation in the overall group. On this premise all wood feedstocks are considered together so that the yield used by the module can be applied to any wood-fed process.



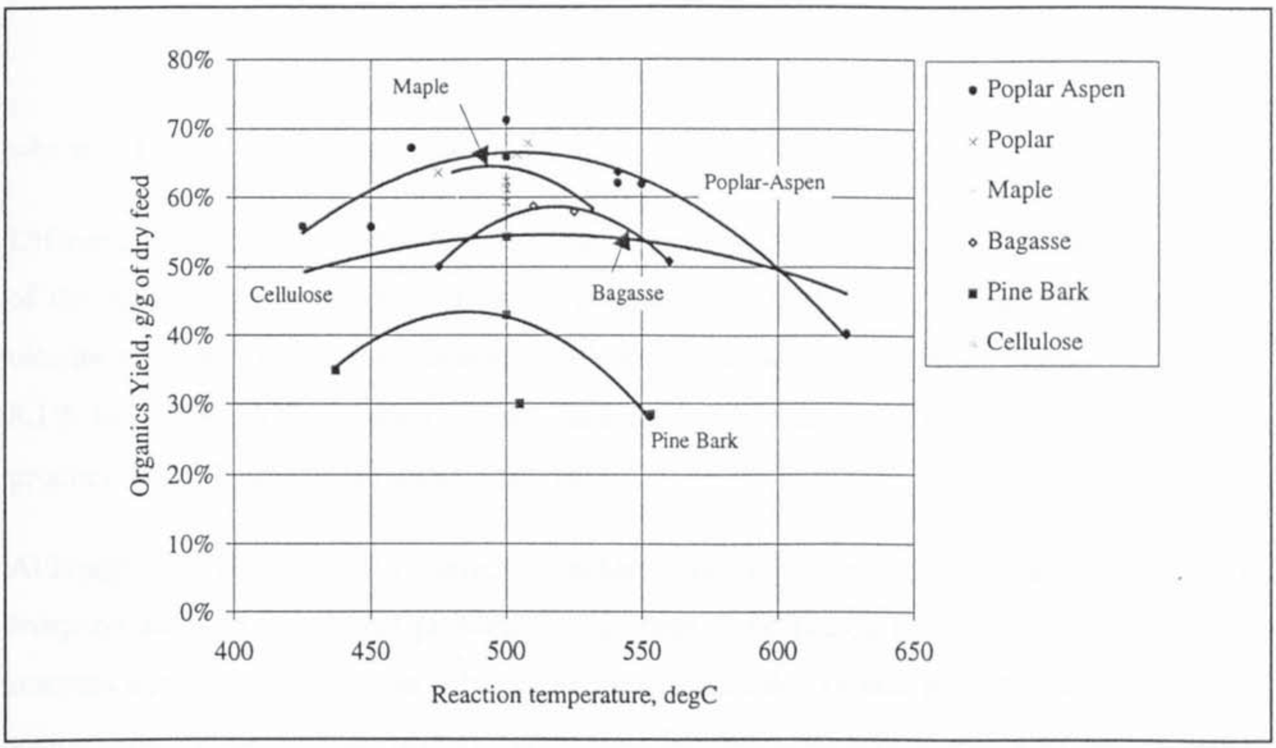


Figure 7.7 - Variation of organics yield with feedstock

All yield data (as a percentage of the dry wood input) is shown together for fluid bed fast pyrolysis in Figure 7.8 and Figure 7.9 against reactor temperature. The condensible organics and reaction water yields are shown in Figure 7.8 and char and off-gas yields are shown in Figure 7.9. Regressions on each of the four yield data are used to predict the products from a fast pyrolysis process operated for maximum organics yield. The regressions give Equations 7.10 to 7.13.

$$\text{Organics,\%} = -16.4 \times 10^{-6}(\mathrm{T})^2 + 0.0164(\mathrm{T}) - 3.47 \tag{7.10}$$

$$\text{Gas,\%} = 8.06 \times 10^{-6}(\mathrm{T})^2 - 0.0071(\mathrm{T}) + 1.64 \tag{7.11}$$

$$\text{Char,\%} = 8.60 \times 10^{-6}(\mathrm{T})^2 - 0.0099(\mathrm{T}) + 2.95 \tag{7.12}$$

$$\text{Water,\%} = 4.70 \times 10^{-6}(\mathrm{T})^2 - 0.0049(\mathrm{T}) + 1.20 \tag{7.13}$$

where T = Reaction temperature, °C

$$\text{Water},\% = 4.70 \times 10^{-6}(T)^2 - 0.0049(T) + 1.20 \tag{7.13}$$

where T = Reaction temperature, °C

Differentiating Equation 7.10 gives the maximum organics yield of 64.9% at 502°C. Yields of the other products at this temperature are shown in Table 7.3. The initial results gave a closure of 97.4%, which was adjusted to 100% by increasing the yield of reaction water from 8.1% to 10.8%. This produce a more realistic water yield since the dry feedstock should produce 10-12% of reaction water [150, 145].

Although the heat fluxes required to achieve rapid heating of the feedstock to these temperatures is a recognised problem [143], there is no reason to assume that commercial reactors cannot be designed to achieve the temperature that should give the maximum liquid yields shown in bench scale performance. Therefore this treatment of the experimental data is used to define the maximum yields that could be expected in commercial systems. The following discussion examines issues that could prevent the process from attaining such yields.

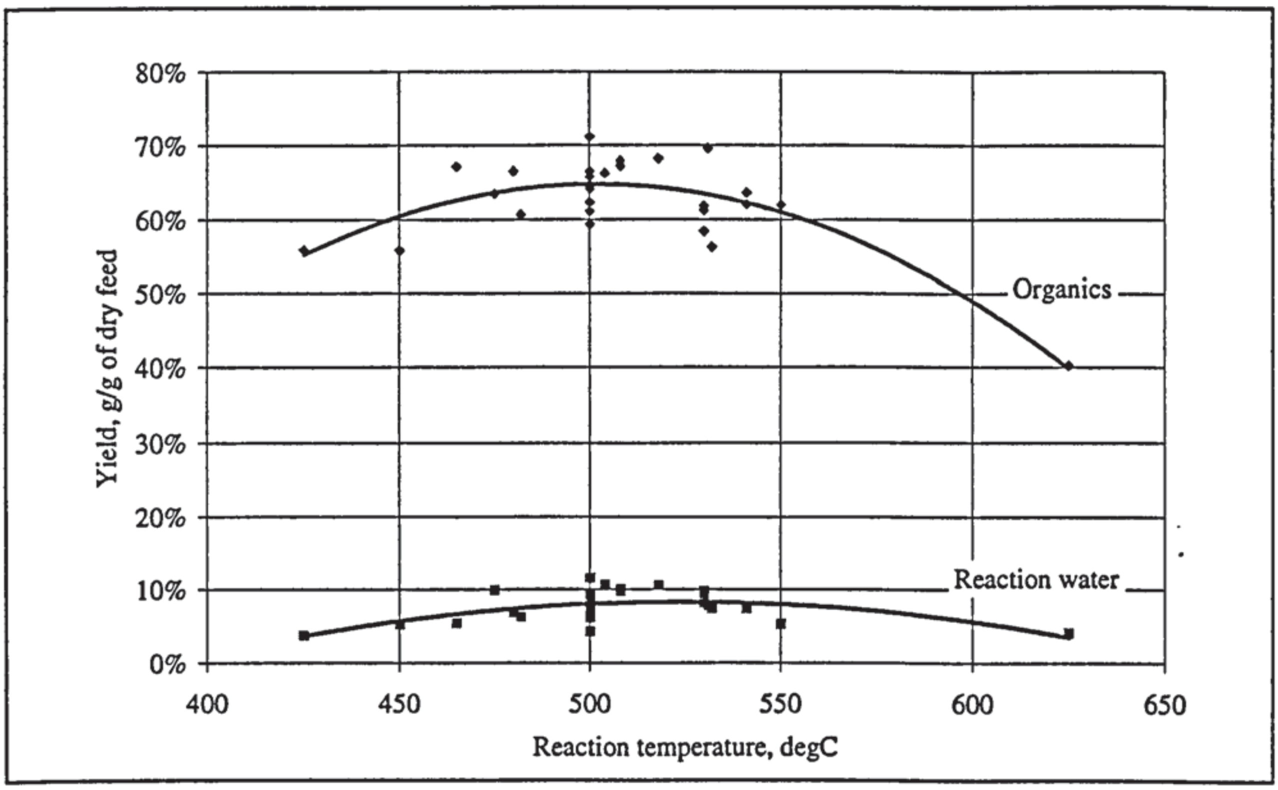


Figure 7.8 - Organics and Water Yields, Fluid Bed Fast Pyrolysis of Wood

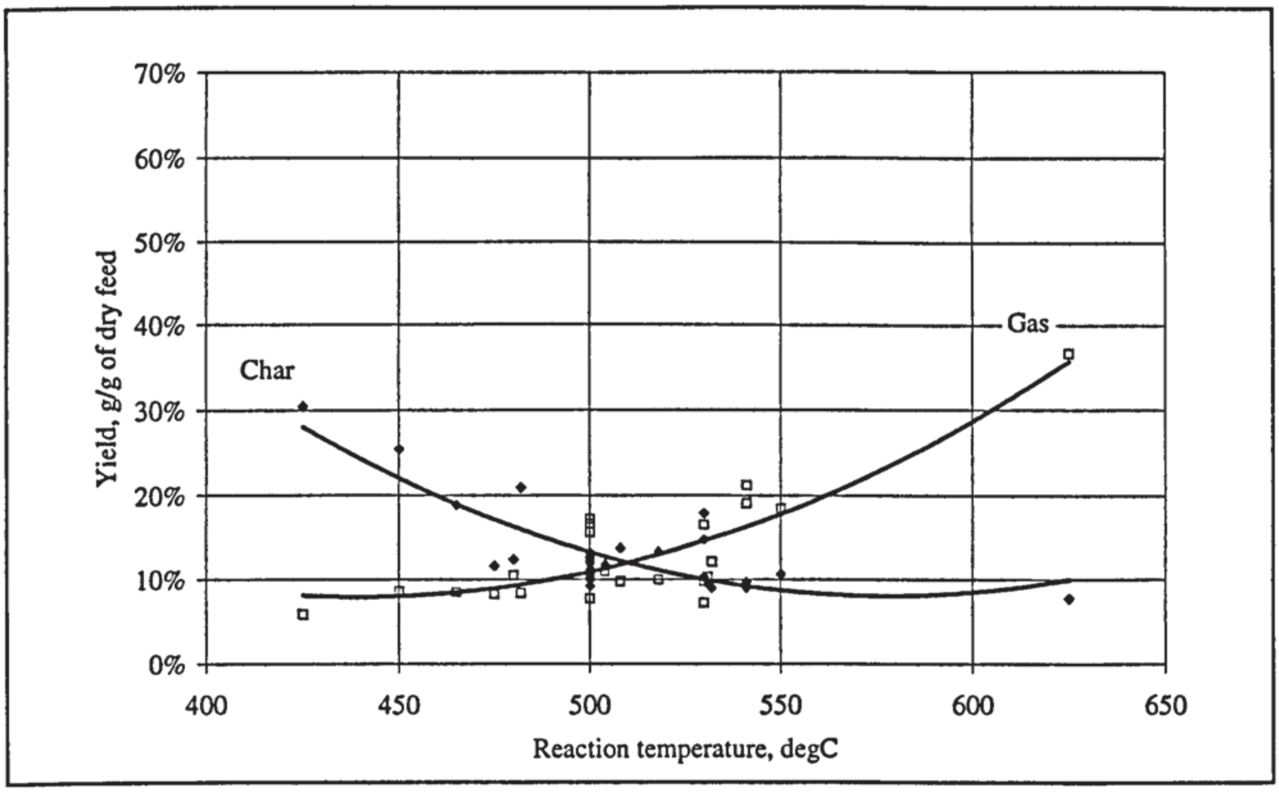


Figure 7.9 - Char and Gas Yields, Fluid Bed Fast Pyrolysis of Wood

Table 7.3 - Ideal fast pyrolysis yields for maximum organics production

Product	Predicted yield	Normalised predicted yield
Organics	64.9%	64.9%
Gas	11.1%	11.1%
Char	13.2%	13.2%
Water	8.1%	10.8%
Total	97.4%	100.0%

The first factor that should be considered is the feed material. The analysis above is based on wood feedstocks, free of bark. Such material would not be available in bulk quantities for energy since it could attract much higher prices as a pulping feedstock. Thus the feed considered here is a whole tree feedstock of much lower value and contaminated by bark. Figure 7.7 shows that the yields for clean wood feedstocks are considerably higher than those for bark. This is due to the higher ash content of the bark, which catalyses reactions in the primary pyrolysis vapours to produce more char and gas [151, 294]. The yields for a wood/bark mixture as used here are predicted in Table 7.4 which assumes that the yields of the two feedstocks are additive (i.e. the two feedstocks do not interact). The calculated figures compare well with the experimental data using poplar containing 15-20% bark [258]. The results are shown graphically Figure 7.10 and on this evidence it is assumed that the level of bark contamination in the feedstock reduces the yield from pure wood feedstocks by 4-5 percentage points.

Another important issue is the time-temperature profile of the organic vapours before they are condensed. Long residence times at high temperatures encourage secondary reactions that form char and crack the condensible vapours to non-condensible gases [145]. The effect is shown in the data Figure 7.11 [147]. Peacocke has concluded from his work that organics yields are not significantly effected if the gas/vapour temperature is maintained below 500°C and the residence time below 2s [145]. Models by Liden and Diebold support this hypothesis [258]. There is a risk in large scale systems that the residence times in the reactor and liquids recovery system cannot be limited to 2s, but there is no evidence to show that the system cannot be designed to meet this criterion. This work will assume that the residence time will be below 2s and hence yields do not need to be adjusted, but there is some uncertainty surrounding this issue.

Table 7.4 - Mixed Feedstock Analysis

Feedstock		Poplar	Bark[a]		Whole tree poplar
Reference		[257]	[259]		[258]
Reactor Conditions					
	Reaction temperature, °C	504	500		504
	Residence time, s	0.5	0.5		0.5
Yields, g/g dry feed					
	Organics	66.2	37		62.9
	Water	10.7	9		9.7
	Char	11.8	33		16.5
	Gas	11	13		11.5
Mass Balance, 100g total dry feed input					
Input, g		85	15	100 (total)	100
Output, g	Organics	56.3	5.6	61.8	62.9
	Water	9.1	1.4	10.4	9.7
	Char	10.0	5.0	15.0	16.5
	Gas	9.4	2.0	11.3	11.5

a *Yields for pine bark are used, no data available for poplar bark*

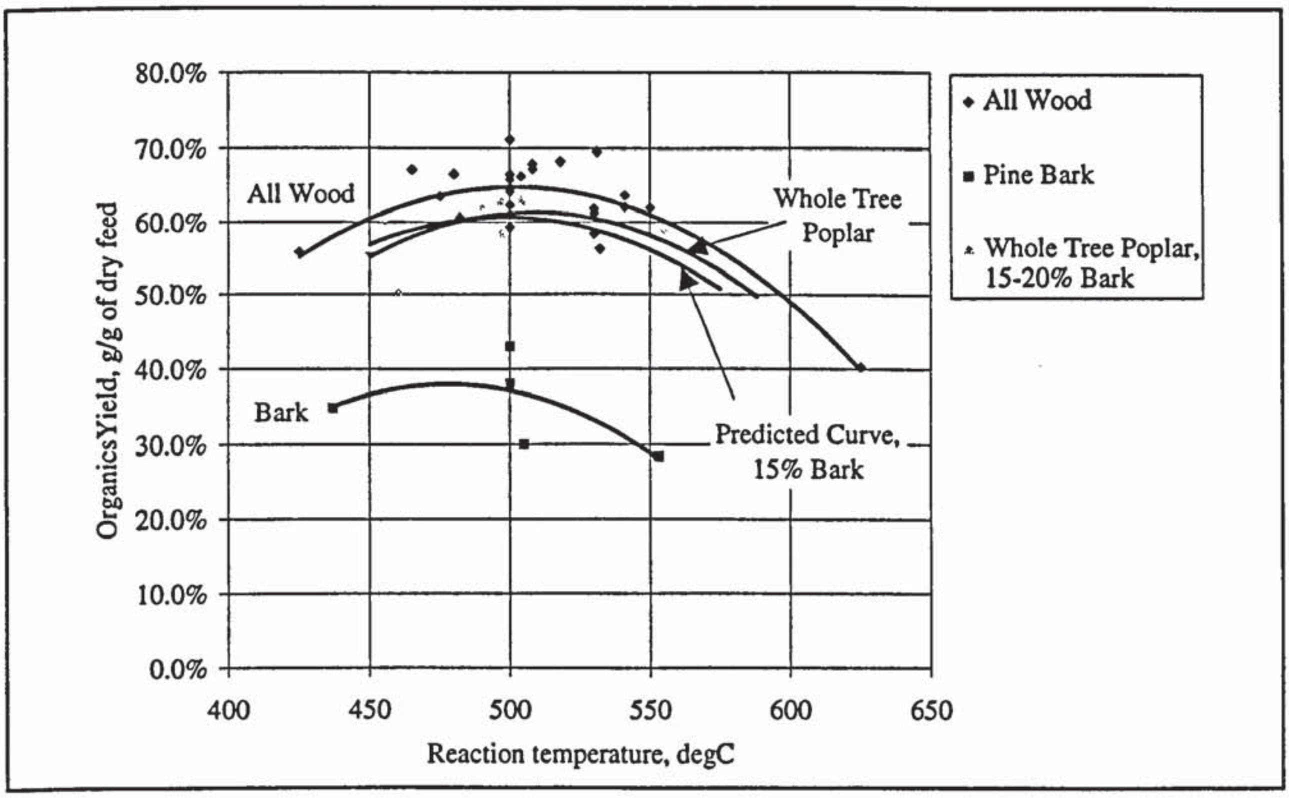


Figure 7.10 - Comparison of Yield Data for Pure Wood and Wood/Bark Mixtures

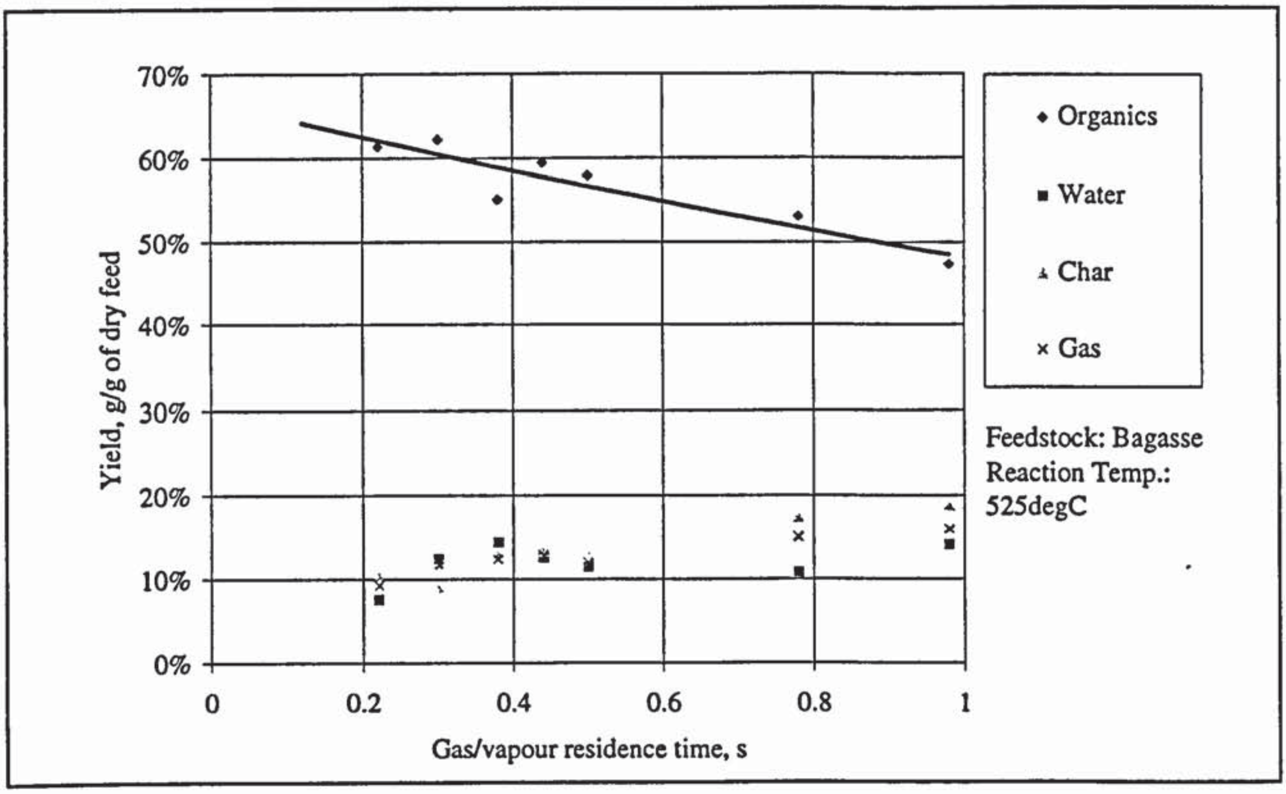


Figure 7.11 - The effect of residence time on organics yield

The liquids recovery and treatment equipment should also be considered. The liquids recovery system has proved particularly problematic for process developers. The Fenosa demonstration plant has not reported yields of above 55% total liquids [160] and this poor yield has been blamed on an inefficient liquids recovery system although higher yields are expected for a re-designed system. The problem is that the pyrolysis vapours exist as an aerosol that is very difficult to capture [143]. Collection by quenching with liquid product is a popular solution but must be carefully designed to avoid clogging due to preferential condensation of the heavier organic compounds in systems that cool the gas/vapour product too slowly. Another option that is investigated is electrostatic precipitation, and this has proved effective at the bench scale [148].

Liquids treatment may be required to reduce the ash and char in the liquid that could cause corrosion, erosion and deposition problems in downstream equipment. While this may be regarded as a fuel quality issue, yields could also be affected. Liquids filtration methods have been unsuccessful except in small batches [152] and hot filtration of the gas/vapour mixture is now under development at NREL [260]. The NREL tests have shown that hot gas filtration can have a severe effect on liquids yields, with initial runs producing yields as low as 38% due to the combined effects of increased residence time, elevated temperatures (to prevent premature condensation of pyrolysis vapours on the filter) and char catalysis.

It can be seen that there are many reasons why it is doubtful that the maximum yields attained in the laboratory will be produced in large scale systems. On the other hand, there is no evidence to show that the design problems noted above cannot be solved. Also, the yields specified in Table 7.3 are based on a regression curve on published data: this has an averaging effect such that the optimum organics yield is not as high as the maximum reported yields. Thus the organic yields predicted by this analysis could go up as well as down. It is highly recommended that the results presented in this work are revisited as more data becomes available, because of the uncertainties involved.

The only adjustment to the original organics yields given in Table 7.3 to account for commercial conditions will be a 5% reduction in yields due to the feedstock, since this is a proven consequence of the bark contamination. This gives an organics yield of 59.9% which is between the 55% yield reported at the Fenosa plant and the typical organics yield of 62% that Ensyn predict for industrial sized facilities [153, 167, 261]. The char yield has been increased by 3% and the gas yield has been increased by 2% to maintain 100% closure. Thus the yields used in the module are as shown in Table 7.5.

Table 7.5 - Fast pyrolysis yields for wood/bark at commercial scales

Product	Yield, g/g dry feed		
	Small scale	**Small scale, normalised**	**Commercial scale, predicted**
Organics	64.9%	64.9%	59.9%
Reaction Water	8.1%	10.8%	10.8%
Char	13.2%	13.2%	16.2%
Gas	11.1%	11.1%	13.1%
Total	94.4%	100.0%	100.0%

7.3.2.2 Pyrolysis liquid heating value

Given the organics yield and the water yield on dry feed in Table 7.5, the lower heating value of the liquid product can be determined and hence the total energy in the liquid product per unit of feed in.

The heating value of many pyrolysis liquids have been found in the literature and are plotted in Figure 7.12 against the moisture content of the pyrolysis liquid [144, 160, 255, 262, 263, 264, 265]. The heating values were generally given on a higher heating value basis, since this is much more easy to determine experimentally. A variety of feedstocks are represented in this chart, since the non-wood feedstocks such as bagasse showed no deviation from the results for wood. Several references gave figures for the heating value of the moisture-free liquid product (at 0% moisture content). This characteristic must be found indirectly since there must always be some moisture present due to the reaction water.

Figure 7.12 shows two regressions for the higher heating value data. The "recorded" line is derived from the heating values that were measured by experiment. The "calculated" line is calculated from the average higher heating value at 0% moisture using Equation 7.14. It can be seen that the lines are in close agreement, confirming the validity of the experimental results. The third line on the chart is a regression on the lower heating values that were calculated from the recorded higher heating values. The lower heating values were found by applying Equation 7.14 to give the higher heating value of the moisture-free liquid, Equation 7.2 to convert the heating value from higher heating value to lower heating value on a dry basis and Equation 4.2 to convert to lower heating value of the wet pyrolysis liquid. The regression of the calculated lower heating values produced Equation 7.15 that is used in the model to calculate the lower heating value of the pyrolysis liquid.

$$HHV_x, GJ/t = HHV_0 \times (1 - x_l) \tag{7.14}$$

$$LHV_x = 21.20 - 25.31(x_l) \tag{7.15}$$

where HHV_x = The higher heating value at x_l, GJ/t

HHV_0 = The higher heating value at x_l=0, GJ/t

LHV_x = Lower heating value at x_l, GJ/t

x_l = Moisture content of the pyrolysis liquid, %wet basis

The overall moisture content of the pyrolysis liquid is required before its lower heating value can be calculated. The moisture content of the feed is given by Equation 7.16.

$$Moisture, x_l\% = \frac{Y_{H20} + \left(\frac{x_f}{1-x_f}\right)}{Y_{H20} + \left(\frac{x_f}{1-x_f}\right) + Y_{org}} \tag{7.16}$$

where x_l = Moisture content of the total pyrolysis liquid, % wet basis

x_f = Moisture content of the feedstock, % wet basis

Y_{H20} = Yield of reaction water from the dry feedstock, %

Y_{org} = Yield of organics from the dry feedstock, %

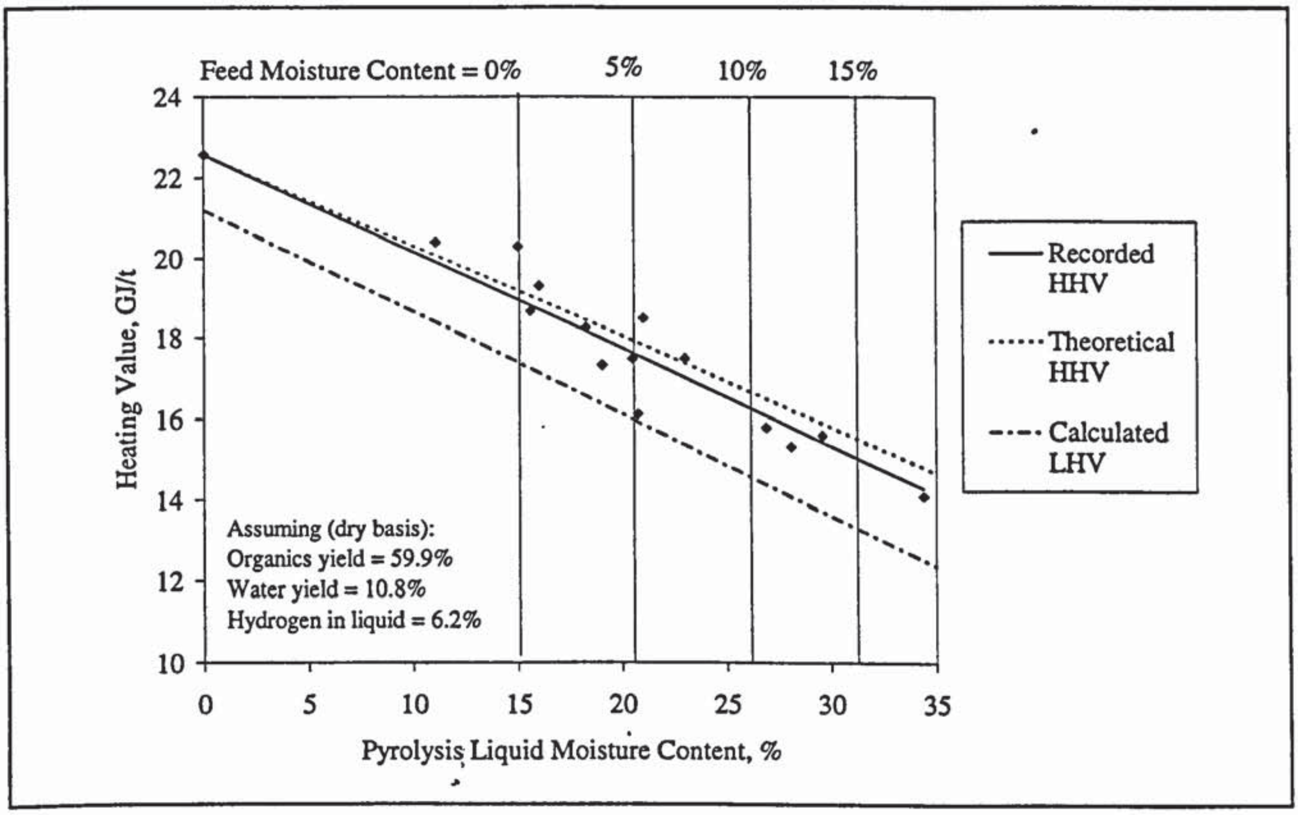


Figure 7.12 - Pyrolysis Liquids Heating Values, Wet Basis

7.3.2.3 Fast pyrolysis conversion efficiency

Once the pyrolysis liquids yield and its heating value is known, the conversion efficiency as defined in Equation 7.9 may be calculated using Equation 7.17.

$$\eta_{conv,pyr} = \frac{Q_{h,conv} \times LHV_{xl}}{E_{th,pret}} \tag{7.17}$$

where $Q_{h,conv}$ = Mass output of intermediate energy carrier (pyrolysis liquid), t/h

LHV_{xl} = Lower heating value of the pyrolysis liquid at its moisture content, GJ/t

$E_{th,pret}$ = Energy input in the prepared feedstock, GJ/h

For the default feed characteristics defined in Section 4.3, a prepared feed moisture content of 10% and the default yields specified in Table 7.5, the energy efficiency of the process is 62.0%. It can immediately be seen that this efficiency is much lower than the conversion efficiency produced in combustion, because of the energy that is lost is the off-gas and char, as well as the energy required for the endothermic pyrolysis reactions. Given the extra energy consumed in grinding the feedstock, the engine generating step will require a substantial efficiency to make up for the losses in the system incurred thus far.

7.3.2.4 Process Energy Requirements

Energy for the process is supplied by combustion of the char and off-gas in an external combustor. While this simplifies the economic analysis, the system could be more cost-effective if the char is sold as a by-product and another energy source is used in the process. Cottam has studied the impact of various scenarios for supplying the reactor energy and concluded that the best option would be to use the off-gas first and make up any shortfall by burning char [150]. In the short term this would appear to be the best solution since there is no established market for the char.

7.3.3 Capital costs

7.3.3.1 Fast Pyrolysis

Bridgwater has compiled plant costs for a number of fast pyrolysis and gasification systems [87, 144]. The costs varied in terms of the extent of the equipment covered and the financial scope of the equipment covered. Each plant cost was analysed by Bridgwater to normalise the costs to the same equipment and financial scope. The equipment coverage was adjusted using ratios given by the figures in Table 7.6, where the atmospheric gasification reactor is assigned a value of 100 and all other costs are relative to this. Capital costs were converted to total plant costs by applying the ratios method presented in Section 4.5.3.

Table 7.6- Relative capital costs for components of gasification and pyrolysis plant

	Atmospheric gasification	Pressurised gasification	Fast pyrolysis
Feed reception, storage and handling	40	40	40
Comminution and screening	20	20	20
Drying	50	50	60
Reactor and feeding system	100	220	100
Wet gas scrubbing and waste water treatment	40		
Hot gas filtration and heat recovery		50	
Liquids recovery			50
TOTAL	250	380	270

This work will use the data that has been manipulated according to Table 7.6. It may be necessary in the light of the current study to readdress the figures in Table 7.6 because of the high grinding costs for fast pyrolysis. These are not reflected in the table and it should be adjusted. The impact of this on the capital cost data for the fast pyrolysis reactor and liquids recovery system is not expected to be severe since the raw data rarely included pretreatment costs.

The data by Bridgwater produces cost relationships that were used by the IEA Pyrolysis Activity [113] and have since been updated by Bridgwater with new information [266]. Some of the data is proprietary and cannot be reproduced but a regression on the data, normalised to a total plant cost for the fast pyrolysis module is shown in Figure 7.13. Equation 7.18 is taken from this chart and is used to calculate the cost of the fast pyrolysis reactor, feeding system and liquids recovery. The cost data used are for systems based on novel technology and it is assumed that the costs are 1st plant costs.

$$TPC_{conv,pyr}, US\$k_{1995} = 50.19 \times \left(Q_{h,pret,dry} \times 1000\right)^{0.6194} \qquad (7.18)$$

where $TPC_{conv,pyr}$ = the total plant cost of the pyrolysis reactor system, $\$k_{1995}$

$Q_{h,pret,dry}$ = the mass flow rate of prepared wood feed into the reactor, odt/h

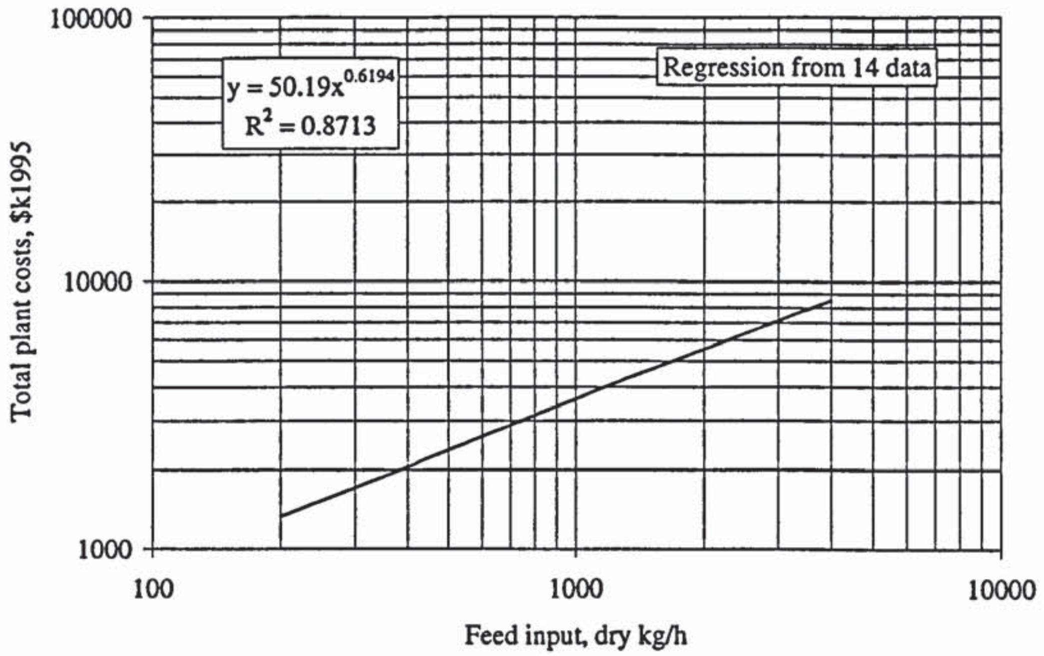


Figure 7.13 - Fast pyrolysis total plant costs

7.3.3.2 Pyrolysis liquids storage

A buffer of pyrolysis liquid product is stored at the fast pyrolysis facility to allow for unplanned shutdowns of the fast pyrolysis system. This will allow the supply of pyrolysis liquid to the engine generators to continue through short interruptions in pyrolysis liquid production and this could give fast pyrolysis an advantage by increasing the reliability of the overall system.

The cost of the pyrolysis liquid storage tanks and transfer pumps have been taken from equipment costs in Garrett [227]. Two tanks and two pumps are used. The tanks are capable of holding one days' production each and the pumps can unload a days' production in 10 hours, either to the engine fuel tanks or to tankers for transport. All equipment is constructed in stainless steel to account for the pH of the pyrolysis liquid although the tanks may in practice be constructed from HDPE to save costs. The equipment costs were converted to total plant costs using the factors method (Section 4.5.4) and are calculated in the model using Equation 7.19.

$$TPC_{Storage} = 119 \times \left(Q_{h,conv}\right)^{0.4045} \tag{7.19}$$

where $TPC_{storage}$ = The total plant cost of the liquids storage and transfer equipment, \$k

$Q_{h,conv}$ = The output flow rate of the pyrolysis liquid conversion product, t/h

7.3.4 Operating costs

7.3.4.1 Labour

Labour requirements are based on the estimates given by Beckman and Graham [194] for the operation of an Ensyn transported bed fast pyrolysis plant. This source is used because it assesses the pyrolysis process only and does not involve pretreatment or an application of the pyrolysis liquid. The labour requirement for plant capacities of 25, 100 and 250 t/d (10% moisture content) are 1, 2 and 3 operators per shift respectively. A regression on this data gives Equation 7.20 that is used to calculate the labour requirement per shift.

$$Labour_{conv,pyr} = 1.04 \times \left(Q_{h,pret,dry}\right)^{0.475} \tag{7.20}$$

where $Q_{h,pret,dry}$ = The input flow rate of prepared feed, odt/h

7.3.4.2 Utilities

Two utilities are required: power for plant items and cooling water for the liquids recovery system. Power requirements have been estimated in work by Beckman [194] and Cottam [150]. The former source suggests a power requirement of 213 kWh/odt; the latter gives 97 kWh/odt. If one assumes the feed characteristics given in Section 4.3 and a 10% moisture content, these figures equate to 4.5% and 2.0% of the thermal input. The first figure is very high, given that the total power consumption for complete combustion and steam cycles is only 4% of the thermal input. Even 2.0% internal consumption seems high: one would expect the power consumption to be lower than in the combustion module since the fan requirements are far lower (there is no oxidant).

More realistic data has been given by Black [267], who has calculated a power consumption of 113-133 kWh/odt for a whole plant including pulverisation, which is a highly energy intensive operation. A second useful figure is given by Diebold [268], who suggests 120 kWh/odt for a large gasoline production system incorporating feed drying, fast pyrolysis, zeolite cracking and refining. One may assume in both cases that the pyrolysis plant consumes a third of the power, resulting in a power consumption of 40 kWh/odt. This equates to nearly 1% of the thermal input to the plant, or half of the power consumption assumed for combustion. This is considered reasonable and a power consumption of 40 kWh/odt is used in the module.

Cooling water usage is a function of the type of liquids recovery system used and can only be estimated very approximately. This uncertainty has little bearing on the model results because the cost of cooling water is very low and the contribution of cooling water to system production costs is minimal. Cooling water usage was estimated by Black [267] to be 11.8 and 10.9 m^3/odt input for two fluid bed fast pyrolysis plant sized at 3.8 odt/h and a 18.9 odt/h. Cottam uses a figure of 9.5 m^3/odt [150]. By contrast, the study by Beckman and Graham suggested a cooling water requirement of 42 m^3/h [194]. Cooling water usage in this module is set at 18.5 m^3/odt of feedstock, being the average of the data presented.

7.3.5 Waste heat available

The availability of waste heat for drying has been studied by Diebold et al. [140] and Cottam [150]. Both studies have concluded that there is ample heat available for drying the feedstock to a 7% moisture content in the flue gases produced when burning the off-gas and char to heat the reactor. In a de-coupled system the only source of waste heat would be the energy in the flue gases. In close-coupled systems the energy in the engine exhaust gases would also be available.

7.4 THE ATMOSPHERIC GASIFICATION MODULE

7.4.1 Feed requirements

Fluid beds are fairly tolerant of variations in feed size [89, 134] and can be expected to operate with a chipped feedstock. For example, the Värnamo IGCC plant uses a wood feedstock with particle sizes between 30-50mm [269]. The Lurgi atmospheric circulating fluid bed process uses feed with particle sizes of under 30 mm [270]. Only the minimum of processing to remove over-size pieces and contaminants such as rocks is required and this is achieved by the screening steps modelled in Chapter 6.

The energy efficiency of gasification and the quality of the fuel gas are improved if the feedstock has a low moisture content. The Värnamo IGCC plant operates with a wood feedstock that has been dried to 10-20% moisture content [269]. The VEGA pressurised system favours a feed moisture content of 15% [271]. Herbert reports that the Lurgi system favours drying to 15% moisture content in rotary dryers [270]. Enviropower suggest gasification feedstocks should be dried to at least 20-30% and are considering drying to 10-15% [109]. This work will assume that the feedstock will be dried to 15% before entering the reactor. Feed drying is included in the pretreatment system modelled in Chapter 6.

7.4.2 Performance

The atmospheric gasification module includes the gasifier feeding mechanism, the gasifier, a catalytic cracker, and the gas cooling and cleaning system. The catalytic cracker is required to crack tars to non-condensible hydrocarbons so that the chemical energy of the tars is not lost during gas cleaning, and is based on the concept developed by TPS and to be installed as part of the UK IGCC THERMIE demonstration concept [144]. The gas is cooled to around 150°C and then scrubbed in a wet gas scrubbed that cools the gas to ambient temperatures and removes the residual tars, particulate, soluble nitrogen compounds and most of the water vapour in the fuel gas.

The performance of a conversion module is measured in terms of its efficiency in turning the energy in the feedstock (on a lower heating value basis) into an intermediate energy carrier for use in a prime mover for electricity generation. In this case the performance of the module is defined in Equation 7.21.

$$\eta_{conv,agas},\% = \frac{\text{Sensible and chemical energy in the cool, cleaned fuel gas}}{\text{Energy in prepared feedstock, lower heating value basis}} \tag{7.21}$$

where $\eta_{conv,agas}$ = the conversion efficiency of the atmospheric gasification module, %

Gasification is a complex process that is influenced by many interdependent variables including the amount of oxidant present; the feedstock composition, morphology and moisture content; reactor temperature and geometry. Ideally, this module would predict the gas characteristics of the cool, clean fuel gas for a given set of process conditions and from that information the sensible and chemical energy of the fuel gas could be calculated. Many models have been developed to aid developers, ranging from basic use of the equations in Table 7.7 [274] through to more complex mechanistic models that include the interactions between the fuel and the operating conditions (e.g. Chern [84], Belleville [272], Shand [273], Maniatis [274]). These models require detailed specification of the operating conditions, details that are rarely available in the literature and that would be beyond the scope of this work to define for the full range of capacities required here. Therefore in this work the energy efficiency of the gasifier has been simplified to a basic mass and energy balance across the gasifier, consistent with the approach used in the combustion module. This considers the gasification reactor and tar cracker as a sub-system with the inputs and outputs shown in Figure 7.14. In practice the details of process conditions in commercial gasifiers are not available in the literature and it would be beyond the scope of this work to attempt to develop a generic set of operating characteristics for the range of capacities to be examined.

Table 7.7 - Principle reactions in gasification

Reaction type	Reaction	Heat of reaction, kJ/mole at 20°C
Heterogeneous	$C + 1/2\,O_2 \rightarrow CO$	+110.6
	$C + O_2 \rightarrow CO_2$	+393.8
	$C + CO_2 \rightarrow 2CO$	-172.6
	$C + H_2O \rightarrow CO + H_2$	-131.4
	$C + H_2O \rightarrow CH_4$	+74.9
Homogeneous	$CO + H_2O \leftrightarrow CO_2 + H_2$	+41.2
	$CH_4 + H_2O \leftrightarrow CO + 3H_2$	+201.9
	$2H_2 + 2CO \leftrightarrow CO_2 + CH_4$	-321.3

The chemical and sensible energy in the fuel gas was found by difference after determining all the other energy fluxes, in a procedure similar to the one adopted in Section 7.2.2. This calculation was carried out in a spreadsheet that is included in Appendix D. The first step in the mass and energy balance is the calculation of the mass inputs. This requires the calculation of the stoichiometric air requirement and the definition of an air factor, which is the ratio of the actual oxidant flow rate to the stoichiometric oxidant flow rate. This was assumed to be 0.3, based on the data produced by Maniatis in his study of the effects of the air factor on gasifier performance [275, 276].

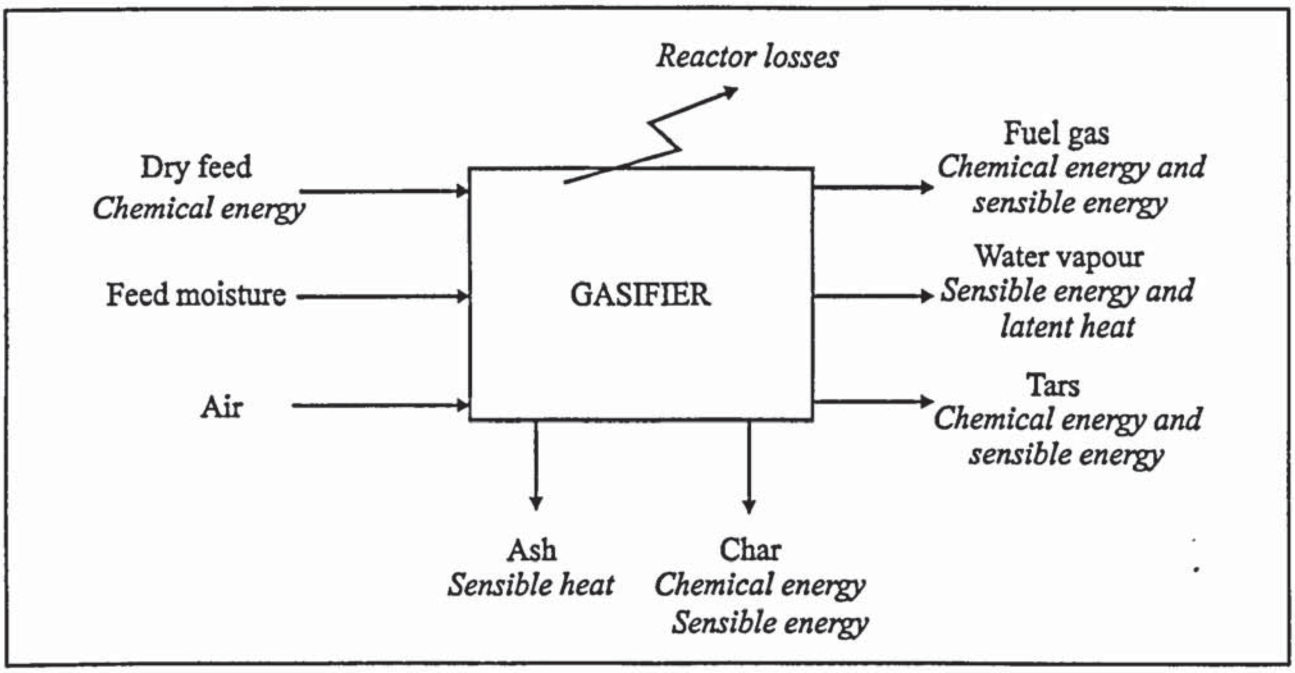


Figure 7.14 - The atmospheric gasifier mass and energy flows

The air factor is one of the most important process parameters and it is a good example of the difficulties that would arise in trying to define the details of the gasifier performance. High air factors increase partial combustion of the char and pyrolysis vapours, resulting in a lower calorific value fuel gas, a higher reactor temperature and an increased volume of fuel gas. The sensible energy of the product gas is increased at the expense of its chemical energy. In an application where the gas can be used hot then a high air factor could be an advantage since tars are cracked at the higher temperatures. In cold gas applications, the designer faces a compromise since a high air factor helps to eliminate tars but increases the sensible energy losses and hence reduces the conversion efficiency. An air factor of 0.3 is close to the value required to give the highest cold gas efficiency found by Maniatis, and any tars remaining in the fuel gas at exit from the gasifier will be cracked in the tar cracker.

Once the amount of air is known, the other mass fluxes can be calculated easily. It is assumed that the ash produced contains 33 %wt of char, which is equivalent to a carbon conversion efficiency of 99.5% [86]. The dolomite fluid bed catalytic tar cracker ensures that the virtually all tars are cracked and the mass balance assumes that only 0.1 %wt of the feedstock emerges as tar [134, 144].

The energy balance is based on the higher heating value of the feedstock, since this represents the total chemical energy in the feedstock. The higher heating value of the feedstock is calculated from its lower heating value using Equation 7.2. The output values used in the energy balance are based on an output temperature from the catalytic cracker of 900°C, selected to match the conditions specified by TPS [144]. The physical and chemical properties for the various species that are shown in Appendix D are standard values from the literature [277, 278]. The energy balance over the gasifier and tar cracker for operation at 1 odt/h is shown in Table 7.8. The reactor losses are assumed to vary logarithmically between 7% at 1 odt/h and 2% at 50 odt/h [86, 144]. The energy in the dry fuel gas is found by difference after all the other energy outputs are determined. This energy is the total chemical energy and sensible energy in the gas at exit from the catalytic cracker at 900°C.

Table 7.8 - Energy balance for the atmospheric gasifier and tar cracker at 1 odt/h

		In	%	Out	%
Feed in (HHV)	GJ/h	20.56	98.21%		
Feed moisture	GJ/h	0.00	0.00%		
Ash	GJ/h	0.00	0.00%		
Oxidant	GJ/h	0.37	1.79%		
Dry fuel gas	GJ/h			17.83	85.15%
Water vapour	GJ/h			1.44	6.87%
Tars	GJ/h			0.02	0.11%
Char	GJ/h			0.17	0.83%
Ash	GJ/h			0.01	0.04%
Losses	GJ/h			1.47	7.00%
Total	GJ/h	20.93	100.00%	20.93	100.00%

Given the energy output from the gasifier and tar cracker, the next stage is to calculate the sensible and latent heat losses that are incurred during gas cooling and cleaning. The gas must be cooled to ambient temperatures to maximise the volumetric efficiency of the engine [279]. Tars must be less than 10-50 mg/Nm^3 [129, 280], which is achieved in the tar cracker. Particulate limits vary between 0.6 mg/Nm^3 [281] to 100 mg/Nm^3 [279]. To achieve these conditions a cyclone, gas cooler and wet gas scrubber are used. The gas cooler will cool the gas to around 150°C. This is high grade heat that could be used elsewhere in the system. Some of this heat is assumed to be used to preheat gasifier air, the rest could be used in feed drying. The wet scrubber will cool the gas further as well as condensing out most of the water in the gas. It is assumed that the gas leaves the wet gas scrubber saturated with water vapour and is reheated slightly to 40°C to produce a gas with a maximum humidity of 80% at entry to the engine [282]. The calculation of the sensible heat losses in the gas requires the specific heat of the gas and an average of published data for fluid bed and circulating fluid bed gas compositions is used. This data is given in Table 7.9. The specific heat of the gas is assumed to be 1.45 kJ/kgK. The residual energy in the fuel gas is its higher heating value. This is converted to the lower heating value by applying the mean of the ratios between LHV and HHV shown in Table 7.9.

Table 7.9 - Fuel gas characteristics for fluid bed and circulating fluid bed gasifiers

Source	Bridgwater [86]	Bridgwater [86]	Maniatis [276]	Maniatis [276]
Dry gas composition, %vol				
H2	18.0%	9.0%	9.9%	8.0%
CO	20.0%	14.0%	16.2%	16.0%
CO2	14.0%	20.0%	16.8%	15.0%
CH4	3.0%	7.0%	5.3%	6.0%
C2H4			2.0%	2.0%
C2H6	1.0%		0.3%	
C3H6			0.1%	
O2			1.4%	
N2	44.0%	50.0%	48.1%	53.0%
	100.0%	100.0%	100.0%	100.0%
LHV, MJ/kg	5.47	4.18	5.26	5.03
HHV, MJ/kg	5.94	4.54	5.67	5.41
Cp, kJ/kgK	1.51	1.44	1.46	1.45
LHV/HHV	0.92	0.92	0.93	0.93

This procedure was carried out for a range of gasifier input capacities using the data sheet given in Appendix D and the results are shown in Figure 7.15. Cold gas efficiencies such as this are rarely published on a lower heating value basis, but since the ratio of LHV to HHV is practically the same in the feed and the product, the efficiency should be almost the same as those given on an LHV basis. Thus the efficiencies that are predicted compare well with data in the literature. The gasifier survey by Bridgwater [144] quotes efficiencies of 67.8% (excluding 10% energy in tars) for the TPS system at Greve in Chianti and 79% for the Ahlstrom atmospheric cfb gasifier. Chern [84] has calculated a cold gas efficiency for a fluid bed of 72% from his modelling work. Maniatis [276] estimates that a cold gas efficiency of 60-70% would be attainable in a commercial gasifier, but this assumes that the chemical energy in tars is lost.

The efficiency of the atmospheric gasification step is therefore calculated using Equation 7.22.

$$\eta_{conv,agas} = \left[-0.00006\left(Q_{h,pret,dry}\right)^2 + 0.0347\left(Q_{h,pret,dry}\right) + 71.10\right] / 100 \qquad (7.22)$$

where $\eta_{conv,agas}$ = the conversion efficiency of the atmospheric gasification module, %

$Q_{h,dry}$ = the dry flow rate of feed into the reactor, odt/h

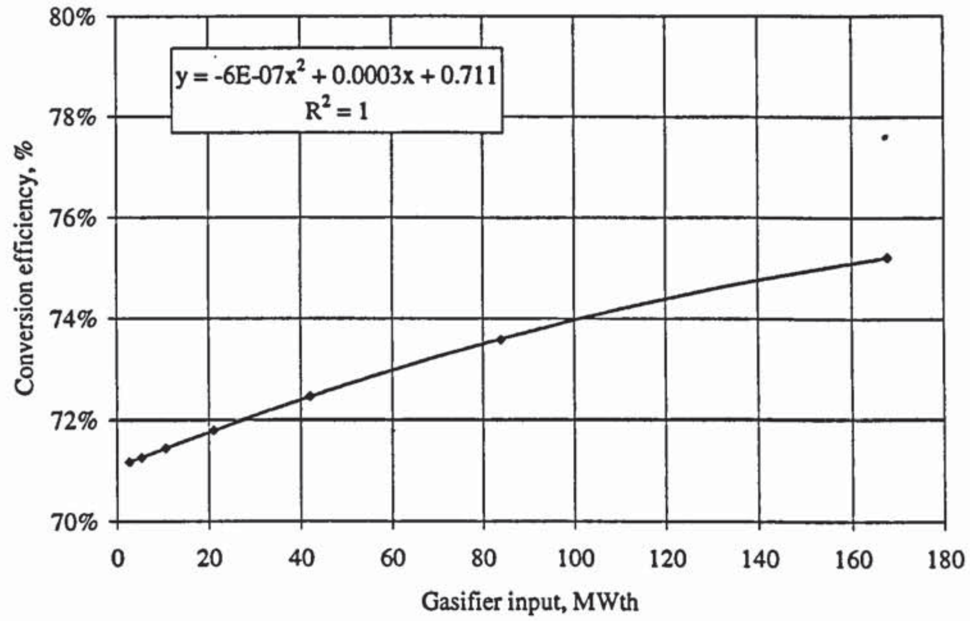


Figure 7.15 - Calculated atmospheric gasifier efficiencies

7.4.3 Capital Costs

Capital costs are taken from the same source used for the fast pyrolysis module (see Section 7.4.3). This data produces the curve shown in Figure 7.16, which is assumed to show 1st plant costs. The curve only gives the costs for the gasifier feeding mechanism, the gasifier and the gas cleaning system. It does not include the cost of the tar cracker and so the costs must be adjusted. The relative costs of the gasifier and the gas cleaning system can be seen in the breakdown of system costs given in Table 7.6. It can be assumed that the catalytic cracker costs the same as the gasifier on the basis that both are fluid beds. However the gasifier cost shown in Table 7.6 includes the biomass feeder, the air blower and many other items that would be only be required once in a two reactor system. Thus it is assumed that the tar cracker adds another 50% to the gasifier cost, or (50/140)% of the module costs. This means that the cost of a system incorporating a tar cracker is (190/140)% of the original cost and the total plant costs for the module are calculated using Equation 7.23.

$$TPC_{conv,agas} = \left[37.91 \times \left(Q_{h,pret,dry} \times 1000\right)^{0.6983}\right] \times \frac{190}{140} \qquad (7.23)$$

where $TPC_{conv,agas}$ = the total plant cost of the atmospheric gasification module, $k_{1995}

$Q_{h,pret,dry}$ = the mass flow rate of prepared wood feed into the reactor, odt/h

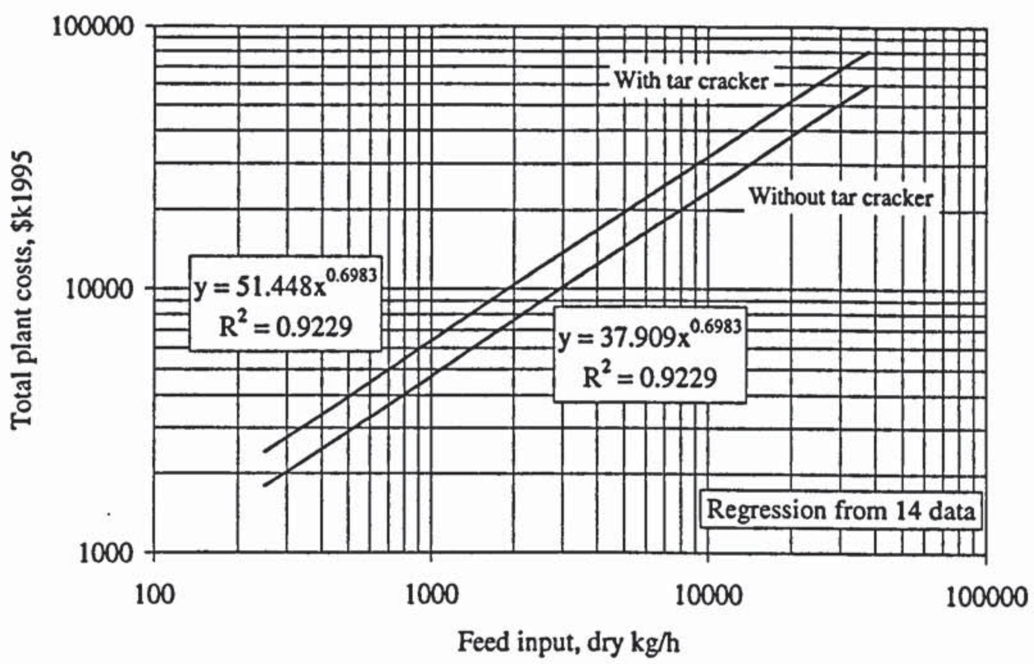


Figure 7.16 - Atmospheric gasification total plant costs

7.4.4 Operating costs

7.4.4.1 Labour

A relationship of the labour requirement for the atmospheric gasification system is derived by comparison with the relationship used for the fast pyrolysis module. The equipment required in both systems is very similar. The atmospheric gasification system is complicated by the addition of the tar cracker, but is simplified by the fact that its heat requirement is generated internally, so precluding the external char combustor used in the fast pyrolysis system. If it is assumed that the extra labour required by the cracker and the labour saved by avoiding an external combustor cancel each other then the labour required for the atmospheric module is the same as for the fast pyrolysis module, and may be calculated using Equation 7.24.

$$\mathrm{Labour}_{\mathrm{conv,agas}} = 1.04 \times \left(Q_{\mathrm{h,pret,dry}}\right)^{0.475} \tag{7.24}$$

where $Q_{h,pret,dry}$ = The input flow rate of prepared feed, odt/h

7.4.4.2 Utilities

The internal power consumption requirements of an atmospheric gasifier have been calculated in the work by Solantausta [113] and McIlveen-Wright [186], but their figures vary too much to give a conclusive figure for use here. Instead the internal power requirement is found by comparison with the fast pyrolysis case. The gasification system must supply more fluidising gas than fast pyrolysis, and has the pressure drop in two reactors to overcome. Conversely, the fast pyrolysis module must meet the extra power requirement of the external combustor, and the two requirements will tend to cancel out. Therefore it will be assumed that the atmospheric gasifier requires the same amount of power as the fast pyrolysis module, and a value of 40 kWh/odt is used.

The cooling water requirement is unlikely to be significant since the make-up water can be taken from the condensate in the scrubber. The treatment of wet scrubber effluent is highly uncertain. One possibility is that most of the scrubber water is recirculated with a bleed to maintain a maximum concentration of tars in the water. The bleed water would be disposed of by incineration, at a significant energy cost [82]. For example, if tars are equivalent to 0.1% of the feed after the cracker and the maximum concentration on the scrubbing water is 2% then each oven dry tonne of feed would produce 50 kg of waste water, and this water would require 0.12 GJ of energy for incineration. Given that the tars yield could be lower and the tars concentration in the scrubbing water could be higher, the uncertainty surrounding this option is very high and effluent treatment has not been included in this work.

A final utility requirement is catalyst for the cracker. Dolomite consumption is calculated at 0.68 t/odt feed input and a cost of 33 $/t [188].

7.4.5 Waste heat availability

The data sheet presented in Appendix D shows that around 4 GJ of high quality heat are available during gas cooling for every dry tonne of feedstock. Some of this energy would be used to preheat combustion air but the remainder is available for drying the feedstock. If the delivered feedstock has a moisture content of 50% and the required moisture content is 15% then the evaporation load is 0.83 twe/odt, requiring 2.94 GJ/odt. Given that there is also considerable heat available in the engine exhaust during generation (see Section 8.4.5), it is clear that there will be sufficient heat available for feed drying. The excess heat could be used for cogeneration, but combined heat and power production is not considered in this work.

7.5 THE PRESSURISED GASIFICATION MODULE

7.5.1 Introduction

The relationships used to model the pressurised gasification module are developed using the same approach that is applied in Section 7.4. In this module the fluid bed or circulating fluid bed gasifier is pressurised. The fuel gas from the gasifier is burned in a gas turbine and does not need to be cooled to ambient temperatures. This means that the gas cleaning system can be designed to maintain temperatures above the tars dew point and so any tars can be burned in the gas turbine with the fuel gas [89]. Gas cleaning in this module is achieved by partial cooling to around 500°C to condense alkali metal vapours onto particulate in the fuel gas, followed by hot gas filtration to remove both particulate matter and alkali metals [117].

7.5.2 Feed constraints

These are exactly the same as for the atmospheric case (see Section 7.4.1).

7.5.3 Performance

The procedure used to define the pressurised gasifier performance is the same as that used in Section 7.4.2, and the data sheet used is presented in Appendix D. The following changes are made:

1. There is no tar cracker and the tar yield is 1% of the dry feed input.
2. The output conditions at the gasifier are 900°C and 25 bar, based on the operating conditions of the Värnamo plant [283].

3. The fuel gas enters the gas turbine at 450°C after gas cooling and hot gas filtration. At these conditions the water vapour in the raw fuel gas is still a vapour and the enthalpy of the water vapour is included in the energy efficiency of the conversion process.

The data sheet was used to produce Figure 7.17, which gives Equation 7.25.

$$\eta_{conv,pgas} = \left[-0.00006\left(Q_{h,pret,dry}\right)^2 + 0.0347\left(Q_{h,pret,dry}\right) + 85.50\right] / 100 \qquad (7.25)$$

where $\eta_{conv,pgas}$ = the conversion efficiency of the pressurised gasification module, %

$Q_{h,pret,dry}$ = the dry flow rate of feed into the reactor, odt/h

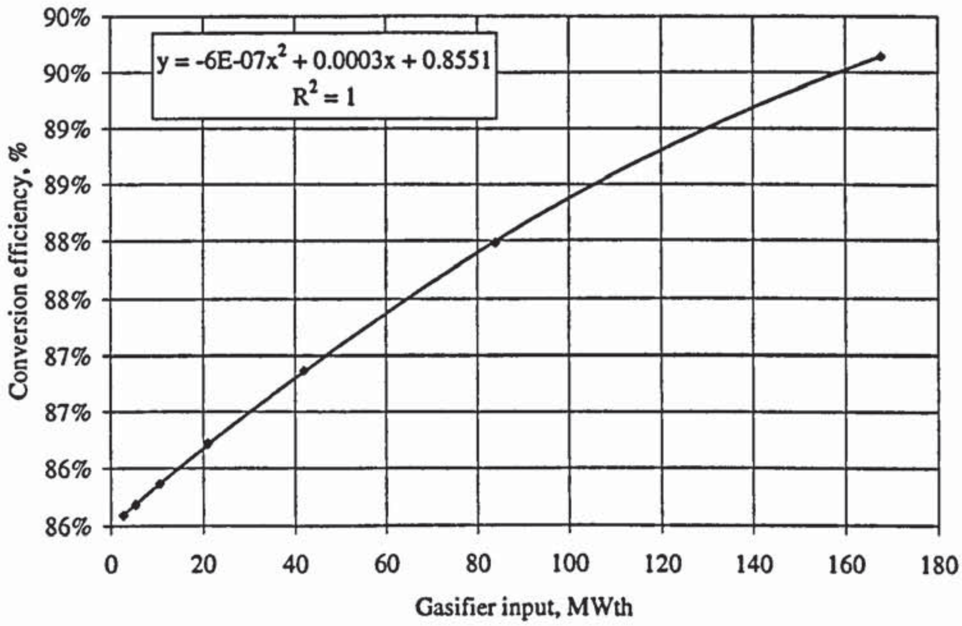


Figure 7.17 - Calculated pressurised gasifier efficiencies

7.5.4 Capital Costs

Total plant costs are based on the data already described in Section 7.3.3.1. This source produced the curve shown in Figure 7.18 and Equation 7.26. Given that this regression is based on only 4 data points, there is some uncertainty about the validity of the capital cost relationship. However, the validity of the relationship is proved in part by the fact that the costs produced are approximately twice the costs for an atmospheric gasifier. Bridgwater has compared this ratio with other atmospheric and pressurised equipment costs given in Garrett and shown that the pressurisation tends to double costs [113]. The four data points used are for novel, 1st plant and the costs for this module are assumed to be 1st plant costs.

$$TPC_{conv,pgas} = 116.18 \times \left(Q_{h,pret,dry} \times 1000\right)^{0.6384} \tag{7.26}$$

where $TPC_{conv,pgas}$ = the total plant cost of the atmospheric gasification module, $\$k_{1995}$

$Q_{h,pret,dry}$ = the mass flow rate of prepared wood feed into the reactor, odt/h

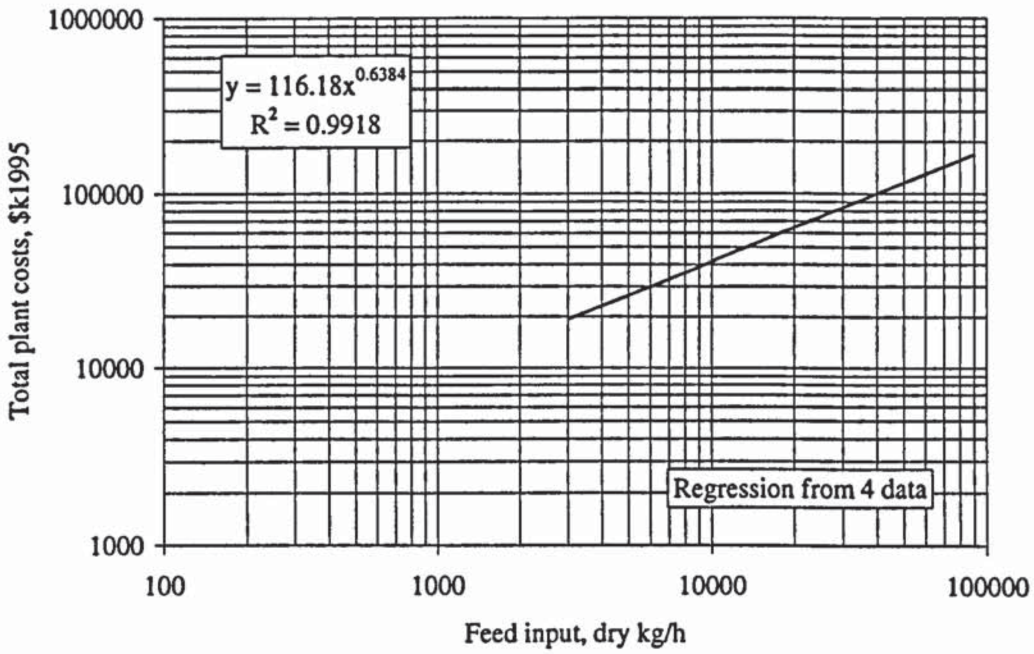


Figure 7.18 - Pressurised gasification total plant costs

7.5.5 Operating costs

7.5.5.1 Labour

The labour requirement for the pressurised gasification module is assumed to be the same as the labour requirement for the atmospheric gasification module, since the equipment required in both are similar. The hot gas filtration used in the pressurised system is likely to require more attention than wet gas scrubbing given its novelty, compensating for the lack of a tar cracker. Thus the labour requirement for the pressurised gasification module is as given by Equation 7.27.

$$\text{Labour}_{\text{conv,pgas}} = 1.04 \times \left(Q_{\text{h,pret,dry}}\right)^{0.475} \tag{7.27}$$

where $Q_{h,pret,dry}$ = The input flow rate of prepared feed, odt/h

7.5.5.2 Utilities

The internal power consumption has been estimated by comparison with the power consumed in the other modules. The main consumer of power is the booster compressor that raises bleed air from the gas turbine compressor to a pressure sufficient to overcome the pressure drop in the fluid bed, the gas filtration system and fuel gas injection to the gas turbine. It is difficult to estimate this requirement since the gas turbine pressure will change with each machine, but it is likely to be significantly more than the fan power requirement for the atmospheric gasifier. The feeding system could also consume substantial power if a lock hopper system is used by virtue of the need to compress the inert gas purge. This power consumption and the expense of the purge (probably nitrogen) has discredited lock hopper systems and it is assumed that a screw feeder is used instead. The gas cleaning systems in both the gasification modules will require power, the former for pumping scrubbing water and the latter in terms of the pressure drop (already noted) and the flushing mechanism required to periodically clean the filters. In summary it will be assumed that the compressor work increases the amount of power required in this module by 50%, and a value of 60 kWh/odt is used. This assumption will be tested by sensitivity analysis because of the uncertainty surrounding this issue.

There are no other utility requirements for this module.

7.5.6 Waste heat availability

The high efficiency of the pressurised gasification system is a consequence of maintaining a high fuel gas temperature through gas cleaning. If the sensible heat of the fuel gas is largely retained, then there is less energy available for use elsewhere in the system. What energy there is available from cooling the gas from 900°C to 500°C is in the form of high grade heat that would be used in raising superheated steam (see Section 8.5) rather than for feed drying. Therefore it can be assumed that no heat would be available from this module for feed drying.

7.6 SUMMARY

The chapter describes the development of four modules that model the cost and performance of four generic conversion technologies. The conversion technologies are:

1. Combustion;
2. Fast pyrolysis;
3. Atmospheric gasification; and
4. Pressurised gasification.

In each case the conversion efficiency of the module, in terms of the amount of energy in the prepared feed that is available as an input to the prime mover in the electricity generating cycle (in the next chapter) has been calculated using a basic analysis of the mass and energy flows through the equipment that is modelled. Capital costs for the combustion module are based on cost data for existing, established technology and is considered 100th plant cost. Capital costs for the other three modules are based on capital cost data for demonstration plant or costs predicted by process developers. These are considered 1st plant costs. Production costs have been taken from the literature where possible and where data is not available production costs are estimated by comparison between the technologies.

Every effort has been made to eliminate the inaccuracies that can arise when developing generic models. However, some compromises have had to be made and key uncertainties will be tested by sensitivity analyses in Chapter 9 to assess their impact. Uncertainties that should be addressed in further work are listed below.

1. The availability of waste heat from the combustion flue gases for feedstock drying should be examined, given the apparent deficit in the current system.
2. The impact on fast pyrolysis yields of changing operating conditions during scale-up must be examined as data becomes available. Particular concerns are the gas/vapour residence time, the efficiency of liquids recovery and the impact of hot gas filtration.
3. The interaction between fast pyrolysis feed moisture content and product quality should be evaluated since the cost of drying and increases in product viscosity may be outweighed by the improvements in product heating value.

4. Effluent quantities, disposal methods and waste water treatment costs from wet scrubbing are very unclear and their impact on the system could be significant. This area should be examined in more detail.

5. The total plant costs for the gasification and fast pyrolysis modules must be examined. The relative costs of the systems should be reassessed given the extra pretreatment required for the fast pyrolysis system. Costs for the tar cracker in the atmospheric gasification module are currently only estimated because of the lack of data. The pressurised gasification costs are based on only 4 data points and should be modified as required when more data is available.

6. All power consumption relationships are uncertain because more definitive relationships would require a close examination of the equipment required in each system and over several representative capacities. Internal power consumption will be examined by sensitivity analysis and if it is a significant factor in the overall production cost then a more thorough analysis of power requirements will be needed.

7.7 NOMENCLATURE

All costs are given in US$$_{1995}$ unless otherwise indicated.

E_{flue} = The energy in the flue gas from combustion, MW_{th}

$E_{th,pret}$ = The energy available in the prepared feedstock, MW_{th} LHV basis

H = Percentage of hydrogen in dry feed, %odt

h_{in}, h_{out} = Enthalpy of water vapour at entry and exit of the feed dryer, kJ/kgK

HHV = Higher heating value, GJ/t. Subscript indicates the moisture content of the feed or product

LHV = Lower heating value, GJ/t. Subscript indicates the moisture content of the feed or product

Q_{H2O} = The quantity of feed moisture to be removed, twe/h

$Q_{h,conv}$ = The mass of pyrolysis liquid produced, t/h

$Q_{h,pret,dry}$ = The dry mass of prepared feed supplied to the conversion reactor, odt/h

T_{in}, T_{out} = Temperature of hot gases at entry and exit of the feed dryer, °C

$TPC_{conv,agas}$ = The total plant cost of the atmospheric gasification module, $\$k_{1995}$

$TPC_{conv,comb}$ = The total plant cost of the combustion module, $\$k_{1995}$

$TPC_{conv,pgas}$ = The total plant cost of the pressurised gasification module, $\$k_{1995}$

$TPC_{conv,pyr}$ = The total plant cost of the fast pyrolysis module, $\$k_{1995}$

x_f, x_l = Feed moisture content (x_f) or pyrolysis liquid moisture content (x_l). Both %wet basis.

x_{in}, x_{out} = Moisture content of feed at entry and exit of the feed dryer, %wet basis

twe = tonnes of water evaporated

Y_{H20} = Yield of reaction water from fast pyrolysis, %odt of feed

Y_{org} = Yield of condensible organics from fast pyrolysis, %odt of feed

$\eta_{conv,agas}$ = the conversion efficiency of the atmospheric gasification module. %LHV in prepared feed

$\eta_{conv,comb}$ = the conversion efficiency of the combustion module. %LHV in prepared feed

$\eta_{conv,pgas}$ = the conversion efficiency of the pressurised gasification module. %LHV in prepared feed

$\eta_{conv,pyr}$ = the conversion efficiency of the fast pyrolysis module. %LHV in prepared feed

8. THE ELECTRICITY GENERATION MODULES

8.1 INTRODUCTION

This chapter describes four modules (sub-models) that calculate the cost and performance of generating electricity from an intermediate energy carrier (steam, pyrolysis liquid or fuel gas). A fifth module is described that calculates the net electricity output to the grid and the costs of grid connection. Four generating cycles are modelled, as summarised in Table 8.1. These generating cycles were introduced in Chapter 2, where the basic characteristics and status of each cycle in biomass systems was given.

Table 8.1 - The electricity generation modules

	Steam Cycle	Liquid Dual-Fuel Engine	Gas Dual-Fuel Engine	Gas Turbine Combined Cycle
See	Section 8.2	Section 8.3	Section 8.4	Section 8.5
Previous module	Combustion, Section 7.2	Fast Pyrolysis, Section 7.3 or Liquid transport, Chapter 5	Atmospheric gasification, Section 7.4	Pressurised gasification, Section 7.5
Energy carrier	Superheated steam	Pyrolysis liquid	Cool, clean fuel gas	Hot, clean fuel gas
Output stream	Gross electricity output			
Next module	Grid connection, Section 8.6			

The methodology for calculating the cost and performance criteria has been described in Chapter 4. This chapter applies this methodology to produce modules that calculate the cost and performance of each generating cycle. Each module description includes the development of relationships for the calculation of the generating efficiency, the total plant costs of the module and the operating costs that are incurred. There is also a discussion in each module about the availability of waste heat for drying the feedstock during pretreatment.

The chapter concludes by describing the grid connection module. The main purpose of this module is to calculate the total internal power consumption for the system and so produce a net electricity output.

8.2 STEAM CYCLE MODULE

8.2.1 Fuel requirements

The intermediate energy carrier used by the steam cycle module is the steam that is raised in the boiler. Combustion and the generation of steam is modelled in the combustion module described in Chapter 7. Only the energy supplied to raise steam is defined in the combustion module. The steam conditions are not specified. These are expected to vary with generating capacity and steam conditions are assumed to be typical of the current state of the art for each capacity.

8.2.2 Generating efficiency

The steam cycle module assumes that a basic Rankine steam cycle with superheat is used. Enhancements of the Rankine steam cycle such as regenerative feed water heating or steam reheat are not usually applied at this scale (1-100 MW_e) because the additional capital cost and cycle complexity do not justify the modest efficiency increases that can be obtained [41]. Full expansion of steam is assumed since power generation only is required and condensing steam turbines offer the maximum power output and generating efficiency. The cycle efficiency will be calculated in two stages. Firstly an ideal cycle efficiency will be calculated by analysis of the Rankine cycle using typical steam conditions for plant of this capacity range. This ideal cycle efficiency will then be modified to account for losses in the steam turbine to give an actual cycle efficiency.

The ideal cycle efficiency is a function of the steam conditions at entry to the steam turbine. Large scale coal-fired steam turbine plant (typically greater than 500 MW_e) can operate at steam inlet conditions of 124-248 bar, 510-566°C, and a condenser pressure of 0.03-0.12 bar [284]. Such high pressures and temperatures are highly demanding of materials, increasing equipment costs to levels only viable at the very large scale where scale economies and high efficiencies are possible. Biomass-based combustors are limited to more modest steam conditions. The size of the condenser and the cooling water it requires can also become critical at very low vacuums, and so a relatively high condenser pressure is used. In this analysis, a fixed condenser pressure of 0.1 bar is used for all cycles.

The ideal Rankine cycle efficiency has been calculated for steam conditions typical in biomass plant, using steam temperatures and pressures given in the literature [71, 74, 229, 240, 285]. The calculations required are standard and can be found in most thermodynamics textbooks. An example of the spreadsheet used is given in Appendix E and calculated ideal cycle efficiencies are shown in Table 8.2.

The losses in the steam turbine (its isentropic efficiency) are a function of the steam pressure and the generating capacity [73]. No matter how well designed the steam turbine components are, there will always be machining tolerances that allow a small amount of leakage between stages and prevent the steam from following its optimum path through the turbine. The minimum tolerances are constant, with the effect that the amount of leakage in a small turbine is higher than in a large turbine. The amount of leakage will increase at higher pressures. Appropriate steam turbine efficiencies are given in Table 8.2, based on data published by Guinn [73]. The product of ideal cycle efficiency and steam turbine efficiency gives the actual cycle efficiency before generator losses, also shown in Table 8.2.

Table 8.2 - Calculated steam cycle efficiencies

Steam conditions in the literature					Calculated efficiencies, %			Gross Output
kg/s	bar	°C	MWth	Source	Ideal	Turbine	Actual	MWe
4	32	385	12	[285]	34.4%	69%	23.8%	2.8
8	21	260	21	[285]	28.8%	72%	20.8%	4.3
8	19	316	22	[285]	30.0%	73%	21.7%	4.7
8	30	441	24	[41]	34.7%	72%	25.0%	5.9
8	45	399	23	[285]	35.8%	71%	25.4%	5.8
10	42	399	30	[74]	35.5%	70%	24.8%	7.2
15	45	399	45	[285]	35.8%	77%	27.5%	12.1
29	45	399	90	[285]	35.8%	84%	30.0%	26.7
29	45	371	90	[285]	35.4%	84%	29.7%	25.8
32	93	513	106	[71]	40.1%	83%	33.3%	34.7
61	88	510	201	[240]	39.9%	89%	35.5%	70.3
71	110	512	232	[229]	42%	89%	37.8%	87.5
76	83	485	245	[285]	39.3%	90%	35.4%	85.2

The data given in Table 8.2 is presented as a graph of cycle efficiency against thermal input in the superheated steam in Figure 8.1. A best fit regression curve to the data is used to give the relationship between steam cycle efficiency and the energy supplied to the steam at the boiler resulting in Equation 8.1.

$$\eta_{gen,steam} = 0.0516\ln\left(E_{th,conv}\right) + 0.0794 \tag{8.1}$$

where $\eta_{gen,steam}$ = gross generating efficiency before internal consumption, %

$E_{th,conv}$ = the energy supplied by the boiler, MW_{th}

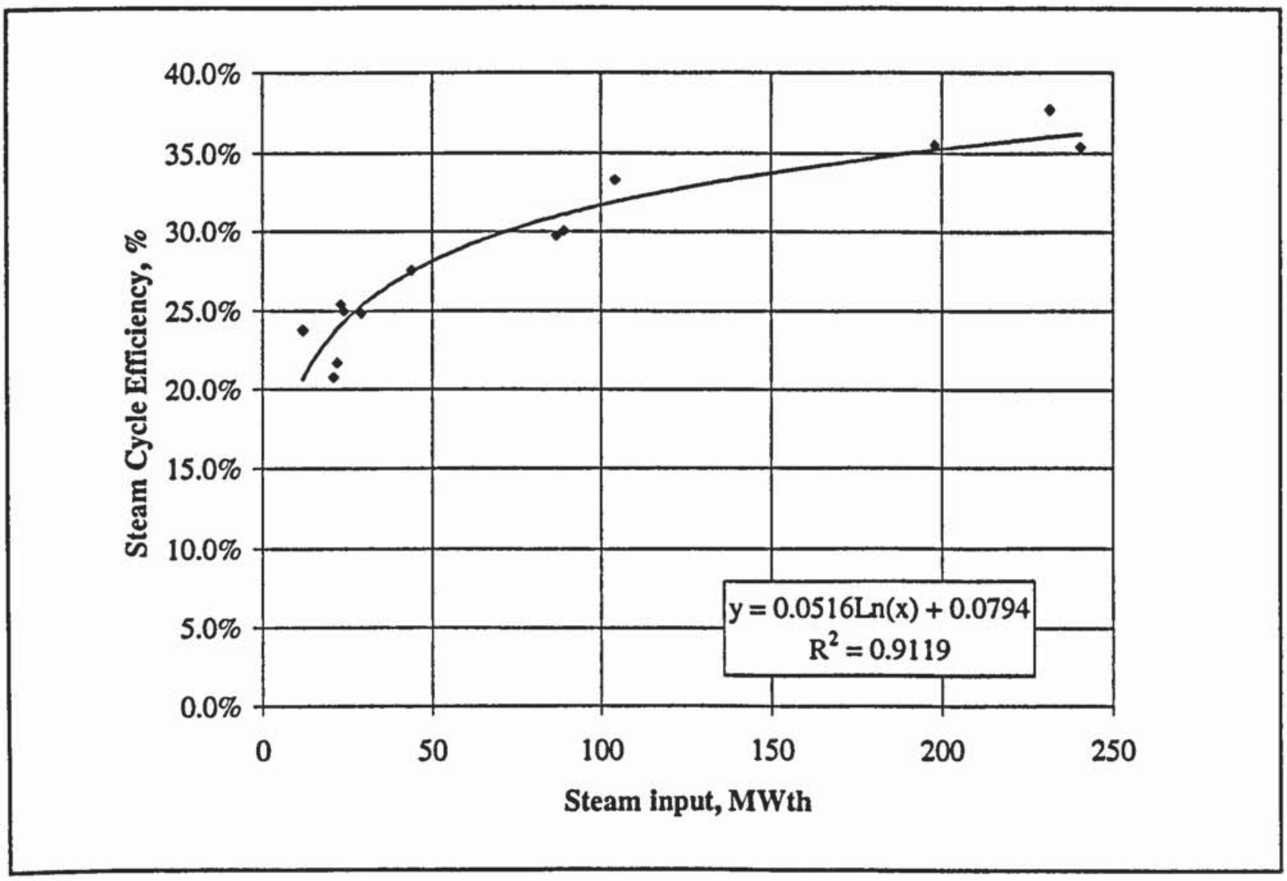


Figure 8.1 - Steam cycle efficiency as a function of boiler energy input

8.2.3 Capital costs

Steam cycle capital costs have been extracted from total combustion plant cost data where the data has split the total costs into pretreatment costs, boiler costs, steam plant and emissions control equipment. This data was converted to total plant costs using the ratios method (the data is shown in Appendix E) and used to produce Figure 8.2. A regression on the data gives Equation 8.2.

$$TPC_{gen,steam},\ \$k_{1995} = 1407\left(E_{e,gross}\right)^{0.695} \tag{8.2}$$

where $E_{e,gross}$ = Gross generator output, MW_e

These capital costs are for established equipment and the learning effect that may be applied to them is minimal. They are assumed to be 100th plant costs (see Section 4.7).

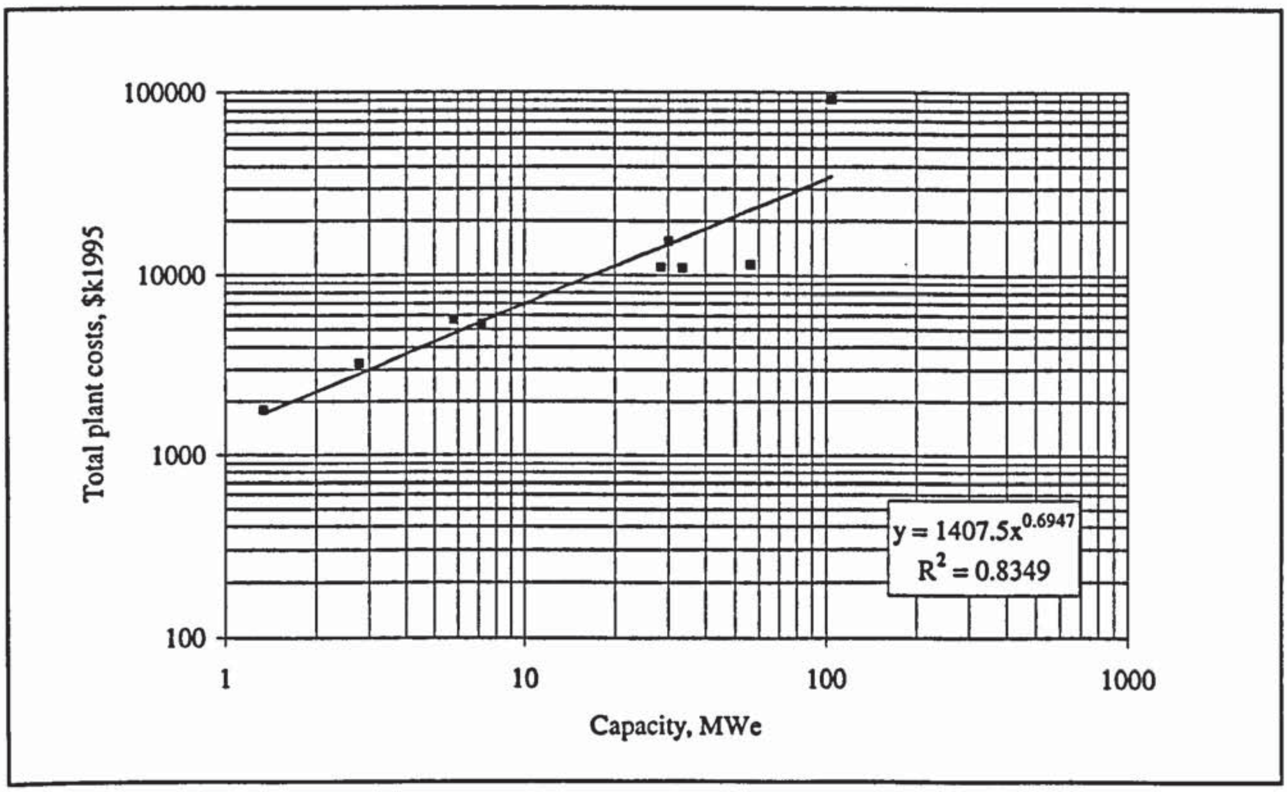


Figure 8.2 - Steam cycle capital cost as a function of gross power output

8.2.4 Operating costs

8.2.4.1 Utilities

Internal power consumption for combustion steam plant was discussed in Section 7.2.4.2. The internal power consumption of the cycles used to produce the capital cost relationship is between 5 and 11% of the gross power output and data by Grass [31] gives an internal power consumption for fluid bed combustors and steam cycles of up to 17.7%. These higher internal power consumption rates are due to the extra fan requirements for the fluid bed. Given that the pretreatment system internal power consumption and the combustor internal power consumption are accounted for separately, the internal power consumption of the steam cycle is set at 4% of the gross power output. This power consumption is not charged to the module but is subtracted from the gross system electricity output by the grid connection module (Section 8.6).

Cooling water is required at a rate of 5 t/MWh and is charged at a rate of 1.7 ¢/t (Section 4.6.5). This has been estimated from an estimate of 5000 t/h cooling water consumption for a 1000 MWe coal combustion plant [286]. This is a considerable extrapolation but the impact of cooling water consumption is very low and any error is insignificant.

Make-up water is required at a rate of 1.5 t/MWh produced (based on an average of data given by Golobic [251] and Tewksbury [240]), at a cost of 1.03 \$/t (see Section 4.6.5).

8.2.4.2 Labour

Labour requirements in the literature for combustion and steam cycle plant were analysed in Section 7.2.4.1. Using the data given in that earlier discussion, the specific labour requirements per MW_e output are presented in Figure 8.3. Just as in Section 7.2.4.1, two functions are used to calculate the labour requirement: the logarithmic function given in Equation 8.3 is used up to 35 MW_e, the power function given in Equation 8.4 is used above 35 MW_e,

$$\mathrm{Labour}_{steam} = \left[-0.1951\ln\left(P_{e,gross}\right) + 0.9298\right] \times P_{e,gross} \qquad (8.3)$$

$$\mathrm{Labour}_{steam} = \left[1.6887\left(P_{e,gross}\right)^{-0.5539}\right] \times P_{e,gross} \qquad (8.4)$$

where $P_{e,gross}$ = Power output from the steam turbine, MW_e.

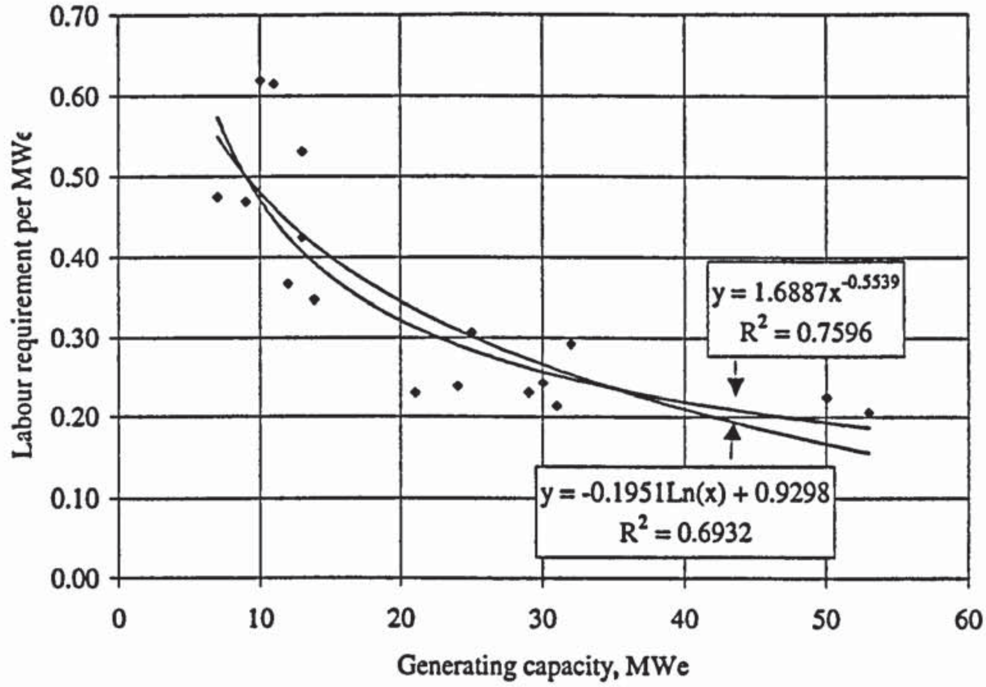


Figure 8.3 - Labour requirements, steam cycle module

8.2.4.3 Maintenance and overheads

Maintenance costs of steam cycles are reported to be 4% of installed capital costs or equivalent to 0.2-0.4 ¢/kWh [250]. A default value of 0.4 ¢/kWh is used in the module.

Overheads are calculated as a percentage of capital costs, as described in Chapter 4. A default of 4% of total plant costs is used.

8.2.5 Waste heat availability

Waste heat for the flue gas dryer (described in Chapter 6) is not produced by this module. Dryers used in combustion and steam cycle systems will use flue gas from the combustor, as described in Section 7.2.5.

8.3 LIQUID-FIRED DUAL FUEL ENGINE MODULE

8.3.1 Fuel requirements

The main energy input to this module is the energy supplied by the pyrolysis liquid output as calculated in the fast pyrolysis module reported in Chapter 7. This fuel is supplemented by an auxiliary diesel fuel that ignites the main charge, required because the pyrolysis liquid ignition characteristics are poor. The level of diesel pilot fuel required is not currently known and will be established in the current demonstrations [152, 170]. This module will assume that 7.5% of the total energy supplied to the engine is provided by the diesel pilot fuel, based on initial test results [287].

The use of pyrolysis liquids at elevated temperatures could result in char deposits in the fuel injection system [110] and for this reason the injection system is flushed by a solvent at the beginning and end of engine runs. A suitable solvent is methanol, used for ten minutes during start-up and shut down as a replacement for pyrolysis liquid [287]. This third fuel will add to the complexity of the fuel storage and injection system, as discussed in Section 8.3.4.

8.3.2 Engine replication

Engines are often used in multiples since this increases the overall generation reliability and their capital costs are virtually independent of scale (see Section 8.3.4). Multiple engines must be used to meet generating capacities in excess of around 40 MW_e because engines with larger capacities are rarely available [288]. A conventional plant operating on diesel might install several engines including spares to meet a changing electricity demand with allowances for machines in scheduled or unscheduled maintenance, as discussed by Singer [80]. Engine redundancy has not been included in this study since it would add extra costs to the engine-based systems only and distort the systems comparisons: scale economies in the steam turbine and gas turbine generating cycles favour the installation of a single machine. However, reliability may be an important issue where the electricity output must be guaranteed to meet contractual obligations. This may demand the installation of spare machines and the effect of this on the systems should be examined in further work.

Following discussions with Ormrod Diesels [287] and ECNZ (the main New Zealand utility) [289] the maximum engine sizes shown in Table 8.3 were specified. The specified power output is shared equally between as many engines as are required to conform to the maximum engine sizes in each capacity range.

Table 8.3 - Maximum Diesel Engine Sizes

Total Generating Capacity, MW_e	Maximum Generating Capacity per Engine, MW_e	Equivalent Energy Input, GJ/h LHV
< 10	5	45
10-20	7.5	67.5
>20	10	90

8.3.3 Generating efficiency

Efficiency data for diesel engines has been collected from the literature and is presented in Figure 8.4 [130, 131, 287, 288, 290, 291]. Diesel engine efficiency is not significantly affected by operation in dual fuel mode when using natural gas [93] and it is assumed in the absence of operational data that the dual fuel operation with a pyrolysis liquid will also have minimal impact on efficiency.

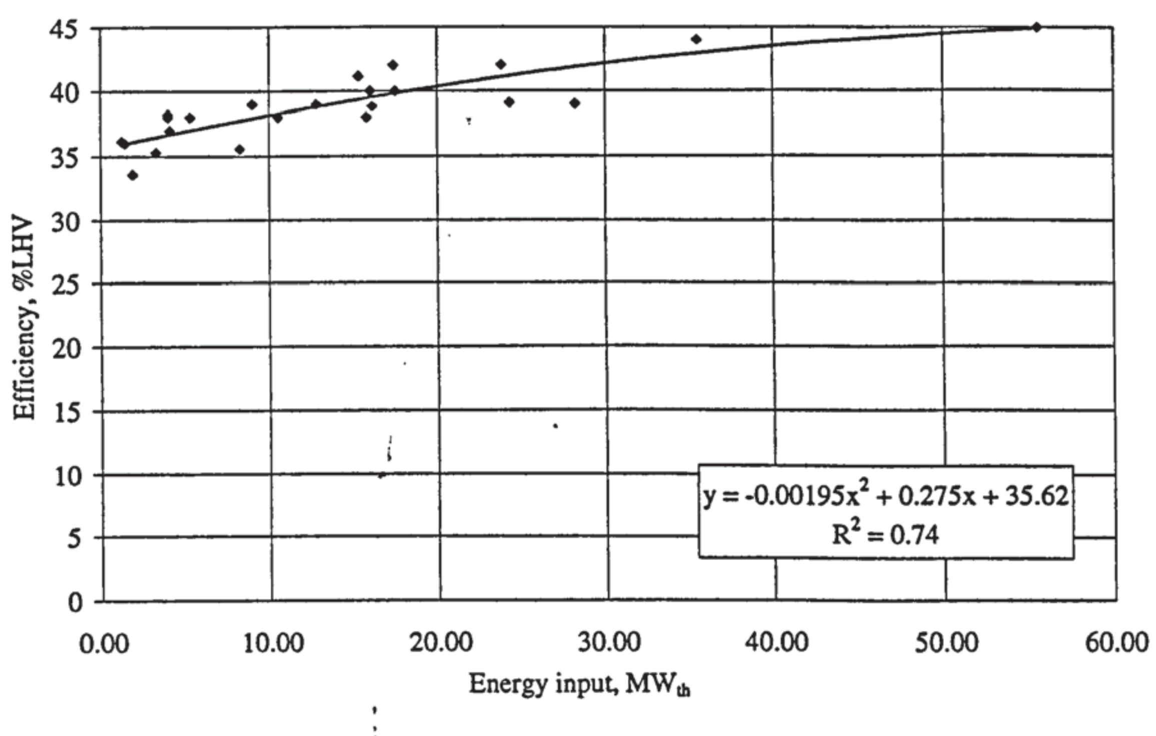


Figure 8.4 - Diesel engine generating efficiencies

The data in Figure 8.4 was converted to an energy input basis using Equation 8.5. Efficiencies were then plotted against engine energy input and a regression curve generated to give Equation 8.6. Equation 8.6 is applicable to a maximum of 70 MW_{th}, allowing the calculation of the single engine capacities shown in Table 8.3. The trend cannot be applied to higher capacities because of the quadratic function.

$$\left(E_{th,conv} + E_{th,aux}\right) = \frac{P_{e,gross}}{\eta_{gen,peng}} \times 3.6 \tag{8.5}$$

$$\eta_{gen,peng} = -0.00195\left(E_{th,conv} + E_{th,aux}\right)^2 + 0.275\left(E_{th,conv} + E_{th,aux}\right) + 35.62 \tag{8.6}$$

where $\eta_{gen,peng}$ = Gross electrical efficiency, % LHV basis

$P_{e,gross}$ = Gross power output, MW_e

$E_{th,conv}$ = Energy supplied by the pyrolysis liquid, MW_{th} LHV basis

$E_{th,aux}$ = Energy supplied by the diesel pilot fuel, MW_{th} LHV basis

8.3.4 Capital costs

Dual fuel diesel engine and generator investment costs were taken from the literature and discussions with manufacturers. Costs vary widely according to the speed and build quality of the engine: low speed, heavy duty engines tend to have higher initial costs offset by lower maintenance costs. Low speed, heavy duty engines were selected where possible because they are more adaptable to unconventional fuels. The capital costs that were identified have been adjusted using the ratios method (see Chapter 4) to give the total plant costs in $US\$_{1995}$ and the results are shown in Figure 8.5 [93, 94, 287, 289, 292]. The basic data is presented in Appendix E.

The costs shown in Figure 8.5 are based on dual fuel engines operating with natural gas and diesel. Adjustments should be considered for:

1. the storage of the pyrolysis liquid in a day tank;
2. the extra complexity imposed by the methanol flush and the probable requirement for special materials and fuel pre-heating to resist corrosive attack from the pyrolysis liquid and lower viscosity [293].

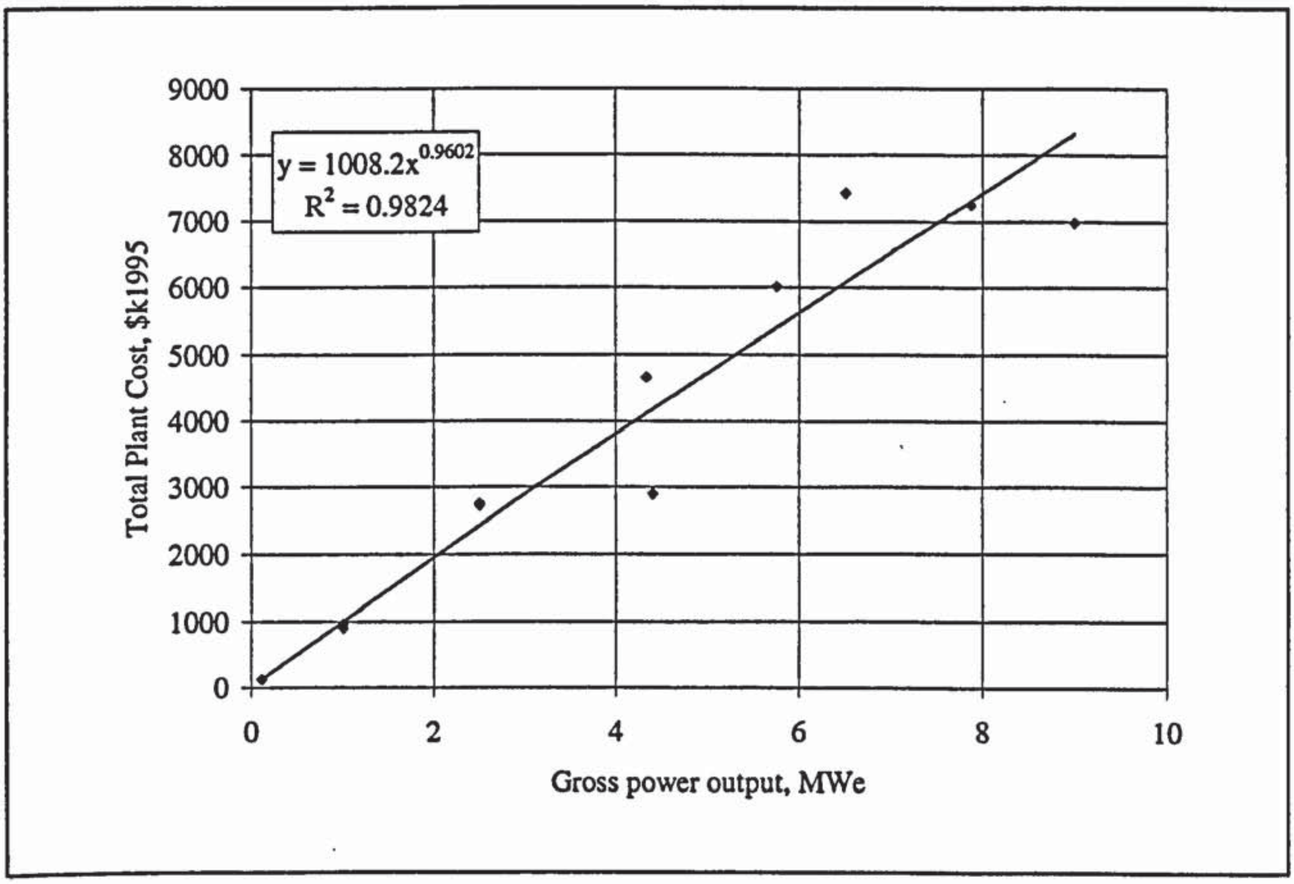


Figure 8.5 - Total plant cost of dual fuel diesel engine generating sets

A nominal additional cost of 10% has been included to allow for these items to give the total plant cost for the dual fuel diesel engine and auxiliary fuels storage, as shown in Equation 8.7.

$$TPC_{gen,peng},\ \$k_{1995} = \left[1008\left(P_{e,gross}\right)^{0.96}\right] \times 1.10 \tag{8.7}$$

where $P_{e,gross}$ = Gross generator output, MW_e

If multiple engines are used it is assumed that plant replication and shared plant items will offer some capital cost savings. Therefore the total plant cost for all engines is given by Equation 8.8 where n is the number of engines used.

$$TPC_{gen,peng},\ k\$_{1995} = TPC_{engine} \times n^{0.9} \tag{8.8}$$

Fast pyrolysis liquid bulk storage is already included in the fast pyrolysis module. The costs of pyrolysis liquids storage are only required when modelling de-coupled fast pyrolysis and diesel engine systems. In such cases, pyrolysis liquids storage costs are included in the engine module and are calculated using the same procedure given in the fast pyrolysis module description (see Section 7.3.3.2)

8.3.5 Operating costs

8.3.5.1 Auxiliary fuels

Diesel oil is charged at a rate of 0.51 \$/l, as used in the pretreatment module [248]. The consumption of diesel is calculated from the energy requirement specified in Section 8.3.1 and using a diesel lower heating value of 42.6 GJ/t and a density of 0.85 kg/l. The methanol requirement is very low since it is only used for short periods to flush the system. Methanol costs have not been included.

8.3.5.2 Utilities

Some of the gross power output from the engine(s) is used in the plant for auxiliaries such as fan motors, the lubrication oil pumps, fuel injection pumps and the control equipment. A factor of 3.0% of the gross power output is used to calculate this internal power consumption [287]. This power consumption is not charged to the module but is subtracted from the gross system electricity output by the grid connection module (Section 8.6). No other utilities are required.

8.3.5.3 Labour

Labour requirements for diesel engine gensets were given by Solantausta in the study reviewed in Section 3.4.3 [113]. Equation 8.9 has been derived from that data.

$$\text{Labour}_{\text{peng}}, \text{operators / shift} = 0.4847\left(P_{e,\text{gross}} - P_{e,\text{gen}}\right)^{0.483} \tag{8.9}$$

where $P_{e,gross}$ = Total gross electricity output of all engines, MW_e

$P_{e,gen}$ = Internal power required by the engines, MW_e

This curve is compared in Figure 8.6 with a annual labour cost data given in a survey of diesel generating plant [291]. Unfortunately this data only showed total labour costs and as such the labour requirement has been derived from the total costs using a labour cost of \$30000/person and 5 shifts in rotation. The data shows that the curve presented by Solantausta is reasonable. The scatter of the data produced from the review is probably due to the assumptions made in converting the costs to the number of operators.

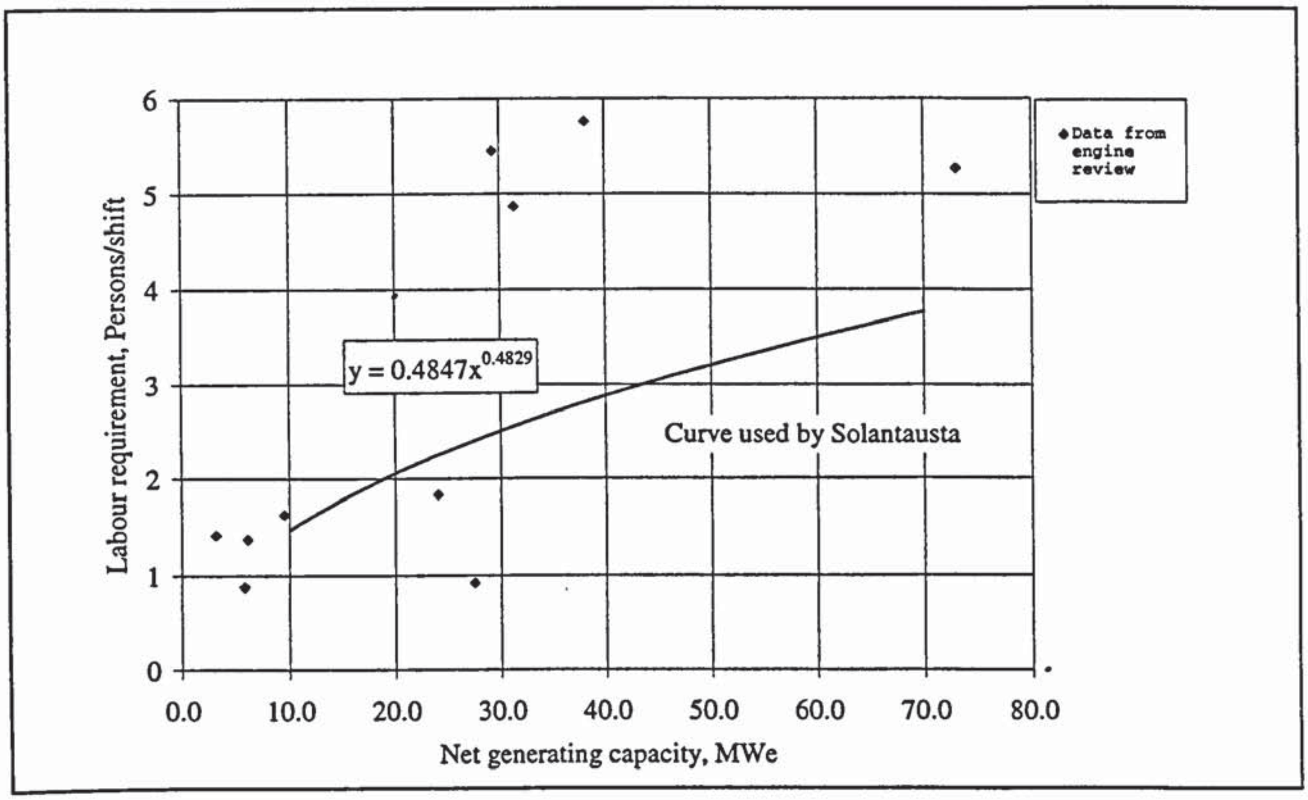


Figure 8.6 - Diesel Plant Labour Requirement

8.3.5.4 Maintenance and overheads

Operating data in the literature [291] gave an average maintenance cost (materials only, labour was excluded and is included in the labour costs above) of 0.67 ¢/kWh. The values were independent of capacity. It can be assumed that the maintenance costs for a dual fuel engine operating on pyrolysis oil would be higher because of the contaminants in the liquid, its viscosity and other deleterious characteristics. Therefore a maintenance cost of 1 ¢/kWh is assumed. Another major operating cost that should be included under maintenance is lubricating oil. This accounted for 4% of the total operating costs given in [291], with an average cost of 0.22 ¢/kWh. Again this cost was independent of capacity. Thus a total maintenance cost of 1.22 ¢/kWh is included.

Overheads are charged at the standard rate defined in Chapter 4, 4% of total plant costs per year.

8.3.6 Waste heat availability

It is assumed that the dryer heat requirement would be met by the flue gases produced from the combustion of the off-gas and char in the fast pyrolysis module (see Section 7.3.5). It would not be possible to use waste heat for the dryer in de-coupled systems anyway.

Since the dryer energy is supplied elsewhere, the waste heat produced by the engine module would be available to supply heat for process uses or district heating. Combined heat and power production has not been considered in this work because it has been rarely applied in UK. However, the latest NFFO order did include a provision for cogeneration and CHP has recently been attracting increasing support in the UK [27]. Further work should examine opportunities for cogeneration carefully as a means of increasing the system revenue and offsetting the electricity production costs.

8.4 GAS-FIRED DUAL FUEL ENGINE MODULE

8.4.1 Introduction

This module calculates the cost and performance of electricity generation using a dual-fuel diesel engine or engines fired by a low heating value gas with a diesel pilot fuel. This module uses many of the relationships defined in Section 8.3, and as such:

- The number of engines required is as specified in Section 8.3.2;
- Diesel fuel charges are as specified in Section 8.3.5.1;
- Utilities requirements are as specified in Section 8.3.5.2;
- Labour requirements are as specified in Section 8.3.5.3; and
- Maintenance and overheads charges are as specified in Section 8.3.5.4.

8.4.2 Fuel requirements

In the gas dual fuel engine generator two fuels are required: the fuel gas and a diesel pilot to ignite the main gaseous fuel. There is more experience of dual fuel operation with low heating value gases than with pyrolysis liquid and the amount of diesel fuel required is widely reported to be equivalent to around 5% of the total energy input [93, 128, 129, 130, 294]. Thus it is assumed that 5% of the energy input is provided by a diesel pilot fuel.

8.4.3 Generating efficiency

The generating efficiency of this module uses the same basic relationship developed for the pyrolysis liquid fuelled engine, with a reduction of 10% to account for operation in dual fuel mode with a low heating value gas [294, 129]. This gives Equation 8.10.

$$\eta_{gen,geng} = \left[-0.00195\left(E_{th,conv} + E_{th,aux}\right)^2 + 0.275\left(E_{th,conv} + E_{th,aux}\right) + 35.62\right] \times 0.9 \quad (8.10)$$

where $\eta_{gen,geng}$ = Gross electrical efficiency, % LHV basis

$E_{th,conv}$ = Energy supplied by the fuel gas, MW_{th} LHV basis

$E_{th,aux}$ = Energy supplied by the diesel pilot fuel, MW_{th} LHV basis

8.4.4 Capital costs

The capital costs are based on the relationship developed for the liquid-fired dual fuel engine module with the following changes:

1. The factor of 10% is not required because the engine injection system does not include methanol or pyrolysis liquid storage.
2. Engine power outputs on low heating gases are de-rated by varying amounts depending on the heating value of the gas. De-rating of engine outputs of between 6% and 50% have been reported, with most engines de-rated by around 20% [33, 129, 128, 133, 294, 295]. Given a de-rating of 20%, an engine must be 25% bigger (i.e.100%/80%) to give the required power output when fired by a low heating value gas.

These two changes give Equation 8.11 that is used to calculate the total plant cost of a single engine using low heating value gas.

$$TPC_{gen,geng}, \$k_{1995} = 1008\left(P_{e,gross} \times 1.25\right)^{0.96} \quad (8.11)$$

where $P_{e,gross}$ = Gross generator output for a single engine, MW_e

Single engine size limits are the same as those presented in Section 8.3.2 and where multiple engines are required Equation 8.8 is used to calculate the total plant cost for all engines.

8.4.5 Waste heat availability

The hot gases for drying in the atmospheric engine and dual fuel engine system could come from two basic sources: heat recovery from the raw fuel gas during gas cooling or the engine exhaust gases. The energy available from fuel gas treatment is discussed in Section 7.4.5.

A typical energy balance for a diesel engine is given by Kauffman [296], and is presented in Table 8.1. It can be seen that a substantial part of the energy in the fuel is emitted in the exhaust gases. Assuming that the exhaust gases leave the engine at 370°C [292] and leave the dryer at 102°C (see Section 6.7.1), the heat available for drying was calculated and compared with the energy required for drying, with the results shown in Table 8.5. There is sufficient energy for drying the feedstock at both 1 MW_e and 100 MW_e, with more surplus energy at 100 MW_e due to the higher gasification efficiency and therefore a lower feed requirement.

Table 8.4 - Heat balance from a 1 MW_e diesel engine

	W/kW_{th}	%
Jacket water	314	11.5
Turbocharger	55	2.0
Lubricating oil	104	3.8
Aftercooling	109	4.0
Exhaust	951	34.7
Radiation	205	7.5
Power	1000	36.5
Total	2738	100.0

Table 8.5 - Waste heat availability for drying from the dual fuel engine exhaust gases

System output	MWe	1.00	100.00
Generation efficiency	%LHV	33%	39%
Engine input	MWth	3.05	253.88
Energy supplied by fuel gas	MWth	2.90	241.19
Gasifier efficiency	%LHV	70%	77%
Energy input to gasifier	MWth	4.13	313.20
	GJ/h	14.85	1127.52
Prepared feed LHV	GJ/odt	16.04	16.04
Feed input required	odt/h	0.93	70.31
Evaporation load	twe/h	0.76	57.90
	GJ/h	2.71	205.72
Energy in flue gas	%LHV in	35%	35%
	GJ/h	3.81	317.15
Energy available	GJ/h	2.96	246.37
Surplus	%	9.2%	19.8%

8.5 GAS TURBINE COMBINED CYCLE MODULE

8.5.1 Fuel requirements

This module assumes that the fuel available will be a pressurised, clean low heating value fuel gas as produced by the Pressurised Gasification module in Section 7.5. No auxiliary fuel is required during normal operation although a back up distillate fuel will be used during start up since the gasifier must be supplied with pressurised air from the gas turbine compressor before it can provide a fuel gas. This distillate requirement is not considered here because the GTCC is assumed to be a base load plant and therefore the start-up fuel would be required very infrequently.

There are many issues that must be considered if a gas turbine is to be adapted for operation with a low heating value gas, including the air and fuel mix required to give the correct turbine inlet temperatures, the control and operation of the gas turbine compressor, any emissions limitations, injection system limitations and possible redesign of the early gas turbine stages to meet the increased flow rates that will arise when firing low heating value gases [110, 111]. Such analyses are beyond the scope of this study and the approach here is more general.

One key simplification is that the energy output from the gasification module is the fuel input to the gas turbine. Strictly, the fuel input to the gas turbine is only the dry fuel gas and the tar vapours, and should not include the enthalpy of the steam. The fuel component of the fuel gas is burned in combustion air in sufficient quantities to give the desired turbine inlet temperature and the expansion of the combustion gases through the turbine stages produces power. The steam does not take part in the combustion process and its effect is simply to increase the mass flow through the gas turbine and thus its output and efficiency: the gas turbine is effectively operating in a partial STIG cycle [111]. In this respect, the steam displaces part of the fuel that would have been required to give a specific output, and therefore in this generic analysis it is reasonable to consider the fuel energy and sensible energy of the total fuel gas stream as a single fuel input.

8.5.2 Generating efficiency

Generating efficiency is calculated using a methodology developed by Maude [81] during the early evaluation of coal IGCC systems. This used a basic flow diagram as shown in Figure

8.7. A spreadsheet was constructed based on this methodology and is included in Appendix E.

The methodology begins with the gasifier, not the clean and partially cooled gas that is the output from the pressurised gasification module ("zgH" in Figure 8.7). This is because the sensible energy lost in cooling the gas during gas cleaning is recovered for use in the steam cycle. From the data sheet developed for the pressurised gasification energy balance, shown in Appendix D, it can be seen that the sensible heat losses in cooling the gas are around 12% of the energy in the gas that leaves the gasifier, a significant amount. A new equation was developed from the data sheet in Appendix D for use in this analysis that calculates the energy stream "gH" in Figure 8.7, and a 12% by-pass ratio was used (the by-pass ratio is the amount of energy recovered in gas cooling that is used directly in the steam turbine).

The gas turbine efficiency is set using an equation derived from performance data for aeroderivative gas turbines taken from data published in Modern Power Systems [124], as shown in Figure 8.8. This assumes that all gas turbines would be capable of operating using low heating value gases, which is currently not the case. In fact there are currently very few gas turbines that are available for operation on low heating value gases, and in reality IGCC system outputs currently rise in steps that correspond to the capacities of the available gas turbines [109]. The continuous curve shown in Figure 8.8 is an approximation required to allow the continuous range of capacities required by this work. In future the variety of available gas turbines should increase so that the stepped relationship approaches a continuous curve.

The steam cycle efficiency uses Equation 8.1 derived above to calculate efficiencies for various energy inputs. The Maude methodology assumes that the total sensible energy in the gas turbine exhaust gases is recovered for the steam cycle. This can not be the case and so the energy in the gas turbine exhaust gases is adjusted by assuming that the gas turbine exhaust is at 500°C (the average of data given in Modern Power Systems [124]) and that the heat recovery steam generator that would raise steam from the exhaust gases operates down to 200°C. Any residual sensible energy in the exhaust gases is available for the dryer.

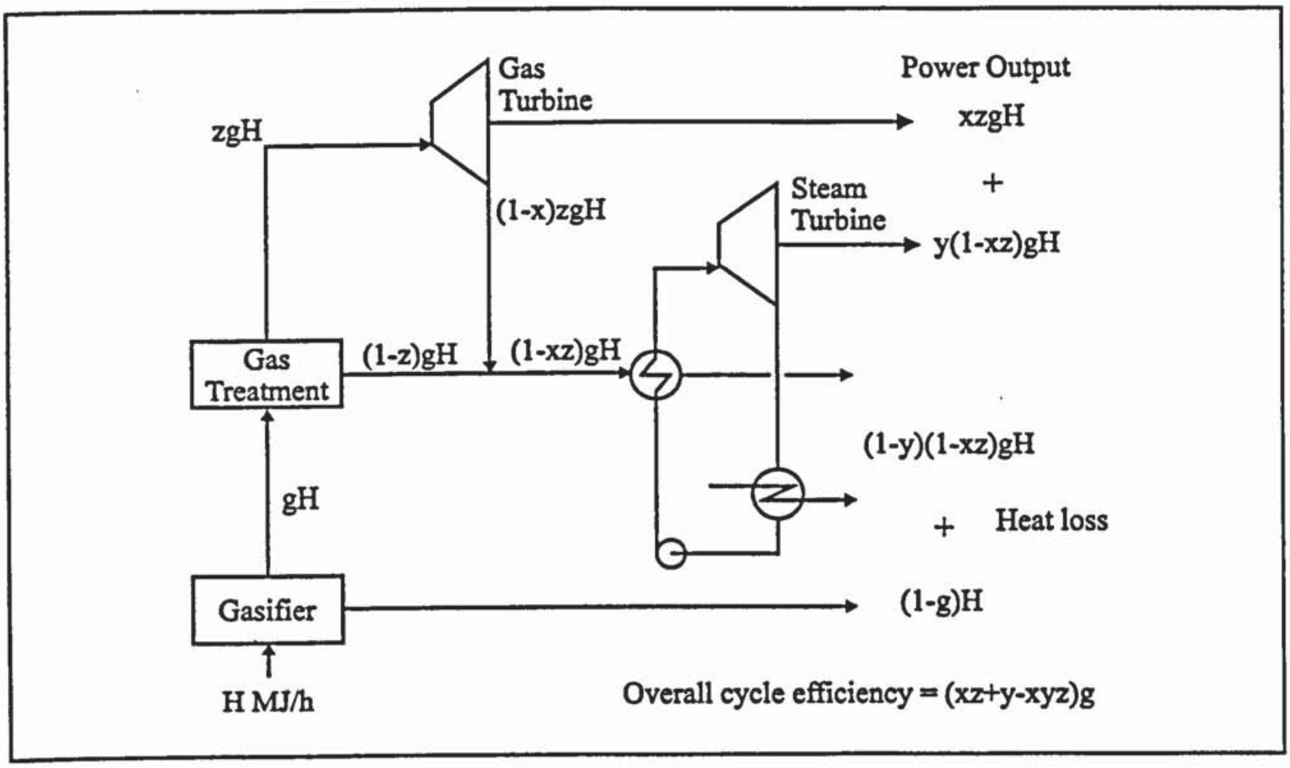


Figure 8.7 - Methodology for calculating the efficiency of the GTCC

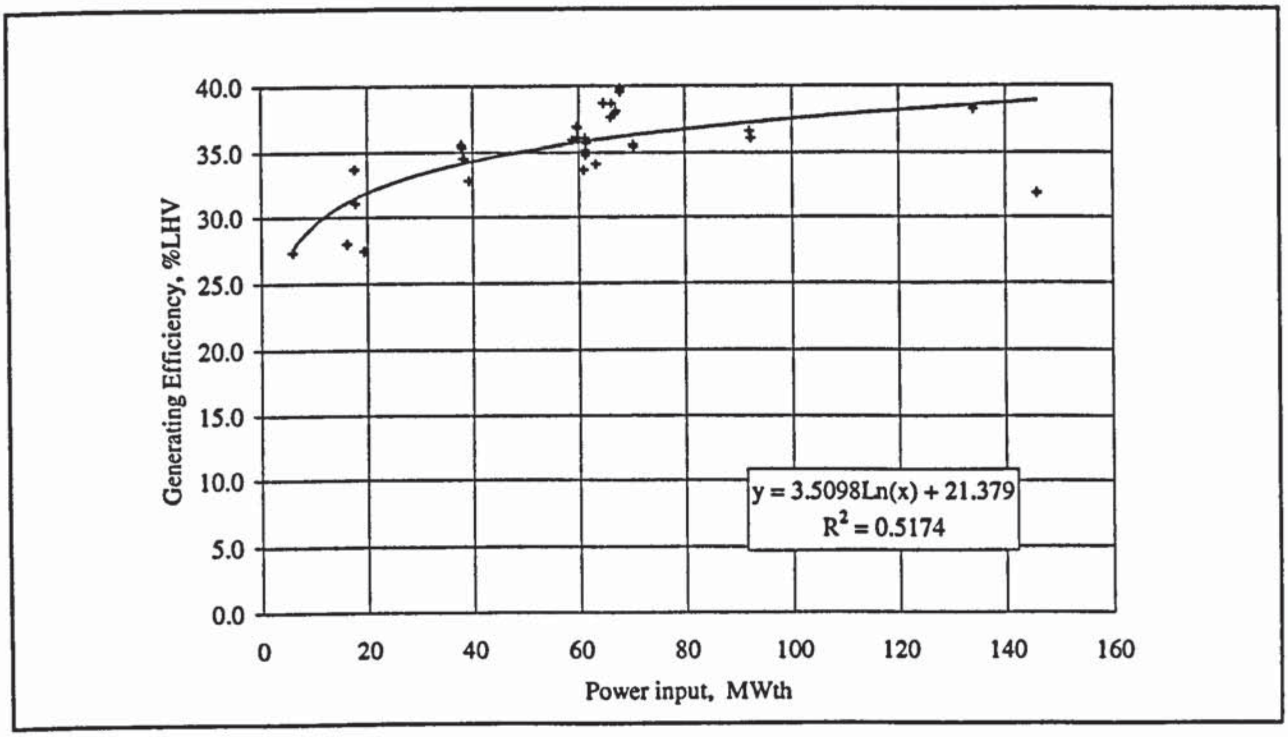


Figure 8.8 - Simple cycle gas turbine efficiencies

The energy efficiency for the module is the ratio of the total power output ("xzgH" and "y(1+xz)gH" in Figure 8.7) to the gas turbine input ("zgH" in Figure 8.7). This ratio was calculated for a variety of capacities using the spreadsheet in Appendix E, producing the points in Figure 8.9. A regression analysis on this data gives the gas turbine combined cycle efficiency that is used in the module, Equation 8.12.

$$\eta_{gen,gtcc} = 0.0589\ln\left(E_{th,conv}\right) + 0.2565 \tag{8.12}$$

where $\eta_{gen,gtcc}$ = The gross generating efficiency of the gas turbine combined cycle, %

$E_{th,conv}$ = The lower heating value of the conversion energy product, MW_{th}

Given this very general analysis of what is in reality a very complex design problem, it is useful at this point to briefly compare the efficiencies produced by this method to the efficiencies produced by more detailed case study analyses. Figure 8.10 gives this comparison, with pressurised gasification IGCC efficiencies (before internal consumption) taken from work by McIlveen-Wright [186] and Solantausta [113] shown as well as the predicted performance of the BIOCYCLE THERMIE project [54]. The line produced by this work will be slightly lower than the total system would produce because the energy saved by drying the feed is not included. This would raise the efficiency by around 3%, and with this in mind it can be seen from Figure 8.10 that the efficiencies agree reasonably well.

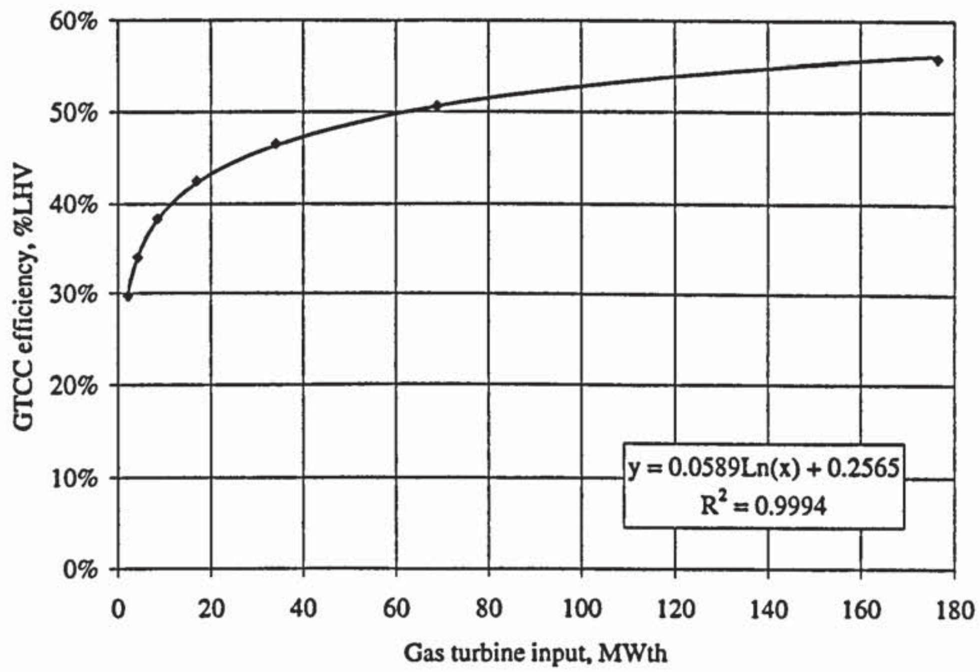


Figure 8.9 - Gas turbine combined cycle generating efficiencies

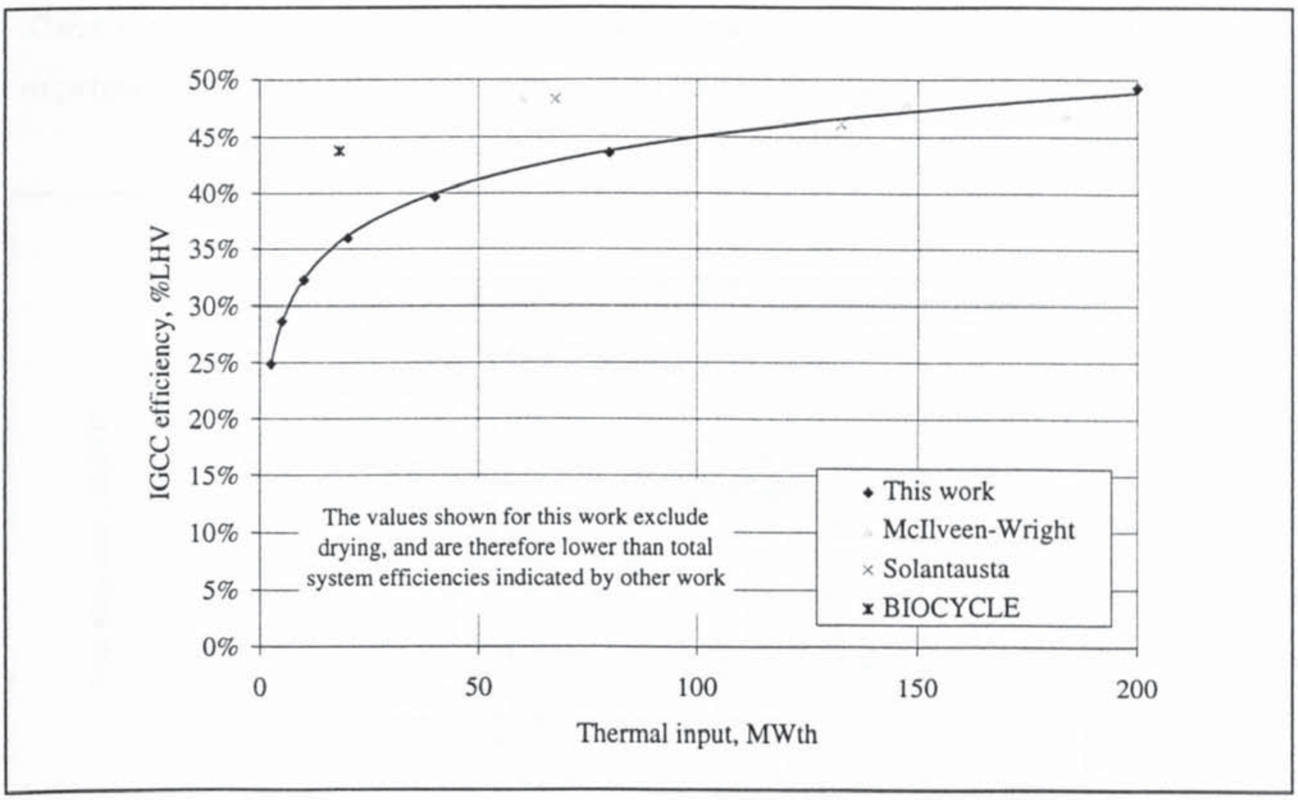


Figure 8.10 - Gross IGCC efficiencies, approximate comparisons with previous work

8.5.3 Capital costs

Capital costs have been taken from the literature for natural gas-fired gas turbine combined cycles. Some gas turbine modifications may be required to adapt the fuel and air injection systems, the combustion chambers and the turbine blading to low heating value gas operation. Gas turbine modification costs have not be included but they are expected to only be a small percentage of the gas turbine, which is in itself only part of the combined cycle. Hence the error introduced by this assumption is small. In any case, it is possible that the gas turbine manufacturer would bear this cost as a research and development expense. Combined cycle cost data is appended in Appendix E [98, 203, 297, 298, 299]. The cost data was adjusted to total plant costs using the ratios method and adjusted to US\$, 1995 basis. The results of normalising the cost data are shown in Figure 8.11, which gives Equation 8.13.

$$TPC_{gen,gtcc},\ \$k_{1995} = 2123 \times \left(P_{e,gross}\right)^{0.8613} \tag{8.13}$$

where $TPC_{gen,gtcc}$ = The total plant cost of the gas turbine combined cycle module, $\$k_{1995}$

$P_{e,gross}$ = The total gross power output, MW_e

These capital costs are based on established equipment and the learning effect that they will experience is minimal. They are assumed to be 100th plant costs.

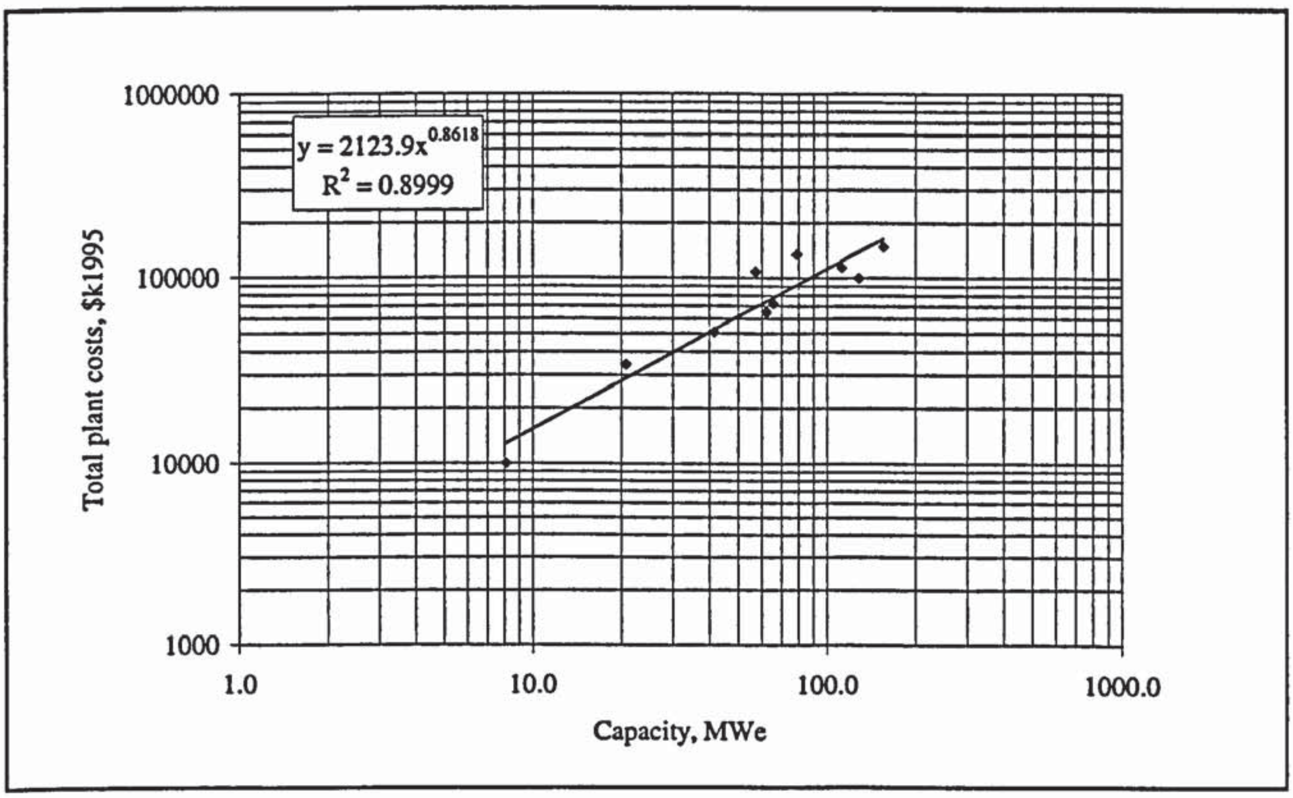


Figure 8.11 - Gas turbine combined cycle total plant costs

8.5.4 Operating costs

8.5.4.1 Utilities

Bressan [300] reports internal power consumption for several gas turbine combined cycles rated at between 42 and 63 MW_e and operated on natural gas. Power consumption for the six cycles ranged from 2.5% to 3.6% of the gross power output, with an average value of 3%. This mean value is used here as the internal power consumption for the gas turbine system.

Cooling water and boiler feed water requirements are calculated using the data for the steam cycle given in Section 8.2.4.1. Since the steam turbine only contributes about a 1/3 of the total power produced by the module, the consumption figures have been divided by 3 to give a cooling water requirement of 1.6 t/MWh and a boiler feed water requirement of 0.5 t/MWh.

8.5.4.2 Labour

Only one reference could be found in the literature that gave labour requirements for a gas turbine combined cycle with a capacity under 100 MW_e. This estimated that the labour requirement at would be 13 operators/year at 55 MW_e and 14 operators/year at 75 MWe []. This converts to 0.23 operators/year/MW_e at 55 MW_e and 0.18 operators/year/MW_e at 75 MW_e. These two data show good agreement with the estimates already made for the steam cycle and the diesel engines, as shown in Table 8.6. It can be expected that the labour required is higher than the labour for the steam cycle since the GTCC is more complex than the steam cycle. It is also reasonable to expect the labour required in the engine based cycles to be greater than the GTCC since multiple engines are used. On the basis of this evidence, the labour for the gas turbine combined cycle is based on the relationships for the steam cycle, with an additional 25% to account for the extra complexity. Thus Equation 8.14 is used at capacities of less than 35 MW_e and Equation 8.15 is used at the higher capacities.

$$\text{Labour}_{gtcc} = \left[-0.1951\ln\left(P_{e,gross}\right) + 0.9298\right] \times P_{e,gross} \times 1.25 \tag{8.14}$$

$$\text{Labour}_{gtcc} = \left[1.6887\left(P_{e,gross}\right)^{-0.5539}\right] \times P_{e,gross} \times 1.25 \tag{8.15}$$

where $P_{e,gross}$ = Power output from the gas turbine combined cycle, MW_e.

Table 8.6 - Comparison of labour requirements for the generation modules

Capacity, MW_e	Labour requirement, operators/year/MW_e		
	Gas turbine combined cycle	Steam cycle	Engine-based cycles
55	0.23	0.18	0.31
75	0.18	0.15	0.25

8.5.4.3 Maintenance and overheads

Reported maintenance costs for gas turbine combined cycles range from 0.7-1.0 ¢/kWh [297, 301]. Given the novel application, a high maintenance cost is likely because of the potential for damage to the gas turbine by contaminants in the gas. Thus the maintenance cost for the gas turbine combined cycle is set at 1.0 ¢/kWh.

Overheads are charged in the same manner as all the other modules, as 4% of the total plant cost for the module.

8.5.5 Waste heat availability

Waste heat for drying the wood will be available from the gas turbine exhaust gases after the heat recovery steam generator. The sensible energy of this gas stream was calculated as part of the procedure for calculating the gas turbine combined cycle efficiency, and a calculation based on this figure can be performed to estimate the surplus or deficit if this stream was used to dry the feedstock. The analysis is very similar to that carried out for the gas-fired dual fuel engine, and the results are shown in Table 8.7. There is a significant deficit due to the low temperature of the gas turbine exhaust after the heat recovery steam generator. Optimising the dryer can help to alleviate most of this deficit. For example, a reduction of the dryer exit temperature to 80°C, which is feasible although there is an increased risk of a plume [242], would reduce the deficit at 1 MW_e to 17.8%. The deficit could be further reduced by installation of a more efficient dryer. The dryer in Table 8.7 requires 3.553 GJ/twe (tonne water evaporated), a dryer with a 10% higher efficiency would use 3.120 GJ/twe, resulting in a deficit of 8.6%. The lower drying energy requirement is within the range of dryer energy consumption reported in Section 6.7.1, but it is not known how such a dryer would effect the costs of the drying stage. It can be concluded that there is a significant risk with the current system of not producing enough heat for drying. It has been shown that minor adjustments to the system can almost eliminate the problem and so the current module will not be adjusted further but it is important to investigate this further in later work to eliminate this uncertainty.

Table 8.7 - Waste heat availability for drying from the gas turbine exhaust gases

System output	MWe	1.00	100.00
Generation efficiency	%LHV	33%	57%
Engine input	MWth	3.04	174.70
Energy supplied by fuel gas	MWth	2.89	165.96
Gasifier efficiency	%LHV	86%	92%
Energy input to gasifier	MWth	3.38	180.82
	GJ/h	12.16	650.94
Prepared feed LHV	GJ/odt	16.04	16.04
Feed input required	odt/h	0.76	40.59
Evaporation load	twe/h	0.62	33.43
	GJ/h	2.22	118.77
Energy in flue gas	GJ/h	2.74	147.74
	GJ/h	1.49	80.44
Surplus	%	-32.8%	-32.3%

8.6 GRID CONNECTION MODULE

8.6.1 Net electricity output

This module calculates a net system capacity in MW_e and a net system output in MWh/y. The two figures are not directly related because some of the pretreatment equipment does not operate continuously.

The net capacity is calculated by subtracted the sum of the power requirements in MW_e of the pretreatment, conversion and generation stages from the gross power output, as shown in Equation 8.16.

$$P_{e,net} = P_{e,gross} - \left(P_{e,pret} + P_{e,conv} + P_{e,gen}\right) \tag{8.16}$$

where $P_{e,net}$ = Power supplied to the grid, MW_e

$P_{e,gross}$ = Power at the generator terminals, MW_e

$P_{e,pret}$ = Power required by the pretreatment module, MW_e

$P_{e,conv}$ = Power required by the conversion module, MW_e

$P_{e,gen}$ = Power required by the generation module, MW_e

In de-coupled systems fast pyrolysis systems the pretreatment and conversion stages are separated from the generator, and so only the generator internal power requirement is subtracted from the gross power output, as shown in Equation 8.17. The power requirement for the pretreatment and fast pyrolysis systems is supplied by the grid in this work, but further work could examine the costs and benefits of generating the power required at the site, possibly with some of the pyrolysis liquid.

$$P_{e,net} = P_{e,gross} - P_{e,gen} \tag{8.17}$$

where $P_{e,net}$ = Power supplied to the grid, MW_e

$P_{e,gross}$ = Power at the generator terminals, MW_e

$P_{e,gen}$ = Power required by the pretreatment module, MW_e

The net electricity output to the grid is calculated in a similar way to the net power output, except that this time the electricity actually consumed is subtracted from the gross output. For convenience in efficiency calculations that include the feed or conversion products, the units used are GJ/y. Equation 8.18 is used in close-coupled systems, Equation 8.19 is used in de-coupled systems.

$$E_{e,net} = E_{e,gross} - \left(E_{e,pret} + E_{e,conv} + E_{e,gen}\right) \tag{8.18}$$

where $E_{e,net}$ = Electrical supplied to the grid, GJ/y

$E_{e,gross}$ = Electrical at the generator terminals, GJ/y

$E_{e,pret}$ = Internal power consumption of the pretreatment module, GJ/y

$E_{e,conv}$ = Internal power consumption of the conversion module, GJ/y

$E_{e,gen}$ = Internal power consumption of the generation module, GJ/y

$$E_{e,net} = E_{e,gross} - E_{e,gen} \tag{8.19}$$

where $E_{e,net}$ = Electrical supplied to the grid in a de-coupled system, GJ/y

$E_{e,gross}$ = Electrical at the generator terminals in a de-coupled system, GJ/y

$E_{e,gen}$ = Internal power consumption of the de-coupled generation module, GJ/y

8.6.2 Capital costs

The connection of the system to the grid must be safe and meet the requirements of the utility in terms of protection for both the grid and the electricity supplier [230]. Grid connection equipment includes electrical control, protection equipment, transformers and switchgear. The costs of grid connection are very case-specific since they are a function of location, the size of plant and the grid voltage at the connection [302]. There is therefore some uncertainty associated with these capital costs. Equipment costs [287, 289, 302] were converted to direct plant costs using the ratios method (see Chapter 4) and the results are shown in Figure 8.12. Regression analysis of the data produced Equation 8.20. Although there is uncertainty about grid connection costs, their proportion of the total system capital costs is small and they should not have a significant impact on the overall system.

$$TPC_{Grid}, \$k_{1995} = 346 \times \left(P_{e,net}\right)^{0.537} \tag{8.20}$$

where TPC_{Grid} = Total plant cost of the grid connection equipment, $\$k_{1995}$

$P_{e,net}$ = Power supplied to the grid, MW_e

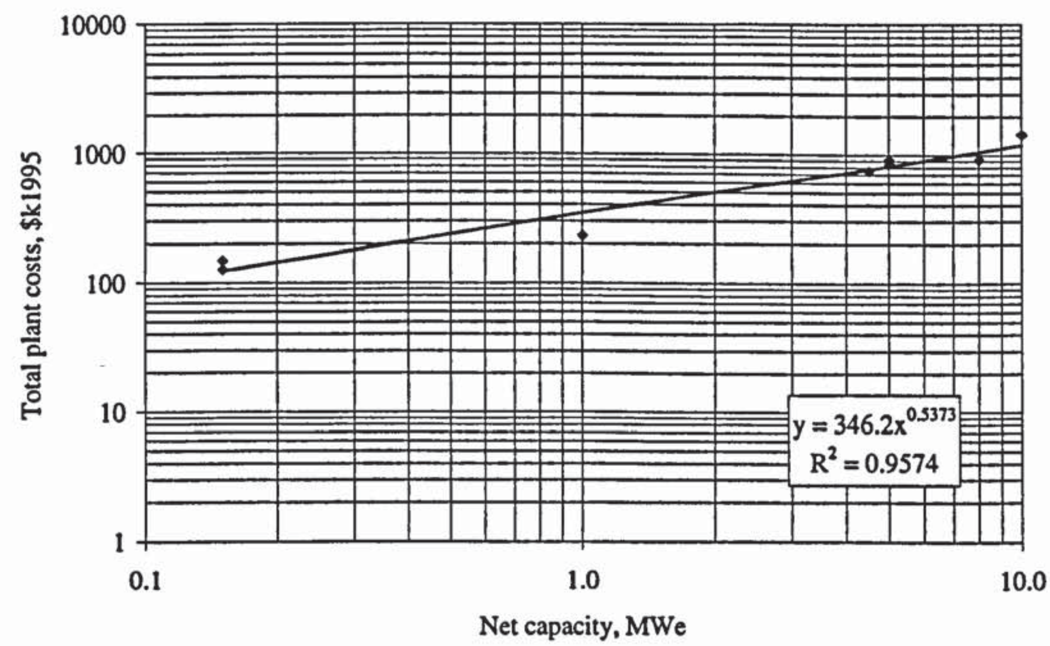


Figure 8.12 - Direct Plant Costs, Grid Connection

8.6.3 Operating costs

The capital-related operating costs of capital amortisation, overheads and maintenance are all calculated according to the methodology defined in Section 4.6. No labour or utilities are charged to this module. It is assumed that any labour would be supplied by the generation module.

8.7 SUMMARY

This chapter describes the development of the four electricity generation modules that are used to calculate the cost and performance of four generating cycles that were selected in Chapter 2 for evaluation as part of an integrated biomass to electricity system.

The modules are based on established technologies and are assumed to be 100th plant cost data. There is some uncertainty regarding the modifications that may be required to fire diesel engines on fast pyrolysis liquids and gas turbines on low heating value gases. The former costs have been increased by 10% to allow for modifications and extra liquid storage. Gas turbine modifications have not been included as they are not expected to be a significant part of the gas turbine combined cycle cost.

The most significant uncertainties in this chapter are introduced during the development of an efficiency relationship for the gas turbine combined cycle. Three simplifications have been introduced: fuel input to the gas turbine include all of the energy output from the gasifier; the combined cycle has been reduced to a very simple flowsheet; and gas turbine performance has been derived from all aeroderivative gas turbines rather than the few machines that have so far been promoted for low heating value gases. These measures were necessary to develop a generic relationship between gasifier output and combined cycle efficiency and a brief comparison has shown an approximate agreement between the resultant relationship and other work. However there is some error introduced and this aspect of the IGCC system model should be re-examined in future work.

The further development of the GTCC module should also include an examination of the energy available for drying. It has been shown that there is a deficit in the current system. It may be necessary to compromise the performance of the gas turbine combined cycle to ensure a dry feedstock by, for example, allowing some energy recovery for drying during gas cleaning and a higher temperature after the heat recovery steam generator.

Finally, all of the systems have opportunities for combined heat and power production that are not explored in this work and would be worth investigating in future. It would be difficult to assess heat availability for export without developing process flow schemes but some approximate calculations could be made by considering the major heat sources that are currently not exploited. The steam turbine condensers are possibilities, as are the engine exhaust gases (there is sufficient energy for drying in the conversion modules to make this energy available).

8.8 NOMENCLATURE

All costs are in US$, 1995 basis

$E_{e,conv}$ = annual electricity consumed in feed conversion, GJ/y

$E_{e,gen}$ = annual electricity consumed in electricity generation, GJ/y

$E_{e,gross}$ = annual electricity generated before internal losses, GJ/y

$E_{e,net}$ = annual electricity exported to the grid after internal losses, GJ/y

$E_{e,pret}$ = annual electricity consumed in feed pretreatment, GJ/y

$E_{th,aux}$ = energy in the auxiliary diesel fuel, MW_{th}

$E_{th,conv}$ = energy in the conversion product, MW_{th}

LHV = lower heating value

$P_{e,conv}$ = power required by the conversion module, MW_e

$P_{e,gen}$ = power required by the generation module, MW_e

$P_{e,gross}$ = power generated before losses, MW_e

$P_{e,net}$ = power supplied to the grid after losses, MW_e

$P_{e,pret}$ = power required by the pretreatment module, MW_e

$TPC_{gen,geng}$ = Total plant cost of the gas-fired dual fuel engine module, $\$k_{1995}$

$TPC_{gen,gtcc}$ = Total plant cost of the gas turbine combined cycle module, $\$k_{1995}$

$TPC_{gen,peng}$ = Total plant cost of the liquid-fired dual fuel engine module, $\$k_{1995}$

$TPC_{gen,steam}$ = Total plant cost of the steam cycle module, $\$k_{1995}$

$\eta_{gen,geng}$ = the gross generating efficiency of the gas-fired dual fuel engine, %

$\eta_{gen,gtcc}$ = the gross generating efficiency of the gas turbine combined cycle, %

$\eta_{gen,peng}$ = the gross generating efficiency of the liquid-fired dual fuel engine, %

$\eta_{gen,steam}$ = the gross generating efficiency of the steam cycle, %

9. RESULTS

9.1 INTRODUCTION

This chapter evaluates the four integrated biomass to electricity systems that have been modelled in the preceding chapters. The four systems and the parameters used to define them are presented in Section 9.2. The comparisons will evaluate the costs and performance of these base case systems in four ways.

1. Variations with capacity are examined in Section 9.3.
2. Learning effects are examined in Section 9.4.
3. Variations with feed cost are examined in Section 9.5.
4. Variations with availability are examined in Section 9.6
5. De-coupled systems are examined in Section 9.7.

The emphasis throughout these evaluations will be the comparison of the systems; the results are compared relative to the established combustion system and not to the conditions in the wider electricity market. Opportunities for the four systems in the electricity market are evaluated in Section 9.8.

The chapter concludes with an examination of the sensitivities and uncertainties of the four systems in Section 9.9.

9.2 SYSTEM DEFINITIONS

The four basic systems that were selected in Chapter 4 for further evaluation are presented in Table 9.1. The labels "Combustion", "Gas-Eng", "IGCC", and "Pyr-Eng" will be used throughout this chapter to refer to the systems shown in Table 9.1.

Each system has a set of conditions that define its characteristics in a base case that will be used throughout the comparisons in Section 9.3. These characteristics are presented in Table 9.2. All the base cases assume that the system is constructed immediately so that no learning effects are included. The base case Combustion system capital costs are considered the 100th plant, because all the capital cost relationships are based on the costs of established equipment. The other systems are 1st plant because the costs of the conversion technologies are based on the costs of first-of-a-kind installations.

The base case Combustion system is presented in all the systems comparisons as a reference point.

Table 9.1 - Systems for evaluation and their modules

Combustion	Gas-Eng	IGCC	Pyr-Eng
• Feed transport • Feed pretreatment • Combustion • Steam cycle • Grid connection	• Feed transport • Feed pretreatment • Atmospheric gasification • Gas-fired dual fuel diesel engine • Grid connection	• Feed transport • Feed pretreatment • Pressurised gasification • Gas turbine combined cycle • Grid connection	• Feed transport • Feed pretreatment • Fast pyrolysis • Pyrolysis liquid transport[a] • Liquid-fired dual fuel diesel engine • Grid connection

a only used in de-coupled fast pyrolysis systems

9.3 CAPACITY VARIATIONS

9.3.1 Overview

Limits of 1-100 MW_e were set in Section 2.2 with the provisos that the lower limit was unlikely to be viable in a dedicated commercial system and the upper limit was probably too high for the UK given the limited availability of feedstock. This section presents an initial comparison of the four systems under base-case conditions over the range 1-100 MW_e. This gives a broad overview of the systems' technical and economic performance and in the process highlights the most interesting capacity range for further analysis.

Figure 9.1 presents the net system efficiencies for the four base cases as a function of capacity. The Gas-Eng and Pyr-Eng systems have a virtually constant efficiency while the IGCC and Combustion systems offer more significant efficiency improvements with increasing capacity. The result of this is that any efficiency advantage offered by the engine-based systems at small scale is eroded as capacity increases, implying that they are more suitable for small scale rather than large scale installations.

Table 9.2 - Base case system characteristics

	Combustion	Gas-Eng	IGCC	Pyr-Eng
Configuration	Close-coupled[a]	Close-coupled[a]	Close-coupled[a]	Close-coupled
No. conversion sites	1	1	1	1
Feedstock (see Section 4.3)				
Cost before transport, \$/odt	40.00	40.00	40.00	40.00
Moisture content as delivered, %wet basis	50%	50%	50%	50%
Lower heating value, dry, GJ/odt	19.3	19.3	19.3	19.3
Feed transport (see Chapter 5)				
Land area limitation, %	5%	5%	5%	5%
Feed loading cost, \$/t	2.60	2.60	2.60	2.60
Feed transport cost, \$/t/km	0.09	0.09	0.09	0.09
Pretreatment (see Chapter 6)				
Delivery schedule, weeks/y	52	52	52	52
Delivery schedule, days/week	5	5	5	5
Delivery schedule, hours/day	10	10	10	10
Dry matter loss in storage, %/month	2.5%	2.5%	2.5%	2.5%
Moisture content change in storage, %/month	0%	0%	0%	0%
Overs in the feedstocks, %	5%	5%	5%	5%
Cost of diesel fuel, \$/l	0.51	0.51	0.51	0.51
Feed conversion (see Chapter 7)				
Prepared feed moisture content, %wet basis[a]	35%	15%	15%	10%
Prepared feed size[a]	Chips	Chips	Chips	Powder
Efficiency, %	Default[b]	Default[b]	Default[b]	Default[c]
Availability, %	90%	90%	90%	90%
Internal power consumption, kW_e/odt/h	65[d]	40	40	60
Generation (see Chapter 8)				
Efficiency, %	Default[b]	Default[b]	Default[b]	Default[b]
Internal power consumption, %/MW_e gross	4 %	3 %	3 %	3 %
Maintenance cost, ¢/kWh	0.4	1.12	1.12	1.0
Diesel consumption, %MW_{th} input	-	5%	-	7.5%
Financial (see Chapter 4)				
Currency[a]	$US\$_{1995}$	$US\$_{1995}$	$US\$_{1995}$	$US\$_{1995}$
Nominal interest rate, %	10%	10%	10%	10%
Inflation rate, %	5%	5%	5%	5%
Labour cost, \$/person/year	30000	30000	30000	30000
No. of shifts used[a]	4	4	4	4
Overheads, %TPC/y	4%	4%	4%	4%
Maintenance cost, %TPC/y (Pretreatment and conversion only)	4%	4%	4%	4%

a *Not a user variable, presented here for information only*

b *The default efficiency is calculated by the module using the relationships given in the module descriptions*

c *The default efficiency of the fast pyrolysis module is fixed to 62.0% and is calculated indirectly from the liquids yields and liquids heating values determined in Section 7.3.2.*

d *The combustion module actually bases internal power consumption on the thermal input. 65 kW_e/odt/h is calculated from 2% of the thermal input, as defined in Section 7.2.4.*

In Figure 9.2 the strong scale economies of the capital costs can be seen. All the systems show dramatic reductions in specific total plant costs up to 20 MW_e, but beyond this capacity the costs are more stable. It is important to consider Figure 9.1 and Figure 9.2 together, since there is a consistent link between higher efficiency and higher capital cost. This interaction will be explored in more detail later.

Electricity production costs for the four base cases are presented in Figure 9.3. This shows that the combustion system consistently produces electricity at a lower cost than the novel systems. This is a major barrier to the development of the novel technologies in biomass to electricity systems, and is brought about because the combustion system enjoys the benefits of learning effects, which are explored later in Section 9.4. If only the 1st plant systems are included it can be seen that Pyr-Eng is the least expensive option at small-scale, with IGCC producing the lowest cost electricity at the larger capacities.

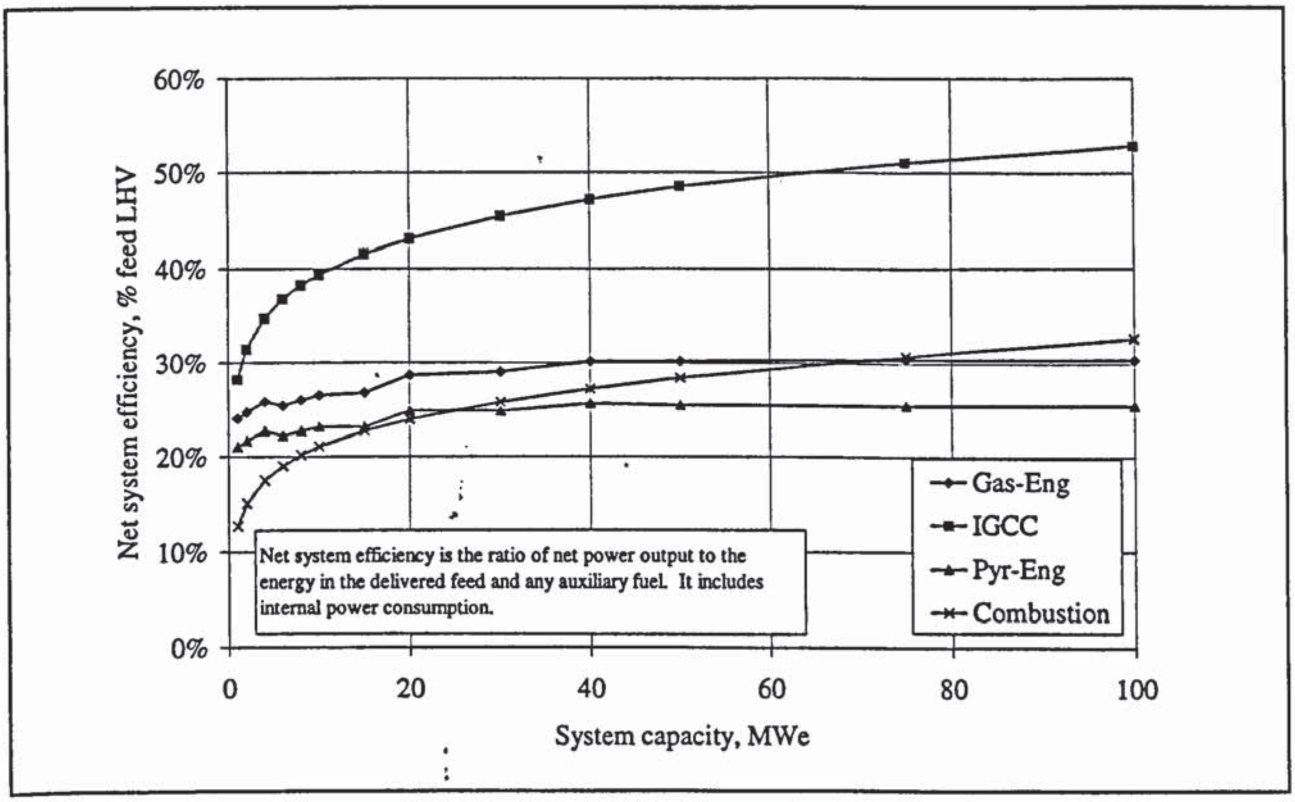


Figure 9.1 - Net system efficiencies, base case systems from 1-100 MW_e

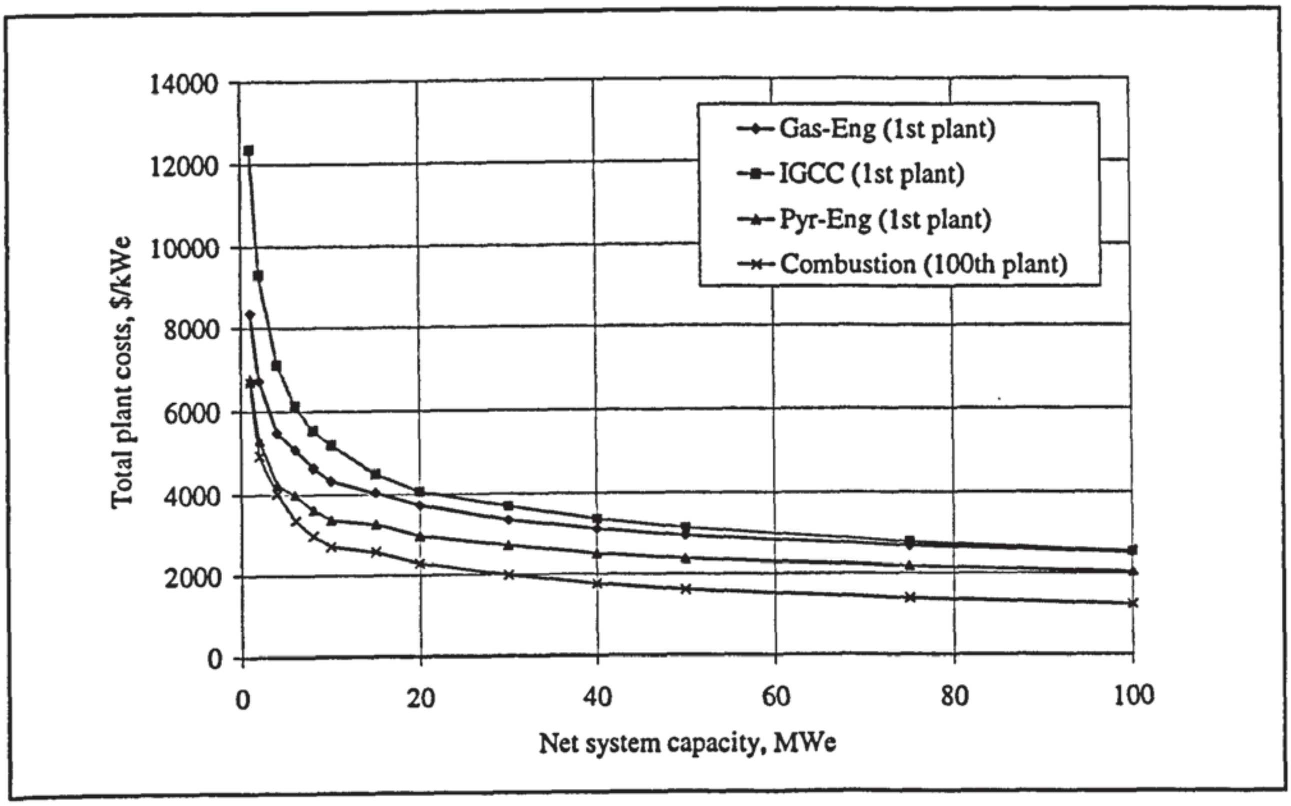


Figure 9.2 - Total plant costs, base case systems from 1-100 MW_e

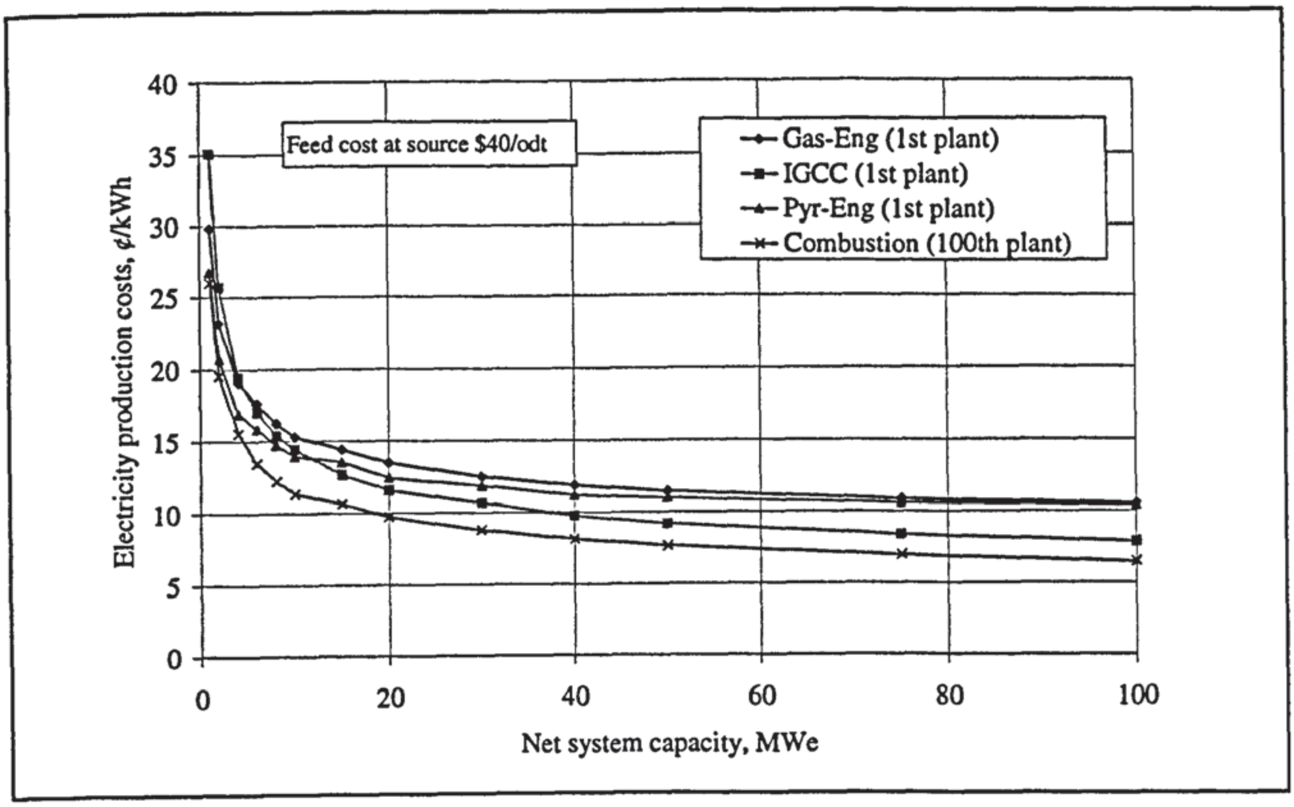


Figure 9.3 - Electricity production costs, base case systems from 1-100 MW_e

The combination of low efficiencies and high total plant costs at small scale results in extremely high electricity production costs that are unlikely to be viable under any circumstances. In view of this, the subsequent evaluations will not consider systems below 5 MW_e. Further evaluations will also be limited to capacities below 20 MW_e, as Figure 9.3 shows that the relative costs of the four systems are established by this point and no new information would be revealed. The 20 MW_e limit is also closer to the likely upper limit in the UK. Full sets of results for the base cases between 5 and 20 MW_e are presented in Appendix G.

9.3.2 System efficiencies at 5-20 MWe

The efficiency of the system dictates the amount of feedstock required to support a specific power output. This not only effects the total cost of feedstock, but it will have consequences throughout the system because all equipment sizes, capital costs and operating costs are also affected. Figure 9.4 presents the overall net system efficiency in the 5-20 MWe range. In this capacity range there is a consistent ranking among the systems: IGCC is the most efficient system, combustion the least. The engine based technologies offer moderate efficiencies with little improvement with scale.

Energy fluxes through the base case systems are compared in Table 9.3 for 5 MW_e and in Table 9.4 for 20 MW_e. The gain in energy between feed storage and entry to the conversion reactor is caused by drying, which increases the lower heating value of the feed. This energy must come from somewhere and it has been assumed in this work that the energy required for drying is available as waste heat. This has been validated in the Gas-Eng and Pyr-Eng systems, but brief examinations of the Combustion and IGCC systems have shown that there is a possibility that insufficient waste heat would be available. If there is an energy deficit the feed may have to be fed into the reactor with a higher moisture content or more energy will have to made available from other parts of the system. In both cases the efficiency of the system would suffer. Time limitations have prevented further analysis of this issue but it is imperative that the waste heat availability is re-examined so that the modules can be modified and this uncertainty is removed.

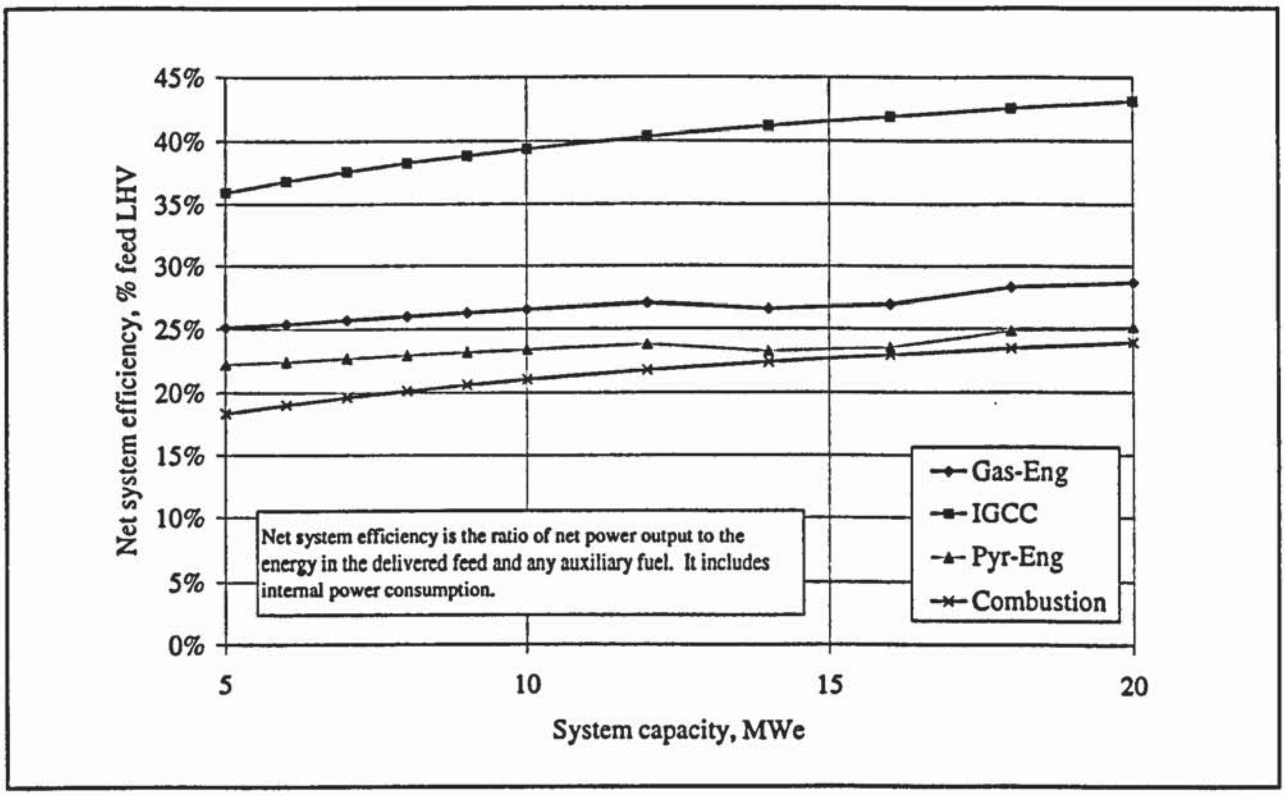


Figure 9.4 - Net system efficiencies, base case systems from 5-20 MWe

Another notable feature is that all the conversion efficiencies show improvements in efficiency with increasing capacity except in the case of fast pyrolysis. The efficiency improvements in combustion and gasification are caused by reduced reactor losses that result in more energy in the hot gases leaving the reactor. In fast pyrolysis there will also be more energy available as losses are reduced, but it is much less clear how this will effect the process. Since the process heat is provided externally the effect may be to decrease the amount of char and off-gas required to fuel the process. The excess char could be sold as a by-product, but revenue from char sales has not been included in this work because the market for the char is uncertain.

Table 9.3 - Energy fluxes and efficiencies in base cases at 5 MW_e

		Gas-Eng	IGCC	Pyr-Eng	Combustion
Energy inputs to system					
Feed delivered	GJ/y LHV	541856	395419	605115	782670
Auxiliary fuel	GJ/y LHV	22537	0	33767	0
Energy fluxes in system					
Feed from storage	GJ/y LHV	532478	388575	594642	769124
Feed at reactor	GJ/y LHV	596357	435191	671640	820835
Intermediate energy	GJ/y LHV	428200	374386	416463	683153
Gross electricity	GJ/y	152857	152912	169640	166373
Internal consumption	GJ/y	10772	10861	27541	22732
Net electricity	GJ/y	142084	142051	142099	143641
Efficiencies					
Conversion [a]	%LHV	71.8%	86.0%	62.0%	83.2%
Generation [b]	%LHV	33.9%	40.8%	37.7%	24.4%
Gross system [c]	%LHV	27.1%	38.7%	26.6%	21.3%
Internal losses [d]	%gross	7.0%	7.1%	16.2%	13.7%
Net system [e]	%LHV	25.2%	35.9%	22.2%	18.4%

a *Conversion efficiency = Intermediate energy / feed at reactor*
b *Generation efficiency = Gross electricity / (Intermediate energy + Auxiliary fuel)*
c *Gross system efficiency = Gross electricity / (Feed delivered + Auxiliary fuel)*
d *Internal losses = Net electricity / Gross electricity*
e *Net system efficiency = Net electricity / (Feed delivered + Auxiliary fuel)*

Table 9.4 - Energy fluxes and efficiencies in base cases at 20 MW_e

		Gas-Eng	IGCC	Pyr-Eng	Combustion
Energy inputs to system					
Feed delivered	GJ/y LHV	1905949	1321136	2154169	2385815
Auxiliary fuel	GJ/y LHV	80966	0	120209	0
Energy fluxes in system					
Feed from storage	GJ/y LHV	1872961	1298270	2116885	2344522
Feed at reactor	GJ/y LHV	2097652	1454018	2390994	2502152
Intermediate energy	GJ/y LHV	1538352	1266889	1482580	2131993
Gross electricity	GJ/y	610324	608404	670449	644422
Internal consumption	GJ/y	39921	38932	99524	73599
Net electricity	GJ/y	570403	569472	570925	570823
Efficiencies					
Conversion [a]	%LHV	73.3%	87.1%	62.0%	85.2%
Generation [b]	%LHV	37.7%	48.0%	41.8%	30.2%
Gross system [c]	%LHV	30.7%	46.1%	29.5%	27.0%
Internal losses [d]	%gross	6.5%	6.4%	14.8%	11.4%
Net system [e]	%LHV	28.7%	43.1%	25.1%	23.9%

a *Conversion efficiency = Intermediate energy / feed at reactor*
b *Generation efficiency = Gross electricity / (Intermediate energy + Auxiliary fuel)*
c *Gross system efficiency = Gross electricity / (Feed delivered + Auxiliary fuel)*
d *Internal losses = Net electricity / Gross electricity*
e *Net system efficiency = Net electricity / (Feed delivered + Auxiliary fuel)*

Finally with regard to Table 9.3 and Table 9.4, it can be seen that the internal energy consumption for the fast pyrolysis and combustion systems is considerably higher than the gasifier-based systems. The cause of the extra power consumption in fast pyrolysis is the grinder, and again this shows that it would be worth investigating ablative pyrolysis processes in further work since this does not require a powdered feedstock. The high power consumption of the Combustion system is due to the extra fluidising air required: given an excess air requirement for combustion of 1.47 and an air factor of 0.3 in gasification, the combustion is required to blow 5 times as much air through the fluid bed as the gasifier.

9.3.3 System capital costs at 5-20 MWe

Total plant costs are given in Figure 9.5 for capacities from 5-20 MW_e. In this capacity range the discontinuities in capital costs can be seen more clearly. These are caused by changing equipment requirements in the pretreatment steps and also engine replication in the engine generating modules.

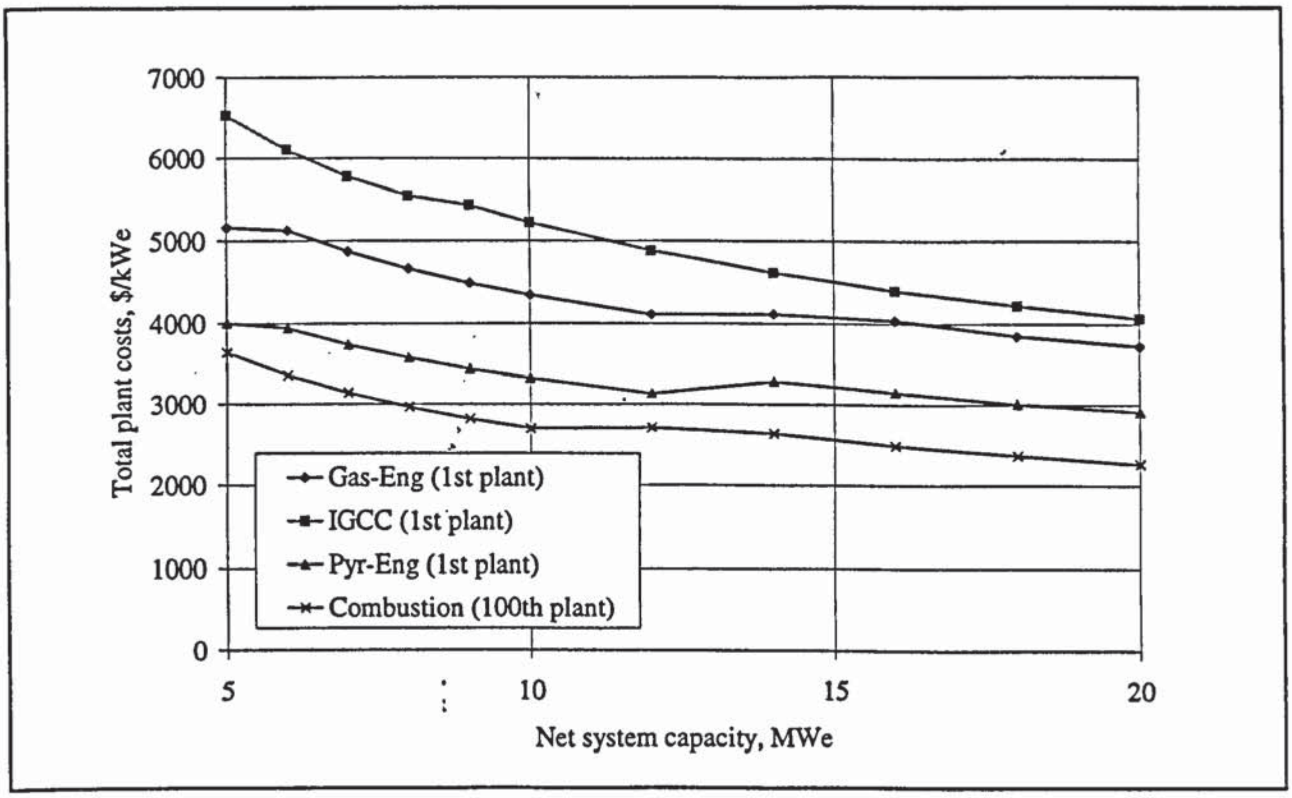


Figure 9.5 - Total plant costs, base case systems from 5-20 MW_e

The scale economies in the various parts of the system are shown in Figure 9.6. This chart and Table 9.5 that accompanies it show that scale economies are more significant in the IGCC and Combustion systems, and that in all systems most economies of scale are gained in the conversion modules.

One concern that is raised by Figure 9.5 and Figure 9.6 is the high capital cost of the Gas-Eng system, mostly attributable to the costs of the atmospheric gasifier and tar cracker. The tar cracker costs have been added using a very simple analysis of the basic gasifier costs and hence there is some uncertainty here that should be addressed by further analysis of these costs. Another useful exercise would be to examine the option of removing the cracker altogether. Tars could be reduced as much as possible by increasing the air factor in the gasifier to raise the temperature and encourage thermal cracking. Raw gas tar levels would still be higher than the levels with a second reactor, and the amount of tars removed by the wet scrubber would be greater. The removal of the tar cracker would therefore reduce capital costs but there are implications on efficiency (by changing the air factor and removing more tars) and production costs (the cost of extra feed and a more contaminated waste water stream).

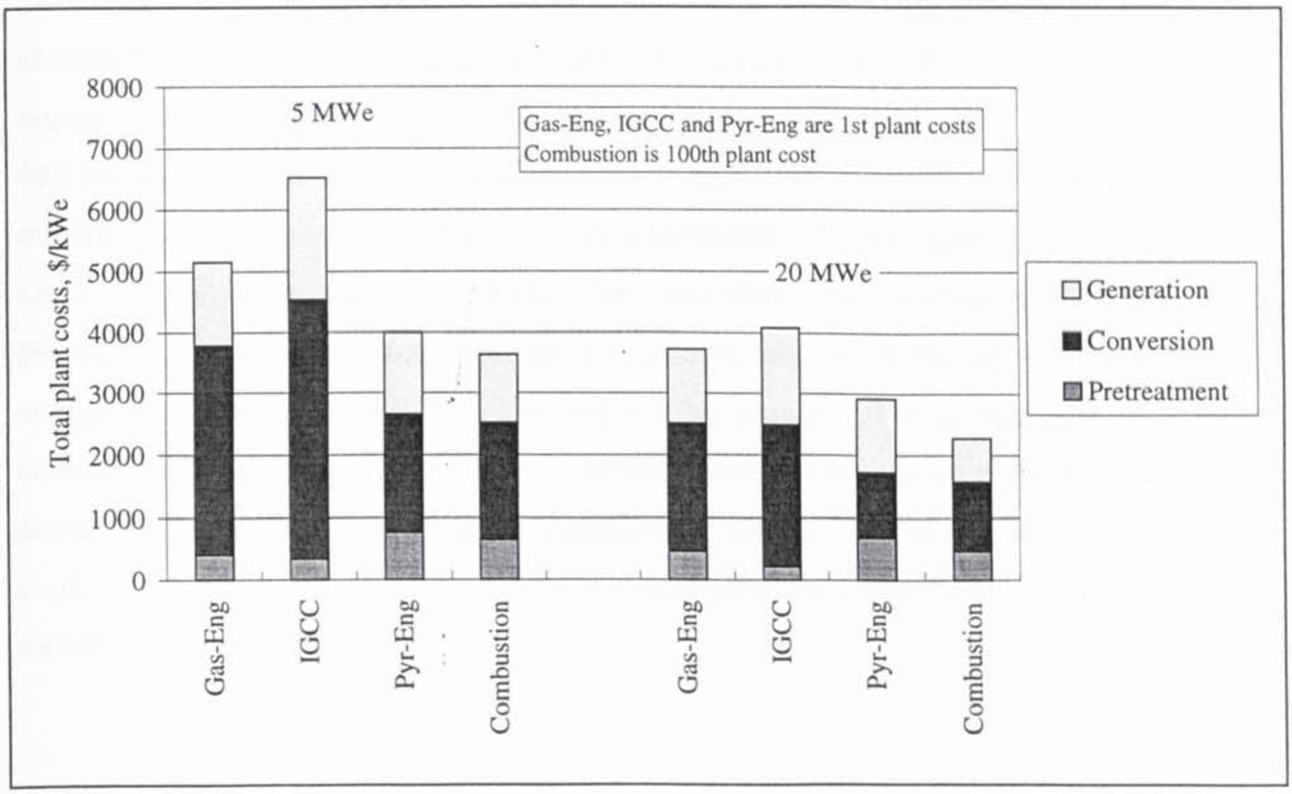


Figure 9.6 - Breakdown of total plant costs, base case systems at 5 and 20 MW_e

Table 9.5 - Breakdown of total plant costs for base cases at 5 and 20 MW_e

	Gas-Eng		IGCC		Pyr-Eng		Combustion	
5 MWe								
Pretreatment	410	8%	348	5%	776	19%	656	18%
Conversion	3375	65%	4195	64%	1879	47%	1859	51%
Generation	1371	27%	1986	30%	1348	34%	1126	31%
TOTAL	5156		6529		4003		3641	
20 MWe								
Pretreatment	465	12%	222	5%	696	24%	474	21%
Conversion	2032	55%	2264	56%	1028	35%	1097	48%
Generation	1227	33%	1584	39%	1193	41%	702	31%
TOTAL	3724		4070		2918		2273	
% Reduction								
Pretreatment	-13%		36%		10%		28%	
Conversion	40%		46%		45%		41%	
Generation	11%		20%		11%		38%	
TOTAL	28%		38%		27%		38%	

One interesting feature of Table 9.5 is the variation in the pretreatment system costs. The differences between the systems at 5 MW_e are easy to understand: they generally reduce with increasing efficiency due to the reduction in the amount of feed that must be processed. The exception is the Pyr-Eng system that carries the extra cost of grinding. At 20 MW_e there appears to by an anomaly: the Gas-Eng pretreatment costs have increased from 410 $/kW$_e$ to 465 $/kW$_e$. There is also considerable variation in the scale economies of the pretreatment modules in each system. Figure 9.7 presents a breakdown of the pretreatment total plant costs against capacity to explain these points. This chart shows the discontinuities already seen in Figure 9.5, and it is the changes to a more expensive feed reception step and the use of pre-storage screening at 20 MW_e that have caused the increase. The pretreatment costs at any scale are highly dependent on the amount of feed processed and therefore the equipment used, and discontinuities can produce some unusual scale effects. This is one of the advantages of explicitly modelling all of the steps in feed pretreatment and a significant novel feature of the model.

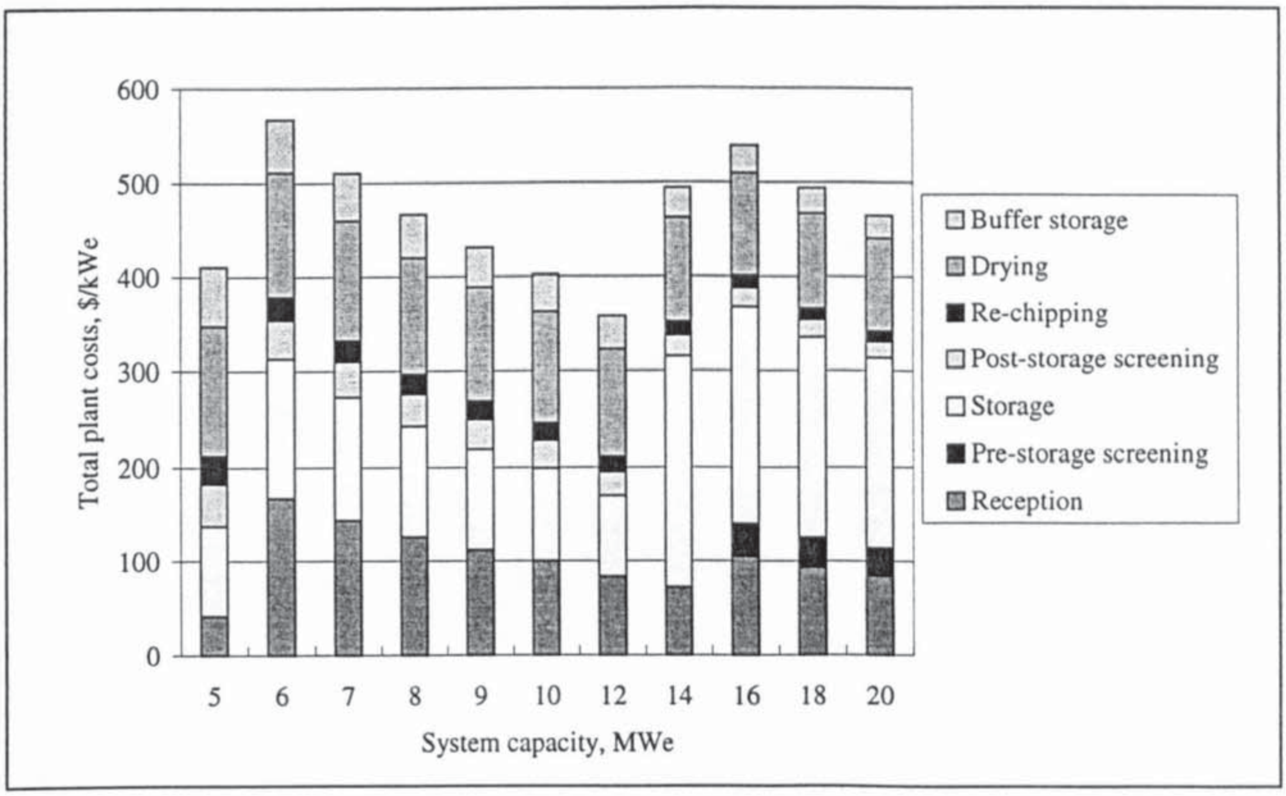


Figure 9.7 - Breakdown of pretreatment total plant costs, Gas-Eng system

If only the total plant costs were considered, a developer would chose Combustion. Conversely, an examination of Figure 9.4 in isolation would suggest the selection of IGCC at any scale. However, a comparison of Figure 9.4 and Figure 9.5 shows that high efficiency is only achieved through higher investment. Ultimately, the interaction between system efficiency and total plant costs can be measured in terms of the electricity production cost but it may sometimes be useful to quickly ascertain which systems offer the best compromise between system efficiency and capital costs. For this reason the "specific efficiency cost" (SEC) is defined. SEC is the ratio of the specific total plant costs to the net system efficiency. The systems that use capital most effectively in converting feedstock to electricity will have a low SEC. Although the SEC only provides a basic tool for comparing systems it is a far more effective measure than the efficiency or capital cost in isolation. Figure 9.8 compares SEC for the base case systems. SEC is too coarse a measure to distinguish between Combustion, IGCC and Pyr-Eng at 5-12 MW_e, but it can be used to show that the Gas-Eng system is unlikely to be competitive at any scale and that the Pyr-Eng system is not likely to be competitive at capacities above 12 MW_e.

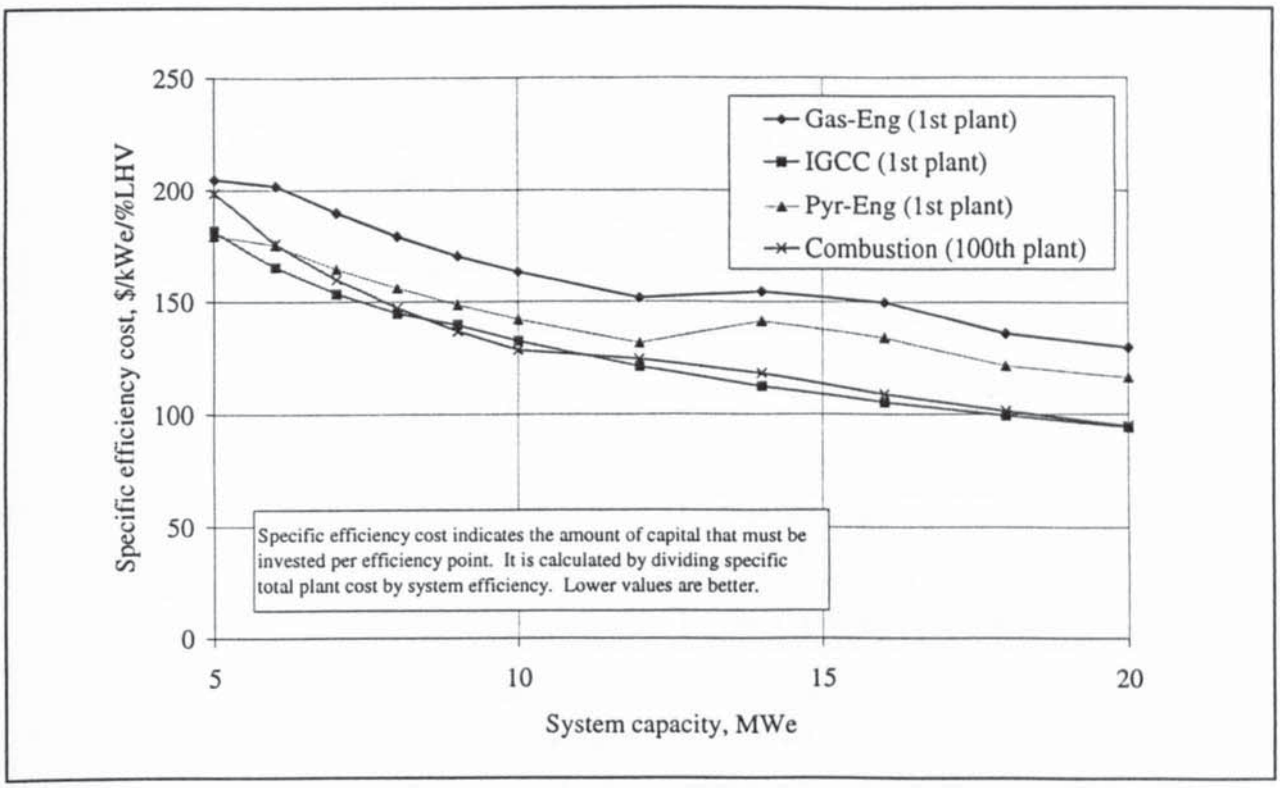


Figure 9.8 - Specific efficiency costs for base cases at 5 and 20 MW$_e$

9.3.4 Electricity production costs at 5-20 MWe

Figure 9.9 gives the electricity production costs for the four base cases from 5-20 MW$_e$. The detail of the production costs is presented in Table 9.6 for the 5 and 20 MW$_e$ limits. The graph shows how the ranking of the four systems change with capacity. The combustion system is by far the most economic in this comparison at all scales. Of the 1st plant systems, Gas-Eng is the least viable at all capacities. The Pyr-Eng system has the lowest production costs at small scale while the IGCC system is better above 12 MW$_e$.

Table 9.6 breaks the production costs down into the feedstock cost and production costs for each module. It can immediately be seen that the transport costs are relatively insignificant, even at 20 MW$_e$, and this confirms the conclusions of Marrison [185] and McIlveen-Wright [186].

Another observation is that the feedstock cost is a much smaller proportion of the IGCC system costs than in the other systems, and that in Combustion the converse is true. This is a result of the efficiencies in each system, and it can be concluded that the Combustion system will be far more sensitive to changes in feedstock cost than the IGCC system.

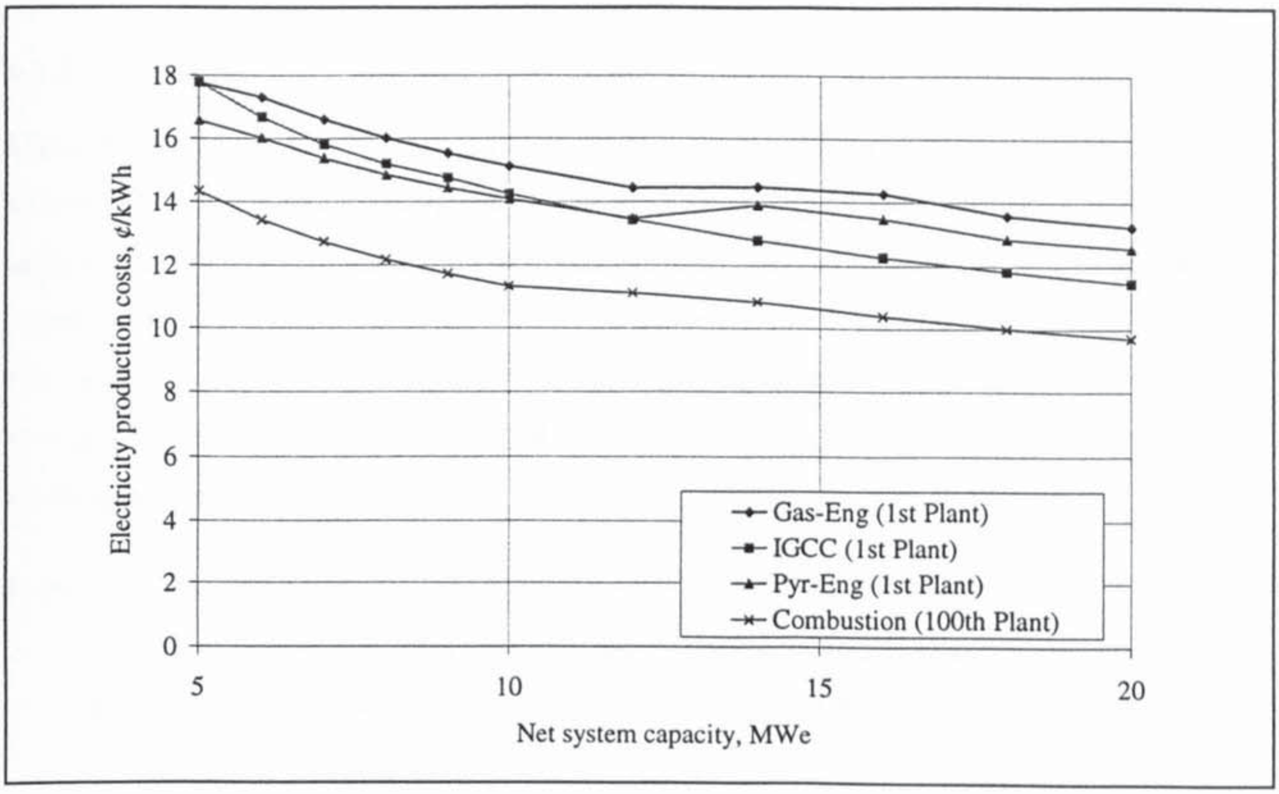


Figure 9.9 - Electricity production costs for base case systems from 5 to 20 MW_e

Table 9.6 - Breakdown of electricity production costs for base cases at 5 and 20 MW_e

System	**Gas-Eng**		**IGCC**		**Pyr-Eng**		**Combustion**	
Status	1st plant		1st plant		1st plant		100th plant	
Production costs at 5 MW_e, $k/y								
Feed before transport	1287	18%	939	13%	1437	22%	1858	32%
Feed transport	217	3%	153	2%	246	4%	328	6%
Feed pretreatment	666	9%	617	9%	1016	16%	906	16%
Feed conversion	3020	43%	3550	51%	1803	28%	1576	28%
Electricity generation	1702	24%	1633	23%	1912	29%	925	16%
Grid connection	129	2%	129	2%	129	2%	129	2%
TOTAL	7020		7020		6542		5722	
Electricity cost, ¢/kWh	17.79		17.79		16.57		14.34	
Production costs at 20 MW_e, $k/y								
Feed before transport	4529	22%	3137	17%	5115	26%	5665	37%
Feed transport	918	4%	598	3%	1060	5%	1197	8%
Feed pretreatment	1915	9%	1021	6%	2726	14%	1943	13%
Feed conversion	7172	34%	7563	42%	3846	19%	3656	24%
Electricity generation	6191	29%	5544	31%	6882	35%	2666	17%
Grid connection	271	1%	271	1%	271	1%	271	2%
TOTAL	20996		18134		19901		15398	
Electricity cost, ¢/kWh	13.24		11.46		12.55		9.71	

9.3.5 Summary

Many of the observations made in this initial evaluation are obvious and have been seen before. They do however demonstrate the models' ability to produce continuous data that adjusts the system performance, equipment requirements and capital costs as the system capacity changes. The breakdown of costs by modules is also a useful way of showing where the system costs are concentrated. If costs are concentrated in a module with a high uncertainty (for example the pressurised gasification module) then there is a greater risk to the developer (see Section 9.9).

It has been shown that:

- the lowest production costs are produced by the Combustion system;
- at low capacities (less than 12 MW_e), the electricity production cost of the Pyr-Eng system comes closest to Combustion;
- at high capacities (above 12 MW_e), the nearest electricity production cost to Combustion is given by the IGCC system;
- the mediocre cost and performance of the Gas-Eng system does not give it any advantage at any scale;
- the pretreatment equipment costs at any capacity are highly dependent on the amount of feed to be handled and processed; and
- system efficiencies are only increased by higher investment. A new ratio has been defined, the specific efficiency cost, that rapidly indicates the cost-effectiveness of each system by showing how much capital is invested per percentage point of system efficiency.

9.4 LEARNING EFFECTS

Learning effects were introduced in Section 4.7 and the concepts and relationships used here are defined in that discussion. This section will apply learning effects to the base case systems to assess their impact.

Two analyses of learning effects are presented. The first selectively applies a learning factor of 20% to the gasification and fast pyrolysis conversion modules only. The second analysis assumes that all the costs in the novel systems are 1st plant costs and applies a learning factor to the whole system accordingly. This is the approach used by Elliott and Booth [24]. The collective application of learning factors is optimistic given the fact that only certain parts of the system are novel, other parts such as the pretreatment steps are based on established equipment with established costs. It is felt that this selective application of the learning factor is more realistic.

The effects of selective learning effects on total plant costs and electricity production costs are shown in Figure 9.10 and Figure 9.11 for the 1st, 10th and 100th plant in each system at 5 MW_e. It can be seen that learning effects applied to the conversion modules alone are sufficient to bring the electricity production costs of all the systems much closer to current Combustion costs by the 10th plant. It is also observed that the effect is more marked in the capital intensive systems, such that the IGCC is transformed from the option with the most expensive production costs to the system that produces the least expensive electricity. By the 100th plant costs, all the systems show a significant advantage over the Combustion system, although this is an unfair comparison because some improvement in the Combustion system could also be expected during this period.

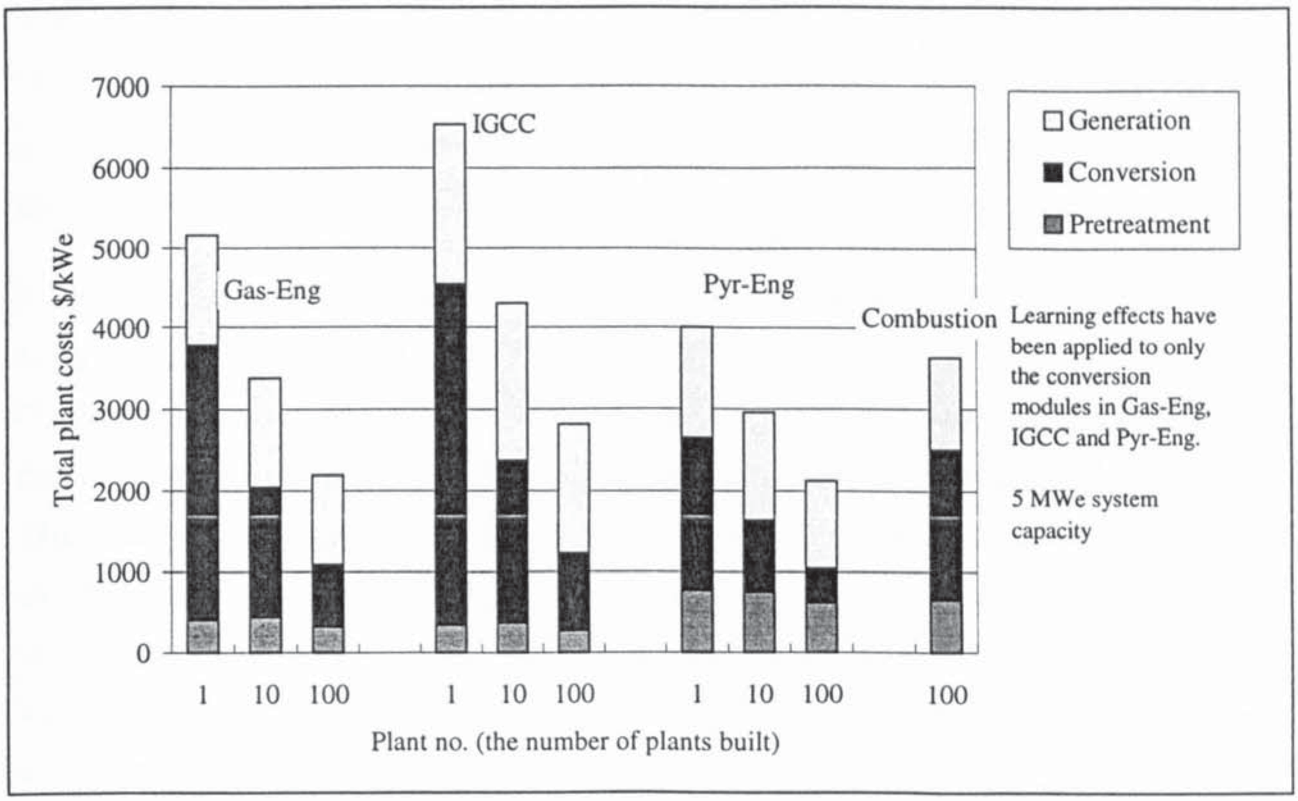


Figure 9.10 - Total plant costs with learning effect applied selectively

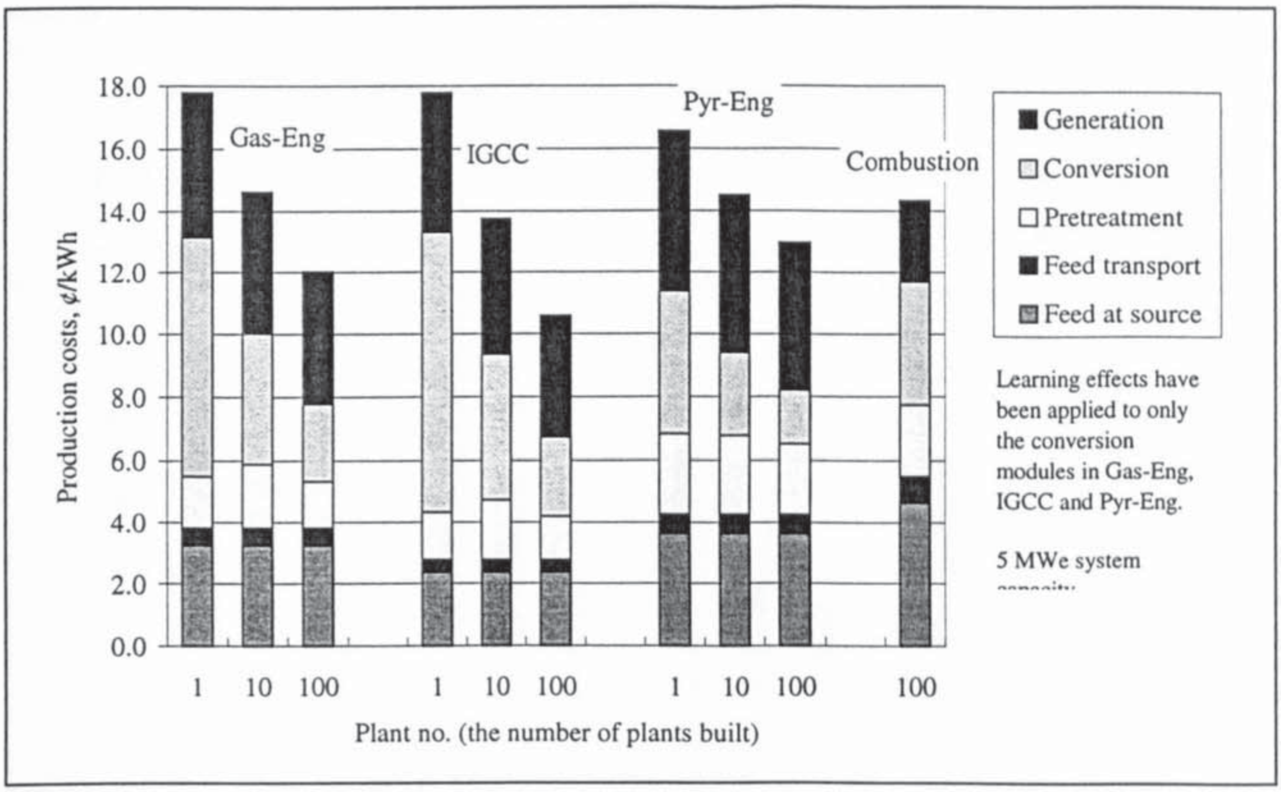


Figure 9.11 - Production costs with learning effect applied selectively

The collective application of learning factors is demonstrated in Figure 9.12 and Figure 9.13. The large savings gained by the 10th plant are questionable and by the 100th plant the costs are unbelievably low. It is doubtful where such improvements would be possible in reality and the application of learning factors on a selective basis is a far more likely scenario.

In summary, it has been shown that if learning effects are taken into account it is more than likely that the Gas-Eng, IGCC, and Pyr-Eng systems will become more economic than the Combustion system. The problem is how the initial investment can be found to construct the first few, highly expensive systems. This is why the Brazilian GEF project and the Targeted THERMIE projects described in Section 2.6.5 are so important. In these projects public money is used to subsidise the high initial capital costs and the industry in general should benefit from the experience gained. If the three THERMIE projects, the GEF project and the Värnamo plant are all successfully implemented then the number of systems will already have doubled twice and the costs should fall accordingly.

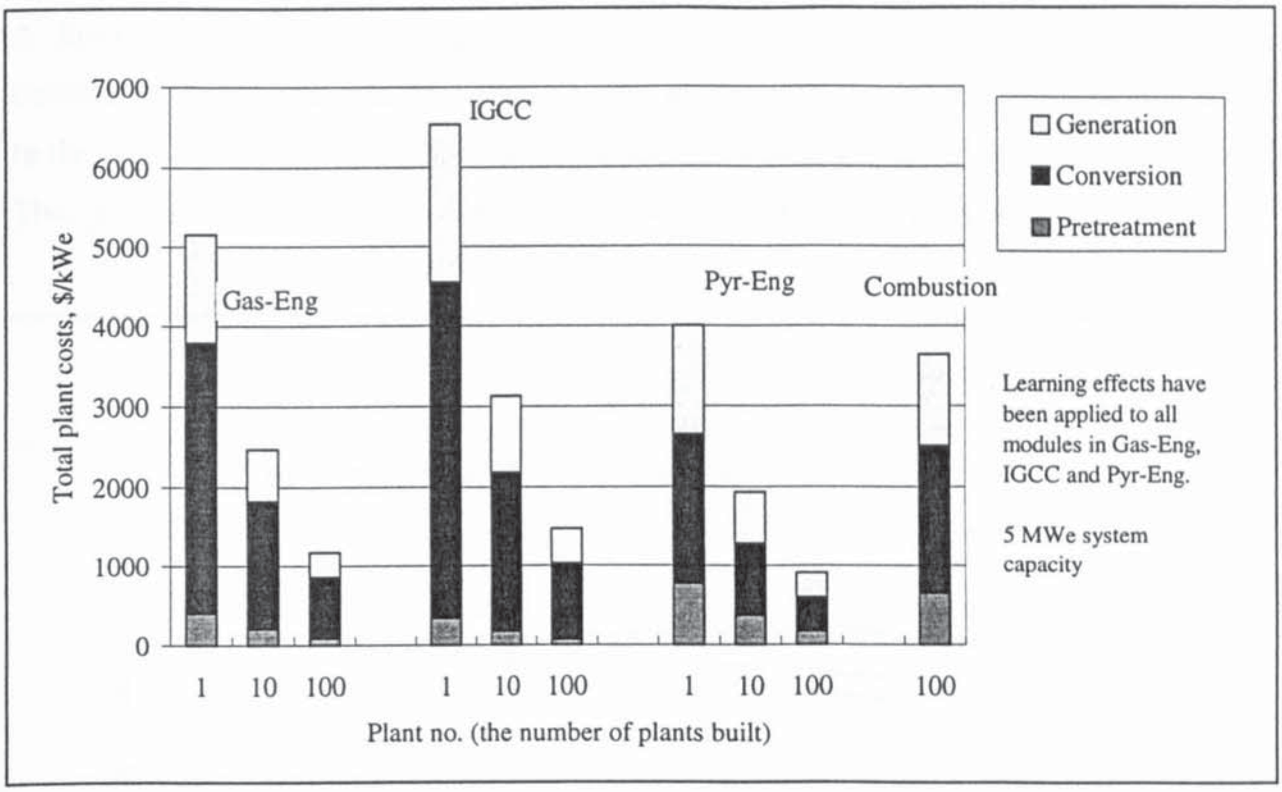


Figure 9.12 - Total plant costs with learning applied collectively

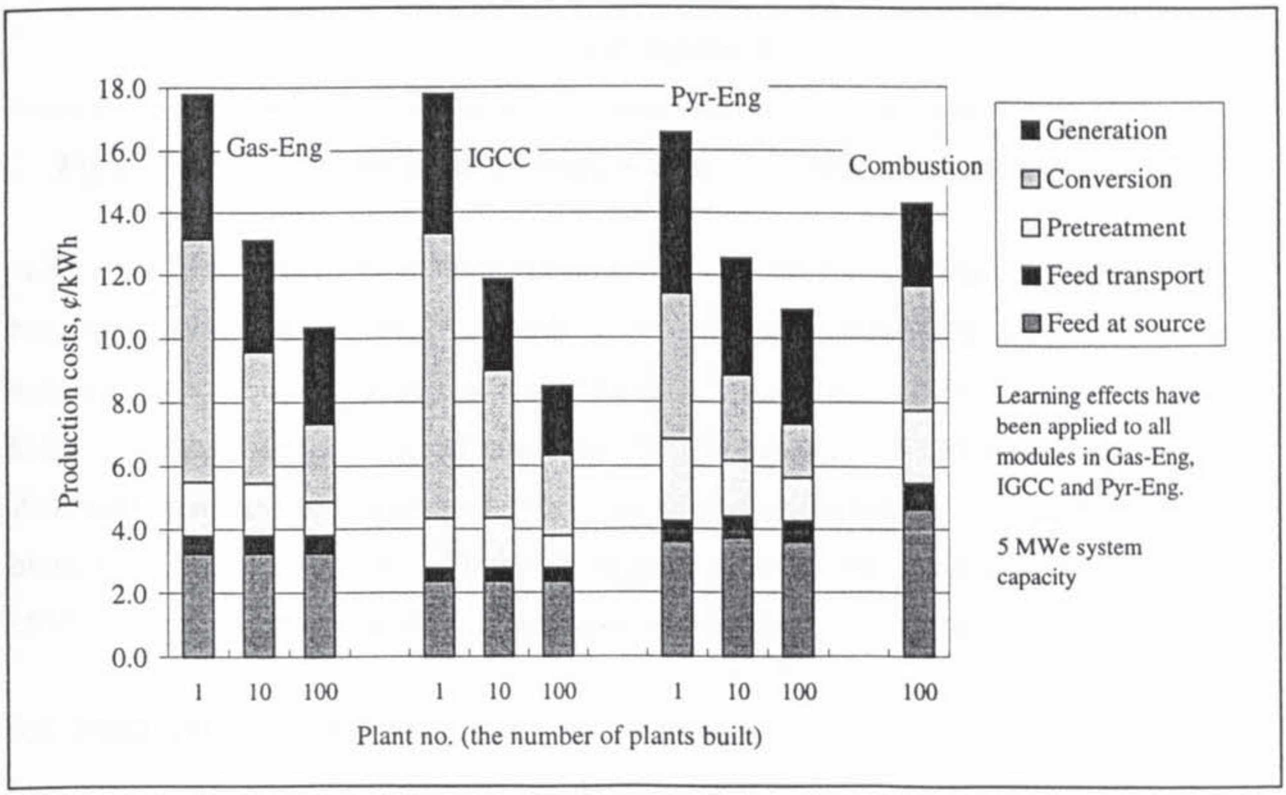


Figure 9.13 - Production costs with learning applied collectively

A further graph showing learning effects is presented in Figure 9.14. This shows the production costs for the four base cases after the selective application of a 20% learning factor to the conversion modules in the Gas-Eng, IGCC and Pyr-Eng systems as a function of scale. This graph should be compared with the analysis using current, 1st plant costs.

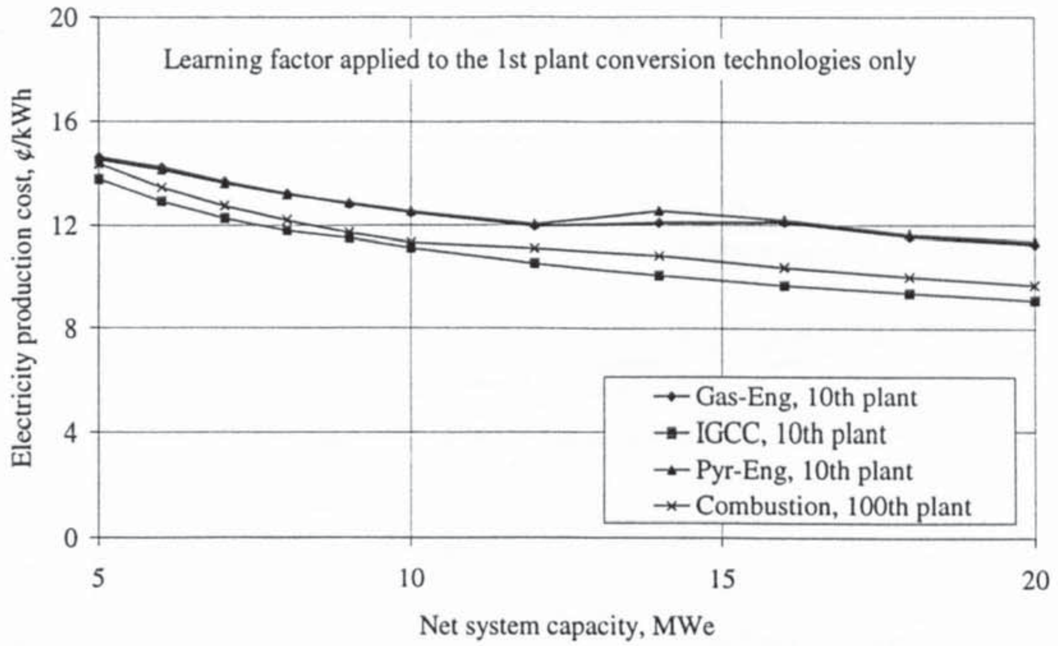


Figure 9.14 - Electricity production costs after selective learning effects, 5-20 MW_e

A final consideration about learning effects is that the potential for efficiency improvements has not been considered. This is far harder concept to apply consistently and is more suited to a case-by-case analysis. It is however reasonable to assume that the financial benefits of learning through reduced capital costs will be compounded by efficiency improvements. Although these may be less dramatic, they may never-the-less help to increase the viability of biomass to electricity systems. Further work could examine the potential in each system for efficiency improvements and their effect on the comparative system costs.

9.5 FEEDSTOCK COSTS

This analysis evaluates the effects on the systems of changing feedstock costs, ranging from feed available at zero cost through to wood chips costing \$80/odt. Given the differences in efficiency between the systems, the cost of the feedstock should have an influence on the selection of the system. Figure 9.15 presents the effect of changing feedstock costs at 5 MW_e, using base case system characteristics. The chart shows that at low feedstock costs, there is no advantage to paying for high efficiency. Thus the low efficiency but low capital cost combustion system increases its lead over the other technologies. This is interesting because it shows that the development of the advanced technologies would not be accelerated by the widespread availability of a low-cost feedstock, since this would encourage the construction of more combustion systems. As feed cost increases, efficiency becomes more important and the IGCC system begins to show benefits.

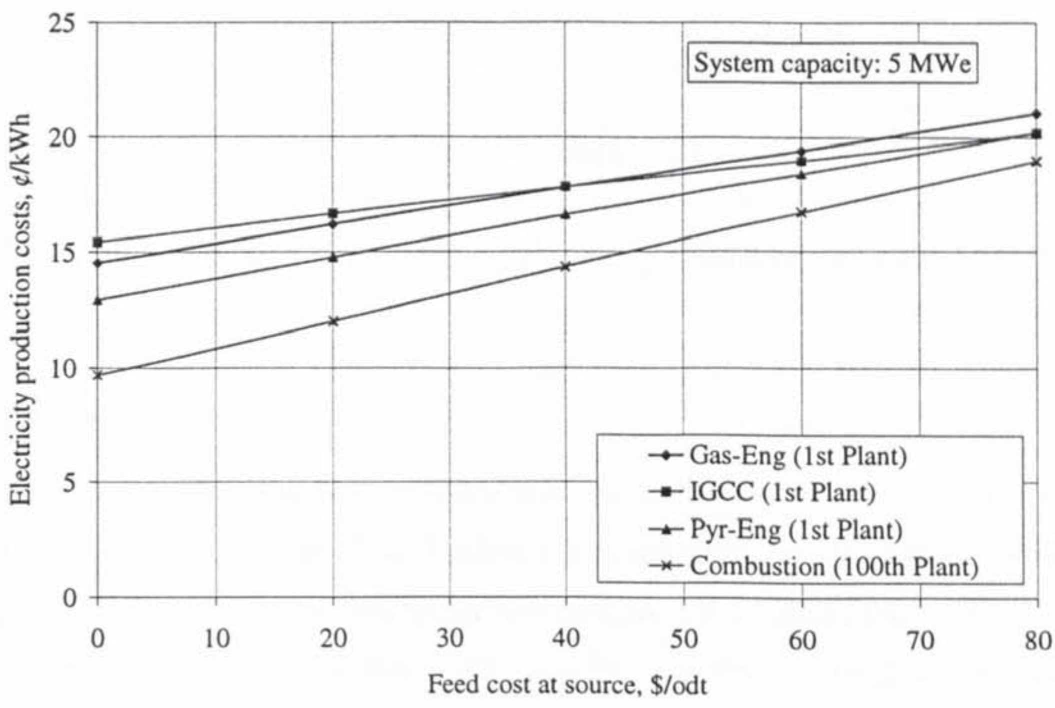


Figure 9.15 - Impact of feed cost on electricity production price at 5 MW_e

The same analysis is carried out at 20 MW_e, with the results shown in Figure 9.16. A comparison in Figure 9.15 reveals that the costs are less sensitive to feed cost at 20 MW_e than they are at 5 MW_e. This may seem surprising, given that it was demonstrated in Table 9.6 that the proportion of production costs attributed to feedcost rises with scale as a result of reducing capital-related expenditure. It is concluded that the extra efficiency achieved at 20 MW_e reduces the amount of feed required so that its influence on cost is reduced. Thus it is more important to ensure a low, fixed feed cost in the small scale systems than at the large scale.

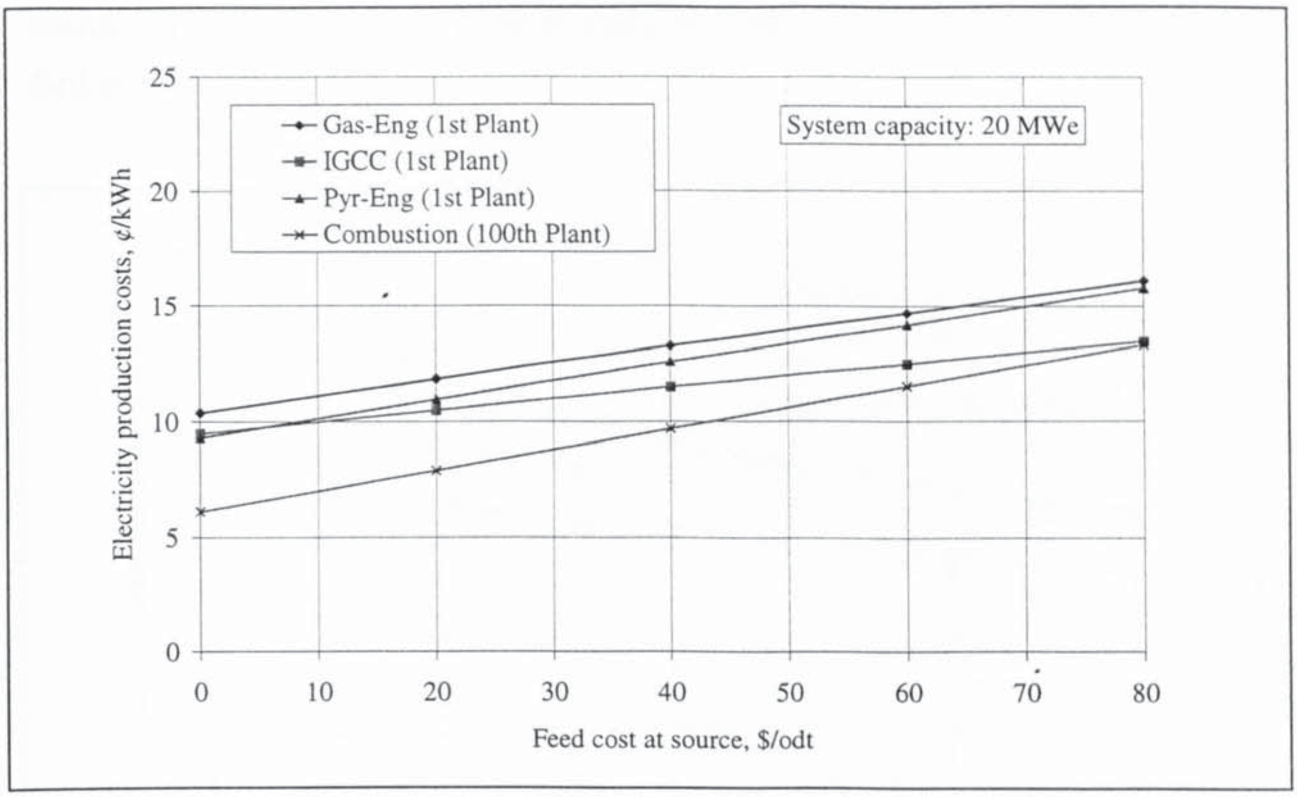


Figure 9.16 - Impact of feed cost on electricity production price at 20 MW$_e$

9.6 AVAILABILITY

All of the above comparison have assumed that the availability in each system is the same, 90% or 7884 h/y of production. This implies a high reliability for all systems. While a 90% or higher availability could be reasonably assumed for Combustion, the novel technologies and their application in the Gas-Eng, IGGC and Pyr-Eng systems suggest that unscheduled shutdowns are more likely to reduce availability.

The effect of changing system availability is shown in Figure 9.17. It can be seen that the IGCC system is most affected by scale, by virtue of its high capital cost. Given the high uncertainty associated with the reliability of the hot gas filters in this system there is a high risk that availability will be much less than 90% until hot gas cleaning is proven. The Gas-Eng and Pyr-Eng systems are less effected by availability because of their lower capital cost. They also have the ability to switch to full diesel operation if necessary and so their would be far less risk in these systems of interrupting electricity supply. The financial implications of full diesel operation have not been tested here but the impact in situations where there is a contractual obligation to meet electricity production targets could be substantial and worthy of further investigation.

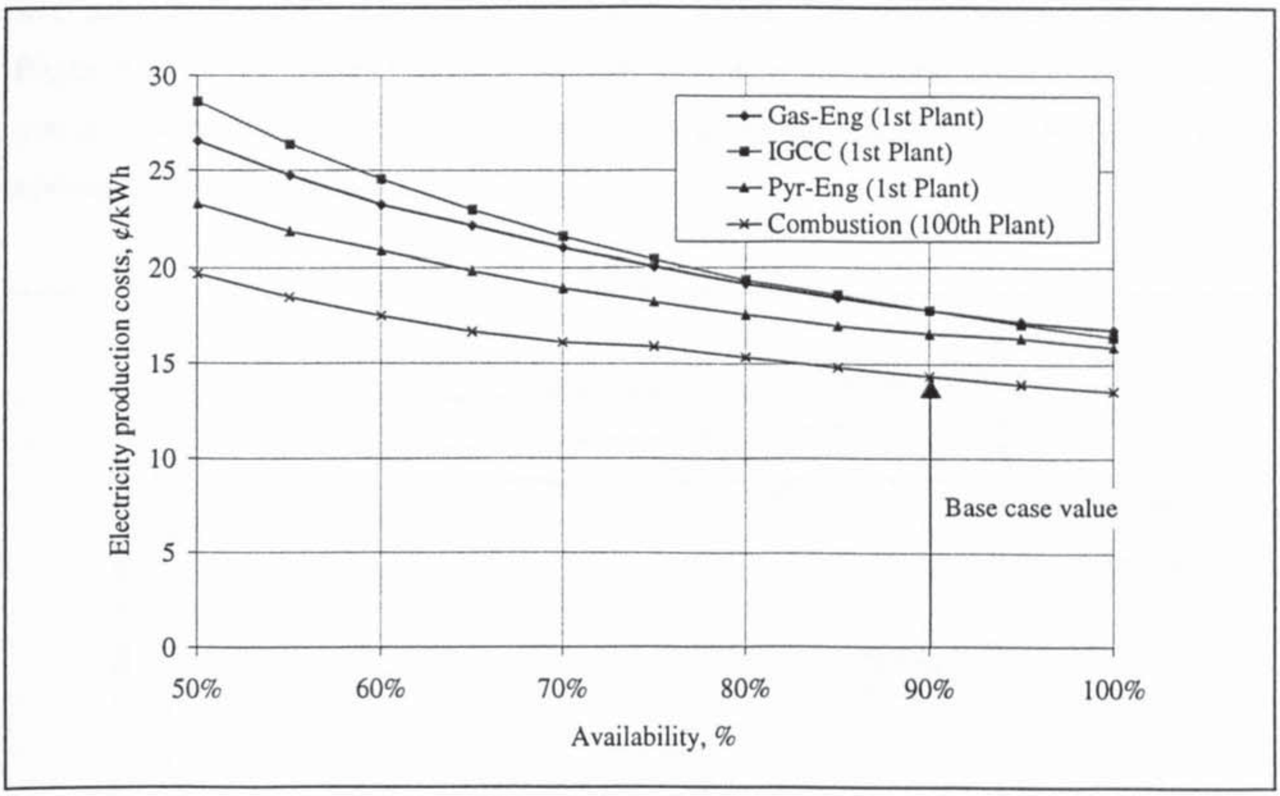


Figure 9.17 - Electricity production cost variations with availability

9.7 DE-COUPLED SYSTEMS

9.7.1 Introduction

The concept of a de-coupled system was introduced in Section 2.7.6. De-coupling is the separation of the conversion and generation system in time and/or space. It is an option that is only available to the fast pyrolysis system, and this section will examine situations that might allow de-coupling to be used to the advantage of fast pyrolysis.

9.7.2 Remote sources

One of the problems with a biomass feedstock is the cost of transporting it to the conversion facility. Although feed transport costs have been shown to be insignificant under current conditions, where feedstocks are available some distance away from a suitable generation site they can become more important. Under these circumstances a close-coupled system located at the generation site would require the transport of the feed over long distances, increasing the delivered cost of the feed. A de-coupled fast pyrolysis unit could operate differently: the pyrolysis unit could be located at the feed source with transportation of the pyrolysis liquid to the generating site. Given the increased heating of the pyrolysis liquid (14 GJ/t compared with 8.4 GJ/t) it is more economic to transport the liquid. This application is demonstrated in Figure 9.18. It can be seen that there is a slight advantage for the de-coupled system at higher transport distances, but the impact is not great given the generally low sensitivity in all systems to feed transportation costs.

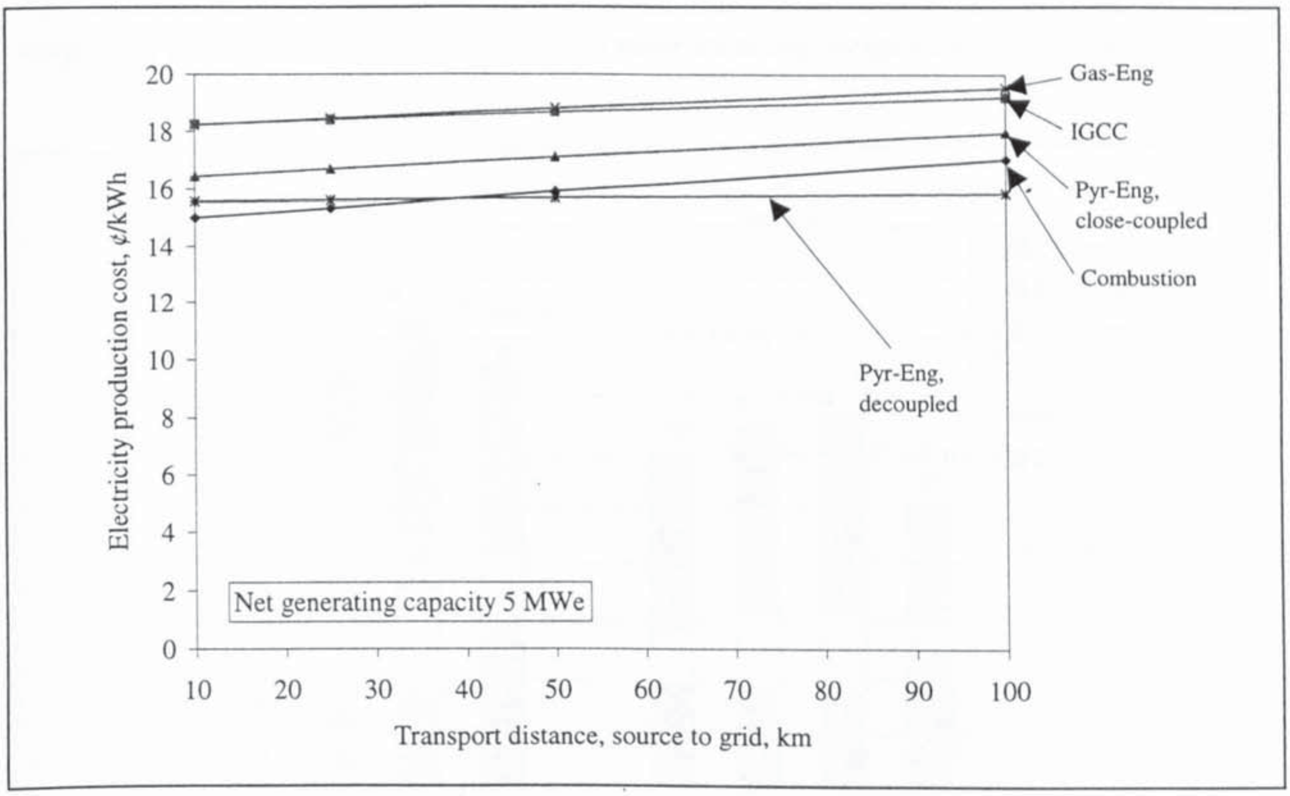


Figure 9.18 - De-coupling where feed source and electricity user are remote

9.7.3 De-coupled Systems - Multiple Pyrolysis Sites

This was another case that was originally examined as a means of reducing feed transport costs. The idea is that where feed is distributed over large areas feed transport costs can be reduced by setting up several small fast pyrolysis units to supply pyrolysis liquid to a central generator. The disadvantage to the system is that overall total plant costs would be increased by the loss of scale economies when building several smaller units rather than a single large unit. An analysis of this concept is presented in Figure 9.19. Two capacities are examined, one producing 5 MW_e at a central generator, the other producing 20 MW_e. In both cases the base case close-coupled Combustion system costs are shown for comparison. In each case three Pyr-Eng configurations are shown, the first a single, close-coupled system, the others de-coupled systems using either two or three fast pyrolysis units to supply the central generator. It can be seen that the de-coupled systems produce more expensive electricity than the close-coupled systems.

One aspect of this system that is not evaluated and that might yet prove to be to the benefit of this configuration is that the risk of problems during pyrolysis is spread among several conversion sites. As a result the supply of pyrolysis liquid could be more reliable and this may make a multiple conversion site system more attractive, despite the higher costs.

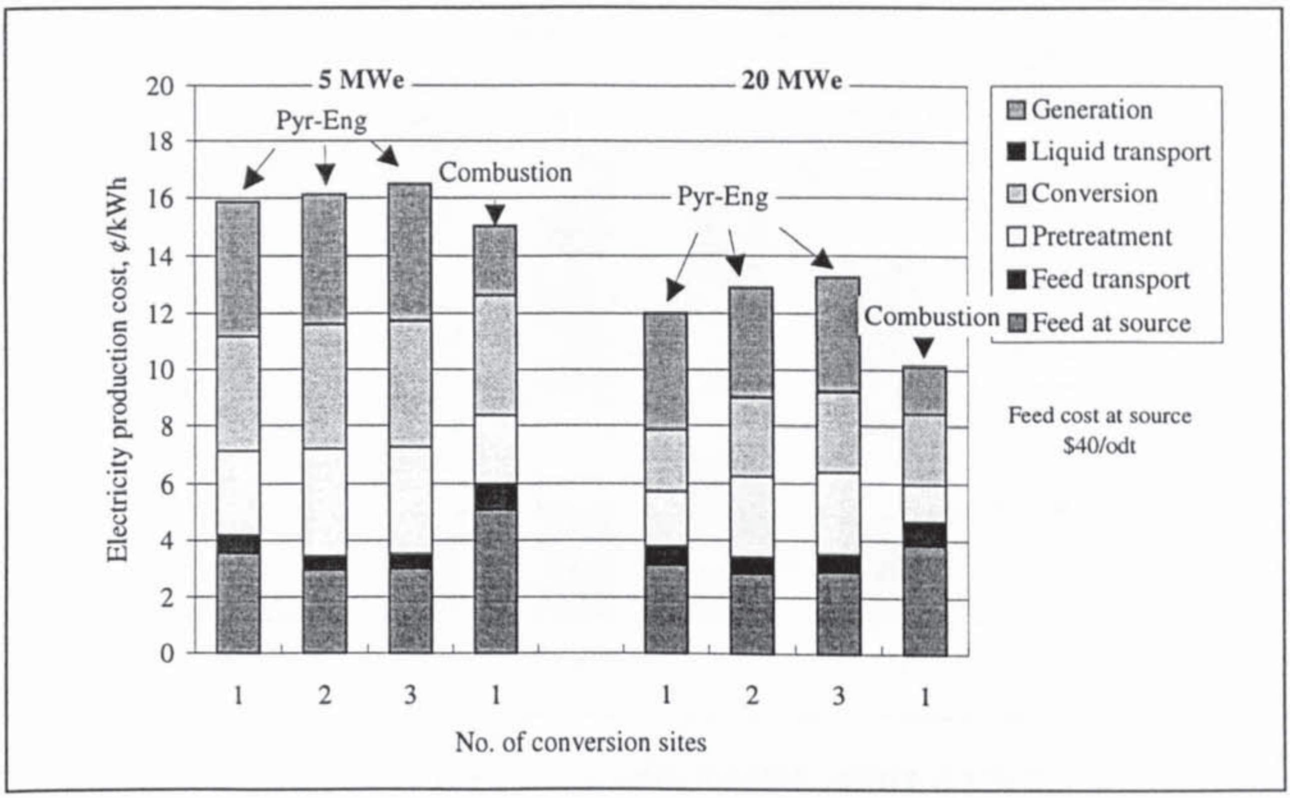


Figure 9.19 - De-coupling using multiple conversion sites

9.7.4 De-coupled Systems - Multiple Generators

It has already been shown in Table 9.5 that the pyrolysis module is the most scale-sensitive part of the fast pyrolysis system. It would therefore be logical in cases where electricity was required at several remote sites to use a large, central pyrolysis plant to supply the generators. Figure 9.20 presents the results of an analysis to show how this system configuration could help the comparative economics of fast pyrolysis. The analysis assumes that 5 MW_e is required at four generating sites within the feed supply area. If close-coupled systems are used then this system will require four complete units comprising pretreatment, conversion and generation stages. This configuration is modelled for the four base case systems by using 4 close-coupled sites and a total electricity output of 20 MW_e. An alternative de-coupled system was modelled using a single fast pyrolysis site that supplies pyrolysis liquid to four 5 MW_e generators. The results in Figure 9.20 are very promising for the de-coupled system. There is a potential risk in the use of one pyrolysis plant to supply all the engines but this can be alleviated by buffer storage of the pyrolysis liquid. This option represents the best opportunity for implementing fast pyrolysis using current costs.

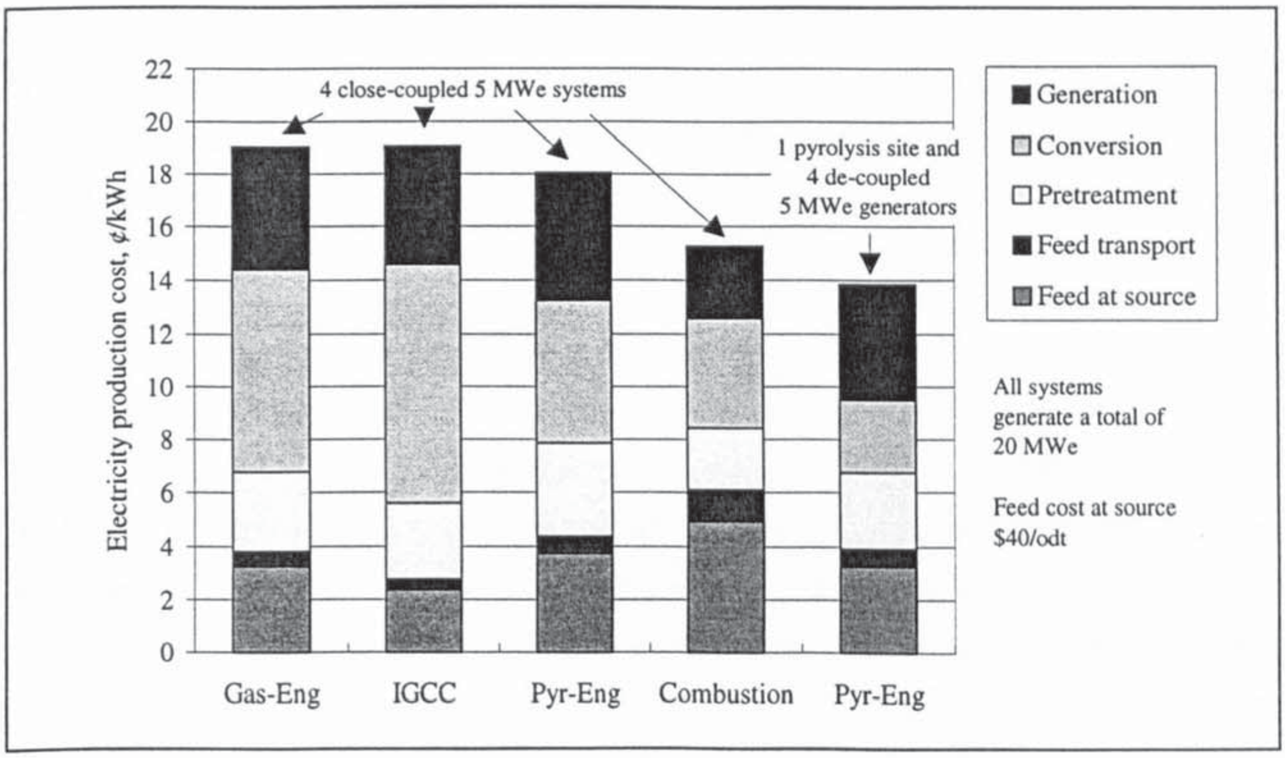


Figure 9.20 - De-coupling using multiple generation sites

9.8 MARKET OPPORTUNITIES

Up to now the systems have been evaluated purely by comparing their relative production costs. This section compares the systems to the absolute reference of the existing electricity market, seeking opportunities for biomass.

Before any comparison can be made, the electricity production costs predicted by the models must be converted to a price that includes a profit element. One of the simplest ways of doing this is to apply is a 10% return on investment. Thus electricity price is calculated by adding

10% of the total plant cost to the annual production costs and dividing the total by the electricity output. There are other more detailed ways of calculating the electricity price based on cash flow analyses but these are not required for this discussion. The electricity price for the base case systems over 5-20 MW_e are shown in Figure 9.21.

For comparison, three reference costs are given. The lowest reference is the UK pool price, which is the average price that utilities receive for electricity supplied to the grid. It must be conceded that biomass will never be able to compete with the costs of the large conventional power stations that are able to deliver electricity to the grid at only 4 ¢/kWh (2.5 p/kWh) and still be profitable. It would require huge increases in fossil fuel prices or legislation such as high carbon taxes to make biomass competitive in this market.

The highest reference shown is the price that will be paid for electricity generated under the 3rd NFFO order using wood IGCC systems. Here it can be seen that the Combustion system could already match such an elevated price at scales above around 20 MW_e. The other systems will require further development to reduce costs before they can be economic, but viability by this measure is a likely prospect.

Several countries including Germany, Denmark and Italy have legislation like the NFFO scheme in place that pays premium rates for electricity generated from renewable sources including biomass [303]. Such schemes are driven by a combination of the pressures described in Chapter 1. There is always the possibility, unlikely at present, that public and political support for subsidising renewable energy could wane due to lack of interest and budget restrictions. Thus although such schemes are valuable opportunities in the short term, it would be far better to find markets that could be penetrated without any subsidies.

One opportunity is indicated by reference to the mean European electricity price for large consumers [248], shown in Figure 9.21. This is higher than the pool price because it includes grid distribution costs and profits for the utility. There are national and regional variations in this price, as shown in the literature [248, 303], and prices can be as high as 13.5 ¢/kWh in Europe. There may be opportunities where the utility electricity price is high for a developer to undercut the utility price and sell electricity produced from biomass directly to a large consumer. Even at 13.5 ¢/kWh, the current systems are uncompetitive and the first opportunities in this area may be where a feed is available at low or negligible cost.

One possibility in the UK is presented by the introduction of a new landfill tax of around £7/tonne of waste. Businesses that must pay to dispose of their waste could use it instead for electricity generation for their own consumption or possibly for export to neighbours. This could be a particularly attractive opportunity for de-coupled pyrolysis ventures, where the waste from several partners could be processed by a single pyrolysis plant and fuel could then be distributed to each partner for electricity generation at their site.

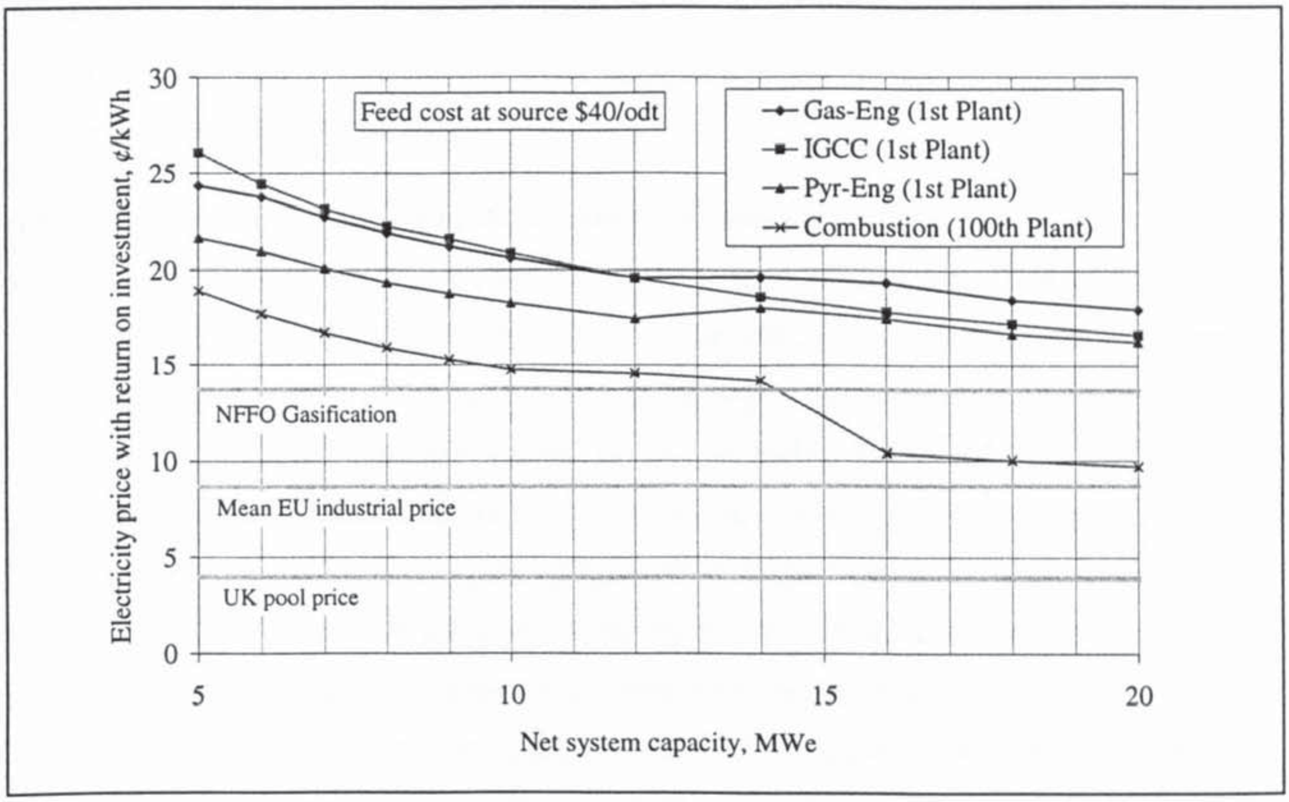


Figure 9.21 - Base case electricity prices with 10% return on investment

9.9 SENSITIVITY ANALYSES

9.9.1 Introduction

The evaluations thus far have taken the results of the models at face value. Sensitivities and uncertainties in the model have not been considered. This section examines the sensitivity of the models to variations in system parameters. This is followed by an examination of the uncertainties that are associated with the most critical system parameters highlighted in the sensitivity analysis. Finally, an assessment of the risk in each system is made on the basis of the relative sensitivities and uncertainties.

9.9.2 Sensitivities

Sensitivity analyses assess the effect of changes to the operating conditions on the system results. A sensitivity analysis is carried out by changing each input parameter in turn by +/- 10% of its base case value. This base case value may be an input parameter such as the feed cost, or a calculated value such as the conversion efficiency or capital costs. The effect of the change is measured in terms of the percentage change on the electricity production cost. A single system capacity of 10 MW_e is used. The full sensitivities analyses for each base case system are presented in Appendix G, while summaries of the 10 most sensitive parameters are shown in Table 9.7 to Table 9.10.

The following observations are made regarding the system sensitivities.

- All the systems are highly sensitive to the system efficiencies. This is unsurprising given that the efficiency of the system effects the amount of feedstock that must be bought and the size of the equipment required to process the feed.
- The most capital intensive system (IGCC) is highly influenced by availability, the percentage of the available operating hours in the year during which the system operates. The least capital intensive system (Combustion) is least sensitive to this parameter.
- All of the systems are sensitive to the capital cost of the conversion process, which represents the largest proportion of the total plant cost in each system.
- Feed related parameters are most important where capital costs are low, hence the influence of feed LHV, moisture content and cost on the Combustion electricity production cost. Feed cost is least important in the IGCC system where the high efficiency of the process reduces the amount of feed required and high capital costs increase the proportion of costs that are spent on capital related costs such as amortisation and overheads.

Table 9.7 - Key sensitivities, Combustion at 10 MW_e

Parameter	Base case value	% change on base case +10% change to parameter	-10% change to parameter
Electricity production cost, ¢/kWh	11.33		
Generation efficiency, %LHV	27.2%	-10.2%	13.0%
Conversion efficiency, %LHV	83.9%	-9.6%	10.7%
Feed LHV when dry, GJ/odt	19.3	-8.0%	8.7%
Feed moisture content, %wet basis	50%	5.83%	-1.32%
Feed cost, $/odt at source	40	3.5%	-3.5%
Combustion TPC, $/$_{kWe}$	1422	2.4%	-2.4%
Project life, years	20	-1.6%	2.0%
Nominal interest rate, %	10%	1.6%	-1.6%
Availability, %	90%	-1.6%	1.5%
Overheads, %TPC/y	4%	1.2%	-1.2%

Table 9.8 - Key sensitivities, Gas-Eng system at 10 MW_e

Parameter	Base case value	% change on base case +10% change to parameter	-10% change to parameter
Electricity production cost, ¢/kWh	15.12		
Generation efficiency, %LHV	35.4%	-5.5%	6.7%
Availability, %	90%	-5.4%	6.6%
Feed LHV when dry, GJ/odt	19.3	-5.2%	6.3%
Conversion efficiency, %LHV	72.4%	-5.1%	6.2%
Conversion total plant cost, $/k$W_e$	2648	3.4%	-3.4%
Feed cost, $/odt at source	40	2.0%	-2.0%
Nominal interest rate, %	10%	2.0%	-1.8%
Project life, years	20%	-1.9%	2.2%
Overheads, %TPC/y	4%	1.4%	-1.4%
Generation TPC, $/kWe	1178	1.1%	-1.1%

Table 9.9 - Key sensitivities, IGCC system at 10 MW$_e$

Parameter	Base case value	% change on base case +10% change to parameter	-10% change to parameter
Electricity production cost, ¢/kWh	14.43		
Availability, %	90%	-6.8%	8.2%
Feed LHV when dry, GJ/odt	19.3	-4.8%	5.8%
Conversion efficiency, %LHV	86.4%	-4.6%	5.6%
Generation efficiency, %LHV	44.4%	-4.6%	5.5%
Conversion TPC, \$/kW$_e$	3089	4.2%	-4.2%
Nominal interest rate, %	10%	2.4%	-2.4%
Project life, years	20	-2.4%	2.8%
Overheads, %TPC/y	4%	1.8%	-1.8%
Generation TPC, \$/kW$_e$	1659	1.7%	-1.7%
Feed cost, \$/odt at source	40	1.5%	-1.5%

Table 9.10 - Key sensitivities, Pyr-Eng system at 10 MW$_e$

Parameter	Base case value	% change on base case +10% change to parameter	-10% change to parameter
Electricity production cost, ¢/kWh	13.93		
Liquids yield, %organics/dry feed	59.9%	-6.25%	7.97%
Generation efficiency, %LHV	44.4%	-6.17%	7.68%
Availability, %	90%	-4.67%	5.67%
Feed cost, \$/odt at source	40	2.51%	-2.58%
Conversion TPC, \$/kW$_e$	1428	2.01%	-1.94%
Project life, years	20	-1.65%	1.87%
Nominal interest rate, %	10%	1.58%	-1.58%
Overheads, %TPC/y	4%	1.22%	-1.22%
Generation TPC, \$/kW$_e$	1154	1.22%	-1.01%
Feed moisture content, %wet basis	50%	1.15%	-1.01%

9.9.3 Uncertainties

Uncertainties should be considered in parallel with the sensitivities, since a combination of high sensitivity and high uncertainty indicates weaknesses in the model or the technology that must be addressed to ensure confidence in the results.

The uncertainties are addressed in four areas that broadly correspond to the key sensitivities that have been highlighted above. These categories are system efficiency, reliability, capital cost and feed characteristics.

One way of measuring the uncertainty associated with the system efficiencies is to compare the results of the model with previous work or established systems. The system efficiencies for the Combustion system are compared with other sources in Table 9.11. The combustion efficiencies calculated come out rather low and are rarely within 10% of the figures given in other sources. However, this can be explained by steam conditions that can be highly variable at any particular scale. There is in fact just as much variation in the efficiencies amongst the external sources. The efficiencies in the model are all based on existing steam cycles for biomass and hence there is considerable confidence that the efficiencies can be achieved. If anything, the steam conditions have been under-estimated since they are based on past rather than current plant data, and it may be worth re-examining them in future.

One uncertainty that may effect the combustion system efficiency is the availability of waste heat for drying. This has already been discussed.

Table 9.11 - Validation of model efficiencies and capital costs, combustion

Source	Net capacity MWe	Feed moisture %wet basis	Efficiency, %LHV Source data[a]	This work[b]	Capital cost, $/kWe Source data	This work
[31]	21.4	29.3	18.6	22.9	-	2081
[31]	25.3	32.1	26.5	23.5	-	1963
[31]	21.6	25.5	19.9	23.0	-	2063
[31]	27.9	24.0	27.2	24.1	-	1867
[195]	5.7	Not known	21.1	18.8	2158[c]	3435
[186]	4.7	50.0	23.1	16.7	3110	3657
[186]	25.5	50.0	25.4	23.1	1900	2071
[186]	51.2	50.0	25.4	26.4	1479	1550
[71]	27.0	24.0	29.0	24.0	-	1892
[71]	46.0	50.0	32.0	25.9	-	1621

a *Efficiencies from [31] have been converted to LHV using the moisture contents given and a hydrogen content of 6% of dry mass.*

b *Efficiencies are calculated from the model, using the feed moisture content and capacity given in the source. Drying has been excluded.*

C *The basis of this cost is not known, it may be lower by virtue of its scope*

In the Gas-Eng system, the efficiency is more easy to predict and the efficiencies calculated by the model were validated in Section 7.4.2. The efficiencies of diesel engines using low heating value gases is established and the data can be applied with confidence. Unfortunately there are few examples of system efficiencies in the literature. Solantausta estimates an efficiency of 33.9% at both 5 and 25 MW_e, this is some points higher than this work would predict but it is noted that the internal power consumption for the whole system is lower than would be expected for the engine alone.

The IGCC system efficiencies calculated by this work are compared with data by other authors in Table 9.12. The agreement between the data is much better than for the Combustion case and only two results were significantly different from the source data. Again the difference amongst the source data is striking, especially when many of the researchers have used the same gas turbine in their analyses. This model does tend to overestimate efficiencies especially at high capacities and this could be addressed in further work, but these capacities are of little relevance to UK applications. One uncertainty that should be addressed is the availability of waste heat for drying as noted above. Although the results are not too bad for the IGCC system, the gas turbine combined cycle efficiency is probably the most uncertain of all the module calculations. It is strongly recommended that this calculation is revisited in further work.

Table 9.12 - Validation of model efficiencies and capital costs, IGCC

Source	Net capacity MWe	Efficiency, %LHV Source data	This work	Difference	Capital cost, $/kWe Source data	This work
[186]	24.7	41.1	44.3	7.8%	4010[a]	3950
[186]	59.7	40.5	49.6	22.5%	3120[a]	2976
[186]	74.2	40.4	51.0	26.2%	2882[a]	2775
[113]	31.9	47.1	45.8	-2.8%	-[b]	-
[113]	59.8	45.1	49.6	10.0%	-[b]	-
[195]	60.3	45.3	49.6	9.4%	1778[c]	2958
[195]	16.4	44.4	42.0	-5.4%	2920[c]	4363
[109]	7.5	42.0	38.0	-9.5%	-	-
[109]	39.0	45.0	47.0	4.4%	-	-
[109]	62.0	48.0	49.8	3.7%	-	-
[109]	72.0	48.0	50.8	5.8%	-	-
[107]	7.2	39.8	37.7	-5.3%	-	-

a *The cost data for the gasifier used in this source in the same as cost data used in this work. All other costs are the calculated from separate data.*

b *All cost data in this source are based on the same data as this work and cannot be used in validation*

c *The basis of these cost data are not known. They may be lower because of their scope.*

The final efficiency to consider is that of the fast pyrolysis system. Although the fast pyrolysis efficiency of 62% used in this work is identical to that used in the IEA study reported in Section 3.4.1, the efficiency is highly questionable for the reasons discussed in Section 7.3.2.1. This must be regarded as a major uncertainty. The efficiency of the engine is far more established and can be treated with confidence.

The next issue to be addressed is system reliability, which is equivalent to availability (although strictly the latter parameter includes both scheduled and unscheduled downtime). This has already been discussed in Section 9.6. The combustion system is based on reliable, established technology and there is every confidence that a 90% availability can be achieved. The reliability of the other systems is far less certain. The greatest uncertainty is associated with the IGCC system, where the gas cleaning system, the operation of the gas turbine and integration of the gasifier and gas turbine compressor have all to be proved. Under these circumstances the assumption of a 90% availability is highly optimistic. The Gas-Eng system is more likely to be reliable, given that all the components of the system are commercial although there are few examples of their integration. A 90% availability in this case is possible if not probable. The reliability of the Pyr-Eng system is more doubtful, given the status of the conversion technology and the novelty of firing engines with pyrolysis liquid. However, the Pyr-Eng system has the advantage that the conversion process and generating cycle are separated by a buffer store of the pyrolysis liquid. Short shutdowns in either the conversion process or the generation cycle will not cause the shutdown of the whole system. It is assumed that this would make the system about as reliable as the Gas-Eng system i.e., a 90% availability is possible but not probable.

The third uncertainty issue is capital cost. Errors are inherent in all cost evaluations, but the uncertainty associated with capital costs in 1st plant systems where there is little or no historical data to rely on is bound to be greater. Most of the cost data has been based on established costs and is reliable to within +/-30%, the typical range for study estimates such as this. The gasification and fast pyrolysis data is subject to the most uncertainty. Particular concerns are the costs of the pressurised gasifier, which is only based on 4 data, and the costs of the tar cracker which has been very simply estimated.

The final uncertainties to consider are associated with the feedstock. The sensitivities highlighted the influence of feed LHV on the system. This is a trivial result since the LHV of woods do not show much variance, as shown by Table 2.4. There is uncertainty about feed costs but this cannot be alleviated by improvements to the model and the only option is to consider how the variations will effect the comparisons between the systems, as carried out in Section 9.5.

9.9.4 Assessment of the relative system risks

Table 9.13 compares the sensitivities and uncertainties in each system by assigning qualitative scores to each system. There is little more that can be done without applying complex risk analysis tools such as Monte Carlo simulations that are beyond the scope of this work. The table helps to assess the relative risk with each system since the lower the total given, the lower the risk. It can be seen that not only does the Combustion system produce the lowest cost electricity, it also carries with it the lowest risk.

Where high sensitivity and high uncertainty are found together the parameter is particularly critical. Thus the IGCC system represents a very high risk because of chances of unforeseen shutdowns are high and the impact of such shutdowns on the production costs are critical. The fast pyrolysis liquids yield is another risk, and system developers would be wise to predict yields conservatively until more experience of commercial scale plant is gained.

Table 9.13 - A comparison of system sensitivities and uncertainties

	Combustion	Gas-Eng	IGCC	Fast-pyrolysis
Efficiency				
Sensitivity[a]	5	4	4	4
Uncertainty[a]	3	2	4	5
Reliability				
Sensitivity	1	4	5	3
Uncertainty	1	3	5	3
Capital cost				
Sensitivity	3	3	4	4
Uncertainty	2	4	5	3
Feed cost				
Sensitivity	3	2	1	3
Uncertainty	3	3	3	3
TOTALS	21	25	31	28

a 1 = low, 5 = high

10. SUMMARY AND CONCLUSIONS

10.1 INTRODUCTION

The aim of this work was to calculate, evaluate and compare the costs and performance of integrated biomass to electricity systems. Integrated biomass to electricity systems include novel technologies that convert solid biomass to an intermediate energy carrier such as fuel gas or liquid fuel that can then be used to generate electricity in high efficiency gas turbines or internal combustion engines.

Specific aims of the project were as listed below.

- The project must compare all three thermochemical conversion technologies: combustion, gasification and fast pyrolysis.
- The project must examine the systems over a continuous 1-100 MW_e range of capacities.
- The project must include a comprehensive examination of feed pretreatment, defined as the processes required at the site before feed conversion to ensure a continuous supply of feedstock to the specifications demanded by the conversion technology.
- The project must include an examination of systems de-coupling. This is the separation in time or space of the feed conversion and electricity generation steps, and is only possible in systems based on fast pyrolysis.
- The project should include an evaluation of learning effects. Such effects are widely expected to reduce the current high costs of the novel processes and experience in their technologies accumulates.

10.2 MODELLING

After a review of the current state of the art in biomass to electricity systems, the following four basic systems were selected for further evaluation on the basis of their near-term commercial potential, their applicability to scale, and the amount of data available for modelling. The combinations of conversion technology and generating cycle that were studied are:

1. Combustion to raise steam for a steam cycle;

2. Atmospheric gasification, producing a fuel gas for a gas-fired dual fuel diesel engine;

3. Pressurised gasification, producing a fuel gas to fire a gas turbine combined cycle; and

4. Fast pyrolysis to produce a liquid fuel for use in a liquid fired dual fuel diesel engine.

All the conversion processes were based on fluid bed or circulating fluid bed reactors. In all cases the feedstock was wood, selected because it is the most popular feedstock currently used and as such there was more information available regarding its handling and processing. This wood was delivered to the conversion site as wood chips, which is consistent with usual industry practice. Feed production methods were not evaluated in this thesis, but their impact on system cost and performance was studied in an associated project for the International Energy Agency, with results given in Appendix A.

It was originally hoped that environmental impacts of the systems could be evaluated in parallel with technical and economic performance. This is an area of growing concern as the systems approach commercial status. To this end, a methodology was developed to quantify the emissions and hazards associated with each system but no satisfactory way could be found to integrate this methodology (described in Appendix B) with the technical and economic calculations.

The technical and economic modelling of the systems was achieved by application of a methodology defined in Chapter 4. This methodology was defined to give consistency between models to ensure legitimate comparisons could be made. The modelling was carried out by breaking up the system into smaller sub-systems, with each sub-systems modelled on a spreadsheet. Each spreadsheet model is referred to as a module and the systems have been modelled by combining several of the following modules:

- Feed transport and pyrolysis liquid transport;
- Feed pretreatment;
- Combustion, atmospheric gasification, pressurised gasification and fast pyrolysis;
- Steam cycles, gas-fired dual fuel diesel engine, gas turbine combined cycle, and liquid fired dual fuel diesel engine; and
- Grid connection

The spreadsheet models have been developed successively, although there have been occasional problems, particularly when developing relationships for the generic performance of highly complex and closely integrated sub-systems such as the gas turbine combined cycle. Another problem with separating the system into discrete stages has been that the availability of heat for drying has been difficult to quantify. It is recommended that both of these topics are examined in more detail in further work.

10.3 RESULTS

The spreadsheet models have been used to evaluate and compare the technical and economic performance of integrated biomass to electricity systems at a generic level over a wide range of scales and under a variety of operating conditions.

The somewhat disappointing overall conclusion from the evaluation is that none of the systems are competitive with conventional large-scale electricity generation from fossil fuels that enjoy the benefits of low fuel costs and strong economies of scale. Generation of electricity to the grid may be possible in the long term if fuel prices rise or environmental taxes are imposed that raise the price of electricity from fossil fuels. Neither option can be expected in the short term in the UK.

There are two main opportunities for electricity generation from biomass:

1. Electricity generation for the grid where incentives or subsidies such as the UK Non-fossil fuel obligation support an elevated electricity price for renewable energy; or
2. Electricity generation for direct sale to the consumer, thus by-passing the costs of distribution.

Incentive schemes for the former case are now in place in many countries in Europe, but they rely on public and political support for their long-term implementation. The latter case is confined to niche markets where the local electricity price for large consumers is high and where, preferably, there is a low cost feedstock available. Under market conditions where the disposal of biomass wastes to landfill incurs a cost the generation of electricity could become a very attractive alternative.

One way of increasing revenue of biomass to electricity schemes is to combine electricity generation with other products. The most obvious product is heat, and it is strongly recommended that combined heat and power schemes are examined as a means of making the small-scale generation of electricity economic. Fast pyrolysis also offers opportunities for chemicals production. Markets for the chemicals are currently uncertain but a more long term aim for this technology could be the co-production of chemicals, heat and electricity.

Given that the absolute costs of electricity from biomass are high, the following conclusions are made with regard to the comparative costs of the four alternative systems.

10.3.1 Variations with scale

- Combustion was the most cost effective technology at all but the very smallest generating capacities, if the analysis is based on current costs for each process and therefore ignores potential learning effects in gasification and fast pyrolysis.
- The capital costs of the pretreatment module cannot be predicted using a simple relationship because of the changes in equipment required and feed specifications at each scale and for each capacity; thus it is important to model explicitly the costs and performance of feed handling and processing before the conversion reactor.
- Of the novel systems, fast pyrolysis was the most economic at scales up to 12 MW_e. IGCC systems were the most economic at scales above 12 MW_e. The atmospheric gasification and gas-fired dual fuel engine could not be supported at any scale, due mainly to the extra costs imposed by the tar cracker.
- The engine-based technologies are relatively expensive in large scale applications due to their low scale-economies and moderate efficiencies. At large scale, only the IGCC system comes close to competing with the conventional combustion system.
- Electricity production costs below 5 MW_e are extremely expensive in all technologies, due to a combination of high capital costs and low system efficiencies.
- Comparisons between systems cannot be carried out on the basis of efficiencies or investment alone since it has been shown that increased efficiencies are only achieved through increased investment. A "specific efficiency cost", the ratio of capital invested to system efficiency, can be used to give an approximate indication of which systems offer the best compromise between system efficiency and capital cost.

10.3.2 Learning effects

- The application of learning effects has a significant effect on the comparison of electricity production costs between systems by reducing the cost of the novel systems in relation to combustion.

- If a 20% learning factor is applied to the novel systems selectively (using learning factors with the conversion technologies only) then by the 10th plant IGCC system electricity production cost falls below the Combustion system.
- The collective application of learning factors to the whole system, as promoted by Elliott and Booth [24] is over-optimistic and suggests unacceptably high reductions in capital costs.
- Although there are potential cost advantages for the novel systems after learning effects are applied, current costs are high and are unlikely to encourage the investment for the initial systems that are required before the benefits of learning can be gained. This highlights the advantage of the current IGCC projects that are part-funded by public money. The same type of partnership between public and private finance should be implemented for fast pyrolysis systems development.

10.3.3 Feedstock costs

- All the systems are sensitive to feed costs, with the least efficient systems being the most affected.
- Low feed costs give an advantage to combustion because its low investment costs become more important than its low efficiency. Although low feed costs improve the viability of biomass to electricity systems in general, policy makers should be aware that feed cost subsidies alone would not encourage the development of the new technologies.

10.3.4 De-coupling

- Systems de-coupling increases the flexibility of the fast pyrolysis based system.
- Where the feedstock and generating site are located remotely, there is a slight cost advantage in siting the pyrolysis plant at the feed source and transporting pyrolysis liquid rather than feedstock.
- The use of multiple pyrolysis sites to supply a single generator is not cost-effective due to the increase in overall capital costs. There could be advantages in terms of spreading the risks associated with fast pyrolysis performance between several plants, but this has not been quantified.
- The use of multiple generators supplied by a single pyrolysis site offers significant reductions in electricity production costs when compared with an equivalent network of multiple, close-coupled sites.

10.3.5 Sensitivities and Uncertainties

- All the models are very sensitive to efficiency, and comparisons between the efficiencies calculated in this work and those calculated in other analyses has highlighted significant uncertainties. Not only were the results from this model different from the results is other work, but there was no clear consensus among the third parties.
- All the models, particularly the capital intensive systems, are very sensitive to system availability. Given the novelty of the gasification and fast pyrolysis systems, there is a possibility that the availability of 90% used in the comparisons may not be achieved. This problem is a particular concern in the IGCC system, which is not only the most sensitive to availability changes, but also the most likely to experience reliability problems during the first few systems.

11. FURTHER WORK

11.1 IMPROVEMENTS TO THE EXISTING MODEL

Suggestions for improving the modules have been given in Chapters 5-8 as the modules were developed. It is suggested that the following issues are addressed urgently to improve the reliability in the models and confidence in the results.

1. Examine the availability of waste heat for drying and adjust the modules (notably in the Combustion and IGCC systems) if it is found that insufficient heat is available.
2. Update the pyrolysis yield relationship as more data become available on the performance of large scale systems.
3. Include the costs of effluent disposal from the atmospheric gasifier wet scrubbing systems.
4. Examine in more detail the costs of a tar-cracker in the atmospheric gasification system and the effects on the system of measures to avoid a tar cracker altogether.
5. Re-evaluate the relationship developed for steam cycle efficiency using the latest state-of-the-art steam conditions and cycle configurations.
6. Examine the assumptions made in the development of the gas turbine combined cycle efficiency relationship. This examination should focus on the effects of considering the whole gasifier output as the fuel and the use of all gas turbines rather than just those that currently have a low heating value gas capability.

11.2 ADDITIONAL AREAS FOR INVESTIGATION

It would be useful to examine the following areas that could offer reductions in electricity production costs and improve the opportunities for electricity from biomass. In all cases it is imperative that the current methodology is maintained so that all results are comparable. If changes are made to the existing methodology then these changes should be applied to the current modules also for consistency.

1. Evaluate other feedstocks. This may require additions and changes to the current pretreatment steps and changes to the current feed conversion modules to account for different feed compositions.

2. Examine configurations of the basic technologies that are more suited to small-scale generation such as grate-fired combustors or downdraft gasifiers. These may offer opportunities at the small-scale that are not available to the fluid bed systems because of their high capital cost at small scale.

3. Examine ablative fast pyrolysis to highlight its benefits and drawbacks in comparison with fluid bed pyrolysis. One benefit is the avoidance of feed grinding but there may be disadvantages in other areas such as product yields.

4. Include equipment in de-coupled fast pyrolysis systems that allow the in-plant generation of power in systems using pyrolysis liquid where the fast pyrolysis site and electricity generation plant are remote from each other. Internal power generation is a likely prospect where the feedstocks are available in very remote locations and the addition of this capability would make the fast pyrolysis sites independent of the existing grid.

5. Add the option of atmospheric gasification and a gas turbine combined cycle.

6. Examine combined heat and power production. This cogeneration of heat for export is one way of increasing the revenue of small scale power plant. This is the key recommendation for further work as it may be the only way to give biomass an increasing share of the electricity market without relying on subsidies.

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Appendix A Techno-economic assessment of biomass to energy

Final report to the IEA Bioenergy Agreement 1992-1994
Published in Biomass and Bioenergy, 9(1-5):205-226

Appendix B Bioenergy Environmental Evaluation Scheme

Extract from Report No. ES 94/2 (FINAL REPORT)

"The nature and control of solid, liquid and gaseous emissions from the thermochemical processing of biomass"

Compiled by Dr K L Mackie
New Zealand Forest Research Institute
Private Bag 3020
Rotorua
New Zealand

Aston University

Appendix C Feed Pretreatment Module Data Sheets

Appendix C .1 Feed reception data sheet

Pretreatment Step: Feed Reception			All items in mild steel						Engineering, design and supervision						15%	
													Management overheads		10%	
Operating Characteristics, all options													Commissioning			5%
Q in	160	t/d											Contingency			10%
Delivery period	10	h/d	for	5	d/week								Contractors' fee			10%
Q out, average	16	t/h							Interest during construction							10%
Q out, design	24	t/h	=	96	m3/h											
Small Scale System																
Unloading time	30	min	for	22	t load											
Option limiting capac	440	t/d theoretical														
Option limiting capac	293	t/d design														
Item	Capacity		EC	Power	Source	Er	Pi	In	El	Ci	St	La	DPC/EC	DPC	IPC	TPC
	Value	units	US$1991	kWh												
Paved unloading area	625	m2	10625	0.00	[Braeul	0.10	0.00	0.00	0.00	0.23	0.00	0.00	1.33	14136	17670	23854
Front end loader	24	t/h	128000	0.00	[Simons	0.00	0.00	0.00	0.00	0.00	0.00	0.00	1.00	128000	128000	172800
TOTALS, 1991 values			138625	0										142136	145670	196654
TOTALS, 1995 values			152363	0										156221	160105	216142
Medium Scale System																
Unloading time	15	min	for	27	t load											
Option limiting capac	1080	t/d theoretical														
Option limiting capac	720	t/d design														
Item	Capacity		EC	Power	Source	Er	Pi	In	El	Ci	St	La	DPC/EC	DPC	IPC	TPC
	Value	units	US$1991	kWh										US$1991	US$1991	US$1991
Truck scale	54	t	30500	0	[Simons	0.13	0.00	0.37	0.32	0.18	0.27	0.00	2.28	69461	86826	117215
Truck dumper	36	t	51500	18	[Simons	0.11	0.00	0.31	0.29	0.16	0.25	0.00	2.12	109180	136475	184241
Receiving bin	216	m3	21438	0	[ALPS]	0.14	0.00	0.42	0.00	0.20	0.00	0.00	1.76	37668	47084	63564
Drag-chain conveyor	108	t/h	128109	55	[Simons	0.09	0.00	0.23	0.23	0.13	0.21	0.00	1.89	242477	303096	409180
Belt conveyor	432	m3/h	36020	6	[Garret	0.12	0.00	0.35	0.37	0.18	0.26	0.00	2.29	82430	103037	139100
TOTALS, 1991 values			267567	79										541215	676518	913300
TOTALS, 1995 values			294083	79										594849	743561	1003807

Appendix C .1 (cont.)

Pretreatment Step: Feed Reception			All items in mild steel						Engineering, design and supervision						15%	
												Management overheads			10%	
Operating Characteristics, all options													Commissioning			5%
Q in	160	t/d											Contingency			10%
Delivery period	10	h/d	for	5	d/week							Contractors' fee				10%
Q out, average	16	t/h								Interest during construction						10%
Q out, design	24	t/h	=	96	m3/h											
Large Scale System																
Unloading time	8	min	for	27	t load											
Train limiting capac.	2025	t/d theoretical														
Train limiting capac.	1350	t/d design														
No. trains required	1															
Item	Capacity		EC	Power	Source	Er	Pi	In	El	Ci	St	La	DPC/EC	DPC	IPC	TPC
	Value	units	US$1991	kWh										US$1991	US$1991	US$1991
(shared items)																
Truck scale	100	t	130000	0	[ALPS,E	0.09	0.00	0.23	0.23	0.13	0.21	0.00	1.89	245635	307044	414510
Belt conveyor	96	m3/h	31218	2	[Garret	0.13	0.00	0.37	0.39	0.18	0.27	0.00	2.34	72909	91137	123034
(per train)																
Truck dumper	54	t	76900	30	[Simons	0.10	0.00	0.27	0.26	0.15	0.23	0.00	2.01	154825	193531	261267
Receiving bin	270	m3	25063	0	[ALPS]	0.14	0.00	0.00	0.00	0.19	0.00	0.00	1.33	33313	41641	56216
Drag-chain conveyor	203	t/h	200034	86	[Simons	0.08	0.00	0.20	0.21	0.12	0.19	0.00	1.80	360010	450013	607517
TOTALS, 1991 values			463215	118										866693	1083366	1462545
TOTALS, 1995 values	.		509119	118										952582	1190727	1607482

Appendix C .2 Pre-store screening data sheet

Pretreatment Step: Pre-store Scree			All items in mild steel						Engineering, design and supervision						15%	
														Management overheads	10%	
Operating Characteristics, all options														Commissioning		5%
Feed delivery rate	720	t/d	at	5	d/week	10	h/d							Contingency		10%
Feed design input rat	108	t/h	=	432	m3/h	at	0.25	t/m3						Contractors' fee		10%
Rate out of store	21.43	t/h	=	85.7	m3/h				Interest during construction							10%
Over-size fraction,	6%															
Over-size fraction,	3%															
Pre-bulk storage																
Re-chip step next to screen																
Effects of extraction on main conveyor capacities are ignored																
Item	Capacity		EC	Power	Source	Er	Pi	In	El	Ci	St	La	DPC/EC	DPC	IPC	TPC
	Value	units	US$1991	kW										US$1991	US$1991	US$1991
Conveyor	432	m3/h	21269	1.6	[Garret	0.14	0.00	0.42	0.35	0.20	0.29	0.00	2.399	51026	63783	86107
Magnet	1.08	m	9642	7.6	[Simons	0.18	0.00	0.54	0.42	0.24	0.33	0.00	2.711	26135	32669	44103
Disk screen	108	m3/h	34128	2.0	[ALPS]	0.13	0.00	0.36	0.38	0.18	0.39	0.00	2.430	82933	103666	139949
Conveyor to re-chip	26	m3/h	15488	0.1	[Garret	0.16	0.00	0.46	0.38	0.22	0.31	0.00	2.516	38969	48711	65760
Main transfer convey	432	m3/h	60999	6.5	[Garret	0.11	0.00	0.29	0.33	0.16	0.24	0.00	2.130	129912	162390	219227
TOTALS, 1991 values			141525	18										328975	411219	555146
TOTALS, 1995 values			155550	18										361577	451971	610160

Appendix C .3 Bulk storage data sheet

Pretreatment Step: Bulk storage			All items in mild steel											Engineering, design and supervision	15%	
														Management overheads	10%	
Operating Characteristics, all capacities														Commissioning		5%
Feed delivered	4000	t/d	for	5	d/week	10	h/d							Contingency		10%
Storage time	3	weeks												Contractors' fee		10%
Storage volume	240000	m3	at	0.25	t/m3									Interest during construction		10%
Reclaim rate	119.0	t/h														
	476.2	m3/h														
Large scale																
Limit	None	m3														
Size of pile	22000	m2														
Item	Capacity		EC	Power	Source	Er	Pi	In	El	Ci	St	La	DPC/EC	DPC	IPC	TPC
	Value	units	US$1991	kW										US$1991	US$1991	US$1991
Conveyor to pile	1600	m3/h	180354	60.7	[Simons	0.35	0.00	0.21	0.21	0.12	0.20	0.00	2.087	376369	470461	635123
Bulldozer	400	t/h	1920000	23.5	[Simons	0.00	0.00	0.00	0.00	0.00	0.00	0.00	1.000	1920000	1920000	2592000
Chain conveyor recla:	39.7		143422	20.4	[Simons	0.09	0.00	0.22	0.23	0.13	0.21	0.00	1.868	267934	334917	452138
Chain conveyor recla:	39.7		143422	20.4	[Simons	0.09	0.00	0.22	0.23	0.13	0.21	0.00	1.868	267934	334917	452138
Chain conveyor recla:	39.7		143422	20.4	[Simons	0.09	0.00	0.22	0.23	0.13	0.21	0.00	1.868	267934	334917	452138
Conveyor to reclaim p	476	m3/h	62807	7.1	[Simons	0.11	0.00	0.29	0.27	0.16	0.24	0.00	2.066	129747	162183	218947
Radial stacker and sc	119	t/h	1730111	104.0	[Simons	0.05	0.00	0.10	0.13	0.07	0.13	0.00	1.473	2549037	3186296	4301499
Front end loader	0	t/h	128000	0	[Simons	0.00	0.00	0.00	0.00	0.00	0.00	0.00	1.000	128000	128000	172800
Conveyor to screen	476	m3/h	37087	3.6	[Garrett	0.12	0.00	0.35	0.31	0.18	0.26	0.00	2.216	82189	102736	138694
TOTALS, 1991 values			4488626	260										5989143	6974429	9415479
TOTALS, 1995 values			4933445	260										6582662	7665589	10348545

Appendix C .3 (cont.)

Pretreatment Step: Bulk storage			All items in mild steel											Engineering, design and supervision	15%	
														Management overheads	10%	
Operating Characteristics, all capacities														Commissioning		5%
Feed delivered	4000	t/d	for	5	d/week	10	h/d							Contingency		10%
Storage time	3	weeks												Contractors' fee		10%
Storage volume	240000	m3	at	0.25	t/m3									Interest during construction		10%
Reclaim rate	119.0	t/h														
	476.2	m3/h														
Small-scale																
Limit	m3															
Item	Capacity		EC	Power	Source	Er	Pi	In	El	Ci	St	La	DPC/EC	DPC	IPC	TPC
	Value	units	US$1991	kW										US$1991	US$1991	US$1991
Paved storage area	44000	m2	748000	0.0	[Simons	0.06	0.00	0.13	0.15	0.09	0.15	0.00	1.579	1181435	1476793	1993671
Front end loader						------included in feed reception-------										
Day bin	5714	m3	212298	0.0	[ALPS]	0.08	0.00	0.19	0.21	0.12	0.19	0.00	1.788	379613	474517	640598
Drag chain conveyor	119	t/h	124533	59.3	[Simons	0.09	0.00	0.23	0.23	0.13	0.21	0.00	1.899	236490	295612	399076
Conveyor to screeninc	476	m3/h	21900	1.8	[Garret(	0.14	0.00	0.41	0.35	0.20	0.29	0.00	2.389	52314	65392	88279
TOTALS, 1991 values			1106731	61										1849851	2312314	3121624
TOTALS, 1995 values			1216407	61										2033170	2541462	3430974
Medium-scale																
Limit	43200	m3														
Size of pile	26667	m2														
Item	Capacity		EC	Power	Source	Er	Pi	In	El	Ci	St	La	DPC/EC	DPC	IPC	TPC
	Value	units	US$1991	kW										US$1991	US$1991	US$1991
Conveyor to pile	1600	m3/h	45027	9.8	[Simons	0.50	0.00	0.32	0.29	0.17	0.25	0.00	2.541	114415	143019	193076
Flinger	400	t/h	4133	32.0	[Simons	0.93	0.00	0.71	0.51	0.29	0.39	0.00	3.840	15870	19838	26781
Front end loader	286	t/h	896000	0	[Simons	0.00	0.00	0.00	0.00	0.00	0.00	0.00	1.000	896000	896000	1209600
Stoker reclaimer	476	m3/h	386831	267.8	[Simons	0.07	0.00	0.16	0.18	0.10	0.17	0.00	1.680	649998	812497	1096871
Conveyor to return	476	m3/h	37087	3.6	[Garret(	0.12	0.00	0.35	0.31	0.18	0.26	0.00	2.216	82189	102736	138694
TOTALS, 1991 values			1369079	313										1758473	1974091	2665022
TOTALS, 1995 values			1504753	313										1932736	2169721	2929124

Appendix C .4 Post-store screening data sheet

Pretreatment Step: Post-store scre			All items in mild steel											Engineering, design and supervision	15%	
														Management overheads	10%	
Operating Characteristics, all options														Commissioning		5%
Feed delivery rate	720	t/d	at	5	d/week	####	h/d							Contingency		10%
Feed design input rat	108	t/h	=	432	m3/h	at	0.25	t/m3						Contractors' fee		10%
Rate out of store	21.43	t/h	=	85.7	m3/h									Interest during construction		10%
Over-size fraction, :	6%															
Over-size fraction, :	3%															
Post-bulk storage, small scale systems																
Re-chip step next to screen																
Effects of extraction on main conveyor capacities are ignored																
Item	Capacity		EC	Power	Source	Er	Pi	In	El	Ci	St	La	DPC/EC	DPC	IPC	TPC
	Value	units	US$1991	kW										US$1991	US$1991	US$1991
Conveyor	86	m3/h	13093	0.3	[Garrett	0.16	0.00	0.49	0.39	0.22	0.32	0.00	2.582	33809	42261	57053
Magnet	0.48	m	5608	4.4	[Simons	0.20	0.00	0.65	0.48	0.27	0.37	0.00	2.965	16631	20789	28065
Disk screen	86	m3/h	27086	1.7	[ALPS]	0.13	0.00	0.38	0.40	0.19	0.41	0.00	2.519	68238	85297	115151
Conveyor to re-chip	21	m3/h	14628	0.1	[Garrett	0.16	0.00	0.47	0.38	0.22	0.31	0.00	2.538	37131	46413	62658
Main transfer conveyo	86	m3/h	13093	0.3	[Garrett	0.16	0.00	0.49	0.47	0.22	0.32	0.00	2.662	34858	43573	58823
TOTALS, 1991 values			73508	7										190666	238333	321750
TOTALS, 1995 values			80792	7										209561.3	261951.7	353634.8

Appendix C .5 Re-chip data sheet

Pretreatment Step: Re-chip			All items in mild steel											Engineering, design and supervision	15%	
														Management overheads	10%	
Operating Characteristics, all capacities														Commissioning		5%
Q in	10	t/h												Contingency		10%
Bulk density	0.25	t/m3												Contractors' fee		10%
														Interest during construction		10%
Item	Capacity		EC	Power	Source	Er	Pi	In	El	Ci	St	La	DPC/EC	DPC	IPC	TPC
	Value	units	US$1991	kW										US$1991	US$1991	US$1991
Surge bin	40	m3	57679	0.0	[Simons	0.11	0.00	0.30	0.28	0.16	0.24	0.00	2.089	120475	150594	203302
Knife hog	10	t/h	39980	65.7	[Carr,S.	0.12	0.00	0.34	0.30	0.17	0.26	0.00	2.193	87690	109613	147978
Transfer conveyor to	40	m3/h	7492	0.2	[Garrett	0.19	0.00	0.59	0.45	0.25	0.35	0.00	2.825	21162	26452	35711
TOTALS, 1991 values			105151	66										229328	286660	386990
TOTALS, 1995 values			115572	66										252054	315067	425341

Appendix C .6 Rotary dryers performance data

Performance Data - Rotary Dryers												
Source		[224]	[224]	[224]	[224]	[242]	[224]	[224]	[224]	[224]	[224]	[224]
Type of dryer		Rotary	Rotary	Rotary	Rotary	Rotary	Rotary	Rotary	Rotary	Rotary	Rotary	Rotary
Size	diam, m	3.7	4.6	3.7	3.7		3.7	3	3.7	4.3		
	length, r	16.8	16.8	12.2	15.2		18.3	12.2	12.2	15.3		
Feed rate in	t/h	20.3	14.4	13.4	19.9	7.6	23.0	13.5	21.8	31.1	6.7	10.9
	odt/h	9.1	6.5	6.0	8.6	3.8	10.3	6.1	9.8	14.0	3.0	4.9
Preferred particle size	mm	50	50	90	90		90	90	90	90	19	19
Maximum particle size	mm	50	75	125	125		125	125	125	125	25	25
Moisture in	%wet	55%	55%	55%	55%	50%	55%	55%	55%	55%	55%	55%
Moisture out	%wet	40%	30%	32%	31%	35%	30%	26%	25%	24%	15%	15%
Heat source		Flue g	Flue g	Flue g	Flue g	Flue g	Flue g	Flue g	Flue g	Flue g	Flue g	Flue g
Inlet temperature	°C	293	355	315	315	250	315	315	315	315	316	316
Exit temperature	°C	93	93	104	99	110	102	104	99.5	99.5	82	82
Evaporation load	twe/h	5.1	5.1	4.6	7.0	1.8	8.2	5.2	8.6	12.6	3.0	4.8
Heat required	GJ/h	19	20	21.5	31.9		36.5	21.2	34	48.5	8.5	13.7
	MJ/twe	3.743	3.885	4.703	4.544	3.490	4.466	4.061	3.952	3.849	2.879	2.861
Pressure drop	bar	0.025	0.025	0.025	0.025		0.025	0.03	0.03	0.03	0.03	0.03
Power consumption	kW					22.4						
Theoretical drying req.												
T in	°C	25										
T evap	°C	100										
cp	kJ/kgK	4.181										
hfg	kJ/kg	2258										
Energy	kJ/kg	2572										
MC Diff		15%	25%	23%	24%	15%	25%	29%	30%	31%	40%	40%
Temp diff		200	262	211	216	140	213	211	215.5	215.5	234	234
Heat requirement		3.743	3.885	4.703	4.544	3.490	4.466	4.061	3.952	3.849	2.879	2.861
Efficiency		68.7%	66.2%	54.7%	56.6%	73.7%	57.6%	63.3%	65.1%	66.8%	89.3%	89.9%

Appendix C .6 (cont.)

Performance Data - Rotary Dryers										
Source		[224]	[224]	[224]	[224]	[224]	[231]	[231]	[231]	[231]
Type of dryer		Rotary	Rotary	Rotary	Rotary	Rotary				
Size	diam, m									3.66
	length, r									9.14
Feed rate in	t/h	20.2	49.0	49.0	39.9	46.3	20.7	25.9	32.0	20.9
	odt/h	9.1	22.0	21.6	18.0	20.8	10.91	13.59	17.27	11.08
Preferred particle size	mm	19	25	25	25	25				
Maximum particle size	mm	25	125	125	125	125				
Moisture in	%wet	55%	55%	55%	55%	55%	47%	48%	46%	47%
Moisture out	%wet	15%	45%	35%	30%	25%	5%	6%	6%	22%
Heat source		Flue g	Flue g	Flue g	Flue g	Flue g	Bark b	Bark b	Bark b	Flue g
Inlet temperature	°C	316	315	315	315	315	385	454	510	
Exit temperature	°C	82	93	93	93	93	124	124	124	
Evaporation load	twe/h	8.8	9.1	14.5	14.5	18.1	9.2	11.5	13.7	6.7
Heat required	GJ/h	25.4	26.84	45.07	43.24	55.52	31.6	38.6	46.1	23.2
	MJ/twe	2.880	2.959	3.107	2.980	3.060	3.436	3.370	3.370	3.463
Pressure drop	bar	0.03	0.025	0.025	0.025	0.025				
Power consumption	kW									33.38
Theoretical drying req.										
T in	°C									
T evap	°C									
cp	kJ/kgK									
hfg	kJ/kg									
Energy	kJ/kg									
MC Diff		40%	10%	20%	25%	30%	42%	41%	40%	25%
Temp diff		234	222	222	222	222	261	331	386	
Heat requirement		2.880	2.959	3.107	2.980	3.060	3.436	3.370	3.370	3.463
Efficiency		89.3%	86.9%	82.8%	86.3%	84.0%	74.8%	76.3%	76.3%	74.3%

Appendix C .7 Rotary dryers cost data sheet

Total plant cost, drying step	Factor	TPC/IP	1.15					
		TPC/DP	1.41					
		TPC/EC	4.13					
Source	[242]	[244]	[224]	[231]	[218]	[243]	[181]	[224]
Desc.	Rotary	Rotary	Rotary	Rotary	Rotary	Rotary	Rotary	Rotary
Includes	Dryer	Dryer	Dryer	Dryer	Dryer	Dryer	Dryer	Dryer
	All an	All an	All an	All an	All an	All an	Convey	All an
Base units	twe/h	twe/h	twe/h	twe/h	twe/h	twe/h	twe/h	twe/h
Base size	1.8	3.4	4	11	16.0	34.7	12.4	40
Published cost	420	287.5	380	2400	7500	2400	743	4200
Basis (1=EC,2=DPC,3=IPC,4=TPC)	2	3	3	3	3	3	1	3
Location	US	US	CA	US	ND	CA	US	CA
Year	1983	1995	1983	1983	1994	1982	1984	1983
Currency	US$	US$	CA$	US$	Dfl	CA$	US$	CA$
Cost Index, source year	98	122	98	99	120	97	99	98
Cost Index, 1995	129	122	129	122	122	129	122	129
Exchange rate, curr per USD	1.41	1.00	1.41	1.00	1.96	1.41	1.00	1.41
TPC now	554	331	409	3401	4476	2607	3781	4516

Appendix C .8 Grinding data sheet

Pretreatment Step: Grinding			All items in mild steel						Engineering, design and supervision						15%	
												Management overheads			10%	
Operating Characteristics, all capacities													Commissioning			5%
Q in	t/h												Contingency			10%
Bulk density	t/m3												Contractors' fee			10%
											Interest during construction					10%
Small scale systems, 4 t/h trains																
No. trains	at	3.9	t/h per train													
Item	Capacity		EC	Power	Source	Er	Pi	In	El	Ci	St	La	DPC/EC	DPC	IPC	TPC
	Value	units	US$1991	kW										US$1991	US$1991	US$1991
(Shared)																
Main transfer convey	564	m3/h	93424	4.2	[Garret	0.10	0.00	0.26	0.25	0.14	0.22	0.00	1.965	183619	229524	309857
Bucket elevator to fe	564	m3/h	21738	11.8	[Garret	0.14	0.00	0.41	0.42	0.20	0.83	0.00	3.000	65214	81518	110049
Feed bin and screw fe	188	m3	12328	1.5	[Simons	0.16	0.00	0.50	0.48	0.23	0.92	0.00	3.283	40474	50592	68300
(per train)																
Coarse pulverizer	3.9	t/h	67369	170.0	[Simons	0.11	0.00	0.28	0.32	0.15	0.67	0.00	2.541	171158	213947	288829
Bucket elevator to fe	15.7	m3/h	5773	0.3	[Garret	0.20	0.00	0.64	0.57	0.27	1.05	0.00	3.730	21532	26915	36335
Feed bin and screw fe	5.2	m3	1086	1.5	[Simons	0.31	0.00	1.11	0.84	0.40	1.42	0.00	5.074	5509	6886	9297
Fine pulverizer	3.9	t/h	104803	256.2	[Simons	0.09	0.00	0.25	0.29	0.14	0.62	0.00	2.393	250766	313457	423167
TOTALS, 1991 values			6572555	15424										16452023	20565029	27762790
TOTALS, 1995 values			7223889	15424										18082404	22603005	30514057
Large scale systems, 8 t/h trains																
No. trains	at	7.8	t/h per train													
Item	Capacity		EC	Power	Source	Er	Pi	In	El	Ci	St	La	DPC/EC	DPC	IPC	TPC
	Value	units	US$1991	kW										US$1991	US$1991	US$1991
(Shared)																
Main transfer convey	564	m3/h	93424	4.2	[Garret	0.10	0.00	0.26	0.25	0.14	0.22	0.00	1.965	183619	229524	309857
Bucket elevator to fe	564	m3/h	26623	11.8	[Garret	0.13	0.00	0.39	0.40	0.19	0.80	0.00	2.908	77422	96778	130650
Feed bin and screw fe	188	m3	12328	1.5	[Simons	0.16	0.00	0.50	0.48	0.23	0.92	0.00	3.283	40474	50592	68300
(per train)																
Coarse pulverizer	7.8	t/h	143336	340.0	[Simons	0.09	0.00	0.22	0.27	0.13	0.59	0.00	2.297	329256	411570	555619
Bucket elevator to fe	31.3	m3/h	9137	0.7	[Garret	0.18	0.00	0.55	0.51	0.24	0.97	0.00	3.450	31518	39398	53187
Feed bin and screw fe	10.4	m3	1737	1.5	[Simons	0.27	0.00	0.95	0.75	0.36	1.30	0.00	4.637	8054	10068	13592
Fine pulverizer	7.8	t/h	222850	512.3	[Simons	0.08	0.00	0.19	0.24	0.12	0.54	0.00	2.174	484452	605565	817513
TOTALS, 1991 values			6919458	15397										15660565	19575706	26427204
TOTALS, 1995 values			7605170	15397										17212513	21515641	29046116

Appendix C.9 Buffer storage data sheet

Pretreatment Step: Buffer storage			All items in mild steel											Engineering, design and supervision	15%	
														Management overheads	10%	
Operating Characteristics, all capacities														Commissioning		5%
Q in	10	t/h												Contingency		10%
Bulk density	0.25	t/m3												Contractors' fee		10%
Total storage capacit	160	m3												Interest during construction		10%
Maximum capacity per	500	m3														
Number of silos	1															
Item	Capacity		EC	Power	Source	Er	Pi	In	El	Ci	St	La	DPC/EC	DPC	IPC	TPC
	Value	units	US$1991	kW										US$1991	US$1991	US$1991
Live-bottom silo	160	m3/h	100565	9	[Simons	####	####	####	####	####	####	####	1.948	195895	244869	330573
TOTALS, 1991 values			100565	9										195895	244869	330573
TOTALS, 1995 values			110531	9										215308	269135	363332.2

Appendix D Feed Conversion Module Data Sheets

Appendix D .1 Combustion efficiency data sheet

COMBUSTION EFFICIENCY CALCULATIONS SHEET										
Mass Balance							**Data**			
		In	%	Out	%		Feed elemental analysis, dry basis			
Feedstock	daf t/h	0.989	9.18%					%C	51.80%	
Feed moisture	t/h	0.538	5.00%					%H	5.70%	
Ash	t/h	0.011	0.10%					%O	40.90%	
Combustion air	t/h	9.230	85.71%					%N	0.10%	
Dry flue gas	t/h			9.701	90.09%			%S	0.00%	
Combustion water	t/h			0.508	4.72%			%Ash	1.10%	
Feed moisture	t/h			0.538	5.00%		Feed MC	%wet bas	35%	
Unburned feed	t/h			0.010	0.09%		Feed HHV	GJ/odt	20.6	
Ash	t/h			0.011	0.10%		Feed LHV	GJ/odt d	19.3	
Total		10.768	100.00%	10.768	100.00%		Feed LHV	GJ/t	11.7	
Energy balance							Feed input	odt/h	1	
		In	%	Out	%		Feed temperature	°C	20	
Feed in (HHV)	GJ/h	20.559					Unburned feed	%daf	1%	
Feed moisture	GJ/h	0					Excess air		147%	
Ash	GJ/h	0								
Combustion air	GJ/h	0					Cp, flue gas	kJ/kgK	1.029	
Dry flue gas	GJ/h			1.547	7.52%		Cp, air	kJ/kgK	1.021	
Water vapour	GJ/h			2.956	14.38%		Cp, ash	kJ/kgK	0.94	
Unburnt feed	GJ/h			0.203	0.99%		hg, water vapour	kJ/kg	2825	
Ash	GJ/h			0.008	0.04%		Ash exit temperatu	°C	800	
Losses	GJ/h			1.018	4.95%		Flue gas exit	°C	175	
Steam (by diff.)	GJ/h			14.827	72.12%					
Total	GJ/h	20.559		20.559	100.00%		Approximate losses	odt/h in	losses	
								1	5%	
Feed in	GJ/h (LH	17.9786						50	1%	
Energy out before	GJ/h			15.845	88.13%		Estimated losses	%feed LH	5.0%	
Steam after losses	GJ/h			14.827	82.47%					

Appendix D .1 (cont.)

COMBUSTION EFFICIENCY CALCULATIONS SHEET										
Stoichiometry								**Drying energy required**		
For	1.000	odt	r requirement is		6.06	t/t as fed		Feed initial MC		50%
	1.538	t			9.32	t/odt		Feed exit MC		35%
	0.989	daf t			9.43	t/daf t		Moisture into drye	t/h	1
								Moisture out of dr	t/h	0.54
	Elemental Input				O_2 required			Dryer load	twe/h	0.46
	%daf t	daf t	kg/kmol	kmol	kmol			Energy required	GJ/twe	3.553
C	52.4%	0.518	12	43.17	43.17				GJ/h	1.640
H	5.8%	0.057	1	57.00	28.50					
O	41.4%	0.409	16	25.56	0.00			Flue gas in stack	°C	175
N	0.1%	0.001	14	0.07	0.07			Flue gas after dry	°C	102
S	0.0%	0.000	32	0.00	0.00			Dry flue gas energ	GJ/h	0.728
		0.985						Water vapour hg in	kJ/kg	2825
			Total O_2 req.		71.7	kmol		Water vapour hg ou	kJ/kg	2676
			O_2 in feed		25.6	kmol		Water vapour energ	GJ/h	0.440
			O_2 from air		46.2	kmol		Total energy avail	GJ/h	1.169
					1.48	t				
	Stoichiometric air requirement				6.34	t		Excess energy	%	-28.7%
			Actual air used		9.32	t				
Flue gas composition										
Dry flue gas properties										
	mw	cp	In flue	In 1 t flue gas						
	kg/kmol	kJ/kgK	t	kJ/kgK						
CO_2	44	0.978	1.90	0.191						
N_2	28	1.049	7.15	0.770						
O_2	32	0.956	0.69	0.068						
SO_2	64									
TOTALS			9.74	1.029						

Appendix D .2 Combustion costs data sheet

TOTAL PLANT COSTS - COMBUSTION PLANT								
	TPC/IPC	1.15						
	TPC/DPC	1.40						
	TPC/EC	4.83						
Source	[285]	[41]	[285]	[285]	[41]	[251]	[41]	[41]
Notes	Stoker	BFBC	Stoker	Stoker	BFBC	Stoker	BFBC	CFBC
		JWP-EPI			PYROPWR		KVAERNER	KVAERNER
Fuel input, MWth	*8*	7	*13.9*	*26*	25	28.0	29	29
Published cost, *1000	1010	1500	3042	5190	5600	5481	10350	11500
Basis (1=EC,2=DPC,3=IPC,4=TPC)	2	2	2	2	2	3	3	2
Location	US	?	?	?	?	CA	?	?
Year	1986	1994	1986	1986	1994	1987	1994	1994
Currency	US$	US$	US$	US$	US$	CA$	US$	US$
Cost Index, source year	94	118	94	94	118	105	118	118
Cost Index, 1995	122	122	122	122	122	129	122	122
Exchange rate, curr per USD	1	1	1	1	1	1.408	1	1
TPC now	1835	2171	5527	9430	8106	5500	12306	16646
Adjustment to fluid bed	10%	0%	10%	10%	0%	10%	0%	0%
TPC adjusted to FBC	2019	2171	6080	10373	8106	6050	12306	16646
TPC/kW in	229	309	398	363	329	196	420	568

Appendix D .2 (cont.)

TOTAL PLANT COSTS - COMBUSTION I								
Source	[41]	[41]	[71]	[71]	[71]	[71]	[71]	[71]
Notes	BFBC	CFBC	Stoker	CFB	Travellii	Travellii	CFB	CFB
	JWP-EPI	KVAERNER	Enkoping	Grenaa	Zurn av.	Zurn av.	Lo	Hi
Fuel input, MWth	73	73	84	87	88.7	88.7	105	105
Published cost, *1000	13500	21000	20600	27230	9195	12023	14025	17490
Basis (1=EC,2=DPC,3=IPC,4=TPC)	2	2	2	2	2	2	2	2
Location	?	?	Sweden	Denmark	US	US	US	US
Year	1994	1994	1992	1992	1992	1992	1992	1992
Currency	US$	US$	USD	USD	US$	US$	US$	US$
Cost Index, source year	118	118	113	113	113	113	113	113
Cost Index, 1995	122	122	122	122	122	122	122	122
Exchange rate, curr per USD	1	1	1	1	1	1	1	1
TPC now	19541	30397	31137	41158	13898	18172	21199	26436
Adjustment to fluid bed	0%	0%	10%	0%	10%	10%	0%	0%
TPC adjusted to FBC	19541	30397	34251	41158	15288	19989	21199	26436
TPC/kW in	267	415	369	472	157	205	203	253

Appendix D .2 (cont.)

TOTAL PLANT COSTS - COMBUSTION I								
Source	[71]	[41]	[71]	[41]	[71]	[240]	[26]	[41]
Notes	Generic	CFBC	CFB	BFBC	Fluid be	Vibratin	*Fluid be*	CFBC
		ABB-CE	Handelov	PYROPWR	Fluid bed			ABB-CE
Fuel input, MWth	130	124	143	147	171	187	*510*	892
Published cost, *1000	13479	16600	28400	22400	34403	20012	33050	74500
Basis (1=EC,2=DPC,3=IPC,4=TPC)	2	3	3	2	2	3	2	2
Location	US	?	Sweden	?	US	US	US	?
Year	1991	1994	1992	1994	1992	1986	1987	1994
Currency	USD	US$	US$	US$	USD	US$	US$	US$
Cost Index, source year	111	118	113	118	113	94	95	118
Cost Index, 1995	122	122	122	122	122	122	122	122
Exchange rate, curr per USD	1	1	1	1	1	1	1	1
TPC now	20741	19737	35261	32423	52000	29869	59420	107836
Adjustment to fluid bed	10%	0%	0%	0%	0%	10%	0%	0%
TPC adjusted to FBC	22815	19737	35261	32423	52000	32856	59420	107836
TPC/kW in	160	159	247	220	304	160	117	121

Appendix D .3 Atmospheric gasification data sheet

ATMOSPHERIC GASIFICATION CALCULATIONS SHEET										
Mass Balance							**Data**			
		In	%	Out	%		Feed elemental analysis, dry basis			
Feedstock	daf t/h	0.99	32.23%					%C	51.80%	
Feed moisture	t/h	0.18	5.75%					%H	5.70%	
Ash	t/h	0.01	0.36%					%O	40.90%	
Oxidant air	t/h	1.89	61.66%					%N	0.10%	
Dry gas	t/h			2.72	88.67%			%S	0.00%	
Reaction water	t/h			0.15	5.02%			%Ash	1.10%	
Feed moisture	t/h			0.18	5.75%		Feed MC	%wet bas	15%	
Tars				0.00	0.03%		Feed HHV	GJ/odt	20.6	
Char	t/h			0.01	0.18%		Feed LHV	GJ/odt c	19.3	
Ash	t/h			0.01	0.36%		Feed LHV	GJ/t	16.0	
Total		3.07	100.00%	3.07	100.00%					
							Feed input	odt/h	1	
Energy balance, at gasifier exit							Feed temperature	°C	20	
		In	%	Out	%		Tars in gas	%feed	0.1%	
Feed in (HHV)	GJ/h	20.56	98.21%				Char in ash	%wt	33%	
Feed moisture	GJ/h	0.00	0.00%				Air factor		30%	
Ash	GJ/h	0.00	0.00%				H to H2	%	100%	
Oxidant	GJ/h	0.37	1.79%							
Dry fuel gas	GJ/h			17.83	85.15%		Gasifier pressure	bar	1	
Water vapour	GJ/h			1.44	6.87%		Gasifier exit	°C	900	
Tars	GJ/h			0.02	0.11%		Air inlet	°C	200	
Char	GJ/h			0.17	0.83%		Water Cp	kJ/kgK	4.18	
Ash	GJ/h			0.01	0.04%		Latent heat	kJ/kg	2256	
Losses	GJ/h			1.47	7.00%		Steam Cp	kJkgK	2.2	
Total	GJ/h	20.93	100.00%	20.93	100.00%		Ash Cp	kJ/kgK	0.94	
							LHV/HHV	%	93%	
Energy in gasifier exit stream			GJ/h	19.29			Dry gas Cp	kJ/kgK	1.45	
Sensible losses, dry gas			GJ/h	3.39			Air Cp	kJ/kgK	1.1	
Sensible losses, water vapour			GJ/h	0.58						
Latent heat losses			GJ/h	0.75			Engine entry	°C	40	
Sensible heat losses, water			GJ/h	0.08			Tars HHV	GJ/t	23.3	
Less tars			GJ/h	0.02			Char HHV	GJ/t	32	
Energy remaining in gas			GJ/h	14.46						
Approximate LHV in gas			GJ/h	13.45			Approximate losses	odt/h in	losses	
								1	7%	
Feed in (LHV basis)			GJ/h	18.87				50	2%	
Energy efficiency			%LHV	71.28%			Estimated losses	%feed LH	7.0%	

Appendix D .3 (cont.)

ATMOSPHERIC GASIFICATION CALCULATIONS SHEET						
Stoichiometry						
For	1.000	odt	: requirement is		1.62	t/t as fed
	1.176	t			1.90	t/odt
	0.989	daf t			1.92	t/daf t
	Elemental Input				O2 required	
	%daf t	daf t	kg/kmol	kmol	kmol	
C	52.4%	0.518	12	43.17	43.17	
H	5.8%	0.057	1	57.00	28.50	
O	41.4%	0.409	16	25.56	0.00	
N	0.1%	0.001	14	0.07	0.07	
S	0.0%	0.000	32	0.00	0.00	
		0.985				
			Total O2 req.		71.7	kmol
			O2 in feed		25.6	kmol
			O2 from air		46.2	kmol
					1.48	t
	Stoichiometric air requirement				6.34	t
			Actual air used		1.90	t
	Stoichiometric water production				0.51	t
Water if proportional to stoichiometric					0.15	t
	Adjusted water production				0.15	t/odt
					0.16	t/daf t

Appendix D .4 Pressurised gasification data sheet

PRESSURISED GASIFICATION CALCULATIONS SHEET									
Mass Balance							**Data**		
		In	%	Out	%		Feed elemental analysis, dry basis		
Feedstock	daf t/h	0.99	32.14%					%C	51.80%
Feed moisture	t/h	0.18	5.74%					%H	5.70%
Ash	t/h	0.01	0.36%					%O	40.90%
Oxidant air	t/h	1.90	61.76%					%N	0.10%
Dry gas	t/h			2.72	88.54%			%S	0.00%
Reaction water	t/h			0.15	5.00%			%Ash	1.10%
Feed moisture	t/h			0.18	5.74%		Feed MC	%wet bas	15%
Tars				0.01	0.33%		Feed HHV	GJ/odt	20.6
Char	t/h			0.00	0.04%		Feed LHV	GJ/odt	19.3
Ash	t/h			0.01	0.36%		Feed LHV	GJ/t	16.0
Total		3.08	100.00%	3.08	100.00%				
							Feed input	odt/h	1
Energy balance, at gasifier exit							Feed temperature	°C	20
		In	%	Out	%		Tars in gas	%feed	1.0%
Feed in (HHV)	GJ/h	20.56					Char in ash	%wt	10%
Feed moisture	GJ/h	0.00					Air factor		30%
Ash	GJ/h	0.00					H to H2	%	100%
Oxidant	GJ/h	0.38							
Dry flue gas	GJ/h			16.96	81.02%		Gasifier pressure	bar	25
Water vapour	GJ/h			2.23	10.64%		Gasifier exit	°C	900
Tars	GJ/h			0.23	1.11%		Air inlet	°C	200
Char	GJ/h			0.04	0.19%		Water Cp	kJ/kgK	4.18
Ash	GJ/h			0.01	0.04%		Latent heat	kJ/kg	2256
Losses	GJ/h			1.47	7.00%		Steam hg	kJ/kg	4150
Total	GJ/h	20.94		20.94	100.00%		Steam hg	kJ/kg	3360
							Ash Cp	kJ/kgK	0.94
Energy in gasifier exit stream			GJ/h	19.42			LHV/HHV	%	93%
Sensible losses, dry gas			GJ/h	1.78			Dry gas Cp	kJ/kgK	1.45
Sensible losses, water vapour			GJ/h	0.26			Dry air Cp	kJ/kgK	1.1
Latent heat losses			GJ/h	0.00			Engine entry	°C	450
Sensible heat losses, water			GJ/h	0.00			Tars HHV	GJ/t	23.3
Energy remaining in gas			GJ/h	17.38			Char HHV	GJ/t	32
Approximate LHV in gas			GJ/h	16.17					
							Approximate losses	odt/h in	losses
Feed in (LHV basis)			GJ/h	18.87				1	7%
Energy efficiency			%LHV	85.69%				50	2%
			%HHV	84.55%			Estimated losses	%feed LH	7.0%

Appendix D .4 (cont.)

Stoichiometry							
For	1.000	odt	requirement is		1.62	t/t as fed	
	1.176	t			1.90	t/odt	
	0.989	daf t			1.92	t/daf t	
	Elemental Input				O2 required		
	%daf t	daf t	kg/kmol	kmol	kmol		
C	52.4%	0.518	12	43.17	43.17		
H	5.8%	0.057	1	57.00	28.50		
O	41.4%	0.409	16	25.56	0.00		
N	0.1%	0.001	14	0.07	0.07		
S	0.0%	0.000	32	0.00	0.00		
		0.985					
			Total O2 req.		71.7	kmol	
			O2 in feed		25.6	kmol	
			O2 from air		46.2	kmol	
					1.48	t	
	Stoichiometric air requirement				6.34	t	
			Actual air used		1.90	t	
	Stoichiometric water production				0.51	t	
Water if proportional to stoichiometric					0.15	t	
		Adjusted water production			0.15	t/odt	
					0.16	t/daf t	

Appendix E Electricity Generation Module Data Sheets

Appendix E .1 Steam cycle performance data sheet

Calculations for basic Rankine cycle with superheat								
Steam conditions, turbine inlet								
	Flow rate	kg/s	76					
	Pressure	bar	83					
	Temperature	°C	485					
	Thermal inpu	MWth	240					
State Point 1 (Turbine entry)					**State Point 3 (Condenser exit)**			
	P1	bar	83			P3	bar	0.1
	T1	°C	485			Tsat	°C	45.8
	Tsat1	°C	233.8			vf*10^-2	m3/kg	0.10099
	Superheat	°C	251.2			hf3	kJ/kg	192
	Condition a	°C	450			sf3	kJ/kgK	0.649
		kJ/kg	3267			q(3-2)	kJ/kg	2165
		kJ/kgK	6.534					
	Condition b	°C	500		**State Point 4 (Boiler inlet)**			
		kJ/kg	3394			P4	bar	0.1
		kJ/kgK	6.703			w(4-1s)	kJ/kg	8.37
	h1	kJ/kg	3356			Isentropic efficiency		0.9
	s1	kJ/kgK	6.652			w(4-1)	kJ/kg	9.30
	q(1-4)	kJ/kg	3155			h4	kJ/kg	201
State Point 2/2s (Turbine exit)					**Results (ideal)**			
	P2	bar	0.1			Power output	MWe	95
	sf2	kJ/kgK	0.649			Spec. steam	kg/kW	2.88
	sg2	kJ/kgK	8.149			Thermal Effy	%	39.3%
	x2s		0.8004					
	hf2	kJ/kg	192					
	hg2	kJ/kg	2584					
	h2s	kJ/kg	2107					
	h(2s-1)	kJ/kg	1249					
	Isen. effy.		0.8					
	h(2-1)	kJ/kg	999					
	h2	kJ/kg	2357					
	x2		0.90					
	s2	kJ/kgK	7.44					

Appendix E .2 Steam cycle costs data sheet

TOTAL PLANT COSTS - STEAM CYCLE									
Factors used to give TPC	TPC/IPC	1.15							
	TPC/DPC	1.40							
	TPC/EC	4.83							
Source	[251]	[240]	[71]	[285]	[285]	[285]	[71]	[71]	[71]
Notes			Enkoping				Grenaa	Hollenba	WTE
								Generic	
Gross output, MWe	7.2	56	30.0	1.3	2.8	5.8	28.4	33.3	104.7
Internal consumption, MWe	0.8	6	1.8	0.1	0.3	0.6	1.7	3.3	4.7
Net output, MWe	6.4	50	28.2	1.2	2.5	5.2	26.7	30	100
%Gross output lost	10.6%	10.7%	6.0%	10.0%	10.0%	10.0%	6.0%	10.0%	4.5%
Published cost, *1000	5215	7708	10300	995	1800	3120	7398	7197	17632
Basis (1=EC,2=DPC,3=IPC,4=TPC)	3	3	2	2	2	2	2	2	1
Location	CA	US	US	US	US	US	Sweden	US	US
Year	1987	1986	1992	1986	1986	1986	1992	1991	1992
Currency	CA$	US$	US$	US$	US$	US$	US$	US$	US$
Cost Index, source year	102	94	113	94	94	94	113	111	113
Cost Index, 1995	129	122	122	122	122	122	122	122	122
Exchange rate, curr per USD	1.408	1	1	1	1	1	1	1	1
TPC now	5387	11505	15568	1808	3271	5669	11182	11074	91945
TPC/kW	841	230	552	1507	1308	1090	419	369	919

Appendix E .3 Gas turbine combined cycle performance data sheet

CALCULATION OF IGCC EFFICIENCIES							
Fuel input	odt/h	40.59					
Lower heating value of fuel	GJ/odt	18					
Energy input	MWth	203	H				
Gasifier efficiency	%	101%	g				
Gasifier energy to gas turbi	%	88%	z				
MWth in fuel gas	MWth	179	zgH				
Gas turbine cycle efficiency	%	40%					
MWe from GT	MWe	71	xzgH				
MWth in turbine exhaust	MWth	108	(1-x)zgH				
MWth from fuel gas	MWth	26	(1-z)gH				
Losses from turbine exhaust	MWth	38%					
Energy supplied to steam tur	MWth	93	(1-xz)gH				
Steam cycle efficiency	%	31%					
Steam cycle output	MWe	29.22	(1-xz)ygH				
Total system output	MWe	100.12					
Overall cycle efficiency		49.3%					
GTCC efficiency		55.9%					
% GT in total output		0.71					
RUN DATA							
Fuel input, odt/h	Fuel input	Dirty gas	Clean gas	GT out	ST out	IGCC Effy	GTCC Effy
	MWth	%	%	MWe	MWe	%	%
0.5	2.5	95.9%	83.9%	0.5	0.1	24.9%	29.7%
1.0	5.0	96.0%	84.0%	1.1	0.3	28.6%	34.1%
2.0	10.0	96.1%	84.1%	2.4	0.8	32.3%	38.4%
4.0	20.0	96.5%	84.4%	5.3	1.9	35.9%	42.5%
8.0	40.0	97.2%	85.1%	11.5	4.4	39.7%	46.6%
16.0	80.0	98.4%	86.1%	25.0	9.9	43.6%	50.6%
32.0	160.0	100.2%	87.7%	54.4	22.2	47.9%	54.6%

Appendix E .4 Gas turbine combined cycle costs data sheet

TOTAL PLANT COSTS - GAS TURBINE COMBINED CYCLE PLANT										
Factors used to give TPC										
TPC/IPC	1.15									
TPC/DPC	1.40									
TPC/EC	4.83									
Source	[297]	[297]	[297]	[298]	[298]	[298]	[203]	[98]	Braeult (B	[299]
Notes	Quote	Quote	Quote	Estimat	Estimat	Estimat	Estimat	Estimat	Estimat	Quote
Gross output, MWe	20.7	41.5	62.2	78.8	57.0	111.9	155.4	128.0	65.6	8.1
Internal consumption, MWe	0.7	1.5	2.2	2.8	2.0	3.9	5.4	4.5	2.3	0.3
Net output, MWe	20.0	40.0	60.0	76.0	55.0	108.0	150.0	123.5	63.3	7.8
%Gross output lost	3.5%	3.5%	3.5%	3.5%	3.5%	3.5%	3.5%	3.5%	3.5%	3.5%
Published cost, *1000	103986	155086	197450	106600	85700	90000	120000	80300	59814	1150
Basis (1=EC,2=DPC,3=IPC,4=	2	2	2	3	3	3	3	3	3	1
Location	Finland	Finland	Finland	US	US	US	US	US	US	UK
Year	1992	1993	1994	1991	1991	1991	1992	1992	1993	1991
Currency	FIM	FIM	FIM	US$	US$	US$	US$	US$	US$	GBP
Cost Index, source year	148	148	148	111	111	111	113	113	116	149
Cost Index, 1995	160	160	160	122	122	122	122	122	122	170
Exchange rate, curr per U	4.584	4.584	4.584	1	1	1	1	1	1	0.632
TPC now	34333	51205	65193	134739	108322	113757	148991	99700	72344	10027
TPC/kW	1657	1235	1049	1711	1901	1016	959	779	1103	1241

Appendix F Integrated Biomass to Electricity System Models

Appendix F .1 Combustion System by Modules, 20 MW_e

	A	B	C	D	E	F
1	**SYSTEM INPUT**					
2	**Parameter**		**Units**	**all sites**	**single site**	
3	User inputs					
4		Feed production rate	odt/y	141625		Input
5		Feed moisture content	%wet	50.0%		Input
6		Feed cost before transport	$/odt	40.00		Input
7		Feed lower heating value	GJ/odt dry	19.3		Input
8	System inputs					
9		No. conversion sites		1		
10	Feed composition					
11		C	%odt	52.0%		
12		H	%odt	5.8%		
13		O	%odt	41.0%		
14		N	%odt	0.1%		
15		S	%odt	0.0%		
16		Ash	%odt	1.1%		
17	Mass flows					
18		Total feed input	odt/y	141625	141625	
19		Total moisture input	t/y	141625	141625	
20		Total wet feed input	t/y	283250	283250	
21		Total ash input	t/y	1558	1558	
22	Energy flows					
23		Lower heating value as delivere	GJ/t	8.42		
24		Energy delivered	GJ/y	2385815	2385815	
25	Feed cost					
26		Total cost to system	$k/y	5665	5665	
27		Specific costs	$/odt	40.00		
28			$/t	20.00		
29			$/GJ	2.37		

Appendix F .1 (cont.)

	A	B	C	D	E	F
1	**FEED TRANSPORT MODULE**					
2	**Parameter**		**Units**	**all sites**	**single site**	
3	User input					
4		Use default distance		TRUE		Input
5		Over-ride distance	km	40		Input
6		Plantation area	%	80%		Input
7		Land area limitation	%	5%		Input
8		Land yield	odt/ha.y	10		Input
9		Feed loading cost	$/t	2.60		Input
10		Feed transport cost	$/t.km	0.09		Input
11		Feed bulk density	odt/m3	0.150		Input
12	System input					
13		Total feed transported	odt/y	141625	141625	
14		Moisture content as transporte	%wet	50.0%		
15		No. of conversion sites		1		
16		No. of generation sites		1		
17	Calculations					
18		Planted land	ha	14162		
19		Plantation area	ha	17703		
20		Total land	ha	354062		
21		Radius total land	km	18.94		
22		Direct distance to conversion s	km	12.78		
23		Actual distance travelled	km	18.07		
24		Distance used in model	km	18.07		
25		Bulk density	t/m3	0.300		
26		Total mass transported	t/y	283250	283250	
27		Loading cost	$k/y	736	736	
28		Transport cost	$k/y	461	461	
29		Total transport cost	$k/y	1197	1197	

Appendix F .1 (cont.)

	A	B	C	D	E	F
1	FEED PRETREATMENT MODULE					
2	**Parameter**		**Units**	**all sites**	**single site**	
3	User input					
4		Delivery days/week	d/week	5		Input
5		Delivery weeks/year	weeks/y	52		Input
6		Delivery hours/day	h/d	10		Input
7		Dry matter losses	%odt/month	2.50%		Input
8		MC change in storage	%wet/month	0%		Input
9		Overs in feed	%	5%		Input
10		Dryer energy required	GJ/twe	3.553		Input
11		Diesel fuel cost	$/l	0.51		Input
12		Current costs for plant no.		100		Input
13		Learning factor		20%		Input
14	System input					
15		Delivered feed input	odt/y	141625	141625	
16		Delivered feed moisture content	%wet	50%		
17		Delivered feed heating value	GJ/odt	19.3		
18		No. conversion sites		1		
19		Bulk density	odt/m3	0.150		
20		Conversion operating hours	h/y	7884		
21		Drying required?		TRUE		
22	Mass flows					
23		Total feed delivered	odt/y	141625	141625	
24			t/y	283250	283250	
25		Delivery days/year	d/y	260		
26		Feed delivered	t/d	1089	1089	
27			t/h	109	109	
28		Storage period	weeks	3		
29		DM losses in storage	%odt	1.7%		
30		MC change in storage	%wet	0.0%		
31		MC out of storage	%wet	50.0%		
32		Feed from storage	odt/y	139174	139174	
33			t/y	278348	278348	
34			t/h	35	35	
35		MC from dryer	%wet	35%		
36		Moisture into dryer	t/h	17.65	17.65	
37		Moisture out of dryer	t/h	9.51	9.51	
38		Drying rate required	twe/h	8.15	8.15	
39		Dried feed	t/y	214114	214114	
40			t/h	27	27	
41	Volumetric flows, per site					
42		Feed bulk density as delivered	t/m3	0.300		
43		Feed bulk density from storage	t/m3	0.300		
44		Feed bulk density as required	t/m3	0.231		
45		Volume delivered	m3/h	363	363	
46		Volume from store	m3/h	118	118	
47		Volume from dryer	m3/h	118	118	
48	Energy flows					
49		Feed heating value, delivered	GJ/t	8.42		
50			GJ/y	2385815	2385815	
51		Feed energy from storage	GJ/t	8.42		
52			GJ/y	2344522	2344522	
53		Drying energy required	GJ/h	29	29	
54			GJ/y	228224	228224	
55		Feed heating value, reactor	GJ/t	11.69		
56			GJ/y	2502152	2502152	
57	Equipment requirements, per site					
58		No. dryers			1	
59		Drying rate per dryer	twe/h		8.1	
60		No. grinders			0	
61		No. buffer storage silos			1	
62		No. front end loaders			1	

Appendix F .1 (cont.)

	A	B	C	D	E	F
63		No. bulldozers			1	
64	Total plant costs					
65		Reception	$k	1687	1687	
66		Screening	$k		650	
67		Storage	$k	4527	4527	
68		Screening	$k	383	383	
69		Re-chipping	$k	236	236	
70		Drying	$k	1470	1470	
71		Grinding	$k	0	0	
72		Buffer storage	$k	532	532	
73		Total cost per site	$k	9485	9485	
74		TPC at n = 0		41772	41772	
75		TOTAL, nth replication	$k	9485	9485	
76	Capital charges					
77		Capital amortisation, fixed	$k/y	1114	1114	
78		Capital amortisation, adjusted	$k/y	729	729	
79		Overheads	$k/y	379	379	
80		Maintenance	$k/y	379	379	
81	Labour costs					
82		Labour req., reception	/shift	0.2	0.2	
83		Labour req., screening	/shift	0.0	0.0	
84		Labour req., storage	/shift	0.8	0.8	
85		Labour req., screening	/shift	0.2	0.2	
86		Labour req., re-chip	/shift	0.3	0.3	
87		Labour req., drying	/shift	0.5	0.5	
88		Labour req., grinding	/shift	0.0	0.0	
89		Labour req., buffer storage	/shift	0.2	0.2	
90		Labour req., total per site	/shift	2.2	2.2	
91		Total labour cost	$k/y	335	335	
92	Utilities					
93		Utilities - diesel	l/y	237250	237250	
94			$k/y	121	121	
95		Installed power, reception	kWe	123	123	
96		Installed power, screening	kWe	18	18	
97		Installed power, storage	kWe	26	26	
98		Installed power, screening	kWe	8	8	
99		Installed power, re-chip	kWe	30	30	
100		Installed power, drying	kWe	70	70	
101		Installed power, grinding	kWe	0	0	
102		Installed power, buffer storage	kWe	20	20	
103		TOTAL	kWe	295	295	
104		Power consumed per site	MWe	0.3	0.3	
105			MWh/y	1439	1439	

Appendix F .1 (cont.)

	A	B	C	D	E	F
1	COMBUSTION MODULE					
2	Parameter		Units	all sites	single site	
3	User input					
4		Availability	%	90%		Input
5		Use default boiler efficiency		TRUE		Input
6		Over-ride efficiency	%	94%		Input
7		No. conversion sites		1		Input
8		Internal power consumption	%MWth	2%		Input
9		Use drying to 35%		TRUE		Input
10		Current costs for plant no.		100		Input
11		Learning factor		20%		Input
12	System input					
13		Total reactor feed	odt/y	139174	139174	
14		Feed LHV	GJ/odt dry	19.3		
15		Moisture content as fed	%wet	35%		
16	Mass flows					
17		Operating hours	h/y	7884		
18		Reactor feed	t/y	214114	214114	
19			t/h	27.2	27.2	
20			odt/h	17.7	17.7	
21	Energy flows					
22		Reactor feed LHV	GJ/t	11.69		
23		Energy input	GJ/y	2502152	2502152	
24			GJ/h	317	317	
25			MWth	88	88	
26		Default boiler efficiency	%		85.2%	
27		Boiler efficiency used	%		85.2%	
28		Energy added to steam	MWth	75	75	
29			GJ/h	270	270	
30			GJ/y	2131993	2131993	
31	Capital cost					
32		Total plant cost, now	$k1995	21932	21932	
33		TPC, 1st plant		96592	96592	
34		Total plant cost, future	$k1995	21932	21932	
35	Operating costs					
36		Capital amortisation, fixed	$k/y	2576	2576	
37		Capital amortisation, adjusted	$k/y	1686	1686	
38		Labour requirement, all sites		1.4	1.4	
39		Total labour cost	$k/y	216	216	
40		Power consumed per site	MWe	1.5	1.5	
41			MWh/y	11844	11844	
42		Overheads	$k/y	877	877	
43		Maintenance	$k/y	877	877	

Appendix F .1 (cont.)

	A	B	C	D	E	F
1	**STEAM CYCLE MODULE**					
2	**Parameter**		**Units**	**all sites**	**single site**	
3	User inputs					
4		Use default efficiency		TRUE		Input
5		Over-ride efficiency		33%		Input
6		Internal power consumption	%	4%		Input
7		Maintenance cost	$/kWh	0.004		Input
8		Current costs for plant no.		100		Input
9		Learning factor		20%		Input
10	System inputs					
11		Thermal input	GJ/y	2131993	2131993	
12		Operating hours	h/y	7884		
13		No. generation sites		1		
14	Energy flows					
15		Energy input	GJ/h	270	270	
16			MWth	75	75	
17		Default efficiency	%		30.2%	
18		Efficiency used	%		30.2%	
19		Gross electricity per site	MWe	23	22.7	
20			MWh/y	179006	179006	
21			GJ/y	644422	644422	
22	Capital costs					
23		Total plant cost per site	$k	12318	12318	
24		TPC, 1st plant		54250	54250	
25		Total plant cost	$k	12318	12318	
26	Operating costs, all sites					
27		Capital amortisation, fixed	$k/y	1447	1447	
28		Capital amortisation, adjusted	$k/y	947	947	
29		Labour requirement per site	/shift	1.46	1.46	
30		Labour cost	$k/y	218	218	
31		Cooling water @ 0.017 $/t	$k/y	15	15	
32		Boiler feed water @ 1.03$/t	$k/y	277	277	
33		Internal power per site	MWe	1	0.9	
34			MW/y	7160	7160	
35		Overheads	$k/y	493	493	
36		Maintenance	$k/y	716	716	

Appendix F .1 (cont.)

	A	B	C	D	E	F
1	GRID CONNECTION					
2	**Parameter**		**Units**	**all sites**	**single site**	
3	User inputs					
4		Current costs for plant no.		100		Input
5		Learning factor		20%		Input
6	System inputs					
7		No. sites		1		
8		Gross capacity	MWe	23	23	
9			MWh/y	179006	179006	
10		Pretreatment power	MWe	0.3	0.3	
11			MWh/y	1439	1439	
12		Conversion power	MWe	1.5	1.5	
13			MWh/y	11844	11844	
14		Generation power	MWe	0.9	0.9	
15			MWh/y	7160	7160	
16	Performance					
17		Total internal consumption	MWe	2.7	2.7	
18			%	11.9%	11.9%	
19			MWh/y	20444	20444	
20			%	11.4%	11.4%	
21		Net capacity	MWe	20.0	20.0	
22			MWh/y	158562	158562	
23			GJ/y	570823	570823	
24	Capital costs					
25		Total plant cost, 1st plant	$k	1729	1729	
26		TPC, 1st plant		7613	7613	
27		Total plant cost	$k	1729	1729	
28	Operating costs					
29		Capital amortisation, fixed	$k/y	203	203	
30		Capital amortisation, adjusted	$k/y	133	133	
31		Labour requirement		0.00	0.00	
32		Total labour cost	$k/y	0	0	
33		Overheads	$k/y	69	69	
34		Maintenance	$k/y	69	69	

Appendix F .2 Gas-Eng System by Modules, 20 MW_e

	A	B	C	D	E	F
1	SYSTEM INPUT					
2	Parameter		Units	all sites	single site	
3	User inputs					
4		Feed production rate	odt/y	113140		Input
5		Feed moisture content	%wet	50.0%		Input
6		Feed cost before transport	$/odt	40.00		Input
7		Feed lower heating value	GJ/odt dry	19.3		Input
8	System inputs					
9		No. conversion sites		1		
10	Feed composition					
11		C	%odt	52.0%		
12		H	%odt	5.8%		
13		O	%odt	41.0%		
14		N	%odt	0.1%		
15		S	%odt	0.0%		
16		Ash	%odt	1.1%		
17	Mass flows					
18		Total feed input	odt/y	113140	113140	
19		Total moisture input	t/y	113140	113140	
20		Total wet feed input	t/y	226279	226279	
21		Total ash input	t/y	1245	1245	
22	Energy flows					
23		Lower heating value as delivere	GJ/t	8.42		
24		Energy delivered	GJ/y	1905949	1905949	
25	Feed cost					
26		Total cost to system	$k/y	4526	4526	
27		Specific costs	$/odt	40.00		
28			$/t	20.00		
29			$/GJ	2.37		

Appendix F .2 (cont.)

	A	B	C	D	E	F
1	FEED TRANSPORT MODULE					
2	**Parameter**		**Units**	**all sites**	**single site**	
3	User input					
4		Use default distance		TRUE		Input
5		Over-ride distance	km	40		Input
6		Plantation area	%	80%		Input
7		Land area limitation	%	5%		Input
8		Land yield	odt/ha/y	10		Input
9		Feed loading cost	$/t	2.6		Input
10		Feed transport cost	$/t.km	0.09		Input
11		Feed bulk density	odt/m3	0.150		Input
12	System input					
13		Total feed transported	odt/y	113140	113140	
14		Moisture content as transporte	%wet	50.0%		
15		No. of conversion sites		1		
16	Calculations					
17		Planted land	ha	11314		
18		Plantation land	ha	14142		
19		Total land	ha	282849		
20		Radius total land	km	16.93		
21		Direct distance to conversion s	km	11.42		
22		Actual distance travelled	km	16.15		
23		Distance used in model	km	16.15		
24		Bulk density	t/m3	0.3		
25		Total mass transported	t/y	226279	226279	
26		Loading cost	$k/y	588	588	
27		Transport cost	$k/y	329	329	
28		Total transport cost	$k/y	917	917	

Appendix F .2 (cont.)

	A	B	C	D	E	F
1	FEED PRETREATMENT MODULE					
2	Parameter		Units	all sites	single site	
3	User input					
4		Delivery days/week	d/week	5		Input
5		Delivery weeks/year	weeks/y	52		Input
6		Delivery hours/day	h/d	10		Input
7		Dry matter losses	%odt/month	2.50%		Input
8		MC change in storage	%wet/month	0%		Input
9		Overs in feed	%	5%		Input
10		Dryer energy required	GJ/twe	3.553		Input
11		Diesel fuel cost	$/l	0.51		Input
12		Current costs for plant no.		100		Input
13		Learning factor		20%		Input
14	System input					
15		Delivered feed input	odt/y	113140	113140	
16		Delivered feed moisture content	%wet	50%		
17		Delivered feed heating value	GJ/odt	19.3		
18		No. conversion sites		1		
19		Bulk density	odt/m3	0.150		
20		Conversion operating hours	h/y	7884		
21		Drying required?		TRUE		
22	Mass flows					
23		Total feed delivered	t/y	226279	226279	
24		Delivery days/year	d/y	260		
25		Feed delivered	t/d	870	870	
26			t/h	87	87	
27		Storage period	weeks	3		
28		DM losses in storage	%odt	1.7%		
29		MC change in storage	%wet	0.0%		
30		MC out of storage	%wet	50.0%		
31		Feed from storage	odt/y	111181	111181	
32			t/y	222363	222363	
33			t/h	28.2	28.2	
34		MC from dryer	%wet	15%		
35		Moisture into dryer	t/h	14.1	14.1	
36		Moisture out of dryer	t/h	2.5	2.5	
37		Drying rate required	twe/h	11.6	11.6	
38		Dried feed	t/y	130802	130802	
39			t/h	17	16.6	
40	Volumetric flows, per site					
41		Feed bulk density as delivered	t/m3	0.300		
42		Feed bulk density from storage	t/m3	0.300		
43		Feed bulk density as required	t/m3	0.176		
44		Volume delivered	m3/h	290	290	
45		Volume from store	m3/h	94	94	
46		Volume from dryer	m3/h	94	94	
47	Energy flows					
48		Feed heating value, delivered	GJ/t	8.42		
49			GJ/y	1905949	1905949	
50		Feed heating value, ex. store	GJ/t	8.42		
51			GJ/y	1872961	1872961	
52		Drying energy required per site	GJ/h	41	41	
53			GJ/y	325317	325317	
54		Feed heating value, reactor	GJ/t	16.04		
55			GJ/y	2097652	2097652	
56	Equipment requirements, per site					
57		No. dryers			1	
58		Drying rate per dryer	twe/h		11.6	
59		No. grinders			0	
60		No. buffer storage silos			1	
61		No. front end loaders			1	
62		No. bulldozers			1	

Appendix F .2 (cont.)

	A	B	C	D	E	F
63	Total plant costs					
64		Reception	$k	1687	1687	
65		Screening	$k	588	588	
66		Storage	$k	4230	4230	
67		Screening	$k	354	354	
68		Re-chipping	$k	218	218	
69		Drying	$k	1977	1977	
70		Grinding	$k	0	0	
71		Buffer storage	$k	492	492	
72		Total plant cost, now	$k	9546	9546	
73		Total plant costs, 1st plant	$k	42042	42042	
74		Total plant cost, future	$k	9546	9546	
75	Capital charges, all sites					
76		Capital amortisation, fixed	$k/y	1121	1121	
77		Capital amortisation, adjusted	$k/y	734	734	
78		Overheads	$k/y	382	382	
79		Maintenance	$k/y	382	382	
80	Labour costs					
81		Labour req., reception	/shift	0.4	0.4	
82		Labour req., screening	/shift	0.1	0.1	
83		Labour req., storage	/shift	2.5	2.5	
84		Labour req., screening	/shift	0.4	0.4	
85		Labour req., re-chip	/shift	0.5	0.5	
86		Labour req., drying	/shift	0.3	0.3	
87		Labour req., grinding	/shift	0.0	0.0	
88		Labour req., buffer storage	/shift	0.2	0.2	
89		Labour req., total	/shift	4.3	4.3	
90		Total labour cost	$k/y	643	643	
91	Utilities					
92		Utilities - diesel	l/y	237250	237250	
93			$k/y	121	121	
94		Installed power, screening	kWe	20	20	
95		Installed power, reception	kWe	123	123	
96		Installed power, storage	kWe	22	22	
97		Installed power, screening	kWe	7	7	
98		Installed power, re-chip	kWe	27	27	
99		Installed power, drying	kWe	98	98	
100		Installed power, grinding	kWe	0	0	
101		Installed power, buffer storage	kWe	17	17	
102		TOTAL	kWe	295	295	
103		Internal power consumption	MWe/site	0.3	0.3	
104			MWh/y/site	1556	1556	

	A	B	C	D	E	F
1	ATMOSPHERIC GASIFICATION MODULE					
2	Parameter		Units	all sites	single site	
3	User input					
4		Availability	%	90%		Input
5		Use default gas'n efficiency		TRUE		Input
6		Over-ride efficiency	%	65%		Input
7		No. gasification sites		1		Input
8		Current costs for plant no.		1		Input
9		Learning factor		20%		Input
10	System input					
11		Total reactor feed	odt/y	111181	111181	
12		Feed LHV	GJ/odt dry	19.3		
13		Moisture content as fed	%wet	15%		
14	Mass flows					
15		Operating hours	h/y	7884		
16		Reactor feed	t/y	130802	130802	
17			t/h	16.6	16.6	
18			odt/h	14.1	14.1	
19	Energy flows					
20		Reactor feed LHV	GJ/t	16.0		
21		Energy input	GJ/y	2097652	2097652	
22			GJ/h	266.1	266.1	
23			MWth	73.9	73.9	
24		Default efficiency	%LHV		73.3%	
25		Actual efficiency	%LHV		73.3%	
26		Energy output	MWth	54.2	54.2	
27			GJ/h	195	195	
28			GJ/y	1538352	1538352	
29	Capital cost					
30		Total plant cost, now	$k	40617	40617	
31		Total plant costs, 1st plant	$k	40617	40617	
32		Total plant cost, future	$k	40617	40617	
33	Operating costs					
34		Capital amortisation, fixed	$k/y	4771	4771	
35		Capital amortisation, adjusted	$k/y	3121	3121	
36		Labour requirement, all sites	/shift	3.7	3.66	
37		Total labour cost	$k/y	548	548	
38		Cooling water @ 0.0017 $/t	$k/y	0.0	0.0	
39		Internal power consumption	MWe	0.6	0.6	
40			MWh/y	4447	4447	
41		Catalyst consumption	t/odt	0.07		
42		Catalyst cost	$k/y	249	249	
43		Overheads	$k	1625	1625	
44		Maintenance	$k	1625	1625	

Appendix F .2 (cont.)

	A	B	C	D	E	F
1	GAS-FIRED DUAL FUEL ENGINE					
2	**Parameter**		**Units**	**all sites**	**single site**	
3	User inputs					
4		Use default efficiency		TRUE		Input
5		Over-ride efficiency		32%		Input
6		Energy supplied by diesel	%	5%		Input
7		Internal energy consumption	%	3%		Input
8		Maintenance cost	$/kWh	0.0122		Input
9		Current costs for plant no.		100		Input
10		Learning factor		20%		Input
11	System inputs					
12		No. generating sites		1		
13		Fuel gas input	GJ/y	1538352	1538352	
14		Cost of diesel	$/l	0.51		
15		Operating hours	h/y	7884		
16	Energy flows					
17		Energy input	GJ/h	195	195	
18		Diesel energy per site	GJ/h	10	10	
19			GJ/y	80966	80966	
20		Total energy input	GJ/h	205	205	
21		No. engines			2	
22		Energy input per engine	GJ/h		102.7	
23			MWth		28.5	
24		Default efficiency	%		37.7%	
25		Efficiency used	%		37.7%	
26		Output per engine	MWe		11	
27		Gross power output	MWe	22	22	
28			MWh/y	169534	169534	
29			GJ/y	610324	610324	
30	Mass flows					
31		Diesel	t/y	1901	1901	
32	Capital costs					
33		TPC, per engine	$k		12210	
34		Total plant cost, now	$k	22785	22785	
35		Total plant costs, 1st plant	$k	100346	100346	
36		Total plant cost, future	$k	22785	22785	
37	Operating costs					
38		Capital amortisation, fixed	$k/y	2676	2676	
39		Capital amortisation, adjusted	$k/y	1751	1751	
40		Labour requirement, per site	/shift	2.10	2.10	
41		Labour requirement, all sites	$k/y	315	315	
42		Diesel cost	$k/y	1140	1140	
43		Overheads	$k/y	911	911	
44		Maintenance	$k/y	2068	2068	
45		Internal power consumption	MWe	0.6	0.6	
46			MW/y	5086	5086	

Appendix F .2 (cont.)

	A	B	C	D	E	F
1	GRID CONNECTION MODULE					
2	**Parameter**		**Units**	**all sites**	**single site**	
3	User inputs					
4		Current costs for plant no.		100		Input
5		Learning factor		20%		Input
6	System inputs					
7		No. generating sites		1		
8		Gross capacity	MWe	21.5	21.5	
9		Gross power output	MWh/y	169534	169534	
10		Power req., pretreatment	MWe	0.3	0.3	
11		Power consumed, pretreatment	MWh/y	1556	1556	
12		Power req., gasification	MWe	0.6	0.6	
13		Power consumed, gasification	MWh/y	4447	4447	
14		Power req., engine	MWe	0.6	0.6	
15		Power consumed, engine	MWh/y	5086	5086	
16	Performance					
17		Total internal consumption	MWe	1.5	1.5	
18			%	7.0%	7.0%	
19			MWh/y	11089	11089	
20			%	6.5%	6.5%	
21		Net capacity	MWe	20	20	
22			MWh/y	158445	158445	
23			GJ/y	570403	570403	
24	Capital costs					
25		Total plant cost, now	$k	1729	1729	
26		Total plant costs, 1st plant	$k	7613	7613	
27		Total plant cost, future	$k	1729	1729	
28	Operating costs					
29		Capital amortisation, fixed	$k/y	203	203	
30		Capital amortisation, adjusted	$k/y	133	133	
31		Labour requirement		0	0	
32		Total labour cost	$k/y	0	0	
33		Overheads	$k/y	69	69	
34		Maintenance	$k/y	69	69	

Appendix F .3 IGCC System by Modules, 20 MW_e

	A	B	C	D	E	F
1	SYSTEM INPUT					
2	**Parameter**		**Units**	**all sites**	**single site**	
3	User inputs					
4		Feed production rate	odt/y	78424		Input
5		Feed moisture content	%wet	50.0%		Input
6		Feed cost before transport	$/odt	40.00		Input
7		Feed lower heating value	GJ/odt dry	19.3		Input
8	System inputs					
9		No. conversion sites		1		
10	Feed composition					
11		C	%odt	52.0%		
12		H	%odt	5.8%		
13		O	%odt	41.0%		
14		N	%odt	0.1%		
15		S	%odt	0.0%		
16		Ash	%odt	1.1%		
17	Mass flows					
18		Total feed input	odt/y	78424	78424	
19		Total moisture input	t/y	78424	78424	
20		Total wet feed input	t/y	156849	156849	
21		Total ash input	t/y	863	863	
22	Energy flows					
23		Lower heating value as delivere	GJ/t	8.42		
24		Energy delivered	GJ/y	1321136	1321136	
25	Feed cost					
26		Total cost to system	$k/y	3137	3137	
27		Specific costs	$/odt	40.00		
28			$/t	20.00		
29			$/GJ	2.37		
30						

Appendix F .3 (cont.)

	A	B	C	D	E	F
1	FEED TRANSPORT MODULE					
2	**Parameter**		**Units**	**all sites**	**single site**	
3	User input					
4		Use default distance		TRUE		Input
5		Over-ride distance	km	40		Input
6		Plantation area	%	80%		Input
7		Land area limitation	%	5%		Input
8		Land yield	odt/ha.y	10		Input
9		Feed loading cost	$/t	2.6		Input
10		Feed transport cost	$/t.km	0.09		Input
11		Feed bulk density	odt/m3	0.150		Input
12	System input					
13		Total feed transported	odt/y	78424	78424	
14		Moisture content as transported	%wet	50.0%		
15		No. of conversion sites		1		
16	Calculations					
17		Planted land	ha	7842		
18		Plantation land	ha	9803		
19		Total land	ha	196061		
20		Radius total land	km	14.09		
21		Direct distance to conversion s	km	9.51		
22		Actual distance travelled	km	13.45		
23		Distance used in model	km	13.45		
24		Bulk density	t/m3	0.3		
25		Total mass transported	t/y	156849	156849	
26		Loading cost	$k/y	408	408	
27		Transport cost	$k/y	190	190	
28		Total transport cost	$k/y	598	598	

Appendix F .3 (cont.)

	A	B	C	D	E	F
1	FEED PRETREATMENT MODULE					
2	Parameter		Units	all sites	single site	
3	User input					
4		Delivery days/week	d/week	5		Input
5		Delivery weeks/year	weeks/y	52		Input
6		Delivery hours/day	h/d	10		Input
7		Dry matter losses	%odt/month	2.50%		Input
8		MC change in storage	%wet/month	0%		Input
9		Overs in feed	%	5%		Input
10		Dryer energy required	GJ/twe	3.553		Input
11		Diesel fuel cost	$/l	0.51		Input
12		Current costs for plant no.		100		Input
13		Learning factor		20%		Input
14	System input					
15		Delivered feed input	odt/y	78424	78424	
16		Delivered feed moisture content	%wet	50%		
17		Delivered feed heating value	GJ/odt	19.3		
18		No. conversion sites		1		
19		Bulk density	odt/m3	0.150		
20		Conversion operating hours	h/y	7884		
21		Drying required?		TRUE		
22	Mass flows					
23		Total feed delivered	t/y	156849	156849	
24		Delivery days/year	d/y	260		
25		Feed delivered per site	t/d	603	603	
26				60	60	
27		Storage period	weeks	3		
28		DM losses in storage	%odt	1.7%		
29		MC change in storage	%wet	0.0%		
30		MC out of storage	%wet	50.0%		
31		Feed from storage per site	odt/y	77067	77067	
32			t/y	154134	154134	
33			t/h	19.6	19.6	
34		MC from dryer	%wet	15%		
35		Moisture into dryer	t/h	9.8	9.8	
36		Moisture out of dryer	t/h	1.7	1.7	
37		Drying rate required	twe/h	8.1	8.1	
38		Dried feed	t/y	90667	90667	
39			t/h	11.5	11.5	
40	Volumetric flows, per site					
41		Feed bulk density as delivered	t/m3	0.300		
42		Feed bulk density from storage	t/m3	0.300		
43		Feed bulk density as required	t/m3	0.176		
44		Volume delivered	m3/h	201	201	
45		Volume from store	m3/h	65	65	
46		Volume from dryer	m3/h	65	65	
47	Energy flows					
48		Feed heating value, delivered	GJ/t	8.42		
49			GJ/y	1321136	1321136	
50		Feed heating value, ex. store	GJ/t	8.42		
51			GJ/y	1298270	1298270	
52		Drying energy required per site	GJ/h	29	29	
53			GJ/y	225498	225498	
54		Feed heating value, reactor	GJ/t	16.04		
55			GJ/y	1454018	1454018	
56	Equipment requirements, per site					
57		No. dryers			1	
58		Drying rate per dryer	twe/h		8.1	
59		No. grinders			0	
60		No. buffer storage silos			1	
61		No. front end loaders			1	
62		No. bulldozers			0	

Appendix F .3 (cont.)

	A	B	C	D	E	F
63	Total plant costs					
64		Reception	$k	1004	1004	
65		Screening	$k	0	0	
66		Storage	$k	972	972	
67		Screening	$k	312	312	
68		Re-chipping	$k	193	193	
69		Drying	$k	1446	1446	
70		Grinding	$k	0	0	
71		Buffer storage	$k	432	432	
72		Total plant cost, now	$k1995	4359	4359	
73		Total plant costs, 1st plant	$k1995	19197	19197	
74		Total plant cost, future	$k1995	4359	4359	
75	Capital charges, all sites					
76		Capital amortisation, fixed	$k/y	512	512	
77		Capital amortisation, adjusted	$k/y	335	335	
78		Overheads	$k/y	174	174	
79		Maintenance	$k/y	174	174	
80	Labour costs					
81		Labour req., reception	/shift	0.4	0.4	
82		Labour req., screening	/shift	0.0	0.0	
83		Labour req., storage	/shift	1.2	1.2	
84		Labour req., screening	/shift	0.4	0.4	
85		Labour req., re-chip	/shift	0.5	0.5	
86		Labour req., drying	/shift	0.3	0.3	
87		Labour req., grinding	/shift	0.0	0.0	
88		Labour req., buffer storage	/shift	0.2	0.2	
89		Labour req., total	/shift	2.9	2.9	
90		Total labour cost	$k/y	435	435	
91	Utilities					
92		Utilities - diesel	l/y	109500	109500	
93			$k/y	56	56	
94		Installed power, reception	kWe	79	79	
95		Installed power, screening	kWe	0	0	
96		Installed power, storage	kWe	17	17	
97		Installed power, screening	kWe	6	6	
98		Installed power, re-chip	kWe	23	23	
99		Installed power, drying	kWe	68	68	
100		Installed power, grinding	kWe	0	0	
101		Installed power, buffer storag	kWe	13	13	
102		TOTAL, per site	kWe	206	206	
103		Power consumed per site	MWe	0.206	0.206	
104			MWh/y	1120	1120	

Appendix F .3 (cont.)

	A	B	C	D	E	F
1	PRESSURISED GASIFICATION MODULE					
2	**Parameter**		**Units**	**all sites**	**single site**	
3	User input					
4		Availability	%	90%		Input
5		Use default gas'n efficiency		TRUE		Input
6		Over-ride efficiency	%	78%		Input
7		No. gasification sites		1		Input
8		Current costs for plant no.		1		Input
9		Learning factor		20%		Input
10	System input					
11		Total reactor feed	odt/y	77067	77067	
12		Feed LHV	GJ/odt dry	19.3		
13		Moisture content as fed	%wet	15%		
14	Mass flows					
15		Operating hours	h/y	7884		
16		Reactor feed	t/y	90667	90667	
17			t/h	11.5	11.5	
18			odt/h	9.8	9.8	
19	Energy flows					
20		Reactor feed LHV	GJ/t	16.0		
21		Energy input	GJ/y	1454018	1454018	
22			GJ/h	184	184	
23			MWth	51	51	
24		Default efficiency	%LHV		87.1%	
25		Actual efficiency	%LHV		87.1%	
26		Energy output	MWth	45	44.6	
27			GJ/h	161	161	
28			GJ/y	1266889	1266889	
29	Capital cost					
30		Total plant cost, now	$k	45283	45283	
31		Total plant costs, 1st plant	$k	45283	45283	
32		Total plant cost, future	$k	45283	45283	
33	Operating costs					
34		Capital amortisation, fixed	$k/y	5319	5319	
35		Capital amortisation, adjusted	$k/y	3480	3480	
36		Labour requirement, all sites		3.1	3.1	
37		Total labour cost		461	461	
38		Internal power consumption	MWe	0.6	0.6	
39			MWh/y	4624	4624	
40		Overheads	$k	1811	1811	
41		Maintenance	$k	1811	1811	

Appendix F .3 (cont.)

	A	B	C	D	E	F
1	GAS TURBINE COMBINED CYCLE MODULE					
2	**Parameter**		**Units**	**all sites**	**single site**	
3	User inputs					
4		Use default efficiency		TRUE		Input
5		Over-ride efficiency		40%		Input
6		Internal power consumption	%	3%		Input
7		Maintenance cost	$/kWh	0.01		Input
8		Current costs for plant no.		100		Input
9		Learning factor		20%		Input
10	System inputs					
11		Operating hours	h/y	7884		
12		No. generating sites		1		
13		Fuel gas input	GJ/y	1266889	1266889	
14	Performance					
15		Energy input	GJ/h	161	161	
16			MWth	45	44.6	
17		Default efficiency	%		48.0%	
18		Efficiency used	%		48.0%	
19		Gross power output	MWe	21	21	
20			MWh/y	169001	169001	
21			GJ/y	608404	608404	
22	Capital costs					
23		Total plant cost, now	$k	29953	29953	
24		Total plant costs, 1st plant	$k	131914	131914	
25		Total plant cost, future	$k	29953	29953	
26	Operating costs					
27		Capital amortisation, fixed	$k/y	3518	3518	
28		Capital amortisation, adjusted	$k/y	2302	2302	
29		Labour requirement		1.78	1.78	
30		Total labour cost	$k/y	267	267	
31		Cooling water at 0.0017 $/t	$k/y	0.5	0.5	
32		Boiler feed water @ 1.03$/t	$k/y	87	87	
33		Overheads	$k/y	1198	1198	
34		Maintenance	$k/y	1690	1690	
35		Internal power consumption	MWe	0.6	0.6	
36				5070	5070	

Appendix F .3 (cont.)

	A	B	C	D	E	F
1	GRID CONNECTION MODULE					
2	**Parameter**		**Units**	**all sites**	**single site**	
3	User inputs					
4		Current costs for plant no.		100		Input
5		Learning factor		20%		Input
6	System inputs					
7		No. generating sites		1		
8		Gross capacity	MWe	21.4	21.4	
9		Gross power output	MWh/y	169001	169001	
10		Power req., pretreatment	MWe	0.2	0.2	
11		Power consumed, pretreatment	MWh/y	1120	1120	
12		Power req., gasification	MWe	0.6	0.6	
13		Power consumed, gasification	MWh/y	4624	4624	
14		Power req., gas turbine	MWe	0.6	0.6	
15		Power consumed, gas turbine	MWh/y	5070	5070	
16	Performance					
17		Total internal consumption	MWe	1.4	1.4	
18			%	6.7%	6.7%	
19			MWh/y	10814	10814	
20			%	6.4%	6.4%	
21		Net capacity	MWe	20	20	
22			MWh/y	158187	158187	
23			GJ/y	569472	569472	
24	Capital costs					
25		Total plant cost, now	$k	1729	1729	
26		Learning factor		20%		
27		Total plant costs, 1st plant	$k	7614	7614	
28		Total plant cost, future	$k	1729	1729	
29	Operating costs					
30		Capital amortisation, fixed	$k/y	203	203	
31		Capital amortisation, adjusted	$k/y	133	133	
32		Labour requirement		0	0.00	
33		Total labour cost		0	0	
34		Overheads	$k	69	69	
35		Maintenance	$k	69	69	

Appendix F .4 Pyr-Eng System by Modules, 20 MW_e

	A	B	C	D	E	F
1	SYSTEM INPUT					
2	Parameter		Units	all sites	single site	
3	User inputs					
4		Feed production rate	odt/y	127874		Input
5		Feed moisture content	%wet	50.0%		Input
6		Feed cost before transport	$/odt	40.00		Input
7		Feed lower heating value	GJ/odt dry	19.3		Input
8	System inputs					
9		No. conversion sites		1		
10	Feed composition					
11		C	%odt	52.0%		
12		H	%odt	5.8%		
13		O	%odt	41.0%		
14		N	%odt	0.1%		
15		S	%odt	0.0%		
16		Ash	%odt	1.1%		
17	Mass flows					
18		Total feed input	odt/y	127874	127874	
19		Total moisture input	t/y	127874	127874	
20		Total wet feed input	t/y	255748	255748	
21		Total ash input	t/y	1407	1407	
22	Energy flows					
23		Lower heating value as deliver	GJ/t	8.42		
24		Energy delivered	GJ/y	2154169	2154169	
25	Feed cost					
26		Total cost to system	$k/y	5115	5115	
27		Specific costs	$/odt	40.00		
28			$/t	20.00		
29			$/GJ	2.37		
30						

Appendix F .4 (cont.)

	A	B	C	D	E	F
1	FEED TRANSPORT MODULE					
2	Parameter		Units	all sites	single site	
3	User input					
4		Use default distance		TRUE		Input
5		Over-ride distance	km	0		Input
6		Plantation area	%	80%		Input
7		Land area limitation	%	5%		Input
8		Land yield	odt/ha.y	10		Input
9		Feed loading cost	$/t	2.6		Input
10		Feed transport cost	$/t.km	0.09		Input
11		Feed bulk density	odt/m3	0.150		Input
12	System input					
13		Total feed transported	odt/y	127874	127874	
14		Moisture content as transporte	%wet	50.0%		
15		No. of conversion sites		1		
16		No. of generation sites		1		
17	Calculations					
18		Planted land	ha	12787		
19		Plantation land	ha	15984		
20		Total land	ha	319686		
21		Radius total land	km	18.00		
22		Direct distance to conversion	km	12.14		
23		Actual distance travelled	km	17.17		
24		Distance used in model	km	17.17		
25		Bulk density	t/m3	0.3		
26		Total mass transported	t/y	255748	255748	
27		Loading cost	$k/y	665	665	
28		Transport cost	$k/y	395	395	
29		Total transport cost	$k/y	1060	1060	

Appendix F .4 (cont.)

	A	B	C	D	E	F
1	FEED PRETREATMENT MODULE					
2	Parameter		Units	all sites	single site	
3	User input					
4		Delivery days/week	d/week	5		Input
5		Delivery weeks/year	weeks/y	52		Input
6		Delivery hours/day	h/d	10		Input
7		Dry matter losses	%odt/month	2.50%		Input
8		MC change in storage	%wet/month	0%		Input
9		Overs in feed	%	5%		Input
10		Dryer energy required	GJ/twe	3.553		Input
11		Diesel fuel cost	$/l	0.51		Input
12		Electricity cost (if purchased	¢/kWh	10		Input
13		Current costs for plant no.		100		Input
14		Learning factor		20%		Input
15	System input					
16		Delivered feed input	odt/y	127874	127874	
17		Delivered feed moisture conter	%wet	50%		
18		Delivered feed heating value	GJ/odt	19.3		
19		No. conversion sites		1		
20		Bulk density	odt/m3	0.150		
21		Conversion operating hours	h/y	7884		
22		Drying required?		TRUE		
23		Intermediate transport distanc	km	0.00		
24	Mass flows					
25		Total feed delivered	t/y	255748	255748	
26		Delivery days/year	d/y	260		
27		Feed delivered	t/d	984	984	
28			t/h	98	98	
29		Storage period	weeks	3		
30		DM losses in storage	%odt	1.7%		
31		MC change in storage	odt/y	0.0%		
32		MC out of storage	%wet	50.0%		
33		Feed from storage	odt/y	125661	125661	
34			t/y	251322	251322	
35			t/h	32	32	
36		Moisture content from dryer	%wet	10%		
37		Moisture into dryer	t/h	15.9	15.9	
38		Moisture out of dryer	t/h	1.8	1.8	
39		Drying rate required	twe/h	14.2	14.2	
40		Dried feed	t/y	139623	139623	
41			t/h	18	18	
42	Volumetric flows, per site					
43		Feed bulk density as delivered	t/m3	0.300		
44		Feed bulk density from storage	t/m3	0.300		
45		Feed bulk density as required	t/m3	0.167		
46		Volume delivered	m3/h	328	328	
47		Volume from store	m3/h	106	106	
48		Volume from dryer	m3/h	106	106	
49	Energy flows					
50		Feed heating value, delivered	GJ/t	8.42		
51			GJ/y	2154169	2154169	
52		Feed heating value, ex. store	GJ/t	8.42		
53			GJ/y	2116885	2116885	
54		Drying energy required per sit	GJ/h	50	50	
55			GJ/y	396865	396865	
56		Feed heating value, reactor	GJ/t	17.12		
57			GJ/y	2390994	2390994	
58	Equipment requirements					
59		No. dryers			1	

Appendix F .4 (cont.)

	A	B	C	D	E	F
60		Drying rate per dryer	twe/h		14.2	
61		No. grinders			3	
62		No. buffer storage silos			1	
63		No. front end loaders			1	
64		No. bulldozers			1	
65	Total plant costs					
66		Reception	$k	1687	1687	
67		Screening	$k	621	621	
68		Storage	$k	4272	4272	
69		Screening	$k	369	369	
70		Re-chipping	$k	228	228	
71		Drying	$k	2343	2343	
72		Grinding	$k	3896	3896	
73		Buffer storage	$k	513	513	
74		Total plant cost, pretreatment	$k	13929	13929	
75		TPC, 1st plant	$k	61346	61346	
76		TOTAL, future plant	$k	13929	13929	
77	Capital charges, all sites					
78		Capital amortisation, fixed	$k/y	1636	1636	
79		Capital amortisation, adjusted	$k/y	1070	1070	
80		Overheads	$k/y	557	557	
81		Maintenance	$k/y	557	557	
82	Labour costs					
83		Labour req., reception	/shift	0.2	0.2	
84		Labour req., screening	/shift	0.0	0.0	
85		Labour req., storage	/shift	0.8	0.8	
86		Labour req., screening	/shift	0.2	0.2	
87		Labour req., re-chip	/shift	0.3	0.3	
88		Labour req., drying	/shift	0.5	0.5	
89		Labour req., grinding	/shift	0.6	0.6	
90		Labour req., buffer storage	/shift	0.2	0.2	
91		TOTAL	/shift	2.8	2.8	
92		Labour cost	$k/y	420	420	
93	Utilities					
94		Utilities - diesel	l/y	237250	237250	
95			$k/y	121	121	
96		Installed power, reception	kWe	123	123	
97		Installed power, screening	kWe	17	17	
98		Installed power, storage	kWe	24	24	
99		Installed power, screening	kWe	8	8	
100		Installed power, re-chip	kWe	28	28	
101		Installed power, drying	kWe	120	120	
102		Installed power, grinding	kWe	1937	1937	
103		Installed power, buffer storag	kWe	19	19	
104		TOTAL	kWe	2276	2276	
105		Power consumed	MWh/y	17032	17032	
106		Close-coupled?		TRUE		
107		Internal power consumption	MWe	2.28	2.28	
108			MWh/y	17032	17032	
109		Electricity cost if purchased	$k/y	0	0	

Appendix F .4 (cont.)

	A	B	C	D	E	F
1	FAST PYROLYSIS MODULE					
2	Parameter		Units	all sites	single site	
3	User input					
4		Availability	%	90%		Input
5		Use default liquids yield		TRUE		Input
6		Over-ride liquids yield	%	53.9%		Input
7		No. conversion sites		1		Input
8		Current costs for plant no.		1		Input
9		Learning factor		20%		Input
10	System input					
11		Total reactor feed	odt/y	125661	125661	
12		Feed LHV	GJ/odt dry	19.3		
13		Moisture content as fed	%wet	10%		
14		Intermediate transport distanc	km	0		
15		Electricity cost if paid	¢/kWh	10		
16	Mass flows					
17		Operating hours	h/y	7884		
18		Reactor feed	t/y	139623	139623	
19			t/h	17.7	17.7	
20			odt/h	15.9	15.9	
21		Default liquid yield	%	59.9%		
22		Default water yield	%	10.8%		
23		Default gas yield	%	13.1%		
24		Default char yield	%	16.2%		
25		Actual liquid yield	%	59.9%		
26		Actual water yield	%	10.8%		
27		Actual gas yield	%	13.1%		
28		Actual char yield	%	16.2%		
29		Dry liquid output	t/h	9.5	9.5	
30		Reaction water output	t/h	1.7	1.7	
31		Feed water output	t/h	1.8	1.8	
32		Total liquids out	t/h	13.0	13.0	
33			t/y	102805	102805	
34		Liquids moisture content	%	26.8%		
35	Energy flows					
36		Reactor feed LHV	GJ/t	17.1		
37		Energy input	GJ/y	2390994	2390994	
38			GJ/h	303	303	
39		Liquids LHV	GJ/t	14.4		
40		Energy out, per reactor	GJ/y	1482580	1482580	
41		Energy efficiency	%LHV	62%	62%	
42	Capital cost					
43		Pyrolysis	$k1995	20124	20124	
44		Storage	$k1995	441	441	
45		Total plant cost, now	$k1995	20564	20564	
46		Learning factor		20%		
47		Total plant costs, 1st plant	$k1995	20564	20564	
48		Total plant cost, future	$k1995	20564	20564	
49	Operating costs, all sites					
50		Capital amortisation, fixed	$k/y	2415	2415	
51		Capital amortisation, adjustec	$k/y	1580	1580	
52		Labour requirement, all sites		3.87	3.87	
53		Total labour cost	$k/y	581	581	
54		Power consumed	MWe	0.6	0.64	
55			MWh/y	5026	5026	1.68
56		Close-coupled?		TRUE		
57		Internal power consumption	MWe	0.6	0.6	
58			MWh/y	5026.4	5026	
59		Electricity cost if paid	$k/y	0	0	

Appendix F .4 (cont.)

	A	B	C	D	E	F
60		Cooling water @ 0.017$/t	$k/y	39.5	39.5	
61		Overheads	$k/y	823	823	
62		Maintenance	$k/y	823	823	

Appendix F .4 (cont.)

	A	B	C	D	E	F
1	LIQUID TRANSPORT MODULE					
2	Parameter		Units	all sites	single site	
3	User input					
4		Use default transport distance		TRUE		Input
5		Over-ride transport distance	km	40		Input
6		Liquid loading cost	$/t	0		Input
7		Liquid transport cost	$/t/km	0.05		Input
8	System input					
9		No. pyrolysis sites		1		
10		No. generating sites		1		
11		Feed supply area radius	km	18		
12		Liquid transported	t/y	102805		
13	Performance					
14		No. multiple sites		1		
15		Sector angle		6.28		
16		Default direct distance	km	0		
17		Default road distance	km	0		
18		Actual distance used	km	0		
19		Liquid loading cost	$k/y	0		
20		Liquid transport cost	$k/y	0		
21		TOTAL liquid transport cost	$k/y	0		

Appendix F .4 (cont.)

	A	B	C	D	E	F
1	LIQUID FIRED DUAL FUEL ENGINE MODULE					
2	Parameter		Units	all sites	single site	
3	User inputs					
4		Use default efficiency		TRUE		Input
5		Over-ride efficiency		35%		Input
6		Availability	%	90%		Input
7		Energy supplied by diesel	%	7.5%		Input
8		Internal energy consumption	%	3%		Input
9		Maintenance cost	$/kWh	0.0122		Input
10		No. generating sites		1		Input
11		Current costs for plant no.		100		Input
12		Learning factor		20%		Input
13	System inputs					
14		Liquid input	t/y	102805		
15		Liquids LHV	GJ/t	14.4		
16		Liquid transport distance	km	0		
17		Cost of diesel	$/l	0.51		
18	Energy flows					
19		Operating hours	h/y	7884		
20		Total liquid energy input	GJ/y	1482580	1482580	
21			GJ/h	188	188	
22		Diesel energy	GJ/h	15	15	
23			GJ/y	120209	120209	
24		Total energy input	GJ/h	203	203	
25		No. engines			2	
26		Energy input per engine	GJ/h		102	
27			MWth		28	
28		Default efficiency	%		41.8%	
29		Efficiency used	%		41.8%	
30		Output per engine	MWe		12	
31		Gross power output	MWe	24	24	
32			MWh/y	186236	186236	
33			GJ/y	670449	670449	
34	Mass flows					
35		Pyrolysis liquid	t/y	102805	102805	
36		Diesel	t/y	2822	2822	
37	Capital costs					
38		TPC, per engine	$k		11864	
39		TPC, all engines	$k	22140	22140	
40		TPC, storage per site	$k	0	0	
41		TPC, now	$k	22140	22140	
42		TPC, 1st plant		97506	97506	
43		TPC, future	$k	22140	22140	
44	Operating costs					
45		Capital amortisation, fixed	$k/y	2601	2601	
46		Capital amortisation, adjustec	$k/y	1701	1701	
47		Labour requirement		2.20	2.20	
48		Labour cost	$k/y	330	330	
49		Diesel cost	$k/y	1693	1693	
50		Overheads	$k/y	886	886	
51		Maintenance	$k/y	2272	2272	
52		Internal power consumption	MWe	0.71	0.71	
53			MW/y	5587	5587	

Appendix F .4 (cont.)

	A	B	C	D	E	F
1	GRID CONNECTION MODULE					
2	Parameter		Units	all sites	single site	
3	User inputs					
4		Current costs for plant no.		100		Input
5		Learning factor		20%		Input
6	System inputs					
7		No. generating sites		1		
8		Gross capacity	MWe	23.6	23.6	
9		Gross power output	MWh/y	186236	186236	
10		Power req., pretreatment	MWe	2.3	2.3	
11		Power consumed, pretreatment	MWh/y	17032	17032	
12		Power req., pyrolysis	MWe	0.6	0.6	
13		Power consumed, pyrolysis	MWh/y	5026	5026	
14		Power req., engine	MWe	0.7	0.7	
15		Power consumed, engine	MWh/y	5587	5587	
16	Performance					
17		Total internal consumption	MWe	3.6	3.6	
18			%	15%	15%	
19			MWh/y	27646	27646	
20			%	15%	15%	
21		Net capacity	MWe	20.0	20.0	
22			MWh/y	158590	158590	
23			GJ/y	570925	570925	
24	Capital costs					
25		TPC, now		1729	1729	
26		TPC, 1st plant		7614	7614	
27		TPC, future	$k	1729	1729	
28	Operating costs					
29		Capital amortisation, fixed	$k/y	203	203	
30		Capital amortisation, adjustec	$k/y	133	133	
31		Labour requirement		0	0.00	
32		Total labour cost	$k/y	0	0	
33		Overheads	$k/y	69	69	
34		Maintenance	$k/y	69	69	

Appendix G Results

Appendix G .1 Combustion Base Case, Results at 5-20 MW.

Results - Combustion System						Settings						
		Number of conversion sites				1						
		Number of generation sites				1	(close-coupled)					
		Calculated costs for plant after				0	replication(s)					
Settings												
Target - calculated	MWe	5	6	7	8	9	10	12	14	16	18	20
Feed cost	$/odt	40	40	40	40	40	40	40	40	40	40	40
Summary												
Electricity cost	¢/kWh	14.34	13.44	12.74	12.18	11.72	11.33	11.12	10.83	10.39	10.02	9.71
Electricity price	¢/kWh	18.90	17.66	16.69	15.92	15.28	14.74	14.53	14.16	13.54	13.02	12.58
Capital cost	$/kWe	3641	3359	3141	2966	2822	2701	2710	2640	2495	2375	2273
Overall efficiency	%LHV in	18.4%	19.0%	19.6%	20.1%	20.6%	21.0%	21.8%	22.4%	23.0%	23.5%	23.9%
Mass flows												
Feed delivered	t/y	92921	107287	121251	134881	148214	161299	186818	212320	236487	260099	283250
Feed from storage	t/y	91312	105430	119153	132547	145649	158508	183584	208645	232394	255598	278348
Feed at reactor	t/y	70240	81100	91656	101959	112038	121929	141219	160496	178765	196614	214114
Energy flows												
Feed delivered	GJ/y	782670	903679	1021300	1136103	1248410	1358624	1573567	1788369	1991933	2190817	2385815
Feed from storage	GJ/y	769124	888039	1003624	1116440	1226803	1335109	1546332	1757416	1957458	2152899	2344522
Feed at reactor	GJ/y	820835	947745	1071101	1191502	1309285	1424873	1650297	1875573	2089064	2297646	2502152
Fuel gas	GJ/y	683153	790208	894630	996896	1097268	1196086	1389702	1584374	1769946	1952266	2131993
Diesel energy input	GJ/y	0	0	0	0	0	0	0	0	0	0	0
Gross electricity out	GJ/y	166373	198381	230325	262221	294055	325859	389366	454627	517991	581226	644422
Net electricity out	GJ/y	143641	172044	200449	228859	257253	285655	342456	400376	457216	514008	570823
Efficiencies												
Conversion efficienc	%LHV	83.2%	83.4%	83.5%	83.7%	83.8%	83.9%	84.2%	84.5%	84.7%	85.0%	85.2%
Generating efficienc	%	24.4%	25.1%	25.7%	26.3%	26.8%	27.2%	28.0%	28.7%	29.3%	29.8%	30.2%
Overall	%LHV	18.4%	19.0%	19.6%	20.1%	20.6%	21.0%	21.8%	22.4%	23.0%	23.5%	23.9%
Electricity												
Gross capacity	MWe	5.9	7.0	8.1	9.2	10.4	11.5	13.7	16.0	18.3	20.5	22.7
Internal requirement	MWe	0.9	1.0	1.1	1.2	1.4	1.5	1.7	2.0	2.2	2.5	2.7
Net capacity	MWe	5.0	6.0	7.0	8.0	9.0	10.0	12.0	14.0	16.0	18.0	20.0
Net capacity per sit	MWe	5.0	6.0	7.0	8.0	9.0	10.0	12.0	14.0	16.0	18.0	20.0
Gross output	MWh/y	46215	55106	63979	72839	81682	90516	108157	126285	143886	161452	179006
Internal consumption	MWh/y	6314	7316	8299	9267	10223	11168	13031	15070	16882	18672	20444
Net output	MWh/y	39900	47790	55680	63572	71459	79349	95127	111215	127005	142780	158562

Appendix G.1 (cont.)

Results - Combustion System						Settings						
		Number of conversion sites				1						
		Number of generation sites				1	(close-coupled)					
		Calculated costs for plant after				0	replication(s)					
Settings												
Target - calculated	MWe	5	6	7	8	9	10	12	14	16	18	20
Feed cost	$/odt	40	40	40	40	40	40	40	40	40	40	40
Summary												
Electricity cost	¢/kWh	14.34	13.44	12.74	12.18	11.72	11.33	11.12	10.83	10.39	10.02	9.71
Electricity price	¢/kWh	18.90	17.66	16.69	15.92	15.28	14.74	14.53	14.16	13.54	13.02	12.58
Capital cost	$/kWe	3641	3359	3141	2966	2822	2701	2710	2640	2495	2375	2273
Overall efficiency	%LHV in	18.4%	19.0%	19.6%	20.1%	20.6%	21.0%	21.8%	22.4%	23.0%	23.5%	23.9%
Capital costs, specific												
Pretreatment	$/kWe	656	571	509	462	424	393	551	592	545	506	474
Conversion	$/kWe	1859	1731	1630	1548	1480	1422	1327	1255	1193	1141	1097
Generation	$/kWe	962	906	861	824	794	767	723	690	661	637	616
Connection	$/kWe	164	151	141	132	125	119	110	102	96	91	86
TOTAL	$/kWe	3641	3359	3141	2966	2822	2701	2710	2640	2495	2375	2273
Operating costs												
Feed production	$k/y	1858	2146	2425	2698	2964	3226	3736	4246	4730	5202	5665
Feed transport	$k/y	328	386	444	502	560	617	733	851	966	1082	1197
Pretreatment	$k/y	906	929	951	971	990	1008	1428	1757	1822	1884	1943
Conversion	$k/y	1576	1758	1929	2091	2246	2394	2674	2945	3193	3430	3656
Liquids transport	$k/y	0	0	0	0	0	0	0	0	0	0	0
Generation	$k/y	925	1061	1191	1318	1442	1562	1796	2027	2246	2458	2666
Connection	$k/y	129	142	154	166	177	187	206	224	241	256	271
TOTAL	$k/y	5722	6422	7095	7746	8378	8994	10573	12050	13198	14312	15398
10% return on capita	$k/y	1820	2015	2199	2373	2540	2701	3252	3696	3992	4275	4546
Feed transport												
Total feed supply ar	ha	116151	134109	151564	168601	185268	201624	233522	265400	295609	325124	354062
Transport distance	km	10.35	11.12	11.82	12.47	13.07	13.64	14.68	15.65	16.51	17.32	18.07
Labour requirements												
Pretreatment	/shift	2.24	2.24	2.24	2.24	2.24	2.24	2.24	2.23	2.23	2.23	2.23
Conversion	/shift	0.79	0.86	0.93	0.99	1.04	1.09	1.18	1.26	1.33	1.39	1.44
Generation	/shift	0.69	0.77	0.85	0.92	0.98	1.04	1.15	1.25	1.33	1.40	1.46
TOTAL	/shift	3.71	3.87	4.01	4.14	4.27	4.38	4.57	4.74	4.89	5.01	5.12
Labour cost	$k/y	557	581	602	622	640	656	686	711	733	752	768
	%prod co	9.7%	9.0%	8.5%	8.0%	7.6%	7.3%	6.5%	5.9%	5.6%	5.3%	5.0%

Appendix G .2 Gas-Eng Base Case, Results at 5-20 MW$_e$

Results - Gas-Eng system					Settings							
				Number of conversion sites	1							
				Number of generation sites	1	(close-coupled)						
				Calculated costs for plant after	0	replication(s)						
Settings		**Results Table**										
Target - calculated	MWe	5	6	7	8	9	10	12	14	16	18	20
Feed cost	$/odt	40	40	40	40	40	40	40	40	40	40	40
Summary												
Electricity cost	¢/kWh	18.21	17.60	16.85	16.25	15.74	15.31	14.60	14.65	14.55	13.84	13.46
Electricity price	¢/kWh	24.79	24.00	22.95	22.10	21.39	20.77	19.77	19.86	19.65	18.71	18.18
Capital cost	$/We	5197	5095	4850	4650	4481	4336	4098	4126	4046	3859	3734
Overall efficiency	%LHV in	25.2%	25.4%	25.7%	26.0%	26.3%	26.5%	27.0%	26.6%	26.9%	28.3%	28.7%
Mass flows												
Feed delivered	t/y	64319	77191	88862	100316	111572	122641	144288	171044	193183	206518	226279
Feed from storage	t/y	63206	75855	87324	98580	109641	120518	141790	168083	189840	202943	222363
Feed at reactor	t/y	37180	44621	51367	57988	64495	70893	83406	98873	111670	119378	130802
Energy flows												
Feed delivered	GJ/y	541760	650181	748488	844964	939773	1033003	1215336	1440702	1627184	1739499	1905949
Feed from storage	GJ/y	532383	638928	735533	830339	923507	1015124	1194301	1415767	1599021	1709392	1872961
Feed at reactor	GJ/y	596251	715577	823772	929951	1034297	1136904	1337576	1585610	1790848	1914461	2097652
Fuel gas	GJ/y	428123	514763	593582	671169	747640	823047	971108	1155137	1308225	1400765	1538352
Diesel energy input	GJ/y	22533	27093	31241	35325	39349	43318	51111	60797	68854	73724	80966
Gross electricity out	GJ/y	152828	185643	215999	246330	276643	306924	367445	428428	490264	550026	610324
Net electricity out	GJ/y	142057	171993	200396	228798	257203	285597	342395	399200	456832	513636	570403
Efficiencies												
Conversion efficienc	%LHV	71.8%	71.9%	72.1%	72.2%	72.3%	72.4%	72.6%	72.9%	73.1%	73.2%	73.3%
Generating efficienc	%	33.9%	34.3%	34.6%	34.9%	35.2%	35.4%	35.9%	35.2%	35.6%	37.3%	37.7%
Overall	%LHV	25.2%	25.4%	25.7%	26.0%	26.3%	26.5%	27.0%	26.6%	26.9%	28.3%	28.7%
Electricity												
Gross capacity	MWe	5.4	6.5	7.6	8.7	9.7	10.8	12.9	15.1	17.3	19.4	21.5
Internal requirement	MWe	0.4	0.5	0.6	0.7	0.7	0.8	0.9	1.1	1.3	1.4	1.5
Net capacity	MWe	5.0	6.0	7.0	8.0	9.0	10.0	12.0	14.0	16.0	18.0	20.0
Net capacity per site	MWe	5.0	6.0	7.0	8.0	9.0	10.0	12.0	14.0	16.0	18.0	20.0
Gross output	MWh	42452	51568	60000	68425	76845	85257	102068	119008	136185	152785	169534
Internal consumption	MWh	2992	3792	4334	4870	5400	5924	6958	8119	9287	10108	11089
Net output	MWh	39460	47776	55666	63555	71445	79333	95110	110889	126898	142677	158445

Appendix G .2 (cont.)

Results - Gas-Eng system						Settings					
			Number of conversion sites			1					
			Number of generation sites			1	(close-coupled)				
			Calculated costs for plant after			0	replication(s)				
Settings		**Results Table**									
Target - calculated	MWe	5	6	7	8	9	10	12	14	16	18	20
Feed cost	$/odt	40	40	40	40	40	40	40	40	40	40	40
Summary												
Electricity cost	¢/kWh	18.21	17.60	16.85	16.25	15.74	15.31	14.60	14.65	14.55	13.84	13.46
Electricity price	¢/kWh	24.79	24.00	22.95	22.10	21.39	20.77	19.77	19.86	19.65	18.71	18.18
Capital cost	$/We	5197	5095	4850	4650	4481	4336	4098	4126	4046	3859	3734
Overall efficiency	%LHV in	25.2%	25.4%	25.7%	26.0%	26.3%	26.5%	27.0%	26.6%	26.9%	28.3%	28.7%
Capital costs, specific												
Pretreatment	$/kWe	451	539	488	448	417	391	350	507	551	506	477
Conversion	$/kWe	3376	3194	3021	2877	2754	2648	2472	2386	2273	2117	2031
Generation	$/kWe	1206	1211	1201	1192	1184	1178	1167	1131	1126	1145	1139
Connection	$/kWe	164	151	141	132	125	119	110	102	96	91	86
TOTAL	$/kWe	5197	5095	4850	4650	4481	4336	4098	4126	4046	3859	3734
Operating costs												
Feed production	$k/y	1286	1544	1777	2006	2231	2453	2886	3421	3864	4130	4526
Feed transport	$k/y	217	266	312	358	404	450	543	661	762	824	917
Pretreatment	$k/y	832	998	1027	1054	1079	1103	1149	1604	2146	2193	2261
Conversion	$k/y	3019	3420	3767	4093	4403	4699	5255	5909	6426	6730	7169
Liquids transport	$k/y	0	0	0	0	0	0	0	0	0	0	0
Generation	$k/y	1701	2037	2345	2650	2952	3253	3848	4424	5021	5612	6186
Connection	$k/y	129	142	154	166	177	187	206	224	241	256	271
TOTAL	$k/y	7186	8408	9382	10327	11247	12145	13886	16242	18459	19744	21330
10% return on capita	$k/y	2598	3057	3395	3720	4033	4336	4917	5776	6474	6947	7468
Feed transport												
Total feed supply are	ha	80399	96489	111078	125395	139465	153301	180360	213805	241479	258147	282849
Transport distance	km	8.61	9.43	10.12	10.76	11.34	11.89	12.90	14.04	14.93	15.43	16.15
Labour requirements												
Pretreatment	/shift	2.45	2.90	2.90	2.90	2.90	2.90	2.90	2.90	4.28	4.28	4.28
Conversion	/shift	2.01	2.19	2.34	2.48	2.61	2.73	2.95	3.20	3.39	3.50	3.66
Generation	/shift	1.08	1.18	1.27	1.36	1.43	1.51	1.65	1.77	1.89	2.00	2.10
TOTAL	/shift	5.54	6.28	6.52	6.74	6.95	7.14	7.50	7.87	9.57	9.78	10.04
Labour cost	$k/y	831	941	978	1011	1042	1071	1125	1181	1435	1467	1506
	%	11.6%	11.2%	10.4%	9.8%	9.3%	8.8%	8.1%	7.3%	7.8%	7.4%	7.1%

Appendix G .3 IGCC Base Case, Results at 5-20 MW.

Results - IGCC system								Settings				
							Number of conversion sites	1				
							Number of generation sites	1	(close-coupled)			
							Calculated costs for plant after	0	replication(s)			
Settings		**Results Table**										
Target - calculated	MWe	5	6	7	8	9	10	12	14	16	18	20
Feed cost	$/odt	40	40	40	40	40	40	40	40	40	40	40
Summary												
Electricity cost	¢/kWh	17.79	16.68	15.82	15.20	14.76	14.25	13.43	12.78	12.26	11.83	11.46
Electricity price		26.06	24.41	23.13	22.23	21.60	20.83	19.59	18.61	17.82	17.17	16.61
Capital cost	$/We	6529	6105	5772	5544	5430	5221	4881	4616	4402	4223	4070
Overall efficiency	%Feed LH	35.9%	36.8%	37.6%	38.3%	38.8%	39.4%	40.3%	41.2%	41.9%	42.5%	43.1%
Mass flows												
Feed delivered	t/y	46947	54942	62776	70477	78643	86108	100775	115142	129255	143148	156849
Feed from storage	t/y	46134	53992	61689	69258	77282	84618	99031	113149	127018	140670	154134
Feed at reactor	t/y	27138	31760	36288	40740	45460	49775	58254	66558	74716	82747	90667
Energy flows												
Feed delivered	GJ/y	395433	462781	528761	593631	662408	725292	848831	969840	1088713	1205734	1321136
Feed from storage	GJ/y	388589	454771	519610	583357	650943	712738	834139	953055	1069870	1184865	1298270
Feed at reactor	GJ/y	435206	509328	581945	653340	729034	798243	934207	1067388	1198217	1327008	1454018
Fuel gas	GJ/y	374399	438599	501615	563682	629606	689989	808903	925747	1040867	1154513	1266889
Diesel energy input	GJ/y	0	0	0	0	0	0	0	0	0	0	0
Gross electricity out	GJ/y	152918	183228	213520	243812	276429	306662	367088	427470	487813	548121	608404
Net electricity out	GJ/y	142057	170449	198847	227265	257143	285538	342328	399116	455903	512687	569472
Efficiencies												
Conversion efficienc	%LHV	86.0%	86.1%	86.2%	86.3%	86.4%	86.4%	86.6%	86.7%	86.9%	87.0%	87.1%
Generating efficienc	%	40.8%	41.8%	42.6%	43.3%	43.9%	44.4%	45.4%	46.2%	46.9%	47.5%	48.0%
Overall	%LHV	35.9%	36.8%	37.6%	38.3%	38.8%	39.4%	40.3%	41.2%	41.9%	42.5%	43.1%
Electricity												
Gross capacity	MWe	5.4	6.5	7.5	8.6	9.7	10.8	12.9	15.1	17.2	19.3	21.4
Internal requirement	MWe	0.4	0.5	0.5	0.6	0.7	0.8	0.9	1.1	1.2	1.3	1.4
Net capacity	MWe	5.0	6.0	7.0	8.0	9.0	10.0	12.0	14.0	16.0	18.0	20.0
Net capacity per site												
Gross output	MWh	42477	50897	59311	67726	76786	85184	101969	118742	135504	152256	169001
Internal consumption	MWh	3017	3550	4076	4596	5357	5868	6878	7876	8864	9843	10814
Net output	MWh	39460	47347	55235	63129	71429	79316	95091	110866	126640	142413	158187

Appendix G .3 (cont.)

Results - IGCC system												
								Settings				
					Number of conversion sites			1				
					Number of generation sites			1	(close-coupled)			
					Calculated costs for plant after			0	replication(s)			
Settings		**Results Table**										
Target - calculated	MWe	5	6	7	8	9	10	12	14	16	18	20
Feed cost	$/odt	40	40	40	40	40	40	40	40	40	40	40
Summary												
Electricity cost	¢/kWh	17.79	16.68	15.82	15.20	14.76	14.25	13.43	12.78	12.26	11.83	11.46
Electricity price		26.06	24.41	23.13	22.23	21.60	20.83	19.59	18.61	17.82	17.17	16.61
Capital cost	$/We	6529	6105	5772	5544	5430	5221	4881	4616	4402	4223	4070
Overall efficiency	%Feed LH	35.9%	36.8%	37.6%	38.3%	38.8%	39.4%	40.3%	41.2%	41.9%	42.5%	43.1%
Capital costs, specific												
Pretreatment	$/kWe	348	315	289	313	380	353	311	281	257	238	222
Conversion	$/kWe	4195	3865	3607	3397	3239	3089	2846	2656	2502	2373	2264
Generation	$/kWe	1822	1774	1735	1702	1686	1659	1615	1578	1547	1521	1498
Connection	$/kWe	164	151	141	132	125	119	110	102	96	91	86
TOTAL	$/kWe	6529	6105	5772	5544	5430	5221	4881	4616	4402	4223	4070
Operating costs												
Feed production	$k/y	939	1099	1256	1410	1573	1722	2016	2303	2585	2863	3137
Feed transport	$k/y	153	182	211	240	272	301	360	419	478	538	598
Pretreatment	$k/y	617	640	662	736	861	878	911	940	969	995	1021
Conversion	$k/y	3550	3917	4258	4578	4904	5191	5729	6229	6698	7141	7563
Liquids transport		0	0	0	0	0	0	0	0	0	0	0
Generation	$k/y	1633	1916	2194	2468	2758	3024	3545	4057	4560	5055	5544
Connection	$k/y	129	142	154	166	177	187	206	224	241	256	271
TOTAL	$k/y	7020	7897	8736	9599	10545	11303	12767	14172	15530	16849	18134
10% return on capita	$k/y	3265	3663	4040	4436	4887	5220	5858	6463	7042	7601	8141
Feed transport												
Total feed supply ar	ha	58683	68678	78470	88097	98303	107636	125969	143927	161568	178935	196061
Transport distance	km	7.36	7.96	8.51	9.01	9.52	9.96	10.78	11.52	12.21	12.85	13.45
Labour requirements												
Pretreatment	/shift	1.55	1.55	1.55	1.55	1.79	1.79	1.79	1.79	1.79	1.79	1.79
Conversion	/shift	1.73	1.87	1.99	2.10	2.21	2.31	2.49	2.65	2.80	2.94	3.07
Generation	/shift	0.81	0.91	1.01	1.10	1.18	1.26	1.39	1.51	1.61	1.70	1.78
TOTAL	/shift	4.09	4.33	4.55	4.75	5.19	5.36	5.67	5.95	6.20	6.43	6.64
Labour cost	for all	20.46	21.65	22.73	23.73	25.93	26.79	28.35	29.75	31.01	32.16	33.20
	$k/y	614	649	682	712	778	804	851	893	930	965	996
	%	8.7%	8.2%	7.8%	7.4%	7.4%	7.1%	6.7%	6.3%	6.0%	5.7%	5.5%

Appendix G .4 Pyr-Eng Base Case, Results at 5-20 MW_e

Results - Pyr-Eng System						Settings						
		Number of conversion sites				1						
		Number of generation sites				1						
		Calculated costs for plant after				0	replication(s)					
Settings		**Results Table**										
Target - calculated	MWe	5	6	7	8	9	10	12	14	16	18	20
Feed cost	$/odt	40	40	40	40	40	40	40	40	40	40	40
Summary												
Electricity cost	¢/kWh	16.57	16.02	15.38	14.86	14.45	14.08	13.48	13.90	13.49	12.85	12.55
Electricity price	¢/kWh	21.64	20.97	20.08	19.36	18.79	18.27	17.43	18.04	17.46	16.65	16.23
Capital cost	$/kWe	4003	3943	3740	3575	3439	3322	3134	3286	3152	3013	2918
Overall efficiency	%LHV in	22.2%	22.4%	22.7%	22.9%	23.2%	23.4%	23.8%	23.2%	23.5%	24.8%	25.1%
Capital costs, specific												
Pretreatment	$/kWe	776	852	784	730	687	651	595	837	789	730	696
Conversion	$/kWe	1879	1753	1640	1547	1469	1402	1293	1238	1167	1079	1028
Generation	$/kWe	1184	1187	1176	1166	1157	1150	1137	1109	1101	1114	1107
Connection	$/kWe	164	151	141	132	125	119	110	102	96	91	86
TOTAL	$/kWe	4003	3943	3740	3575	3439	3322	3134	3286	3152	3013	2918
Operating costs												
Feed production	$k/y	1437	1725	1987	2245	2499	2749	3239	3875	4372	4662	5115
Feed transport	$k/y	246	302	354	407	460	513	620	764	881	950	1060
Pretreatment	$k/y	1016	1189	1248	1304	1380	1431	1529	2357	2498	2601	2726
Conversion	$k/y	1803	2010	2186	2351	2506	2653	2926	3257	3501	3638	3846
Liquids transport	$k/y	0	0	0	0	0	0	0	0	0	0	0
Generation	$k/y	1912	2287	2631	2970	3306	3639	4298	4988	5638	6250	6882
Connection	$k/y	129	142	154	166	177	187	206	224	241	256	271
TOTAL	$k/y	6542	7655	8561	9443	10328	11172	12818	15466	17131	18357	19901
10% return on capita	$k/y	2001	2366	2618	2860	3095	3322	3761	4601	5043	5424	5836
Feed transport												
Total feed supply are	ha	89801	107819	124213	140321	156178	171795	202416	242201	273259	291347	319686
Transport distance	km	9.10	9.97	10.70	11.38	12.00	12.59	13.66	14.95	15.88	16.39	17.17
Labour requirements												
Pretreatment	/shift	1.97	2.21	2.21	2.21	2.36	2.36	2.36	2.65	2.65	2.80	2.80
Conversion	/shift	2.12	2.31	2.47	2.62	2.76	2.88	3.12	3.40	3.60	3.71	3.87
Generation	/shift	1.13	1.24	1.34	1.42	1.51	1.58	1.72	1.86	1.98	2.09	2.20
TOTAL	/shift	5.22	5.77	6.02	6.25	6.62	6.83	7.20	7.91	8.23	8.60	8.87
Labour cost	$k/y	783	865	903	938	993	1024	1080	1186	1235	1290	1331
	%prod co	12.0%	11.3%	10.5%	9.9%	9.6%	9.2%	8.4%	7.7%	7.2%	7.0%	6.7%

Appendix G .4 (cont.)

Results - Pyr-Eng System						Settings						
		Number of conversion sites				1						
		Number of generation sites				1						
		Calculated costs for plant after				0	replication(s)					
Settings		**Results Table**										
Target - calculated	MWe	5	6	7	8	9	10	12	14	16	18	20
Feed cost	$/odt	40	40	40	40	40	40	40	40	40	40	40
Summary												
Electricity cost	¢/kWh	16.57	16.02	15.38	14.86	14.45	14.08	13.48	13.90	13.49	12.85	12.55
Electricity price	¢/kWh	21.64	20.97	20.08	19.36	18.79	18.27	17.43	18.04	17.46	16.65	16.23
Capital cost	$/kWe	4003	3943	3740	3575	3439	3322	3134	3286	3152	3013	2918
Overall efficiency	%LHV in	22.2%	22.4%	22.7%	22.9%	23.2%	23.4%	23.8%	23.2%	23.5%	24.8%	25.1%
Mass flows												
Feed delivered	t/y	71841	86256	99370	112257	124942	137436	161933	193760	218607	233077	255748
Feed from storage	t/y	70597	84763	97650	110314	122780	135058	159130	190407	214823	229043	251322
Feed at reactor	t/y	39221	47090	54250	61285	68211	75032	88406	105782	119346	127246	139623
Pyrolysis liquid	t/y	28878	34673	39944	45124	50224	55246	65093	77887	87875	93691	102805
Energy flows												
Feed delivered	GJ/y	605115	726530	836994	945537	1052387	1157626	1363959	1632045	1841326	1963210	2154169
Feed from storage	GJ/y	594642	713956	822508	929172	1034172	1137591	1340352	1603798	1809457	1929232	2116885
Feed at reactor	GJ/y	671640	806404	929012	1049487	1168084	1284894	1513910	1811468	2043758	2179041	2390994
Liquid from pyrolysi	GJ/y	416463	500026	576051	650754	724292	796722	938728	1123235	1267270	1351155	1482580
Diesel energy input	GJ/y	33767	40543	46707	52764	58726	64599	76113	91073	102752	109553	120209
Gross electricity out	GJ/y	169640	205751	239153	272463	305709	338876	405056	475344	541706	604810	670449
Net electricity out	GJ/y	142099	171998	200414	228818	257225	285620	342421	400461	457297	514123	570925
Efficiencies												
Conversion efficienc	%LHV	62.0%	62.0%	62.0%	62.0%	62.0%	62.0%	62.0%	62.0%	62.0%	62.0%	62.0%
Generating efficienc	%	37.7%	38.1%	38.4%	38.7%	39.0%	39.3%	39.9%	39.1%	39.5%	41.4%	41.8%
Overall	%LHV	22.2%	22.4%	22.7%	22.9%	23.2%	23.4%	23.8%	23.2%	23.5%	24.8%	25.1%
Electricity												
Gross capacity	MWe	6.0	7.2	8.4	9.6	10.8	11.9	14.3	16.7	19.1	21.3	23.6
Internal requirement	MWe	1.0	1.2	1.4	1.6	1.8	1.9	2.3	2.7	3.1	3.3	3.6
Net capacity	MWe	5.0	6.0	7.0	8.0	9.0	10.0	12.0	14.0	16.0	18.0	20.0
Net capacity per sit	MWe	5.0	6.0	7.0	8.0	9.0	10.0	12.0	14.0	16.0	18.0	20.0
Gross output	MWh/y	47122	57153	66431	75684	84919	94132	112515	132040	150474	168003	186236
Internal consumption	MWh/y	7650	9376	10761	12124	13468	14793	17398	20801	23447	25191	27646
Net output	MWh/y	39472	47777	55671	63560	71451	79339	95117	111239	127027	142812	158590

Appendix G .5 Combustion Base Case, Sensitivities at 10 MW.

Sensitivities, Combustion							
	Base case production	11.33	¢/kWh				
	Capacity	10	MWh				
				10%		-10%	
		Base case valu	¢/kWh	% change	¢/kWh	% change
Financial							
	Project life	20	y	11.15	-1.63%	11.55	1.95%
	Nominal interest rate	10	%	11.51	1.63%	11.15	-1.63%
	Inflation rate	5	%	11.23	-0.89%	11.43	0.89%
	Overheads	4	%TPC/y	11.47	1.22%	11.19	-1.22%
	Maintenance	4	%TPC/y	11.43	0.89%	11.23	-0.89%
	Labour cost	30000	$/y/pers	11.40	0.65%	11.25	-0.73%
Feedstock							
	Feed moisture content	50	%	11.99	5.83%	11.18	-1.32%
	Feed cost	40	$/odt	11.73	3.50%	10.93	-3.50%
	Feed LHV when dry	19.3	GJ/odt	10.42	-8.05%	12.32	8.70%
Feed transport							
	Land area limitation	5	%	11.32	-0.08%	11.34	0.08%
	Feed loading cost	2.6	$/t	11.38	0.41%	11.27	-0.49%
	Feed transport cost	0.09	$/t/km	11.36	0.24%	11.30	-0.24%
Feed pretreatment							
	Delivery days	5	d/week	11.33	0.00%	11.70	3.25%
	Delivery weeks	52	weeks/y	11.33	0.00%	11.70	3.25%
	Delivery hours	10	h/d	11.33	0.00%	11.45	1.06%
	Dry matter losses	2.5	%/month	11.32	-0.08%	11.34	0.08%
	Moisture content char	0	%/month	-	-	-	-
	Overs in feed	5	%	11.33	0.00%	11.33	0.00%
	Diesel cost	0.51	$/l	11.34	0.08%	11.32	-0.08%
	Internal power	0.18	MWe	11.34	0.08%	11.31	-0.16%
	Total plant cost	393	$/kWe	11.44	0.98%	11.21	-1.06%
Feed conversion (Combustion)							
	Availability	90	%	11.15	-1.63%	11.50	1.46%
	Conversion efficiency	83.9	%	10.24	-9.59%	12.55	10.73%
	Internal power	2%	MWth	11.25	-0.73%	11.27	-0.57%
	Total plant cost	1422	$/kWe	11.60	2.36%	11.05	-2.44%
Electricity generation (Steam cycle)							
	Generation efficiency	27.2	%	10.18	-10.16%	12.80	13.01%
	Internal power	4	% MWe	11.37	0.33%	11.29	-0.33%
	Maintenance	0.004	$/kWh	11.37	0.33%	11.29	-0.33%
	Total plant cost	767	$/kWe	11.43	0.89%	11.22	-0.98%
Grid connection							
	Total plant cost	119	$/kWe	11.35	0.16%	11.31	-0.16%

Appendix G .6 Gas-Eng Base Case, Sensitivities at 10 MW_e

Sensitivities, Gas-Eng						
Base case production	15.31	¢/kWh				
Capacity	10	MWh				
			10%		-10%	
	Base case valu		¢/kWh	% change	¢/kWh	% change
Financial						
Project life	20	y	15.02	-1.89%	15.65	2.22%
Nominal interest rat	10	%	15.61	1.96%	51.03	233.31%
Inflation rate	5	%	15.16	-0.98%	15.47	1.05%
Overheads	4	%TPC/y	15.53	1.44%	15.09	-1.44%
Maintenance	4	%TPC/y	15.47	1.05%	15.15	-1.05%
Labour cost	30000	$/y/pers	15.44	0.85%	15.17	-0.91%
Feedstock						
Feed moisture content	50	%	15.45	0.91%	15.19	-0.78%
Feed cost	40	$/odt	15.62	2.02%	15.00	-2.02%
Feed LHV when dry	19.3	GJ/odt	14.51	-5.23%	16.28	6.34%
Feed transport						
Land area limitation	5	%	15.30	-0.07%	15.32	0.07%
Feed loading cost	2.6	$/t	15.34	0.20%	15.27	-0.26%
Feed transport cost	0.09	$/t/km	15.33	0.13%	15.29	-0.13%
Feed pretreatment						
Delivery days	5	d/week	15.30	-0.07%	15.30	-0.07%
Delivery weeks	52	weeks/y	15.31	0.00%	15.30	-0.07%
Delivery hours	10	h/d	15.31	0.00%	15.30	-0.07%
Dry matter losses	2.5	%/month	15.32	0.07%	15.30	-0.07%
Moisture content cha	0	%/month	-	-	-	-
Overs in feed	5	%	15.31	0.00%	15.31	0.00%
Diesel cost	0.51	$/l	15.39	0.52%	15.22	-0.59%
Internal power	0.18	MWe	15.32	0.07%	15.30	-0.07%
Total plant cost	391	$/kWe	15.39	0.52%	15.23	-0.52%
Feed conversion (Atmos. gasification)						
Availability	90	%	14.48	-5.42%	16.32	6.60%
Conversion efficienc	72.4	%	14.53	-5.09%	16.26	6.21%
Internal power	40	kW/odt	15.34	0.20%	15.28	-0.20%
Total plant cost	2648	$/kWe	15.83	3.40%	14.79	-3.40%
Electricity generation (Dual fuel engine)						
Generation efficienc	35.4	%	14.47	-5.49%	16.33	6.66%
Internal power	3	% MWe	15.34	0.20%	15.27	-0.26%
Maintenance	0.012	$/kWh	15.44	0.85%	15.18	-0.85%
Total plant cost	1178	$/kWe	15.48	1.11%	15.14	-1.11%
Diesel auxiliary	5	%LHV in	15.34	0.20%	15.28	-0.20%
Grid connection						
Total plant cost	119	$/kWe	15.34	0.20%	15.29	-0.13%

Appendix G .7 IGCC Base Case, Sensitivities at 10 MW_e

Sensitivities, IGCC							
	Base case production	14.25	¢/kWh				
	Capacity	10	MWh				
				+10%		-10%	
		Base case valu		¢/kWh	% change	¢/kWh	% change
Financial							
	Project life	20	y	13.91	-2.36%	14.65	2.84%
	Nominal interest rate	10	%	14.60	2.43%	13.91	-2.36%
	Inflation rate	5	%	14.07	-1.25%	14.45	1.39%
	Overheads	4	%TPC/y	14.51	1.80%	13.99	-1.80%
	Maintenance	4	%TPC/y	14.43	1.25%	14.07	-1.25%
	Labour cost	30000	$/y/pers	14.37	0.83%	14.13	-0.83%
Feedstock							
	Feed moisture content	50	%	14.35	0.69%	14.17	-0.55%
	Feed cost	40	$/odt	14.47	1.52%	14.03	-1.52%
	Feed LHV when dry	19.3	GJ/odt	13.57	-4.78%	15.07	5.75%
Feed transport							
	Land area limitation	5	%	14.24	-0.07%	14.24	-0.07%
	Feed loading cost	2.6	$/t	14.27	0.14%	14.25	0.00%
	Feed transport cost	0.09	$/t/km	14.26	0.07%	14.25	0.00%
Feed pretreatment							
	Delivery days	5	d/week	14.25	0.00%	14.25	0.00%
	Delivery weeks	52	weeks/y	14.25	0.00%	14.25	0.00%
	Delivery hours	10	h/d	14.25	0.00%	14.24	-0.07%
	Dry matter losses	2.5	%/month	14.25	0.00%	14.25	0.00%
	Moisture content char	0	%/month	-	-	-	-
	Overs in feed	5	%	14.26	0.07%	14.25	0.00%
	Diesel cost	0.51	$/l	14.26	0.07%	14.24	-0.07%
	Internal power	0.16	MWe	14.26	0.07%	14.24	-0.07%
	Total plant cost	337	$/kWe	14.32	0.49%	14.18	-0.49%
Feed conversion (Press. gasification)							
	Availability	90	%	13.28	-6.79%	15.43	8.25%
	Conversion efficiency	86.4	%	13.59	-4.64%	15.05	5.61%
	Internal power	60	kW/odt	14.28	0.21%	14.22	-0.21%
	Total plant cost	3089	$/kWe	14.85	4.23%	13.65	-4.23%
Electricity generation (GTCC)							
	Generation efficiency	44.4	%	13.59	-4.64%	15.04	5.54%
	Internal power	3	% MWe	14.28	0.21%	14.22	-0.21%
	Maintenance	0.01	$/kWh	14.26	0.07%	14.24	-0.07%
	Total plant cost	1659	$/kWe	14.49	1.66%	14.00	-1.73%
Grid connection							
	Total plant cost	119	$/kWe	14.27	0.14%	14.23	-0.14%

Appendix G .8 Pyr-Eng Base Case, Sensitivities at 10 MW_e

Sensitivities, Pyr-Eng						
Base case production	13.93	¢/kWh				
Capacity	10	MWh				
			+10%		-10%	
	Base case valu	¢/kWh	% change	¢/kWh	% chang
Financial						
Project life	20	y	13.83	-0.73%	14.32	2.82%
Nominal interest rate	10	%	14.28	2.53%	13.84	-0.66%
Inflation rate	5	%	13.94	0.06%	14.18	1.80%
Overheads	4	%TPC/y	14.23	2.17%	13.89	-0.30%
Maintenance	4	%TPC/y	14.17	1.73%	13.94	0.06%
Labour cost	30000	$/y/pers	14.19	1.88%	13.93	-0.01%
Feedstock						
Feed moisture content	50	%	14.22	2.09%	13.92	-0.08%
Feed cost	40	$/odt	14.41	3.47%	13.70	-1.68%
Feed LHV when dry	19.3	GJ/odt	14.06	0.93%	14.06	0.93%
Feed transport						
Land area limitation	5	%	14.08	1.08%	14.07	1.01%
Feed loading cost	2.6	$/t	14.10	1.22%	13.98	0.35%
Feed transport cost	0.09	$/t/km	14.08	1.08%	14.04	0.79%
Feed pretreatment						
Delivery days	5	d/week	14.06	0.93%	14.05	0.86%
Delivery weeks	52	weeks/y	14.06	0.93%	14.05	0.86%
Delivery hours	10	h/d	14.06	0.93%	14.05	0.86%
Dry matter losses	2.5	%/month	14.06	0.93%	14.05	0.86%
Moisture content char	0	%/month	-	-	-	-
Overs in feed	5	%	14.06	0.93%	14.06	0.93%
Diesel cost	0.51	$/l	14.14	1.51%	13.97	0.28%
Internal power	1.27	MWe	14.18	1.80%	13.94	0.06%
Total plant cost	662	$/kWe	14.19	1.88%	13.93	-0.01%
Feed conversion (Fast pyrolysis)						
Availability	90	%	13.40	-3.78%	14.86	6.66%
Prepared feed moistu	10	%	14.18	1.80%	13.94	0.06%
Liquids yield	59.9	%	13.18	-5.37%	15.18	8.98%
Internal power	60	kW/odt	14.06	0.93%	14.02	0.64%
Total plant cost	1402	$/kWe	14.34	2.96%	13.79	-1.02%
Electricity generation (Dual fuel engine)						
Generation efficiency	39.3	%	13.19	-5.30%	15.14	8.69%
Internal power	3	% MWe	14.09	1.15%	14.02	0.64%
Maintenance	0.01	$/kWh	14.20	1.95%	13.91	-0.15%
Total plant cost	1150	$/kWe	14.23	2.17%	13.92	-0.08%
Diesel auxiliary	5	%LHV in	14.09	1.15%	14.03	0.72%
Grid connection						
Total plant cost	119	$/kWe	14.08	1.08%	14.03	0.72%