TECHNICAL AND ECONOMIC SIMULATION OF BIOMASS PYROLYSIS PROCESSES FOR FUELS AND ELECTRICITY.

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Doctor of Philosophy

THE UNIVERSITY OF ASTON IN BIRMINGHAM

January 1995

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DEPARTMENT OF CHEMICAL ENGINEERING AND APPLIED CHEMISTRY.

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SUMMARY

There is considerable concern over the increase effect of fossil fuel usage on the environment and this concern has resulted in an effort to find alternative, environmentally friendly energy sources. Biomass is an available alternative resource which may be converted by flash pyrolysis to produce a crude liquid product that can be used directly to substitute for conventional fossil fuels or upgraded to a higher quality fuel. Both the crude and upgraded products may be utilised for power generation.

A computer program, BLUNT, has been developed to model the flash pyrolysis of biomass with subsequent upgrading, refining or power production. The program assesses and compares the economic and technical opportunities for biomass thermochemical conversion on the same basis. BLUNT works by building up a selected processing route from a number of process steps through which the material passes sequentially. Each process step has a step model that calculates the mass and energy balances, the utilities usage and the capital cost for that step of the process. The results of the step models are combined to determine the performance of the whole conversion route. Sample results from the modelling are presented in this thesis.

Due to the large number of possible combinations of feeds, conversion processes, products and sensitivity analyses a complete set of results is impractical to present in a single publication. Variation of the production costs for the available products have been illustrated based on the cost of a wood feedstock. The effect of selected macroeconomic factors on the production costs of bio-derived-diesel and gasoline are also given.

KEYWORDS : Thermochemical; Upgrading; Modelling; Fuels; Biomass

DEDICATION.

To the memory of my Grandfather Joe Bradley.

The world was never made;
It will change, but will not fade.
So let the wind range;
For even and morn
Ever will be
Through eternity.
Nothing was born;
Nothing will die,
All things will change.

Extract from a poem by Alfred, Lord Tennyson

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All the people who provided metaphysical sustenance during the years at Aston - especially Fiona, Rita, Alison and Val.

Dr. Rosanna Maggi who taught that a 00 number is not required to partake of beverages which are *shaken not stirred.*

And finally but not in any way least, my Family and Friends for their considerable moral support, encouragement and numerous pachyderm gifts.

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CHAPTER 1.
INTRODUCTION.

1.1. Introduction.

There has been considerable popular interest and concern over the effect fossil fuels usage has on the environment. With the concern over holes in the ozone layer, decreasing fossil resources and the increasing carbon dioxide levels in the atmosphere (0.5% increase per year) [1] contributing to the Greenhouse Effect, there has been a concerted move towards finding environmentally friendly renewable energy sources. There are several alternatives including nuclear, wind, wave, solar and tidal technologies. Nuclear energy, although used extensively in France and other countries, is viewed by the general populace with concern as the result of lack of information and the accidents at Three Mile Island (USA) in 1979 and Chernobyl (Soviet Union) in 1986. Wind, wave, tidal and solar are increasingly popular sources of power but cannot replace the existing use of fossil fuels for transportation fuels and chemicals.

Biomass or plant matter is an available alternative resource which may be utilised for the production of fuels or power. It contributes over 2% of the European Unions (EU) primary energy needs [2], mainly as firewood and charcoal and about 60% of it's industrial needs as timber. Beyond the year 2000, agro-forestry biomass will have a potential production rate of around 600 million tonnes of dry material per year and could provide up to 10% of the EU's energy needs of 100 MTOE/yr as a cost effective, sustainable, environmentally acceptable energy source. Biomass has a low sulphur content [3] hence reducing the problem of acid rain and a heat of combustion comparable with lignite coal [4]. Biomass is also effectively carbon dioxide neutral in that new biomass can absorb carbon dioxide at a greater rate than old biomass releases it on combustion [5]. Biomass may be utilised directly as a solid fuel or converted to either liquid or gaseous fuels [6 , 7] which may be substituted for fossil fuels in many applications.

At present there is considerable interest in the production of liquid fuels as substitutes for fossil transportation fuels and for other applications such as the production of electricity in remote areas. The processes by which biomass may be converted to liquid products include biochemical conversion to ethanol [8], gasification coupled with a synthesis step to methanol [9] and liquefaction [10] or pyrolysis [11] to a bio-crude oil although of different qualities. Of the possible biomass conversion processes for liquid products, flash pyrolysis currently appears to be the most attractive due to the ease with which the product may be transported and stored, the compactness of the pyrolysis process and the energy content of the pyrolysis liquid. The products from biomass flash pyrolysis may be upgraded into substitutes for fossil fuel diesel or gasoline, utilised directly in boilers, engines or turbines [12] or used as a feedstock to produce potentially valuable chemicals [13]. This work therefore focuses on the use of biomass flash pyrolysis for the production of liquid fuels.

Pyrolysis has been used for centuries for the production of charcoal from biomass with low temperatures and long reaction times to maximise charcoal yields. The oil crisis of the 1970's sparked interest in producing liquid fuels from non-fossil sources which led to renewed interest in biomass pyrolysis. Several flash pyrolysis processes were developed in the 1980's including the fluidised bed system of Waterloo (Canada) [14], the transport system of Ensyn Technologies Inc. (Canada) [15] (detailed in Section 2.4.) based on the research carried out at the University of Western Ontario, Georgia Tech. (USA) [16] and the vortex reactor system of NREL (USA) [17] (detailed in Section 2.6.). The common features of all these processes are high heat transfer, moderate temperatures (450-550°C) and short vapour residence times (<1 second) [18]. Although biomass flash pyrolysis plants are not as yet operating commercially for liquid fuel production, Ensyn Technologies Inc. (Canada) commercially produce Rapid Thermal Processing (RTP) pyrolysis plants for the production of chemicals [19].

Recent flash pyrolysis innovations include projects funded under the European Commission's JOULE programme [20] which include the transport reactor system developed by CRES (Greece) [21 , 22], the rotating cone reactor developed by the University of Twente (The Netherlands) [23 , 24], the entrained flow system developed by Egemin (Belgium) [25] and the fluidised systems developed by Union Electrica Fenosa (Spain) [26] (detailed in Section 2.5.), and LNETI now INETI (Portugal) [20] and the ablative pyrolyser developed by Aston University (United Kingdom) [27].

There have been several technoeconomic studies carried out [28 , 29 , 30 , 31 , 32] on biomass pyrolysis and upgrading technologies but the results are not comparable on a consistent basis due to the different basic assumptions utilised. In order for the results to be compatible financial criteria such as plant life, feedstock cost, throughput and utilities costs must be consistent. At present, biomass derived liquid fuels appear to be more expensive to produce than fossil fuels and consequently the circumstances under which it will be economically attractive to utilise biomass are uncertain. The development of biomass conversion and utilisation systems will have many benefits including:

- better management of agricultural land no longer required for food production;
- improvements in the environment and reductions in the biomass derived product cost by utilising wastes and residues;
- developing industrial markets;
- securement of long-term energy supplies;
- making a positive contribution to the Greenhouse effect by reducing carbon dioxide emissions; and
- providing opportunities for the socio-economic growth of the less developed regions of Europe particularly those in the South.

The possible introduction of fiscal incentives such as the levying of sulphur and carbon taxes on fossil fuels or capital subsidies for biomass converters [33] will effectively lower the cost of biomass derived fuels with respect to the fossil equivalents making them more competitive (See Chapter 9 for further details).

1.2. Aim of the project.

The aim of this project was to compare selected biomass flash pyrolysis conversion, upgrading and power production technologies in terms of cost and performance, to develop new relationships and models for these technologies and to identify promising technologies and areas of further research.

The selected technologies were analysed to determine the critical process parameters and their effect on the mass and energy balances. Step models were developed for each process incorporating mass and energy balances, utility requirements and capital costs. The models were assembled to form complete processing routes within a dedicated computer program. The program predicts the quantity and quality of the potential products from the processing routes, carries out energy balances and cost estimations and allows the selected technologies to be compared on a similar and consistent basis. The program has been given the acronym **BLUNT** which stands for **B**iomass **L**iquefaction and **U**pgrading by **N**ovel **T**echnologies. Full details on the structure and development of BLUNT can be found in Chapter 4.

1.3. Scope of the project.

BLUNT starts with a prepared biomass feedstock and finishes with a crude, upgraded or refined liquid product or electricity from biomass conversion. The biomass conversion routes modelled in BLUNT are illustrated in Figure 1.1.

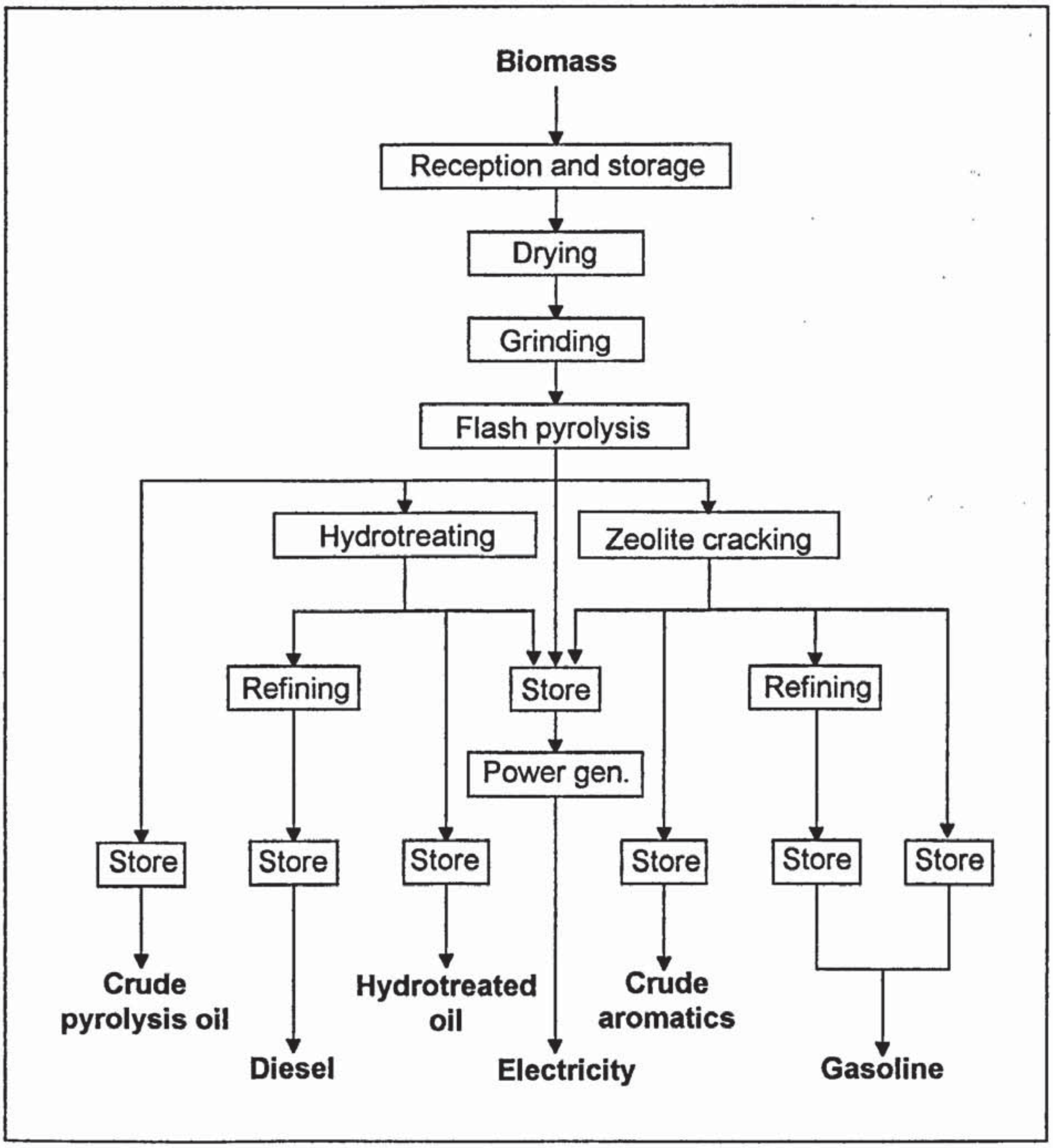


Figure 1.1. Conversion routes modelled in BLUNT.

The raw biomass feedstock must be prepared or pretreated to ensure that it meets the feed requirements for the flash pyrolysis process. The pretreatment steps include feed reception at the plant, storage, drying and size reduction. The pretreatment of the raw biomass prior to conversion by flash pyrolysis is modelled and discussed in Chapter 5.

There are five basic conversion technologies modelled in the program - flash pyrolysis (reviewed in Chapter 2 and modelled in Chapter 6), upgrading by hydrotreating (reviewed in Chapter 3 and modelled in Chapter 7) or zeolite cracking (reviewed in Chapter 3 and modelled in Chapter 8), refining of the upgraded product (modelled in Chapter 8) and power production (modelled in Chapter 8) from the crude or upgraded product.

In addition to modelling the pretreatment and conversion technologies mentioned above, BLUNT includes models to allow the impact of macroeconomic factors on the product cost to be assessed. The factors which are modelled are sulphur and carbon taxes on fossil fuels, irrigation water pricing and capital subsidies. The background, development and modelling of these factors are detailed in Chapter 9.

Economic results from process simulations are given in Chapter 10 and Chapter 11 gives the conclusions both economic and otherwise which have been formed from the work. Recommendations for further work are given in Chapter 12.

CHAPTER 2.

FLASH PYROLYSIS TECHNOLOGIES.

2.1. Introduction.

Pyrolysis, the process traditionally utilised to produce charcoal, produces variable quantities and qualities of charcoal, liquids and gases from biomass, peat or coal [7]. The terms fast or flash pyrolysis are used in this thesis to describe the regime where the formation of char is minimised by rapid heating and the primary product is a liquid [34]. The process typically requires vapour residence times of less than 1 second and moderate reaction temperatures of 400 - 600°C [3]. The organic liquid produced by flash pyrolysis can be used as boiler, furnace or engine fuel or upgraded to produce transport fuels [35].

This Chapter describes the technologies, products and process characteristics of flash pyrolysis which have been considered in the derivation of the models included in BLUNT. It is not intended as a review of flash pyrolysis as this has been thoroughly documented elsewhere [7, 11] but as an introduction to the technologies, an identification of the important features as regards modelling and as a background to the flash pyrolysis models described in Chapter 6. Section 2.2. gives information on the organic liquid, off-gas and char products produced from flash pyrolysis and Section 2.3. considers the process variables which affect flash pyrolysis. The main variables or factors each of which are considered in this thesis are:

- the feedstock (discussed in Section 2.3.1.),
- the feed heating rate in the reactor (discussed in Section 2.3.2.),
- the vapour residence time in the reactor (discussed in Section 2.3.3.) and
- the pyrolysis temperature (discussed in Section 2.3.4.).

Sections 2.4. - 2.6. consider the three flash pyrolysis technologies which, together with the Generic process (described in Chapter 6), are modelled in the program. The three flash pyrolysis technologies modelled were selected because they are all operating at demonstration or commercial level. The flash pyrolysis technologies selected were the Ensyn RTP process (detailed in Section 2.4.), the Union Fenosa process (detailed in Section 2.5.) and the NREL vortex process (detailed in Section 2.6.). Section 2.7. summarises the key issues identified in this Chapter.

2.2. The products from flash pyrolysis.

2.2.1. The liquid product.

The liquid product from flash pyrolysis is an organic liquid known as pyrolysis oil or bio-oil or bio-crude oil and may be a fairly viscous or free flowing substance depending on the process. Typical physical properties for biomass pyrolysis oils may be found later in Tables 2.2. and 2.5. for the products from the Ensyn RTP [19] and NREL vortex [17] processes respectively. The organic liquid has a high oxygen content, up to 45wt% (total liquid), is polar and may contain up to 35wt% of water. Other properties of biomass pyrolysis liquids include a density of approximately 1.2kg/dm^3, pH between 2.8 - 3.8, heating values between 14 - 18.5MJ/kg and an elemental composition claimed to be close to that of the feedstock [36]. The relationship between feedstock and bio-oil composition in relation to hydrogen, oxygen and nitrogen contents is discussed further in Chapter 6.

The water content of the pyrolysis oil is determined by the feed moisture content and the reaction water formed from cracking of the chemical constituents of the biomass. The water content has several effects including reducing the heating value and viscosity, affecting the pH, influencing the chemical and physical stability of the oil; and reducing waste water disposal and hence potential pollution problems [3]. The water is difficult to remove since significant physical and chemical changes are produced in the liquid when evaporated or distilled at temperatures around 100°C [3]. Pyrolysis liquids are thus normally utilised on a "wet" or "as produced" basis. Information on the chemical and physical properties and the related storage problems of bio-oils have been published [3, 37 , 38 , 39 , 40 , 41]. However, these details are not strictly related to the modelling of biomass flash pyrolysis processes and so have not been included in this thesis.

The liquid product may contain a high level of particulate char and ash carried over from the reactor. The char particle sizes are dependent on the cyclone separation efficiency but ranges up to 1000µm have been reported for Ensyn RTP oils [42]. The efficiency of the separation process and hence the amount of recovered char could have a potential impact on the energy efficiency of the flash pyrolysis conversion process if the char is utilised as fuel [43] (See Chapter 6 for further details). High levels of suspended solids or large size particles could affect the downstream processing or utilisation of the bio-oil for example, in utilising the bio-oil for power generation the turbine injection nozzles could become blocked. There has been little research carried out on secondary separation or recovery of the suspended char from the condensed liquid. Cold filtration of the pyrolysis oil to remove suspended solids has only been partially successful [42] and utilisation of the liquid with suspended char may be necessary until separation techniques have been developed.

Crude pyrolysis oils are not miscible with fossil derived oils and must be converted or upgraded to give a product capable of being assimilated into a conventional fuel marketing infrastructure. The background and technologies for upgrading biomass pyrolysis liquids is discussed in Chapter 3 and modelled in Chapters 7 and 8. The possible alternatives to upgrading biomass pyrolysis liquids are dedicating the liquid to a specific application such as boiler or turbine fuels or to construct a unique bio-oil storage, distribution and utilisation system. The latter alternative would be expensive and not viable under the present fossil fuel based economic climate. Combustion testing of crude pyrolysis liquids has been carried out in North America [44] and Europe [45] with few reported problems. Wood derived pyrolysis oil has also been successfully tested in a pilot-injection diesel engine in Finland [46].

2.2.2. The solid char product.

Solids separation from the primary product stream is normally carried out by a cyclone downstream of the reactor to minimise reaction time. The char from flash pyrolysis has a low ash content and relatively high heating value. The char has a bulk density from 150 - 300kg/m^3 and a heating value of around 30GJ/tonne [36]. The char may be combusted to provide energy for the pyrolysis [43] or feed drying processes or may be used as a substitute for activated carbon. For further details on char use as fuel for the pyrolysis process see Chapter 6.

2.2.3. The gaseous product.

The gaseous product from flash pyrolysis has a density of 1.2kg/m^3 and is a medium value fuel gas with a heating value ranging from 15 - 22MJ/Nm^3 [36]. The gas is composed of hydrogen, carbon dioxide, carbon monoxide, methane and both saturated and unsaturated hydrocarbons [47]. The heating value of the gas is enhanced if it is used and kept hot from the sensible heat and relatively high organic content [3]. If the pyrolysis system incorporates a fluidised bed, a portion of the pyrolysis off-gas may be recycled as the fluidising gas. This has not yet been demonstrated and there are several potential problems including condensation or deposition of organic compounds if the off-gas cools while circulating through the reactor system. It is also not known if recirculating the off-gas will have any affect on the quality of the bio-oil product. Alternatively, the off-gas may be combusted to provide heat for feed drying, reducing the external energy demand of the drying process [43]. The use of pyrolysis off-gases for fuel is discussed further in Chapter 6.

2.3. Factors which affect the flash pyrolysis process and products.

2.3.1. The feedstock.

The composition, moisture content and particle size of the feedstock affect the flash pyrolysis process and the yield and composition of the products. There has been a considerable amount of work carried out on the flash pyrolysis of the components of biomass particularly cellulose and lignin which has been reviewed elsewhere [48 , 49]. However, as there is little correlation between product yields and characteristics and feed characteristics, the impact of this basic work on process model development is minimal. And since separation of the chemical components of biomass prior to pyrolysis has not been seriously proposed, only the pyrolysis of whole feedstocks is considered in this thesis.

The pyrolysis of a biomass particle involves heat transfer to the particle, chemical reactions within the particle and escape of volatiles from the particle. If the particle is wet the pyrolysis process is delayed as the water must be removed from the particle by evaporation before pyrolysis commences [50]. An increase in the moisture content of a biomass particle has been shown to delay the onset of pyrolysis and to lower the initial rate of temperature rise within the particle [51]. Wet or raw wood typically contains between 40 - 60% water (wet basis) which is normally reduced to 5 - 10% (wet basis) by drying prior to pyrolysis [6]. The drying step decreases the energy demand on the pyrolyser and reduces the moisture content of the crude pyrolysis oil product. Drying of biomass prior to thermochemical conversion is discussed in Chapter 5 which deals with the pretreatment of biomass. Since crude pyrolysis oil is hydrophilic, the feed moisture content will also affect the water content and hence the physical properties of the product as mentioned in Section 2.2.1. The moisture content of the feedstock is also a major factor in the pyrolyser energy balance [43] as will be demonstrated in Chapter 6.

In addition to the chemical properties which affect biomass pyrolysis (feedstock chemical composition and water content), the physical property of biomass particle size affects the rate of heat and mass transfer. The time required for a biomass particle to lose 95% of it's total weight loss has been shown to increase with particle size [47]. In addition, the carbon content of the char obtained from pyrolysis increases with biomass particle size possibly due to secondary interactions between the volatile and gaseous products within the hot char [47].

2.3.2. Heating rate.

Flash pyrolysis is characterised by rapid heating rates [11] and the rate of heating is significant in the ratio between char and liquid products with slower heating increasing the formation of char [48]. Figure 2.1. shows the effect of the heating rates employed in three different thermochemical technologies on the liquid, off-gas and char product yields. Flash pyrolysis processes, for example the Waterloo fluidised bed system [6], have heating rates typically between 10 - 200°C/second [52] and high liquid yields with low char and gas yields, as shown in Figure 2.1. Carbonisation processes, such as the Bio-Alternative system [53] have low heating rates and slow or conventional pyrolysis processes, such as the Alten system [54], have medium heating rates, typically in the region of 0.1 - 1°C/second [53].

The heating rate of the flash pyrolysis system is influenced both by the reactor type and the particle size of the feedstock. The systems most commonly used for flash pyrolysis are fluidised bed (e.g. the University of Waterloo system [6]), ablative (e.g. the NREL vortex system [17]) and transport (e.g. the Ensyn RTP system [19]) reactors. Fluidised bed pyrolytic reactors utilise a hot, inert upflowing gas to fluidise the feedstock particles and the reactor bed. The solids are held in suspension by the upward flow of the inert gas promoting high mass and heat transfer rates and good mixing. Ablative pyrolysis systems involve contacting the biomass particles with a hot surface and "melting" the feedstock. Heat transfer occurs through a thin liquid film between the hot surface and particle which is then vaporised. Ablative systems have very high heating rates and can use larger feed particles than fluidised systems as the feed particle size does not have such an influence on the heating rate [55]. Transport reactor systems contact the feed material with a hot solid heat carrier and a fluidising gas resulting in gas-solid heat transfer from the fluidising gas and solid-solid heat transfer from the hot fluidising solid. There is also ablation of the feed particles by the solid heat carrier.

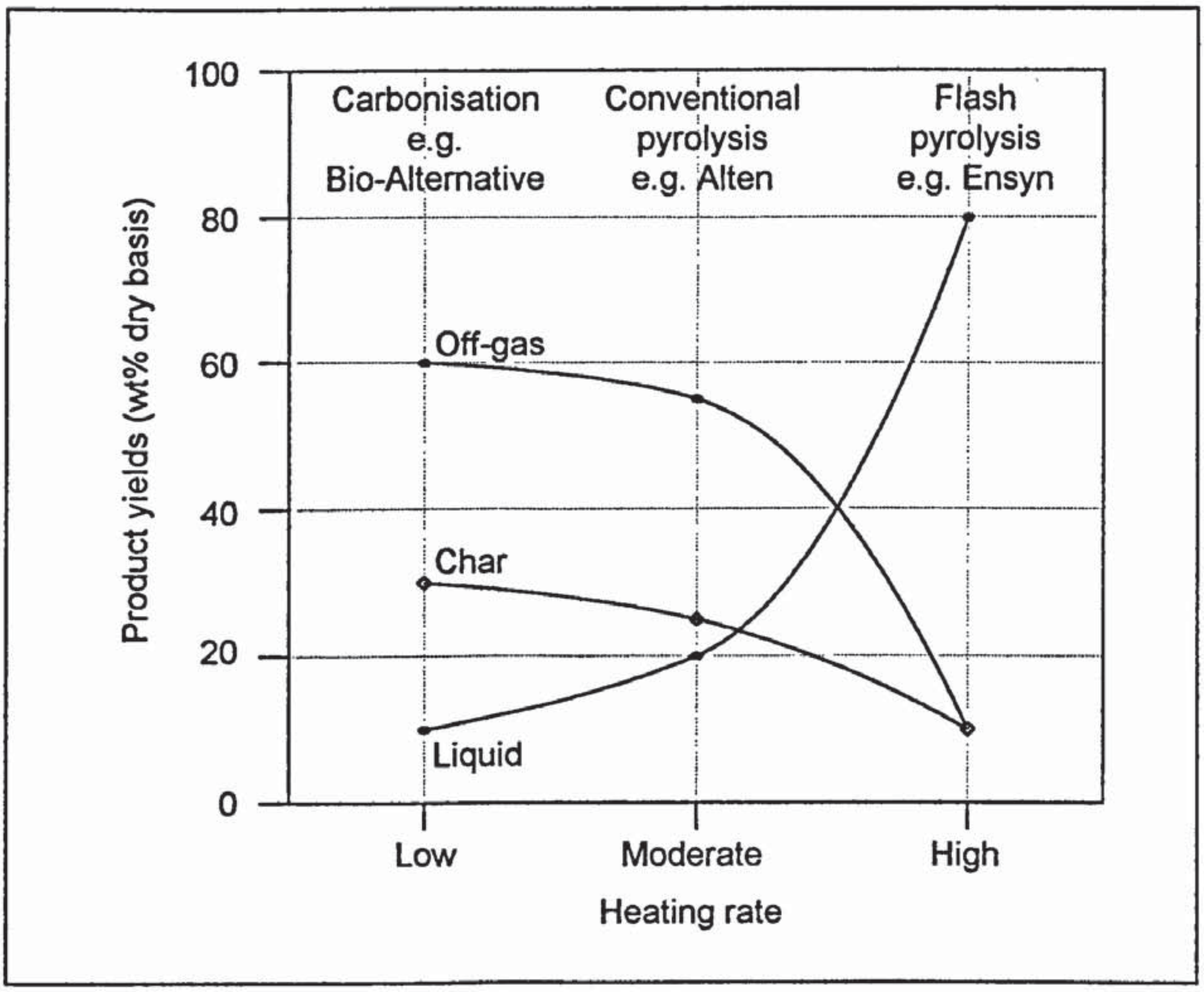


Figure 2.1. The effect of heating rate on product yields for three thermochemical processes [6, 53, 54].

A disadvantage of systems such as fluidised beds which rely mainly on gas-solid heat transfer is that the biomass particles have to be small to ensure rapid heating rates and hence high liquid yields. Transport and ablative pyrolysis processes both involve ablation of the feed to remove the char formed on the outside of the feed particle and so expose fresh biomass for reaction. In addition, the use of a solid heat transfer medium leads to a higher heat transfer and more compact reactor system than a fluidised bed design due to the higher heat capacity of solids than that of gases.

2.3.3. Vapour residence time.

For maximum liquid yields the vapour residence time in the pyrolysis reactor must be short to prevent secondary cracking of the liquid products to form gaseous products. The vapour residence time also affects the chemical composition of the liquid product including the carbon to hydrogen ratio and the oxygen content [56]. There is little published data for the effect of residence time on product yields at the low temperatures (400 - 600°C) of interest and so the effect of residence time on product yields has not been modelled in this thesis. However, Ensyn have published data for the flash pyrolysis of IEA Poplar at 650°C [48] and residence times between 0 -900ms which is shown in Figure 2.2.

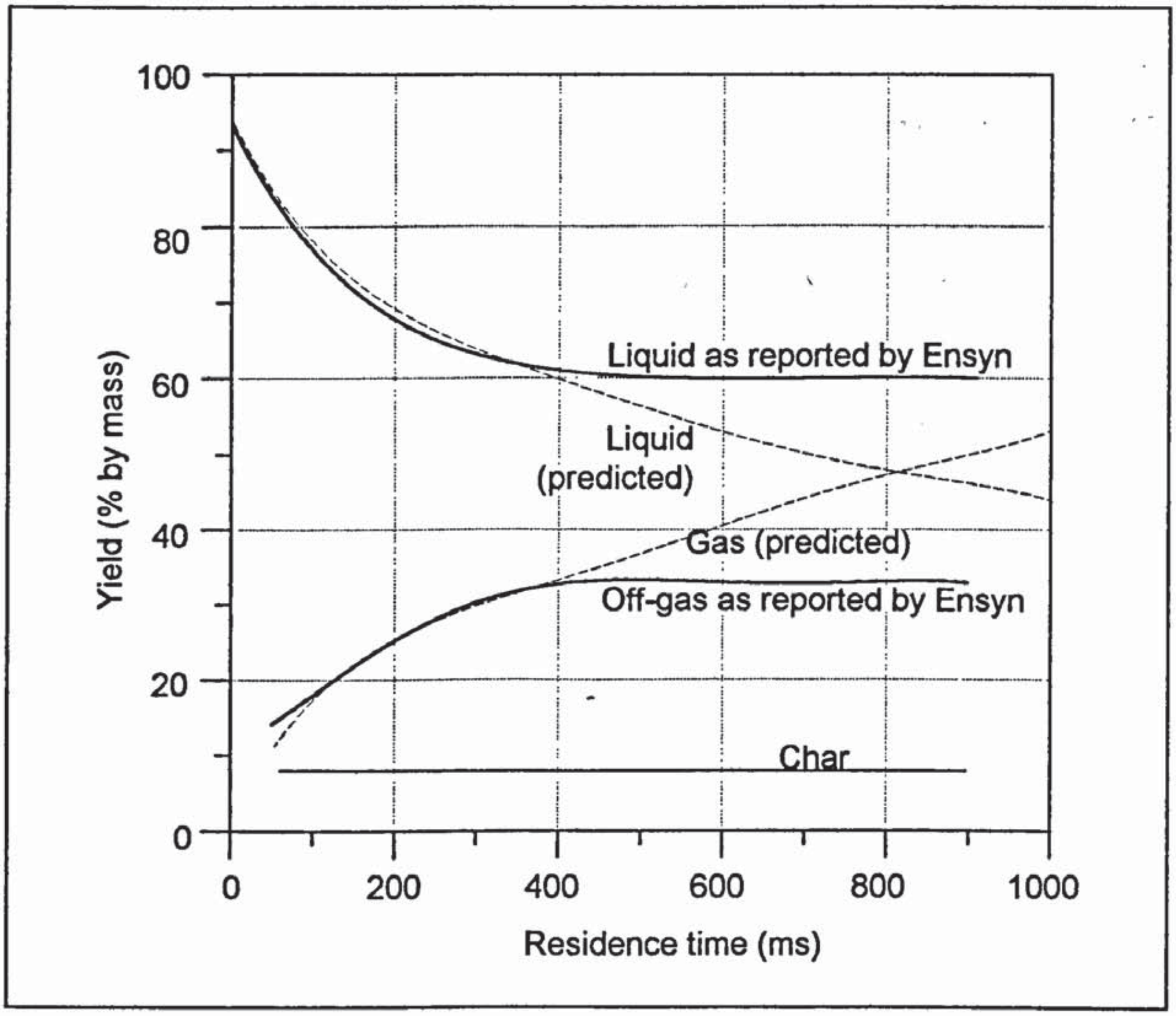


Figure 2.2. The effect of vapour residence time on flash pyrolysis liquid product yields for an isothermal system at 650°C [48].

The reported Ensyn liquid yield line shown in Figure 2.2. is shown as horizontal at a residence time around 400ms and maintains a constant yield between 400 and 900ms. The Ensyn process is representative of an isothermal system due to the temperature in the reactor being maintained by recycling and reheating of the solid heat carrier. For an isothermal system the liquid yields would be expected to decrease and the gas yields increase with time due to thermal cracking of the vapours to form secondary products. The liquid and gas yields proposed by Ensyn in Figure

2.2. are more representative of an adiabatic system, e.g. a fluidised bed, where the pyrolysis temperature fluctuates than an isothermal system.

Figure 2.3. illustrates the effect of pyrolysis temperature on the rate of cracking of pyrolysis vapours. As can be seen from the Figure, the rate of vapour cracking increases with increasing pyrolysis temperature. In an isothermal system if the pyrolysis temperature is above 500°C the vapours crack to form secondary products and below 400°C the vapours condense. If the temperature is maintained at 450°C for more than one second, the lignin which has not already depolymerised does so. This relationship between pyrolysis temperature and vapour residence time for an isothermal system generates the proposed plots of how the Ensyn data should look which are included in Figure 2.2.

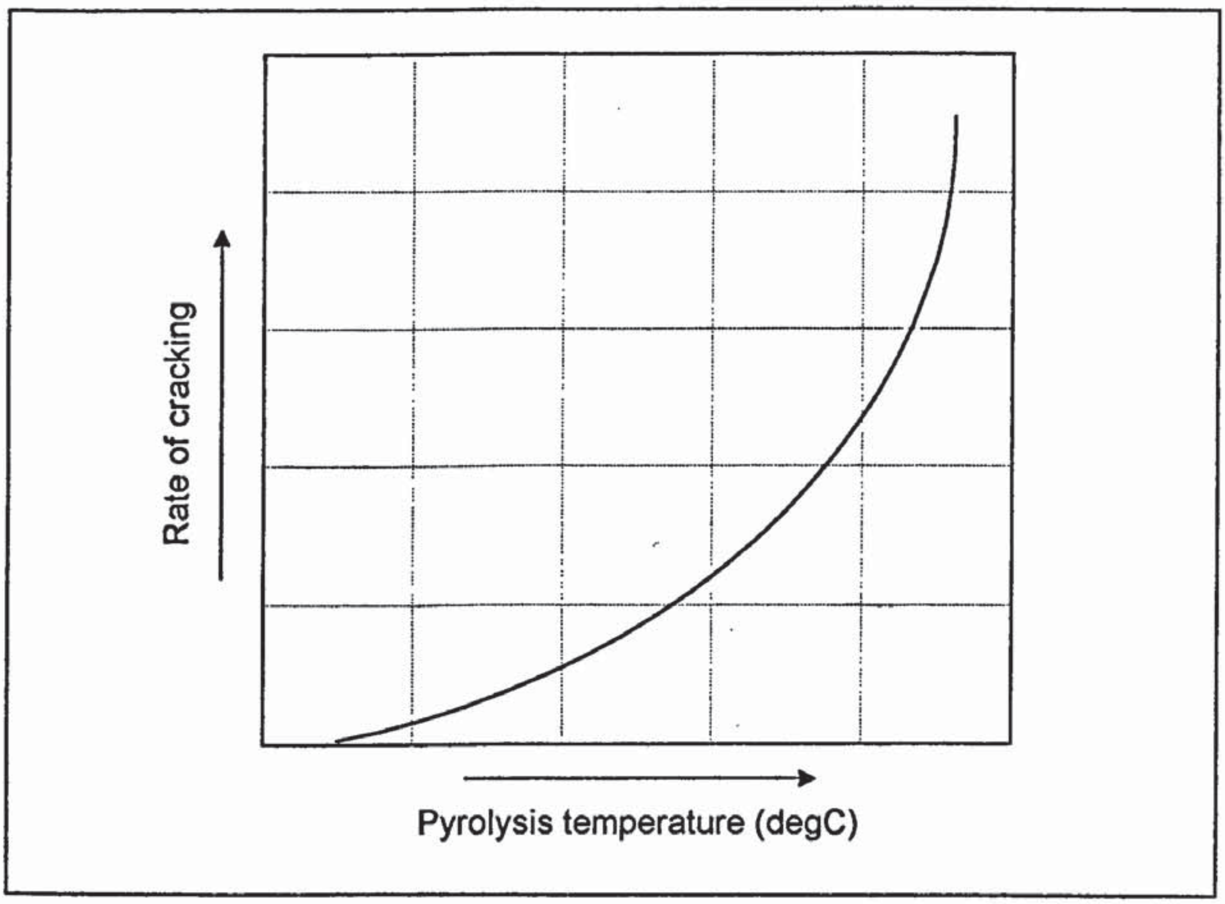


Figure 2.3. The effect of temperature on the rate of cracking during the flash pyrolysis of biomass.

For an adiabatic system the pyrolysis temperature fluctuates as the reactor contents cools down and are then reheated by a fresh influx of heat. This fluctuation in pyrolysis temperature results in a continuous cycling between vapour cracking and condensation effectively stabilising the rate of vapour cracking and producing the plots proposed by Ensyn in Figure 2.2.

2.3.4. Reactor temperature.

The product distribution from flash pyrolysis is markedly influenced by the pyrolysis temperature [57]. Typical product distribution curves for Aspen-Poplar from the Waterloo flash pyrolysis system are shown in Figure 2.4. [58] for vapour residence times between 0.52 - 0.616 seconds.

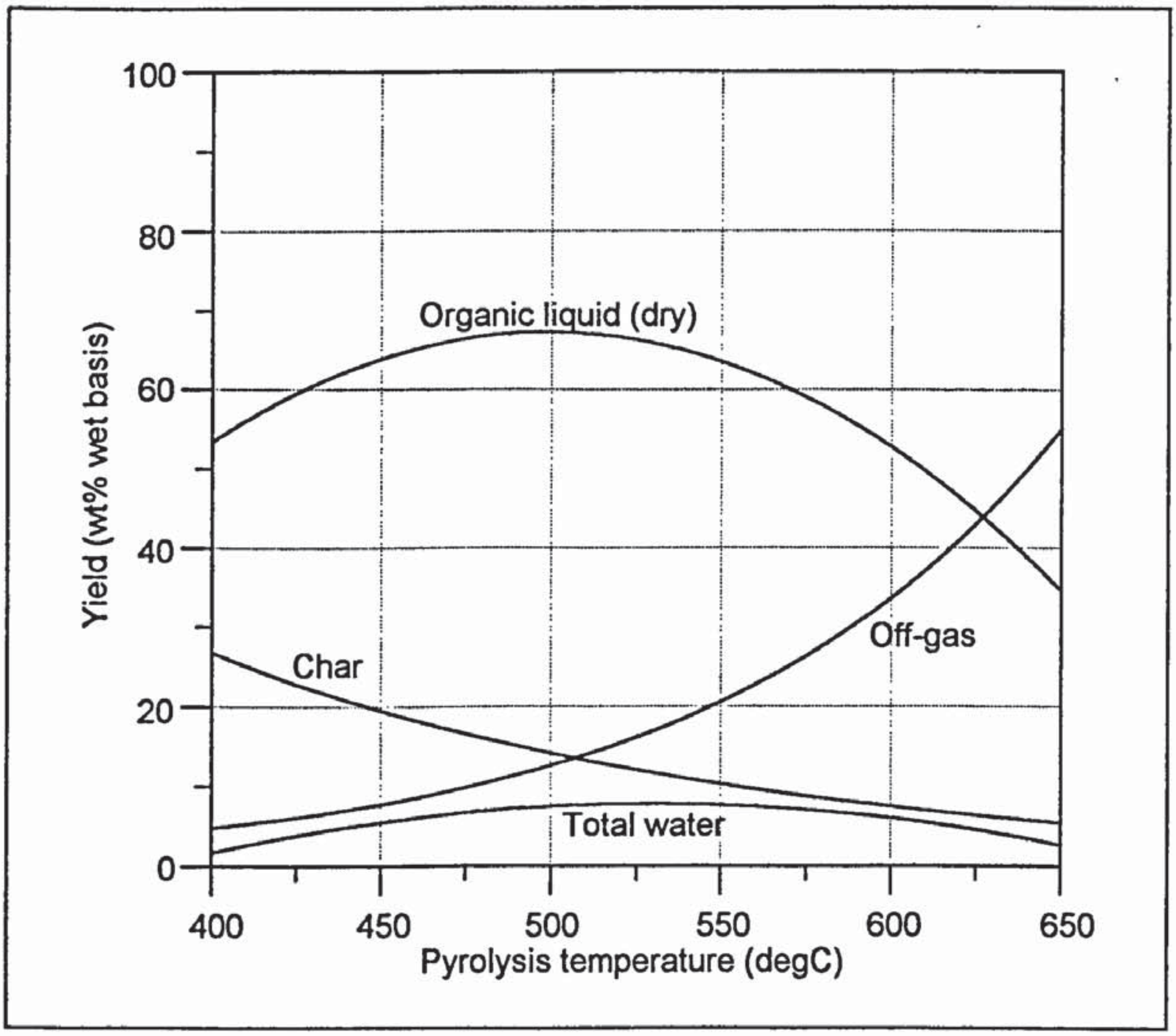


Figure 2.4. The effect of pyrolysis temperature on the product yields from Aspen-Poplar [58].

As can be seen from Figure 2.4. the organic liquid yield is represented by a domed curve with the maximum yield around 500°C. Above 500°C the organic yields decrease as the primary vapours are cracked to form gaseous products. The quantity of char formed from flash pyrolysis decreases with increasing temperature and the gas yield increases as secondary reactions occur. The water yield appears to increase slightly as the temperature rises up to 500°C after which the yield decreases. The total water yield (feed and reaction water) shown in Figure 2.4. is not representative of the Waterloo system showing only a 8wt% yield (wet basis) at 500°C when the expected yield should be around 17.5wt% (wet basis) [59]. The error in water yield may be due to its determination by Karl-Fischer which is known to give low results with pyrolysis liquids [59].

In addition to affecting the relative proportions of the products, the pyrolysis reaction temperature alters the chemical composition of the pyrolysis oil product. With increasing processing temperature the oxygen content and the hydrogen to carbon ratio of the liquid product declines as illustrated in Figure 2.5. [60] These changes represent the conversion of highly oxygenated pyrolysis compounds to more thermally stable, less oxygenated, species with more aromatic structures.

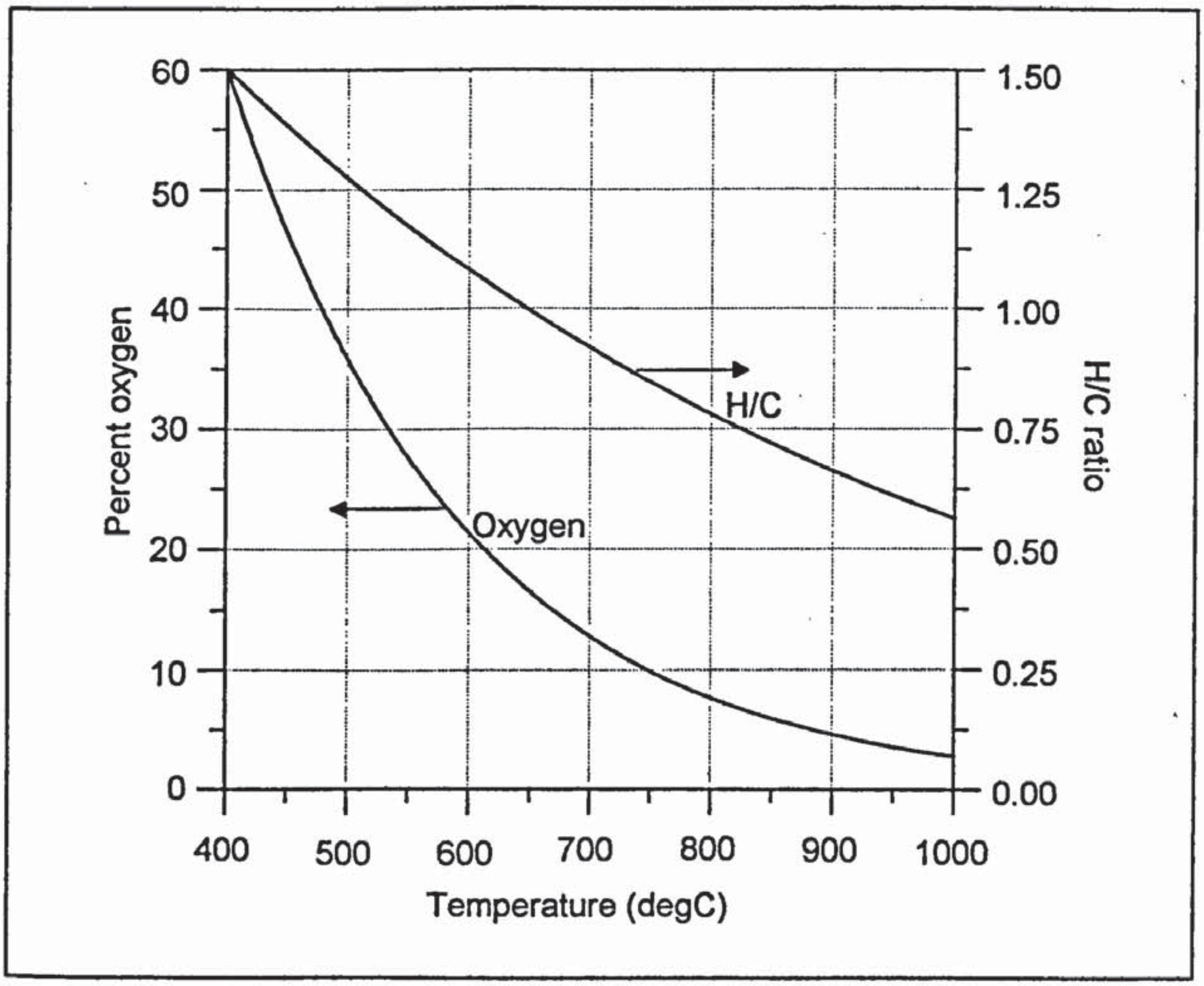


Figure 2.5. Elemental composition of dry liquids produced over a range of temperatures [60].

2.4. The Ensyn RTP pyrolysis process.

Ensyn Technologies Inc. (Ottawa, Canada) was incorporated in 1984 to develop and market flash pyrolysis technology for the production of chemicals. The Ensyn process utilises technology developed from the Ultrapyrolysis work carried out at the University of Western Ontario [15, 61 , 62] in the late 1970's. The Ensyn recirculating, transported-bed system [63] is designed to allow rapid heat transfer to the feedstock with control of the short residence time. Biomass and fossil fuel feedstocks have been processed in Rapid Thermal Processing (RTP) plants with capacities from 0.3 - 1000 kg/hour, reactor residence times of 0.03 - 1.5 seconds and temperatures between 400 - 950°C [19].

2.4.1. Process description.

In the Ensyn system [64], feed passes from a hopper, through metering and feed screws into the base of the reactor where it is mixed with the solid heat carrier. The product leaves the reactor and passes through a primary cyclone where the solid heat carrier is removed. The heat carrier is reheated before being recirculated to carry heat to fresh feedstock. A second cyclone removes the char and inorganic fines consisting of ash and entrained sand from the primary product stream. The secondary cyclone may be coupled to an in-plant recovery system which utilises the char for process heat generation. The product stream passes through condensers, filters and demisters where the liquid product is condensed and recovered.

Further details on the Ensyn system are not available due to commercial confidentiality. A flowsheet for the Ensyn process is given in Figure 2.6.

2.4.2. Feedstocks processed.

The RTP reactors require feedstocks with maximum particle sizes of 6mm and moisture contents below 10% (wet basis) [19]. Biomass feed materials which have been processed in the Ensyn plants include wood (for example, Red Maple and IEA Poplar), cellulose, lignins, pulping residues and agricultural residues [65].

2.4.3. Product yields and properties.

The maximum total liquid yields obtained from dried poplar with a 10wt% moisture content is 83wt% (dry feed basis) at 500°C and 250ms [66]. Table 2.1. gives typical product yields for the Ensyn RTP process [67].

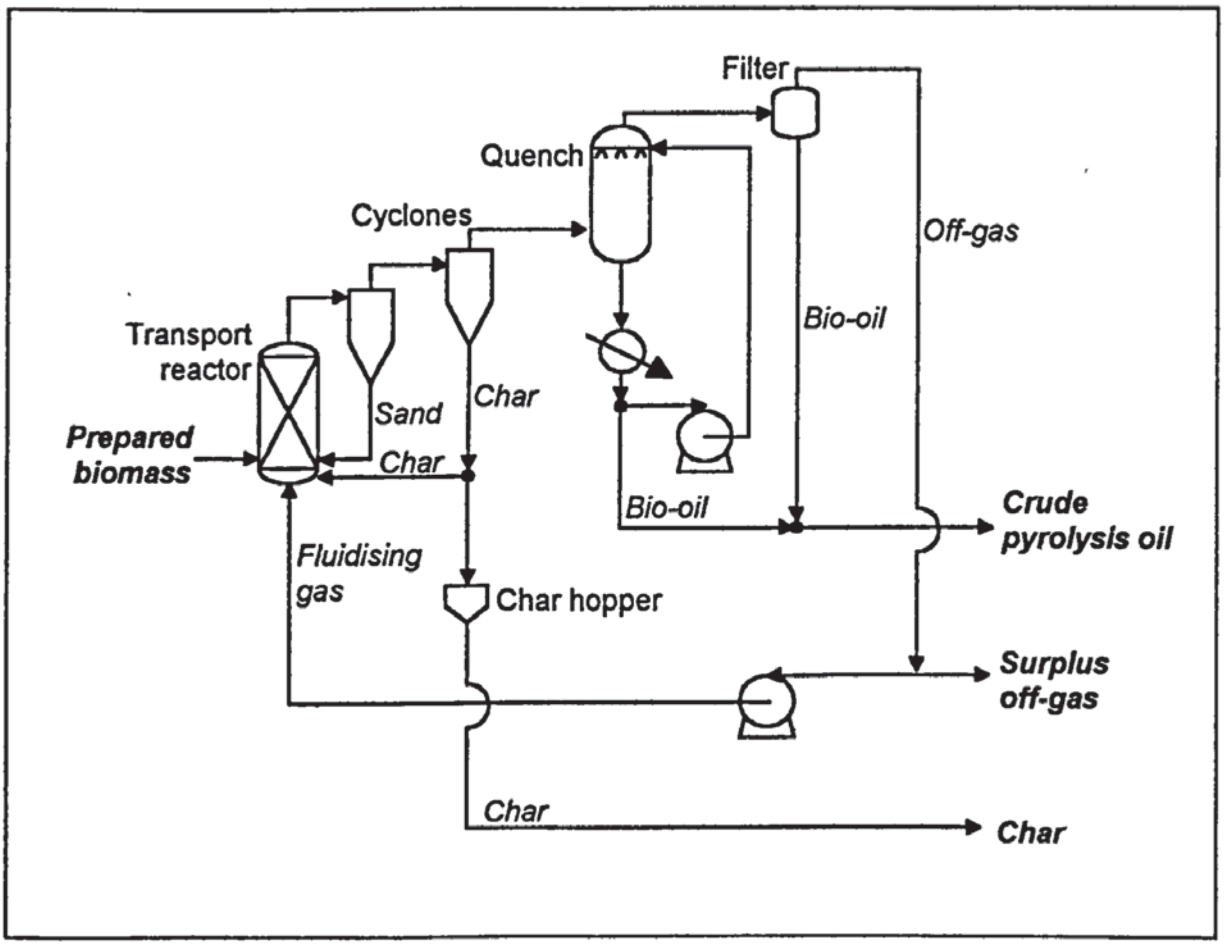


Figure 2.6. Simplified flowsheet for the Ensyn process [64].

Table 2.1.
Product yields for the Ensyn RTP pyrolysis process at 525°C and vapour residence time of 300ms [67].

	Product yield (wt% on dry feed)
Pyrolysis oil	74
Char	15
Off-gas	11

The products of interest from the Ensyn process include liquid fuel oils for use in boiler and turbines, speciality chemicals, commodity chemicals, polymers including copolymers and resins, petrochemicals including ethylene and other olefins and liquid transportation fuels. Table 2.2. gives some properties of wood derived pyrolysis oils from the Ensyn process [3, 68].

Table 2.2.
Ensyn RTP bio-oil characteristics [68].

Property	**Typical values**
Moisture content (%)	16
Acidity (pH)	2.5
Specific gravity/density (kg/l)	1.21
Higher heating value, dry basis (MJ/kg)	23
Higher heating value as produced (MJ/kg)	19.3
Viscosity (cSt @ 40°C)	51
Pour point (°C)	-23
C/H ratio	9.1
Elemental analysis (%)	
Carbon	56.4
Hydrogen	6.2
Nitrogen	0.2
Sulphur	< 0.01
Ash	0.1
Oxygen (by difference)	37.1
Chemical composition (% by mass)	
Water	20-30
Lignin fragments	15-20
Aldehydes	10-20
Carboxylic acids	10-15
Carbohydrates	5-10
Ketones	1-5
Phenols	2-5
Alcohols	2-5

2.4.4. Product use.

A 25 tonnes/day Ensyn RTP system has been in commercial operation since 1992. The crude liquid product produced is delivered to existing extraction processes where speciality chemicals are isolated and recovered for commercial sale [19]. The residue after chemical extraction is regularly fired in a 2 GJ/hr fire-tube, single-pass Kewanee boiler normally fuelled by natural gas. A two-phase air-atomisation nozzle assembly with large fuel passages was designed and fitted to the boiler to allow the use of filtered pyrolysis oil [69].

A series of flame tunnel tests were conducted on Ensyn RTP oil to compare their combustion and heat transfer performance to commercial grade heating oils. The filtered bio-oil was preheated in-line to 90°C and completed a three-hour test burn with flue gas oxygen levels maintained at 5.5% [70]. The thermal heat transfer and gas temperature profiles of the bio-oil were comparable with those of fossil fuels. However, the particulate and carbon monoxide levels in the bio-oil combustion flue gas were higher than for fossil fuel flue gases and were probably due to incomplete burnout of the char.

Pyrolysis oil from the RTP III system has been tested in a single cylinder Petter engine with 5% cetane enhancer to simulate dual fuel firing. Although there was a greater ignition delay than for diesel, the bio-oil burned rapidly. The combustion duration and exhaust emissions are reported to be similar to conventional diesel [51, 71]. The oil has also successfully fired in a dual fuel mode in a modified four cylinder industrial diesel engine.

2.4.5. Summary.

Ensyn are the only company who commercially produce pyrolysis reactors with the performance guarantee of a 70% liquid yield. A 25 tonnes/day pyrolysis plant is being constructed and the construction of two 100 tonnes/day plants in the USA is being negotiated [72]. The technology utilised in the Ensyn process, a circulating fluidised bed or transport reactor, has been demonstrated in the petrochemical industry and no major problems are anticipated in constructing larger scale plants. There has been little experience in utilising recycled off-gas as the fluidising gas and there may be problems (as discussed in Section 2.2.3.). The minimisation of char and ash in the liquid product has still to be addressed.

2.5. The Union Fenosa flash pyrolysis process.

Union Fenosa (Spain) have constructed a 200kg/hour biomass flash pyrolysis pilot plant [73] which has been in operation since 1992. The process is based on the fluidised bed flash pyrolysis process developed by the University of Waterloo (Canada) [6,14]. The Fenosa plant encountered several difficulties on start up and modifications to equipment were required. However, the plant has been operating for 300 hours since July 1993 and is currently processing 160kg/hour of dry biomass [74].

2.5.1. Process description.

The wood feedstock is dried utilising hot air produced by a propane burner. The feed is classified with oversized particles being returned for regrinding.

The prepared feed passes through a cyclone to remove it from the gas stream before being conveyed to the reactor feed hopper. The feed is pneumatically fed to the fluidised bed reactor which is cylindrical with a fine sand bed located above the nozzle plate. The bed is fluidised by recycled process gas which is heated by a propane burner. The reactor operates at 500°C with a fluidising gas temperature of 750°C and vapour residence time of 0.5 - 1.0 seconds [74]. The product stream leaves the reactor at a temperature of 350 - 400°C and passes through a cyclone for removal of the char. The pyrolysis vapours and gas are cooled on leaving the cyclone to just above the liquid dew point using three water cooled heat exchangers in series. The liquid product is recovered and stored while some of the off-gas is returned to the reactor for use as fluidising gas and the remainder is removed from the system. A simplified flowsheet for the Union Fenosa process is given in Figure 2.7. [73].

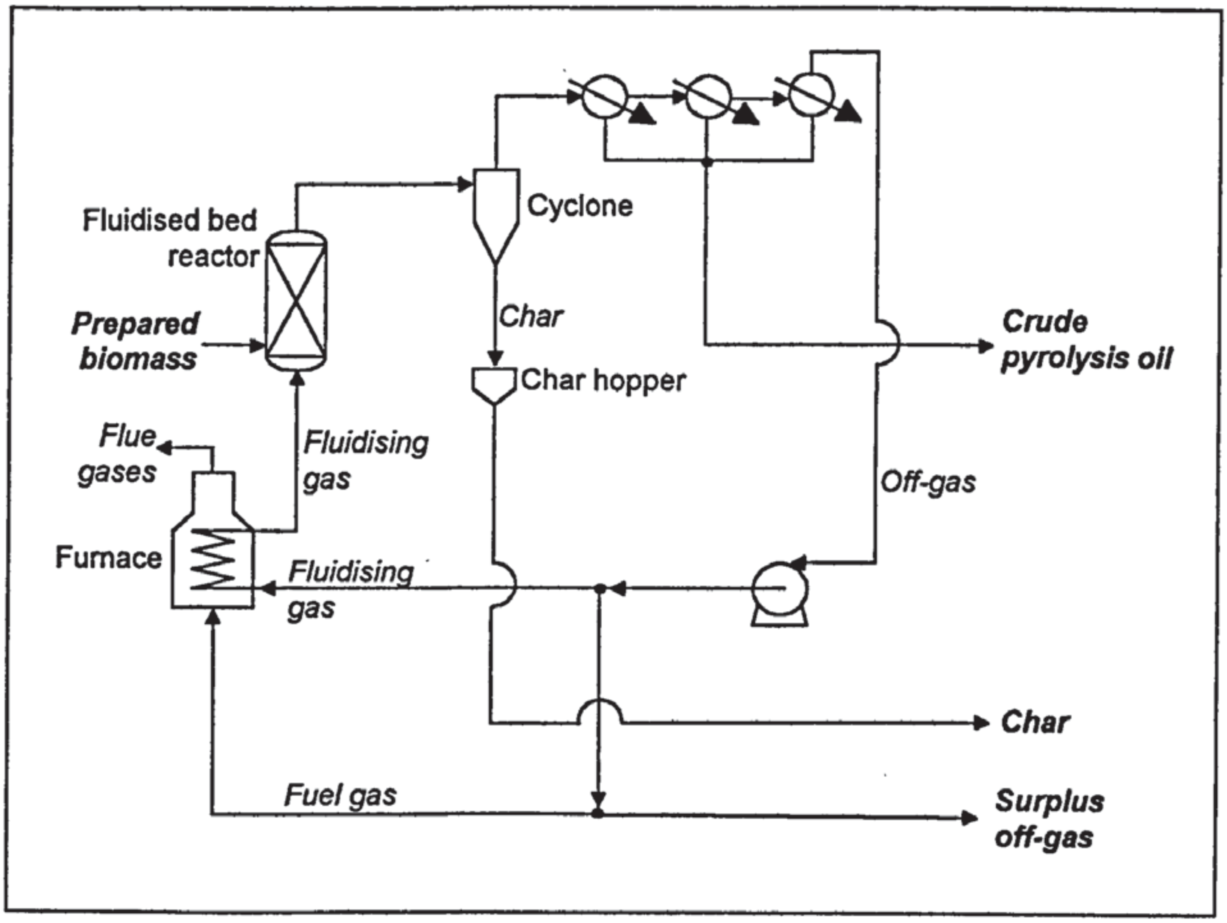


Figure 2.7. Simplified flowsheet for the Union Fenosa pyrolysis process [73].

2.5.2. Feedstocks processed.

To date wood waste and forestry biomass have been processed with raw moisture contents between 30 - 45% [74].

2.5.3. Product yields and properties.

The product yields reported for the process are given in Table 2.3.

Table 2.3.
Product yields for the Union Fenosa flash pyrolysis process [73].

	Product yield (wt% on dry feed)
Pyrolysis oil (crude, wet)	60.0
Char	15.0
Off-gas (by difference)	25.0

2.5.4. Product use.

There were early operating problems at low throughputs which delayed the analysis and use of the Fenosa oils. Consequently, apart from the early preliminary product characteristics, there have been no published results or applications of the Union Fenosa pyrolysis oil.

2.5.5. Summary.

The Union Fenosa process is expected to be able to operate at 2 tonnes/day without any problems. At present the system is being optimised to provide design data for scale up. Samples of the bio-oil produced are being collected for characterisation and testing.

2.6. The NREL vortex pyrolysis process.

The National Renewable Energy Laboratory, NREL, have been developing an ablative flash pyrolysis system since 1980 [18]. The NREL system consists of a vortex reactor, utilising solid conductive heat transfer, which can process 20 - 25kg/hr of feedstock. The technology is being scaled up by Interchem Industries Inc. [13, 75 , 76] to convert 32.7 dry tonnes per day of sawdust into fuel oil replacements and charcoal. Construction of the Interchem plant started in Missouri in 1990 however, it was transferred to Kansas where it has undergone testing and redesign including a new reactor system designed by NREL.

2.6.1. Process description.

The feedstock enters the system where it is entrained and mixed with a high temperature, high velocity carrier gas stream, typically nitrogen or steam. The feed particles enter the vortex reactor tangentially at speeds of over 100m/s [77] and are forced to the reactor wall by high centrifugal forces. Raised helical ribs along the wall of the reactor force the particles into tighter helical paths than would normally occur and increases the contact of the particles with the reactor wall. The reactor wall temperature is limited to a maximum of 625°C to ensure production of a liquid film between the wall of the reactor and the particle, which then vaporises and leaves the reactor. Higher wall temperatures result in solid coke formation on the wall which prevents rapid particle movement and quickly fouls the heat-transfer surface of the reactor. The 625°C wall temperature results in vapour exit temperature and residence time of 525°C and 0.3 seconds respectively and gives the maximum yield of vapours for most biomass feedstocks [18]. A schematic diagram of the process is shown in Figure 2.8. [18].

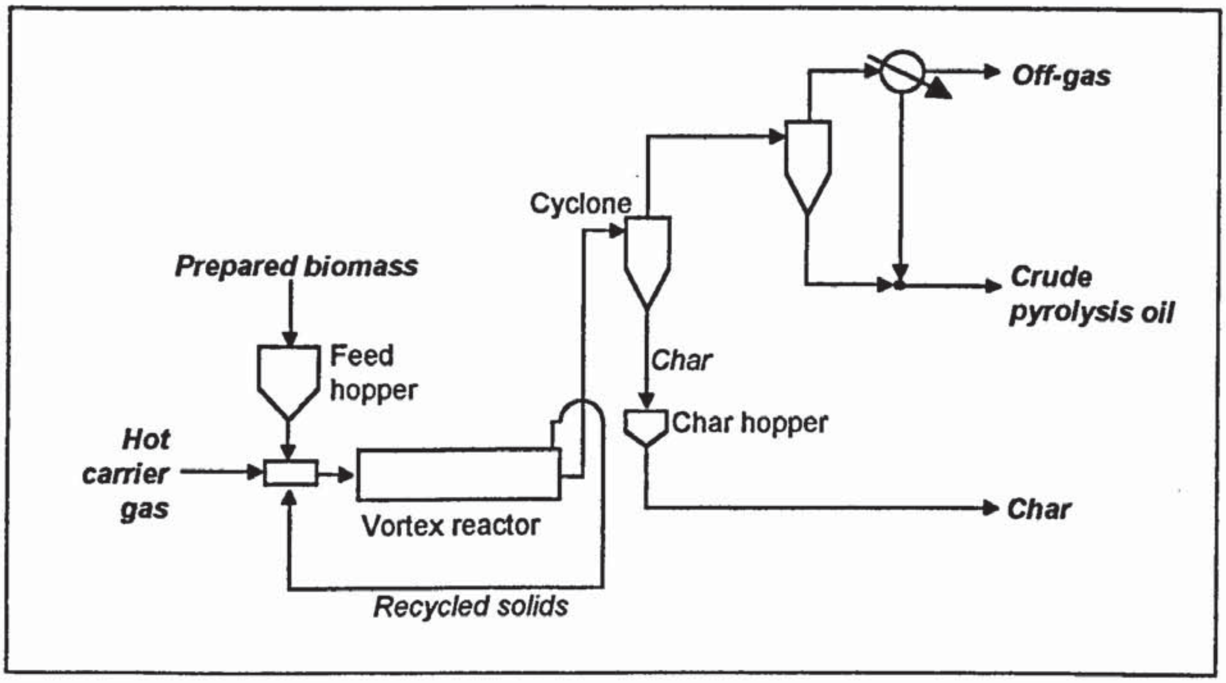


Figure 2.8. Simplified flowsheet for the NREL vortex pyrolysis process [18].

An insulated recycle loop at the exit of the reactor returns partially pyrolysed feedstock and any large char particles to the reactor entrance. Fine char particles (< 50 micrometers in size), gases and vapours leave the reactor through an axial exit. The product stream passes through a char cyclone, maintained above 125°C, where the char is removed as a dry powder [78]. The vapours pass to the first heat exchanger where the condensed liquids and water are retained in the receiver. The cooled gas stream then passes through a series of heat exchangers before being flared.

2.6.2. Feedstocks processed.

Oak, southern pine and switchgrass have been processed to generate oils for combustion testing. The feedstocks were milled and screened through a 2.175mm (1/8 inch) sieve and dried at 105°C to moisture free conditions.

2.6.3. Product yields and properties.

Typical product yields from the NREL vortex system are given in Table 2.4.

Table 2.4.
Product yields for the NREL vortex pyrolysis process [30].

	Product yield (wt% on dry feed)
Pyrolysis oil (wet)	74.0
Char	14.0
Off-gas (by difference)	12.0

Table 2.5. gives some properties of wood derived pyrolysis oils from the NREL system.

Table 2.5.
NREL fast pyrolysis wood-oil characteristics [66].

	Dry basis	**As produced**
Moisture content (%)		15
H/C ratio (atomic)	1.78	1.26
O/C ratio (atomic)	0.68	0.55
Density (g/cm^3 @ 55°C)	1.23	
Heating value, dry basis (MJ/kg)	22.3	20.3
Viscosity (cp. @ 30°C)	90	90
Ash (%)		0.05
Elemental analysis (wt% on feed)		
Carbon	48.6	54.4
Hydrogen	7.2	5.7
Nitrogen	< 0.1	< 0.1
Oxygen	44.2	39.8

2.6.4. Product use.

The vapour stream from the reactor (prior to being condensed) has been cracked using zeolite catalysts to produce aromatics and gasoline [76]. Further details of the zeolite cracking process can be found in Chapter 3.

2.6.5. Summary.

The NREL vortex system is surface area controlled and hence requires a very large reactor system the scaling up of which is difficult. The NREL technology having been demonstrated at research scale is being further developed by Interchem for polyphenol recovery. The IEA ALPS Group [30] have used the NREL system as the basis for the design and costing of a flash pyrolysis with zeolite cracking plant.

2.7. Summary.

The key parameters for flash pyrolysis with regards to modelling are residence time, temperature, feed water content, feed CHO, feed particle size and the flash pyrolysis technology utilised. Apart from residence time and feed particle size there is sufficient available data for the identified parameters to be included in flash pyrolysis models. Further details of the flash pyrolysis models developed from this project can be found in Chapter 6. Biomass flash pyrolysis oils may be utilised directly in an engine or turbine for power generation (See Chapter 9 for further details) or may be upgraded to give higher quality and more marketable products (See Chapter 3 for further details). The condition of the pyrolysis oil and it's oxygen content are important in the upgrading process.

CHAPTER 3.

UPGRADING TECHNOLOGIES.

3.1. Introduction.

This Chapter describes the technologies, products and identifiable process characteristics which have been considered in the derivation of the upgrading models included in BLUNT. It is not intended as a review of upgrading technologies as this has been documented elsewhere [79, 80, 81] but as an introduction to the technologies, an identification of the important features as regards modelling and as a background to the upgrading models described in Chapters 7 and 8.

As described in Chapter 2, crude pyrolysis oil is relatively unstable and has other chemical properties which limit its use [3]. Upgrading is a process by which a higher quality or more stable fuel may be produced and falls into two categories or severities of upgrading. These are mild or limited upgrading to remove the instability of pyrolysis oils, normally referred to as stabilisation and upgrading to produce transportation fuels, involving severe upgrading followed by refining. Upgrading involves several reactions which result in changes in the properties of the crude pyrolysis oil. These changes [81] include a reduction in the high oxygen content [3]; a decrease in the average molecular weight and viscosity; an increase in the thermal stability; and an increase in the miscibility with fossil fuels and immiscibility with water. Pyrolysis oil upgrading technologies are based either on orthodox hydrotreating technology or the use of zeolite cracking.

Zeolite cracking uses shape selective zeolite catalysts [82] to convert the oxygenated compounds derived from biomass materials into a product rich in alkylated aromatic compounds boiling in the gasoline range [3]. The cracking of pyrolysis vapours is currently carried out on a mono-functional ZSM-5 catalyst, as used in the MOGD (Methanol to Olefins, Gasoline and Diesel) process [83] and conceptually follows equation {1} [3].

$$C_6H_8O_4 \longrightarrow 4.6CH_{1.2} + 1.4CO_2 + 1.2H_2O \qquad \{1\}$$

Equation {1} is a hydrogen limited reaction with the aromatic product being limited by the availability of hydrogen for aromatics and water formation and the oxygen present in the crude oil is rejected as both carbon dioxide and water. Aromatic yields of 15% in a hydrogen limited environment have been achieved with a projected increase to

23% if olefinic gas byproducts are alkylated and recovered [84]. The upgrading of pyrolysis vapours by zeolite cracking is currently at the research stage [79] and is described in Section 3.2.

Hydrotreating involves reacting crude condensed pyrolysis oil with hydrogen to remove oxygen and make the oil less unstable. The process for hydrotreating pyrolysis oils is conceptually based on established refinery processes for desulphurisation [85] and includes hydrogenation, deoxygenation and refining mechanisms. The products from pyrolysis oil full hydrotreatment are hydrocarbons in the gasoline and diesel range [3]. The hydrotreating process can be characterised by equation {2} [3] which is a carbon limited system.

$$C_6H_8O_4 + 6H_2 \longrightarrow 6CH_2 + 4H_2O \qquad \{2\}$$

The upgrading of crude pyrolysis oils by hydrotreatment is carried out over conventional sulphided cobalt/molybdenum or nickel/molybdenum catalysts. Product yields of 35wt% on wet liquids have been achieved for a deoxygenation extent of 98% in a continuously operated carbon limited system [79]. Pyrolysis oil hydrotreating is currently at the research stage [79] and is described in Section 3.3. Section 3.4. gives a summary of the salient features from this Chapter as related to modelling of the upgrading processes.

3.2. Zeolite cracking.

3.2.1. Introduction.

Upgrading biomass pyrolysis derivatives by zeolite cracking involves the direct contact of the primary pyrolysis vapours as they are produced, before repolymerisation or other reactions occur, with the zeolite catalyst [3]. The oxygenated compounds contained in the pyrolysis vapours (which include cyclopentanone, cyclopentenone, furfural, phenol, carbohydrate and carboxylic acid derivatives [86]) are cracked releasing the oxygen as carbon oxides. Gaseous olefins are a by-product of the upgrading of pyrolysis vapours and may be oligomerised and aromatised over zeolite catalysts, releasing hydrogen for hydrocarbon forming reactions. The catalytic cracking of pyrolysis vapours eliminates the condensation step and the potential for water phase separation. However, an aqueous condensate stream is produced by the process, which is normally dealt with by incineration [87].

The use of zeolite catalysts have been demonstrated in the Mobil Olefin to Gasoline and Distillate (MOGD) process [83, 88, 89] which was developed to catalytically convert light olefinic compounds to high quality gasoline and distillate. Using the Mobil ZSM-5 catalyst, light olefins are shape selectively oligomerised to higher molecular weight, diesel range *iso*-olefins. The MOGD process was fully developed following large scale test runs in a Mobil refinery in 1981 [89]. Section 3.2.2. summarises the research which been carried out on the zeolite cracking of biomass derived products.

3.2.2. Research carried out on zeolite cracking biomass derived products.

Zeolite cracking has been successfully carried out on liquefaction oils [90, 91, 92, 93, 94] and tall oil [95, 96, 97] demonstrating that it may be used on biomass derived products. Most of the zeolite cracking work has been carried out on primary vapours due to the perceived problems caused by heating condensed pyrolysis oil although there has been some work on condensed [98] and revapourised primary oils [99, 100]. Investigations on related chemicals and model compounds derived from biomass has been carried out [86] and later extended to pyrolysis oils although there were poor hydrocarbon yields and rapid deactivation of the catalyst. Other work includes co-feeding pyrolysis oil with methanol [101] to improve the hydrocarbon yields.

The most extensively reported work on zeolite cracking of primary pyrolysis vapours has been carried out by the National Renewable Energy Laboratory or NREL (USA) [77, 102, 103, 104] using Mobil's ZSM-5 catalyst [83]. It was discovered that for wood the potential products over zeolite catalysts were organics, coke, water and carbon oxides [102]. The ratio of organic (BTXs and alkenes) to inorganic (carbon dioxide, carbon monoxide and water) products was found to increase with an increase in temperature, indicating an increased yield of hydrocarbons products [84].

Deactivation of the zeolite catalyst due to coking was of major concern in the NREL work [84]. Biomass derived pyrolysis vapours have a low hydrogen content which increases the formation of coke and decreases the final yields of hydrocarbon products. The studies at NREL demonstrated very high initial coking rates at the beginning of experimental runs [84]. However, despite the initial coking sufficient catalyst activity remained to convert the pyrolysis vapours before complete deactivation occurred. The presence of steam appeared to decrease the coking rate

[101, 84]. Trials were also conducted investigating coke removal by introducing oxygen into the reactor and burning the coke off the catalyst. The results from the trials indicated that this was a possible way of limiting catalyst deactivation.

A design for a wood-to-gasoline process [77], using an NREL vortex pyrolyser (as described in Section 2.6.) was developed as a result of the trials and is described in Section 3.2.3. An assessment of the process for gasoline production has been completed by the IEA Bioenergy Agreement [30] and this has been utilised to develop the zeolite cracking models described in Chapter 8.

3.2.3. The NREL wood-to-gasoline process.

The NREL wood-to-gasoline process [77] is illustrated in Figure 3.1. The pyrolysis system consists of a vortex reactor (as used the NREL flash pyrolysis process in Section 2.6.) located in the hot region of a furnace. A steam superheater is located above the pyrolysis reactor and above the steam superheater an incinerator for disposal of the phenolic containing waste water is situated. The boiler utilised to generate carrier steam at 1031kPa is located above the incinerator. The product vapours from the pyrolyser pass through a cyclone for removal of the char before passing into the zeolite cracker.

The zeolite cracker utilises a riser to contact the fresh zeolite catalyst with the pyrolysis vapours at 400-550°C. The coked catalyst, after being separated from the process stream and stripped of residual hydrocarbons, is oxidatively regenerated in a fluidised bed at 550°C in a dilute oxygen environment. The regenerated catalyst is cooled and used as a molecular sieve to strip the C_2^+ hydrocarbons from the by-product fuel gas before being recycled to the cracker. The removal of C_2^+ hydrocarbons from the by-product stream before combustion increases the potential gasoline yield by 26 wt%. The main product stream is then cooled and condensed in air-cooled heat exchangers and the hot air produced in this step is utilised to preheat and dry the wet biomass feedstock. The condensed steam and water soluble organics are recycled to the boiler to produce the carrier steam and to concentrate the organics prior to incineration. The thermal energy required for the process is provided by combustion of the char and gas by-products. The product from the process has a high octane number from the presence of methylated benzenes. It is suitable for distillation and blending with lower octane gasolines in a refinery.

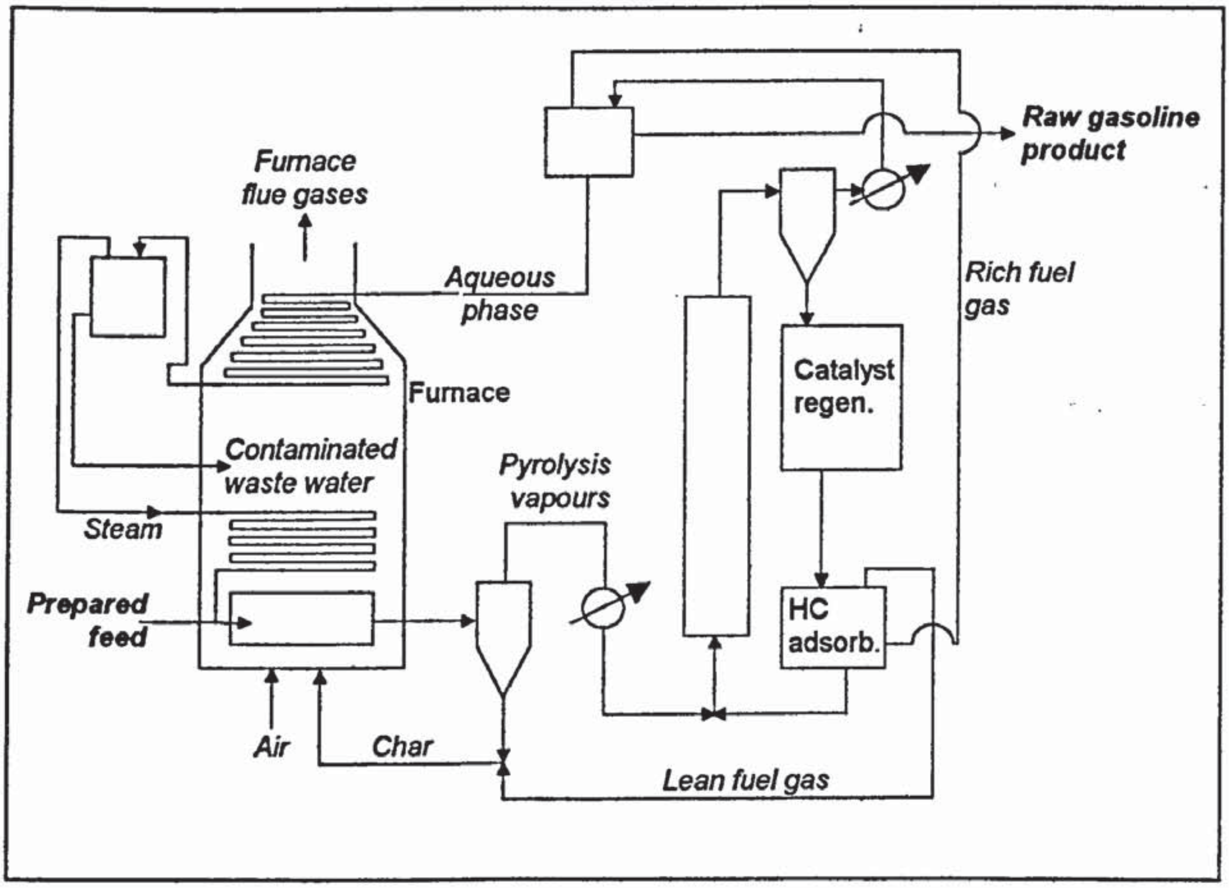


Figure 3.1. Flowsheet for the NREL wood to gasoline process.

3.2.4. Summary.

Zeolite cracking is still in the research stage and information on specific work is scarce. The process gives a highly aromatic product which could be assimilated into an existing refinery structure to produce transportation fuels. The direct contact of the primary vapours with the zeolite catalyst appears to be the best processing route as it avoids potential problems related to reheating or revaporising the condensed oil. The nature of the pyrolysis vapour feed has caused problems with catalyst deactivation and coking although these have to some extent been overcome with the addition of steam or oxygen into the catalytic reactor.

3.3. Hydrotreating.

3.3.1. Introduction.

Upgrading by hydrotreatment involves deoxygenation of the condensed pyrolysis oil using hydrogen in the presence of a catalyst. The conventional cobalt/molybdenum or nickel/molybdenum catalysts used in the petrochemical industry and a variety of related catalysts have been successfully utilised with pyrolysis oils. There has been concern that the use of sulphided catalysts in hydrotreating will increase the sulphur content of the upgraded bio-oil thus reducing one of the potential benefits of biomass derived fuels. However, the sulphided catalysts used in hydrotreating are utilised in the petrochemical industry for desulphurisation and will preferentially allow desulphurisation over deoxygenation [105] thus not adding any additional sulphur to the upgraded product. The longest known continuous run of a pyrolysis oil hydrotreatment process is 8 days [106] although there was substantial deterioration in catalyst activity. Post processing examination of the catalyst suggested that the water in the bio-oil may have attacked the support and reduced the surface area through agglomeration [79]. Other hypotheses include alkali attack, coking and loss of sulphide from the catalyst [79].

When flash pyrolysis oils are heated they tend to polymerise at high temperatures and form a solid material or breakdown to form coke and plug the reactor [107]. To improve stability, primary hydrotreatment or stabilisation of these oils is carried out at a lower temperature in the presence of hydrogen and the catalyst before polymerisation of the oxygen containing compounds can occur [87]. A pyrolysis oil hydrotreating process therefore conventionally consists of two stages: a preliminary stabilisation of the crude oil at 250 - 275°C followed by the deoxygenation or upgrading stage operating at 350 - 400°C [79]. Both stages of the process take place at pressures of 70 -100 bar which is required to maintain the high hydrogen partial pressure.

The reduction in oxygen content and volatility make the hydrotreated oils less water soluble than crude bio-oils resulting in an aqueous phase contaminated with organics at high deoxygenation extents [108]. The more severe the hydrodeoxygenation the less contaminated the waste water becomes as the bio-oils become more hydrophobic. Carbon losses to the waste water are as low as 0.1 - 2% when the hydrotreated product contains less than 1% oxygen [108]. Any alkali metals present in the crude pyrolysis oil are also stripped out of the upgraded oil by the waste water stream [107].

The main drawback of hydrotreating is the significant amount of hydrogen required which equates to 700 litres of hydrogen per kilogram of pyrolysis oil [109] for full hydrotreatment. An excess of typically 100 - 150% is required for processing due to hydrogenation of components such as light aromatics. To put this in perspective, a 1000 tonnes/day biomass processing plant will require about 50 tonnes/day of hydrogen for complete hydrotreating. The hydrogen may be generated in a number of ways including recovery and/or regeneration of the process off-gases. The options for hydrogen generation in a hydrotreating plant are shown in Figure 3.2. and include purchase, gasification of biomass or char, steam reforming of the process off-gases or using refinery surpluses [38].

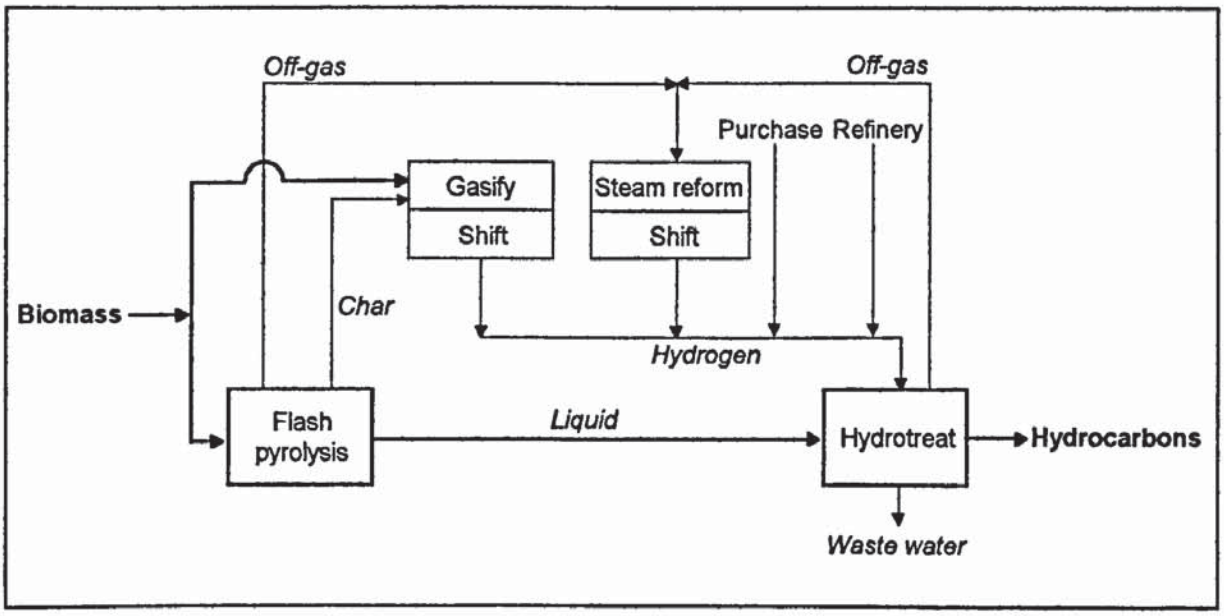


Figure 3.2. Hydrogen production and catalytic processes in hydrotreating [38].

Table 3.1. summarises various hydrogen generation options which could be utilised, the status of the technology and the relative costs [38]. The hydrogen sources could be utilised singly or in combination to give the most efficient and cost effective process. The most attractive routes are to either utilise the surplus hydrogen available at fuel value at refineries or to reform the product gas "in-situ".

One benefit of hydrotreating is that the processing site may be decoupled from the pyrolysis plant allowing for example, one central hydrotreating plant to be fed by several small localised pyrolysis plants or satellite hydrotreating plants to be close-coupled to satellite pyrolysis processes or a hydrotreating plant to be inside a non-dedicated refinery fed from multiple satellite pyrolysers [38]. The optimum route depends on several factors including the productivity of biomass feed, the cost of each operation, the yield of each operation and the value of the product and has not been evaluated due to a dearth of information and the site-specific nature of the variables.

Table 3.1.
Hydrogen generation options [38].
(Basis: relative costs are on a scale of 1 to 10 where 1 is the lowest and 10 the highest cost)

Alternative	Status	Relative cost
Reform/shift product gas 'in-situ' with modified catalyst	Not attempted	2
Reform/shift product gas externally, possibly with membrane separation	Feasible but unproven	6
Gasify biomass, shift CO to H_2 and scrub out the CO_2, possibly with membrane separation	Feasible but unproven	10
Gasify charcoal, then as for biomass	Feasible but unproven	8
Use existing surplus H_2 in a conventional refinery	Commercially available	2
Convert fossil fuel by steam reforming or partial oxidation with shift and CO_2 removal	Commercially available	6
Electrolyse water using solar or conventionally produced electricity	Commercially available	8
Purchase hydrogen in bulk	Commercially available	10
Recovering surplus hydrogen for recycle	Commercially available	4

For some applications, such as gas-turbine fuel, complete deoxygenation may not be required to give a fuel with suitable chemical and physical properties. Partial hydrotreatment would involve partial deoxygenation of the crude bio-oil to give a more stable liquid which would be more readily substituted for conventional hydrocarbon fuels. However, there is uncertainty in the selectivity of the hydrotreating and hydrogenation of the compounds in the crude bio-oil and the properties of the partially upgraded oil. A 15% oxygen content partially upgraded product would have a similar oxygen content as a pressure liquefaction product and potentially similar properties, for example high viscosity and limited water miscibility [10, 110]. These properties would make it difficult to use in most applications although the potential cost savings on the reduced hydrogen consumption and capital costs for a partial hydrotreating process with respect to complete hydrotreating could still make the option attractive.

3.3.2. Research carried out on hydrotreating biomass derived liquids.

Hydrotreating has been carried out on biomass liquefaction oils [111 , 112], solvolysis oils [113] and black liquor from the pulping industry [114 , 115] using conventional sulphided cobalt-molybdenum catalysts. Hydrocarbons and phenolics with low oxygen and sulphur contents were obtained and the processes demonstrated that hydrotreating can be successfully carried out on biomass derived products. The upgrading of lignin extracted from flash pyrolysis oil has been investigated [116] in a non-isothermal reactor using a sulphided commercial Cobalt-Molybdenum catalyst to produce a hydrocarbon liquid consisting of 62% aliphatics and 38% aromatics with about 50% boiling in the gasoline range [116].

A trickle bed reactor has been used for hydrotreating for temperatures between 350 - 400°C, pressures of 5 - 10MPa and over a Pt/Al_2O_3 catalyst [117]. It was found that coking decreased at higher pressures and the concentration of all the other fractions increased due to an increase in the hydrogen concentration in the liquid. The non-volatile phase of the feedstock cracked to produce phenolic compounds which led to the formation of hydrocarbons [117].

The characterisation and upgrading of pyrolysis oils over cobalt-molybdenum (CoMo) and nickel-molybdenum (NiMo) catalysts have been studied using a batch reactor at 12MPa and temperatures between 270 - 400°C [118 , 119] using a hydrogen donor solvent. The elemental composition of the products oils obtained from hydrotreatment using the two different catalysts were similar. However, the nitrogen and oxygen contents of the products using the NiMo were slightly lower, with increased hydrogen consumption and increased formation of gaseous products indicating that the NiMo catalyst causes higher hydrocracking activity [119]. There were indications from the trails that some polymerisation was taking place during the heating up period before the catalysts became active, reducing the quality of the product and causing coking and deactivation of the catalyst. The process was modified to allow preheating of the feed oil at 300°C to stabilise the reactive fractions prior to hydrotreatment at 400°C [120]. This two-stage hydrotreating process gave a 70% reduction in the oxygen content of the feed oil when processed alone and a 85% reduction when co-processed with a hydrogen donor solvent (tetralin) [121].

The use of oxygenated model compounds and conventional pyrolysis oils to investigate hydrotreating mechanisms has been carried out [122 , 123 , 124 , 125]. The work investigated the requirements for stabilisation of the crude oil at low temperatures and for full hydrotreatment to produce a high grade fuel, the influence of the main parameters i.e. temperature, nitrogen in the form of ammonia, water, hydrogen sulphide (for sulphided catalysts) and catalyst type. The results were encouraging indicating that the crude bio-oil could be stabilised at low temperatures by elimination of the unsaturated double bonds present in some of the reactive compounds (e.g. olefins, aldehydes, ketones and carboxylic groups) [125]. The presence of water was found to have a limited effect and the presence of nitrogen compounds (present in small quantities in bio-oil) to have a significant effect on catalyst activity [122]. An increase in the partial pressure of hydrogen sulphide which may be required to keep the catalyst sulphided was shown to dramatically decrease catalyst activity [122].

Veba Oel [106, 126] have conducted preliminary trials on pyrolysis oil from the Ensyn RTP process (See Section 2.4.) and achieved product yields between 30 - 35% for temperatures of 350 - 370°C, liquid space velocities of 0.25 - 0.8 kg/kg hour and deoxygenation extents between 88 - 99.9% [106]. At 99.9% deoxygenation the product oil was "water-white" and consisted of 40% naphtha, 40% middle distillate and 20% vacuum gas oil. There were considerable problems with the Veba trials including complete catalyst deactivation over an 8 day period, gum-like deposits in the feeding tubes and coking in the reactor. It is unclear whether Veba utilised a single or two-stage hydrotreating process but it is likely that it was a single stage process. Some of the problems e.g. excessive coking encountered by Veba may have been due to the quality of the feed bio-oil which had a high solids content [106]. Despite the problems the Veba work generated some interesting data on the organic content of waste water streams, product oil density and water content of the upgraded oil which was considered when the hydrotreating model was written (See Chapter 7 for details).

Extensive work on pyrolysis oil hydrotreatment has been carried out by Battelle PNL (USA) [109, 111, 127 , 128 , 129] and this is the work considered in the modelling of hydrotreatment in Chapter 7 in addition to some of the information generated from the Veba trials. A fixed bed continuous feed catalytic reactor and commercial hydrotreating catalysts were utilised to investigate the catalytic hydrotreating of biomass derived liquid products [127, 128]. Battelle discovered that if an initial step below 280°C in the presence of the catalyst was carried out the formation of coke in the reactor was minimised [127]. This discovery lead to the development of a two-stage hydrotreating process the first stage at 250-280°C, 2010 psig and liquid space velocities of 0.62 (vol oil/vol cat hr) and the second at 370-400°C, 2030psig and a liquid space velocity of 0.11. Conventional sulphided cobalt-molybdenum catalysts were utilised for both stages and it was noticed that the hydrogen consumption of the first or stabilisation stage was about 10% of the second, higher temperature stage [109]. Although sulphided catalysts were used in the hydrotreating trials there was little evidence of the incorporation of sulphur into the product oils [129].

A hydrotreating pilot plant for upgrading biomass pyrolysis oils was commissioned under a European Commission Research Contract to be designed by Battelle [130] based on their hydrotreating work. This plant consists of a short residence time, continuous two-stage stabilisation and deoxygenation hydrotreating process, with a hydrogen surplus and incorporating a membrane separation and hydrogen recycling system to minimise the hydrogen requirements. The Battelle pilot plant design is described in Section 3.3.3.

3.3.3. The Battelle hydrotreating plant.

The Battelle pyrolysis oil hydrotreating plant incorporates two hydrotreatment reactors in series. The first reactor is used to provide a low temperature, 250 - 300°C, catalytic treatment of the pyrolysis oil and is significantly smaller than the second reactor. The oil leaving the first stage reactor is more stable and can be suitably upgraded at a higher temperature, 350°C, in the second stage reactor. Operating pressures for the system are in the 170-190 bar range. The process is designed to produce a 98% deoxygenated product form the crude pyrolysis oil entering the plant. The flowsheet for the Battelle hydrotreating pilot plant is shown in Figure 3.3. The upgraded product from the second stage reactor is cooled to 70 - 80°C to condense most of the water. A hydrogen recycle is utilised to minimise the amount of hydrogen required for operation. The hydrogen is separated from the product oil and water in a high-pressure separation vessel at about 134 bar.

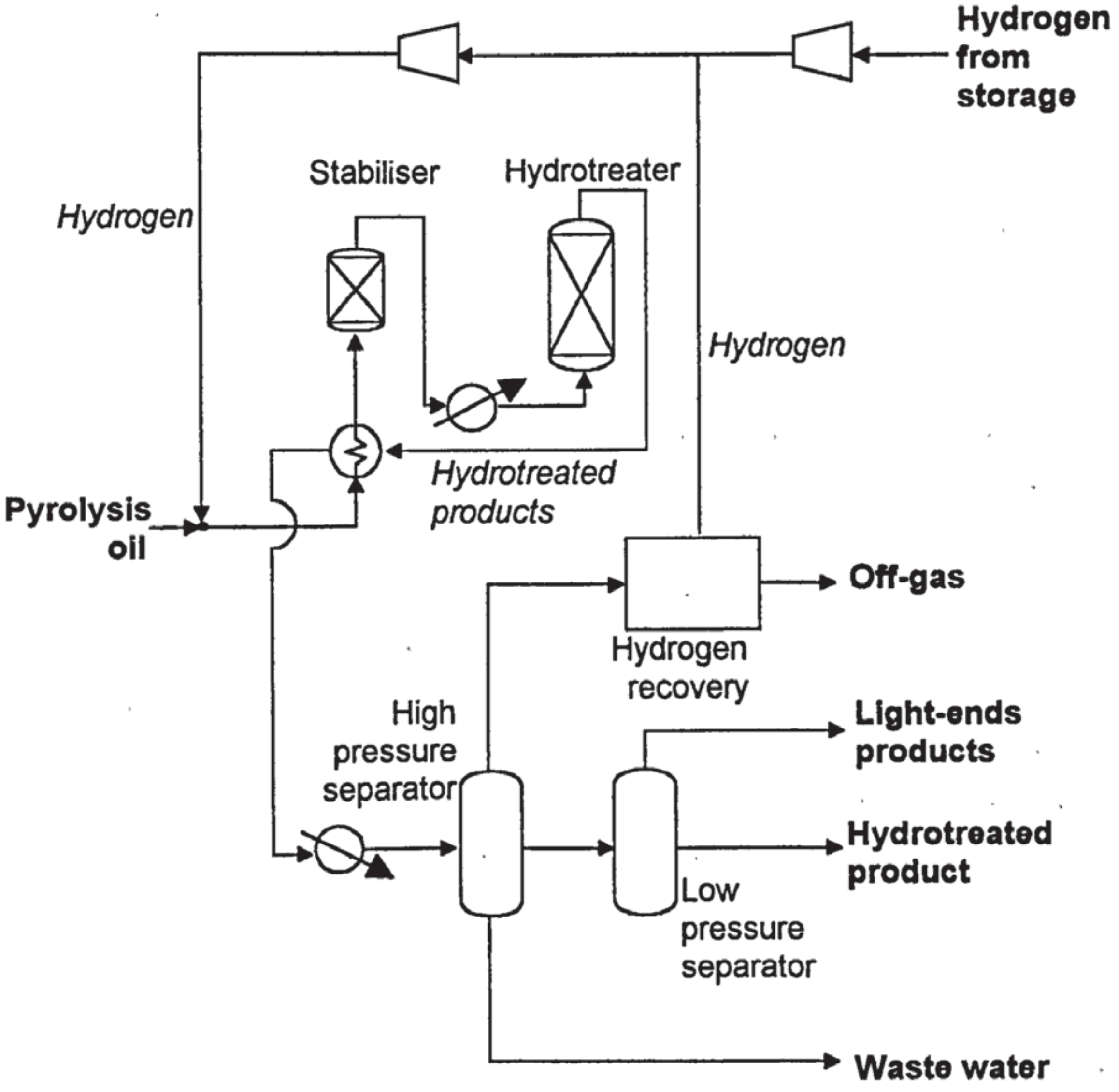


Figure 3.3. Batelle hydrotreating pilot plant design

Carbon dioxide, water and light organics are removed from the hydrogen rich vapour in a membrane separation unit. The purified hydrogen is combined with make-up hydrogen at 20°C and 1 bar and compressed to the reactor pressure. The remaining off-gas can be used as fuel gas or flared.

3.3.4. Summary.

Upgrading biomass pyrolysis oils by hydrotreatment has been carried out at research level and some trials have been carried out in a conventional refinery. The use of a two-stage process has been identified by several groups as the best option for limiting polymerisation, coking and catalyst deactivation during hydrotreatment. The use of model oxygenated compounds to investigate hydrotreating has lead to some understanding of the reactions mechanisms and process parameters including water, nitrogen and sulphur. Results indicate that the choice of catalyst can improve product yields with sulphided NiMo being favoured by some groups and CoMo by others. The potentially high cost of hydrogen for the hydrotreating process could be reduced by using refinery hydrogen and recovering the surplus hydrogen from the hydrotreating process.

3.4. Summary.

Upgrading is utilised to increase the stability or enhance the chemical or physical characteristics of a bio-oil. Partial upgrading performed using mild hydrotreatment on the condensed pyrolysis oil produces a more stable and hence more marketable product. Alternatively, full upgrading using severe hydrotreatment on the condensed pyrolysis liquid or zeolite cracking on the pyrolysis vapours produces products which are miscible with conventional or fossil derived organic chemicals and may be further refined to give transportation fuels. Although research has been carried out on both zeolite cracking and hydrotreating there is insufficient published information for analyses of the critical process parameters to be carried out as for flash pyrolysis in Chapter 2.

At present, zeolite cracking of pyrolysis oils has only been performed on laboratory scale equipment and there is insufficient data available to determine the critical process parameters. The zeolite cracking models which have been developed based on the NREL work and included in the program are described in Section 8.

Hydrotreating has been performed on biomass pyrolysis oils with some success by several groups including Battelle PNL and Veba Oil. The pyrolysis oil oxygen content and the deoxygenation extent or severity of the hydrotreating process are clearly process parameters for a hydrotreating process. The hydrotreating model which has been developed from the Battelle design and which is included in the program is described in Chapter 7.

CHAPTER 4.

DEVELOPMENT OF THE COMPUTER PROGRAM.

4.1. Introduction.

This Chapter describes the development, structure and operation of the computer package, BLUNT developed to technoeconomically model and assess the selected biomass conversion technologies of flash pyrolysis (described in Chapter 2 and modelled in Chapter 6) and upgrading by hydrotreatment or zeolite cracking (described in Chapter 3 and modelled in Chapters 7 and 8 respectively). BLUNT has been designed for use by researchers in the biomass field and those in other fields who wish to understand biomass thermochemical conversion technologies. The program has been designed to give Users a grasp of the main process variables, products and costs and to allow the technologies considered to be assessed on a consistent basis. BLUNT has been designed to be as User-friendly as possible and incorporates 'Help' screens with details on the technologies and hints on program operation. The program can be run on any IBM compatible computer with Microsoft Windows Version 3.0 or later and when installed will run like any other Windows application, allowing the User to use either the keyboard or a mouse for entering data and switching between screens. For details on the operation of BLUNT see the BLUNT Operating Manual in Appendix A and Section 4.7. A copy of BLUNT is an integral part of this Chapter and has been included at the end of this thesis.

The biomass conversion models designed for BLUNT and detailed elsewhere in this thesis incorporate factors known to influence product quality and quantity and hence process performance such as pyrolysis reaction temperature (See Chapter 6 for further details). The result is a flexible User controlled program which incorporates the latest information and technology for flash pyrolysis and upgrading processes. The development of a dedicated computer program was necessary because commercially available simulation packages, for example Process, ChemCad or ASPEN, do not either consider the required technologies or perform economic analyses. The only similar package available for biomass conversion processes is AMBLE [28, 131] which was written at Aston University under an European Commission Research Contract and concentrates on biomass gasification processes with only simple models for biomass flash pyrolysis and upgrading. AMBLE is reviewed in Section 4.2.

The development stages which resulted in the program are described in Section 4.3. Section 4.4. describes how the component parts of BLUNT interact to form the program and Sections 4.5. and 4.6. describe the two main components of BLUNT, the Dynamic Link Library (Section 4.5.) and the Graphical User Interface (Section 4.6.). Section 4.8. looks briefly at the operation of BLUNT and Section 4.9. summarises the main points arising from this Chapter.

4.2. The AMBLE simulation program

4.2.1. What AMBLE does.

AMBLE (the Aston Model of Biomass to Liquids for Energy) [28, 131] models the thermochemical and biochemical conversion of selected feedstocks to produce liquid fuels. The program aims to provide a technical and economic simulation of the processes to allow them to be assessed consistently. The processes and feedstocks modelled by AMBLE are the thermochemical conversion of wood, straw and refuse by gasification, liquefaction and flash pyrolysis and the biochemical conversion of wood, wheat and sugar beet by acid or enzyme hydrolysis followed by fermentation. The liquid products modelled in the program include paraffinic and gasoline blending stocks, fuel oil, methanol, fuel alcohol and ethanol. AMBLE operates for feedstock inputs between 200 and 2000 dry ash free (d.a.f.) tonnes/day.

4.2.2. How AMBLE works.

AMBLE works by breaking down a selected conversion process into a series of process steps representing a particular pretreatment or conversion process through which the material being processed passes sequentially [131]. Each process step is a self-contained unit including all the operations and processes necessary for it to function. The results from the process steps making up a conversion route are combined to give the performance of the whole conversion process [28].

AMBLE operates on an IBM compatible computer under the MSDOS environment using a write-to-screen format i.e. the program User is prompted by AMBLE to select the run data from lists on the screen by typing in the appropriate data. The required information for a simulation run is the feedstock, the product, the feed input in d.a.f. tonnes/day and the conversion process utilised. When all the required data has been entered the program runs the selected process step models, performs the calculations to derive the product yields, production costs and capital costs and displays a summary of the results on the screen. Listings of the results, including a mass balance summary, capital cost summary, utilities usage summary and a breakdown of the annual costs can be printed at the end of a simulation run. To run another simulation the program User must quit and then reinitialise the program.

4.2.3. The limitations of AMBLE.

AMBLE has several limitations that affect the way in which it is used and are described in this Section. Since the AMBLE code was completed in 1990 there have been considerable changes in the use of computers, software and the expectations of a program User. Data entry and operation of AMBLE follows a straight line principle i.e. the program User starts at the beginning of the program, enters all the required data, the program executes and the results are displayed. However, if a User decides on another simulation run with slightly altered data all of the previous steps must be carried out again. This method of data entry and program operation is fine if the program User only requires a single set of data but if several sets are required in which just a single variable is altered, for example feed input to the plant, the program must be operated from scratch repeatedly.

The combinations of feed, product and conversion route allowed by AMBLE are strictly controlled with not all of the feeds being able to produce a particular product [31] due to lack of information or restrictions on the process. However, if a feed/product combination is selected which AMBLE does not cater for, the program crashes and the program User is returned to MSDOS. It is not acceptable when running a program for it to crash whenever a mistake is made and this could have been allowed for when the program was written. The problem could have been overcome by either writing a sub-routine allowing the User to reselect a product or feedstock or by informing the User of the possible products/conversion routes which may be utilised with a particular feedstock when this feedstock is selected.

Although AMBLE simulates biomass flash pyrolysis and upgrading processes the program was originally set up to model gasification process which it does in some depth. The flash pyrolysis, direct liquefaction and biochemical conversion models were included in AMBLE to allow a comparison of gasification with other technologies. The models for flash pyrolysis and upgrading included in the program consist of very simple algorithms representing the mass balances and the utilities usage and take no account of process variables.

AMBLE is constructed from step models representing conversion processes which were originally built as "large as possible to minimise complexity and maximise ease of use and flexibility" (quote from AMBLE's constructor) [131]. However, large blocks of code or procedures consisting of several hundred lines of source code are difficult to check and create debugging problems. If there are bugs in the source code, removing them would require a major overhaul and possibly a total rewriting and restructuring of the AMBLE code from scratch. For any program to be operated an executable file must be generated and this is done by compiling the source code of the program to link all of the procedures and sub-routines into a single file. However, programs with large blocks of code are difficult to compile due to the requirements of a co-processor and large amounts of memory. AMBLE was originally structured with discrete step models representing conversion processes so that new models could be slotted in as required. In practise this is not feasible as the large blocks of code in AMBLE make the addition of new models and re-compilation of the program very difficult.

4.3. The development of BLUNT.

4.3.1. Background.

The work described in this thesis was originally designated as additional step models and conversion processes which would be incorporated into AMBLE adding to the existing scope of the program. However, due to the limitations of AMBLE including the existing number of feedstocks, products and conversion processes modelled and the size of the resulting source code this was not considered feasible and it was decided to build a new stand-alone computer program.

There are a lot of uncertainties and gaps in the knowledge of biomass conversion processes. This sparcity of data and the unconventional nature of biomass conversion processes meant that a specialised chemical engineering design package such as ASPEN was not the best tool for the modelling. The basic concept of AMBLE i.e. the way in which it breaks down a conversion route into process steps was considered an attractive way in which to construct a specialised program with the data available. In addition, features could be built in which addressed the problems encountered with AMBLE and incorporated the latest in design and computational software.

AMBLE was designed to run in a MSDOS environment because at the time of it's construction most computer packages were still MSDOS based. However, since then computers have become more widely utilised and most present day computer users do not use DOS itself directly. Instead they operate their packages in an interactive environment such as Microsoft Windows or the desktop on an Apple Macintosh. where they have a visual display of what the computer offers. Programs written for such an environment tend to have interactive tools i.e. features such as menus which can be opened by clicking with a mouse and buttons which can be pressed. Such an environment is known as a Graphical User Interface (GUI) where the program User can pictorially see and control an environment. The Windows Program Manager itself is a classic example of a GUI. Having operated AMBLE and seen the problems related to data input and display it was decided to incorporate a GUI into the new program.

To maximise the amount of time available for process modelling, it was decided to utilise the part of the AMBLE source code which performed the financial calculations and controlled the calling of the step models. This code was modified to form the basic framework for a program into which the new step models could be placed when developed for debugging and checking.

4.3.2 The choice of programming language.

The code for the program was written in Pascal which is a highly structured programming language. Pascal is an easy language in which to program and performs complex calculations such as those used in mass and energy balances quickly. It was decided to leave the source code in Pascal because the time taken to translate the source code into another language could be more effectively spent on developing process step models. However, since it had been decided to incorporate a graphical interface for data entry and results display the graphical abilities of Pascal had to be assessed.

The version of Pascal utilised for the program development, operates and compiles programs which run in the Windows environment. However, it's graphical abilities are poor and not very well documented [132]. To develop a graphical interface using Pascal each item on the screen has to be written in code, by hand. For example, to have a box on the screen which opens to show a list of feedstocks, code has to be written which draws the box, sizes it and then positions it on the screen. Once this has been done the code which controls the information which goes into the list has to be written and this results in a very large amount of code simply to add a list box to a screen. The finished quality of the graphics on the screen was poor with the text boxes and other items having fuzzy edges and it was decided that Pascal was not the best package for writing the graphical interface.

After consultation with software specialists Visual Basic was recommended and was considered for the development of the GUI. Visual Basic was a new software tool, developed specifically for operation in the Windows environment. The quality of the graphics on the screen was excellent with clear, crisp edges to the text boxes. In addition, it had one advantage over Pascal in that the controls for basic graphics are prewritten making for simple programming. Visual Basic operates differently to most conventional programming software packages and so further details on Visual Basic are included in Section 4.3.3.

The only draw back with Visual Basic was that although calculations could be done the processing was slower than Pascal and complex calculations could not be performed. However, Visual Basic has been designed to allow it to be integrated with other packages such as programs written in C, Excel and Pascal. It was therefore decided to utilise the best features of the two packages and to perform the calculations in Pascal and have the data entry and results in Visual Basic. This led to the development of BLUNT, the structure of which is described in Section 4.4.

4.3.3. Visual Basic.

The Visual Basic programming system [133 , 134] allows the creation of applications that fully exploit the use of Graphical User Interfaces and accelerates the design process by providing the appropriate tools for different aspects of GUI development. Visual Basic applications run under the Microsoft Windows operational environment and allow data exchange to and from other applications. A GUI is created for an application in Visual Basic by drawing objects on the screen and then setting the properties which refine their appearance and behaviour. The GUI is made to react to the User by writing code that responds to events that occur in the interface e.g. code can be written to react to the mouse being clicked on the screen.

The buttons, menus and text boxes on a Graphical Interface are called controls in Visual Basic. When the interface has been drawn in Visual Basic, the controls which have been placed in a blank window will automatically recognise the actions users take such as mouse movements and button clicks. However, code must be written before the controls react to the actions of program users. This is different to conventional programming where code must be written to design the controls and to make controls respond to events.

Programs in conventional programming languages e.g. Pascal, Basic and C++ run from the top down i.e. execution starts from the first line and moves with the flow of the program to different parts as required. A Visual Basic program works completely differently. The core of a Visual Basic program is a set of independent pieces of code which respond only to the events they have been told to recognise. The design of a Visual Basic application follows three simple steps:

1. the window or screen that the User sees is customised to include controls such as text boxes, labels and buttons;
2. the events, e.g. mouse clicks, that the controls on the windows recognise are determined;
3. the event procedures (controlling code) are written to respond to the events.

For example, to set up a button which switches from a main screen in a program to a 'Help' screen the first step is to draw a button on the Visual Basic Screen. Then, the event to which to button will react is determined - in this case the button being clicked by the mouse. The event procedure is written to tell the Visual Basic Screen to open the 'Help' screen when the button is clicked. In the same way a button can be set up on the 'Help' screen which closes the screen when clicked and returns the User to the preceding screen.

When a Visual Basic application is running it cycles through four steps:

1. The application monitors the windows and controls in each window for all the events that each control can recognise for example, mouse movements, mouse clicks and keystrokes.

2. When an event is detected it examines the application to see if an event procedure has been written for that event.
3. If an event procedure has been written the code that makes up the event procedure is executed and the program cycles back to (1).
4. If an event procedure has not been written Visual Basic waits for the next event and goes back to (1).

For example, if a program has been executed which has a screen containing one button designated as *HELP* to switch the User to a 'Help' screen and the 'Help' screen has single button designated *RETURN* which switches the User back to the previous screen. Both buttons are set up to respond to a mouse click and no other controls are available on the screens. The program will remain inactive unless the mouse is clicked on the *HELP* button. If that occurs, the program considers the event procedure for a mouse click on *HELP* and executes the code for the procedure as in the previous example i.e. switches the User to the 'Help' screen. When the code has been executed the program again waits for an action such as a mouse click on *RETURN* and so forth. If the User clicks anywhere else on the screen, no action will be taken by the program because an event procedure has not been written for such an action.

4.4. The structure of BLUNT.

Due to the two different programming languages which supply the computational and graphical aspects of BLUNT the program is composed of three parts :

- a Pascal Dynamic Link Library (DLL) which contains the step models for the pretreatment and conversion steps, the general procedures which link and are utilised by these models and the financial calculations;

- a Visual Basic Graphical User Interface (GUI) which allows the program User to enter data, run the DLL, view the results from a simulation run and consider macroeconomic factors; and

- text data files which contain data such as possible conversion routes and component information required by the program.

The structure of the program and the interaction between the parts of the program is shown in Figure 4.1. The Graphical User Interface allows the User to enter the data required for running the program. This information consists of the feed, product, conversion route, feed input in d.a.f. tonnes/day and the values of the financial and step model variables for which defaults are included in BLUNT (See Section 4.6. for further details). When the program User is satisfied with the information entered, the User presses a button marked *RUN* which sends the information to and initiates the Dynamic Link Library.

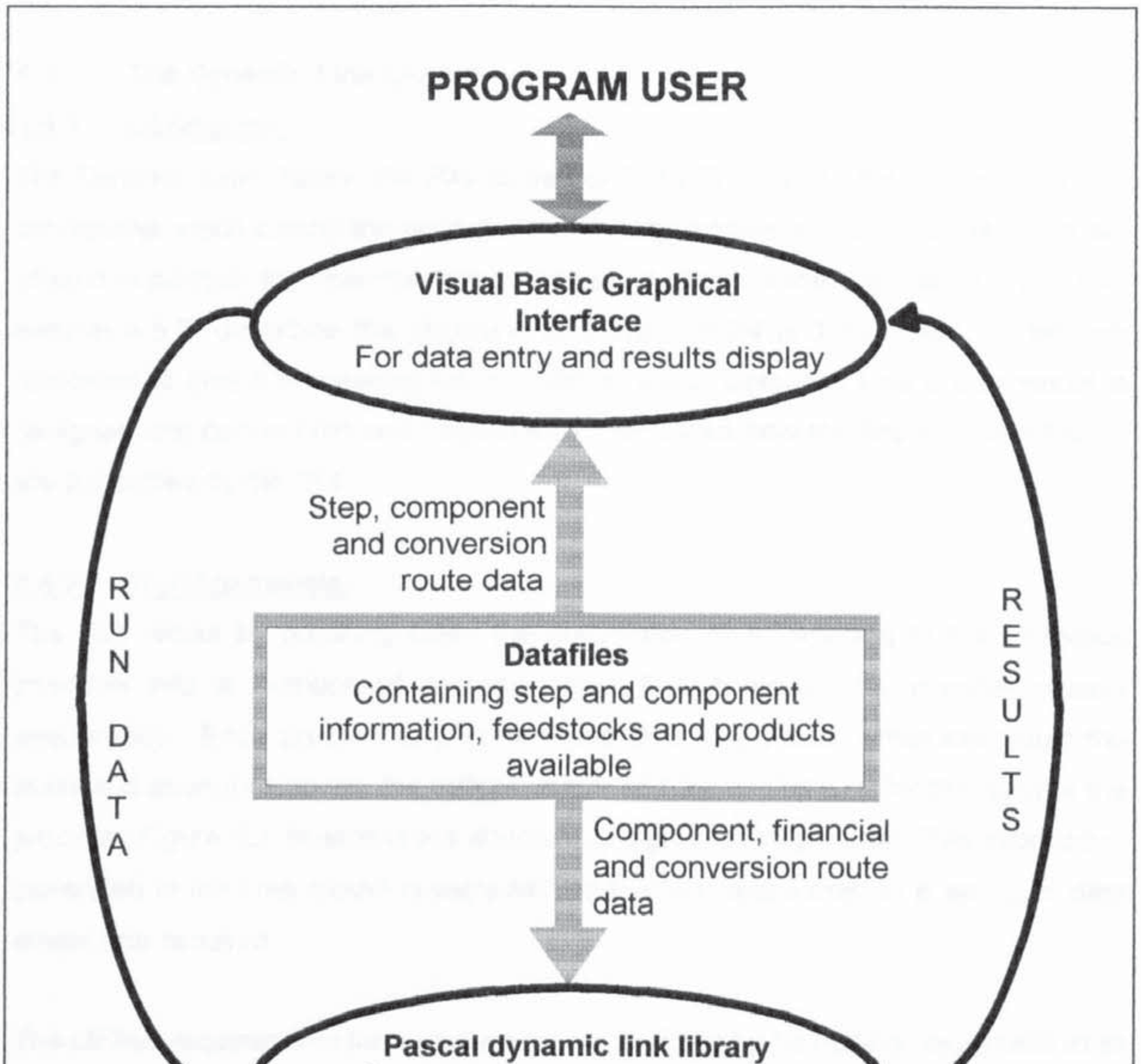


Figure 4.1. The structure of BLUNT and the interactions between the various components.

The Dynamic Link Library runs the pretreatment and conversion step models, performs the financial calculations and sends the results back to the Graphical User Interface for display. The results generated consist of the product cost, capital cost, capital cost breakdown, utilities usage summary, annual cost breakdown and the additional cost summary. Both the Graphical User Interface and the Dynamic Link Library access the text data files for information. The Pascal Dynamic Link Library is described in Section 4.5. and the Visual Basic Graphical Interface is described in Section 4.6. The text data files which are used by both the DLL and GUI are described in Section 4.7.

4.5. The Dynamic Link Library.

4.5.1. Introduction.

The Dynamic Link Library, the Pascal part of BLUNT, contains the step models, the procedures which control the operation of the step models and procedures which are utilised to perform the financial calculations used to determine the cost of a process. Section 4.5.2. describes the structure of a step model and how step models are combined to give a processing route. Section 4.5.3. describes how a step model is designed and constructed and Section 4.5.4. describes how the financial calculations are performed by the DLL.

4.5.2. The step models.

The DLL works by breaking down the conversion route selected in the Graphical Interface into a number of process steps though which the material passes sequentially. Each process step is modelled in a step model which calculates the mass and energy balances, the utilities usage and the capital cost for that step of the process. Figure 4.2. illustrates the structure of a process step model. The information generated in the step model is extracted by the DLL and stored in a series of data arrays until required.

The utilities requirements for a step model are handled by "importing" requirements or "exporting" surpluses into a utilities "pool", controlled by the executive program, where it is debited or credit respectively at the market price. In this way resources from one step can be exported and utilised by another as would be the case in a conversion plant.

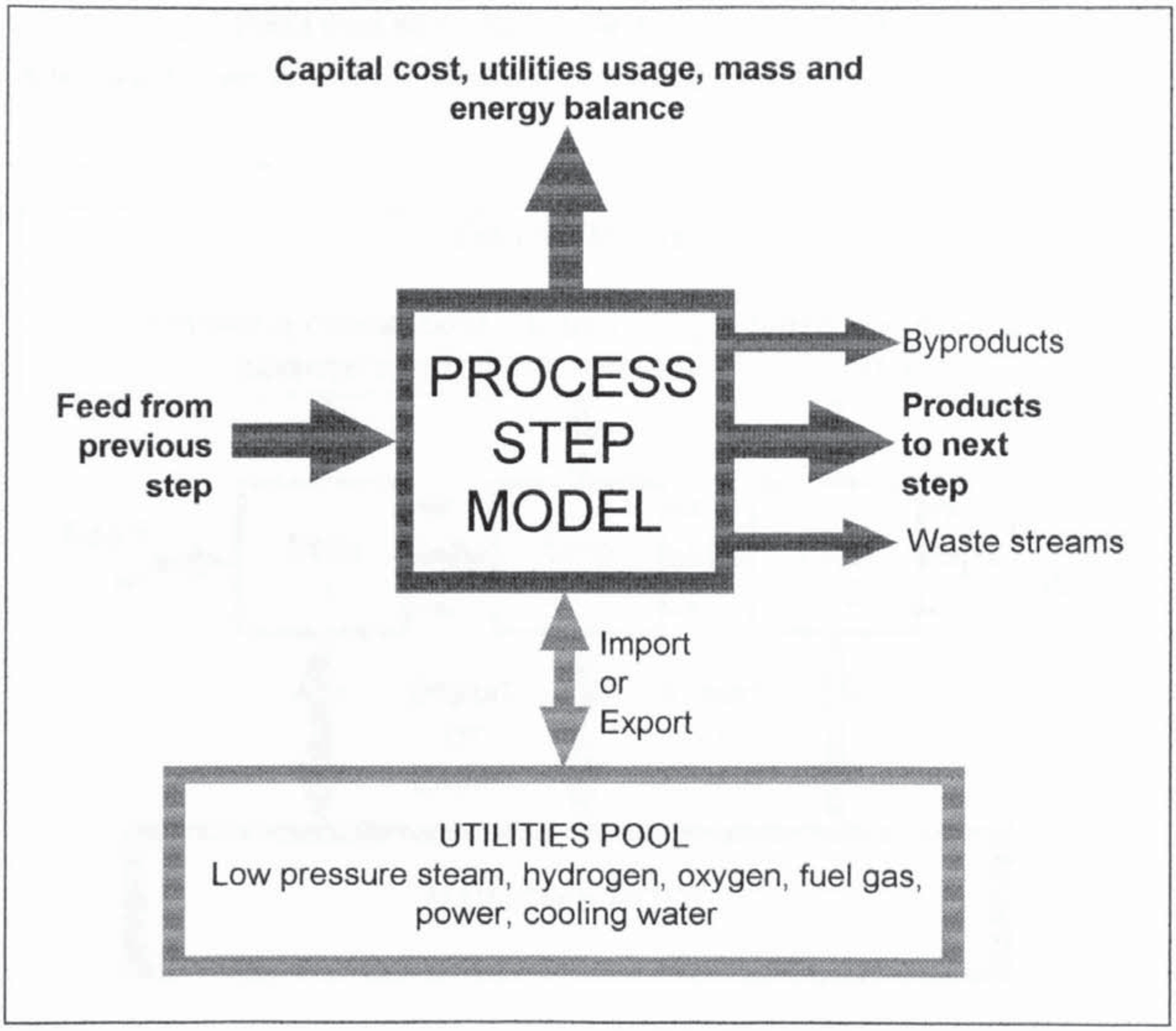


Figure 4.2. The structure of a process step model.

For example, off-gas is produced in the flash pyrolysis step model (See Chapter 2 for details on the products from flash pyrolysis) which can be "sold" to the utilities pool as fuel gas and may be "bought" from the utilities for use as fuel in the wood drying step. There are other methods by which fuel gas or hydrogen could be generated for use in the utilities e.g. gasification or low cost hydrogen could be purchased if the conversion plant was operating in a refinery. However, these options are very site specific and would complicate a system which although simple works very well and allows alternative utility sources to be modelled by altering the costs of the utilities.

For the step models to be combined to form a processing route there must be continuity of information. The information which passes from step to step is the quantity and composition of the material being processed with the 'products' from one step becoming the 'feed' for the succeeding step as shown in Figure 4.3. The information generated from each step model and stored in the data arrays is used to determine the performance of the whole conversion route.

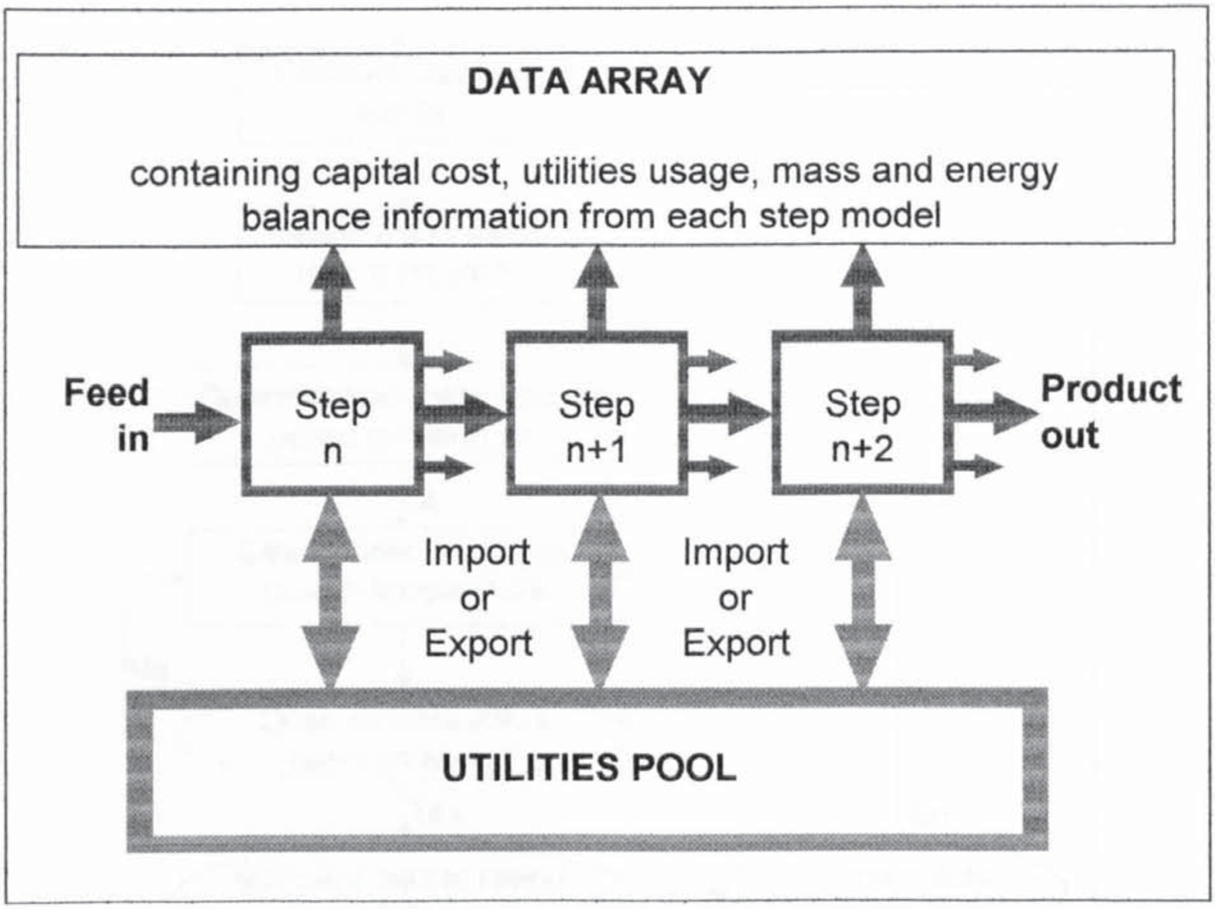


Figure 4.3. The combination of process step models to form a conversion route.

4.5.3. The design of a step model.

The design and development of the step models representing the biomass conversion processes is integral to the development of the computer program. The stages in the development of a step model representing a biomass conversion or upgrading process are illustrated in Figure 4.4. Parts (a) and (b) The design stages are as follows:

1. Collect information on the process to be modelled including a process description, the processing conditions, capital cost and significant process variables. Information can be gathered from conducting literature surveys and contacting the designers or developers of the process.
2. Develop a flowsheet for the process detailing all conversion, separation or heat exchange processes; the process, by-product and waste streams; and any additional process inputs required such as steam or process water.

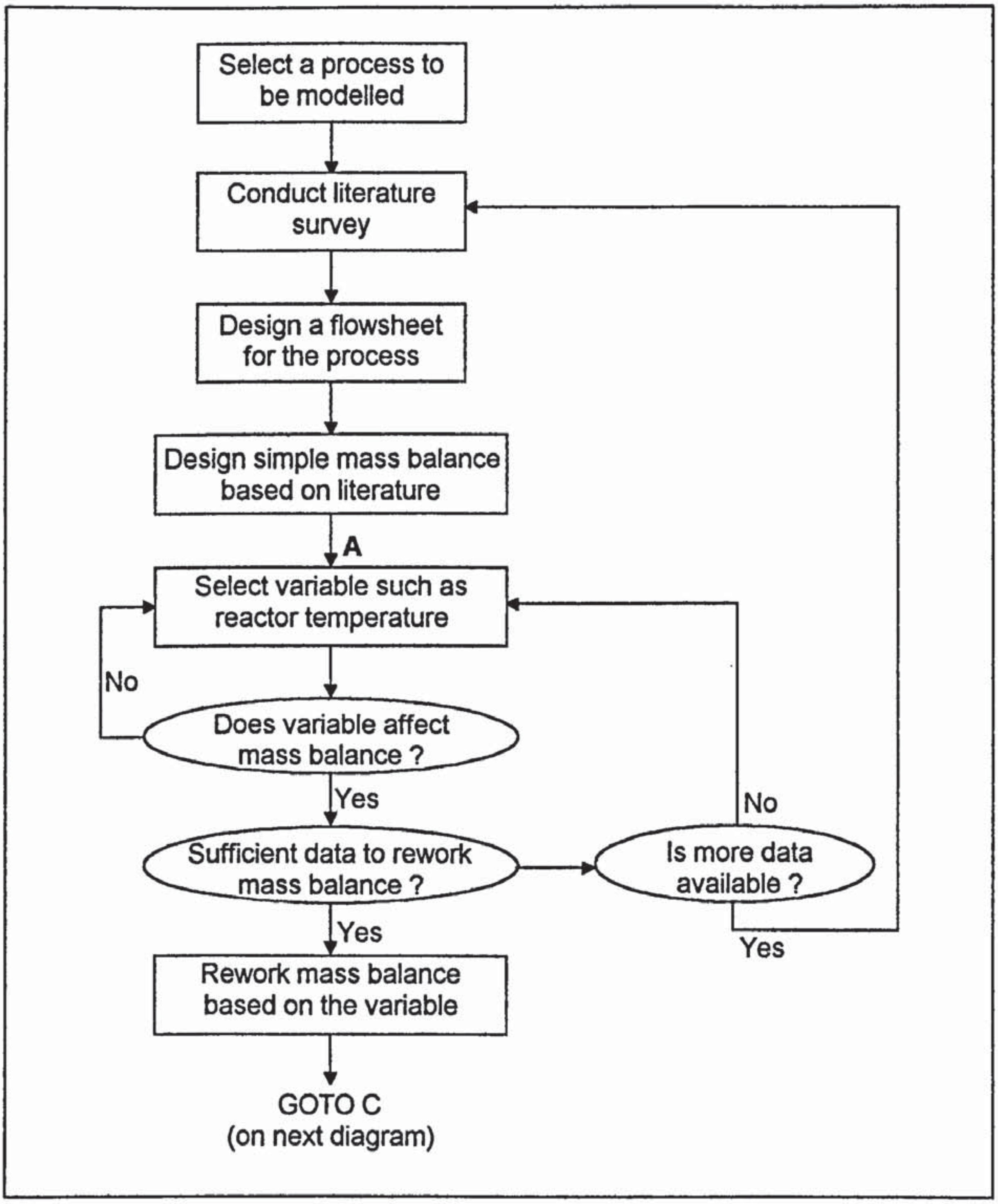


Figure 4.4(a). The stages in the design of a process step model.

3. Construct a mass balance for the process based on the flowsheet and the process information collected.
4. Select one of the process variables, indicated from the collected information, which is known to affect the mass balance for the process e.g. the pyrolysis reaction temperature which alters the organic, reaction water, off-gas and char product yields [6].

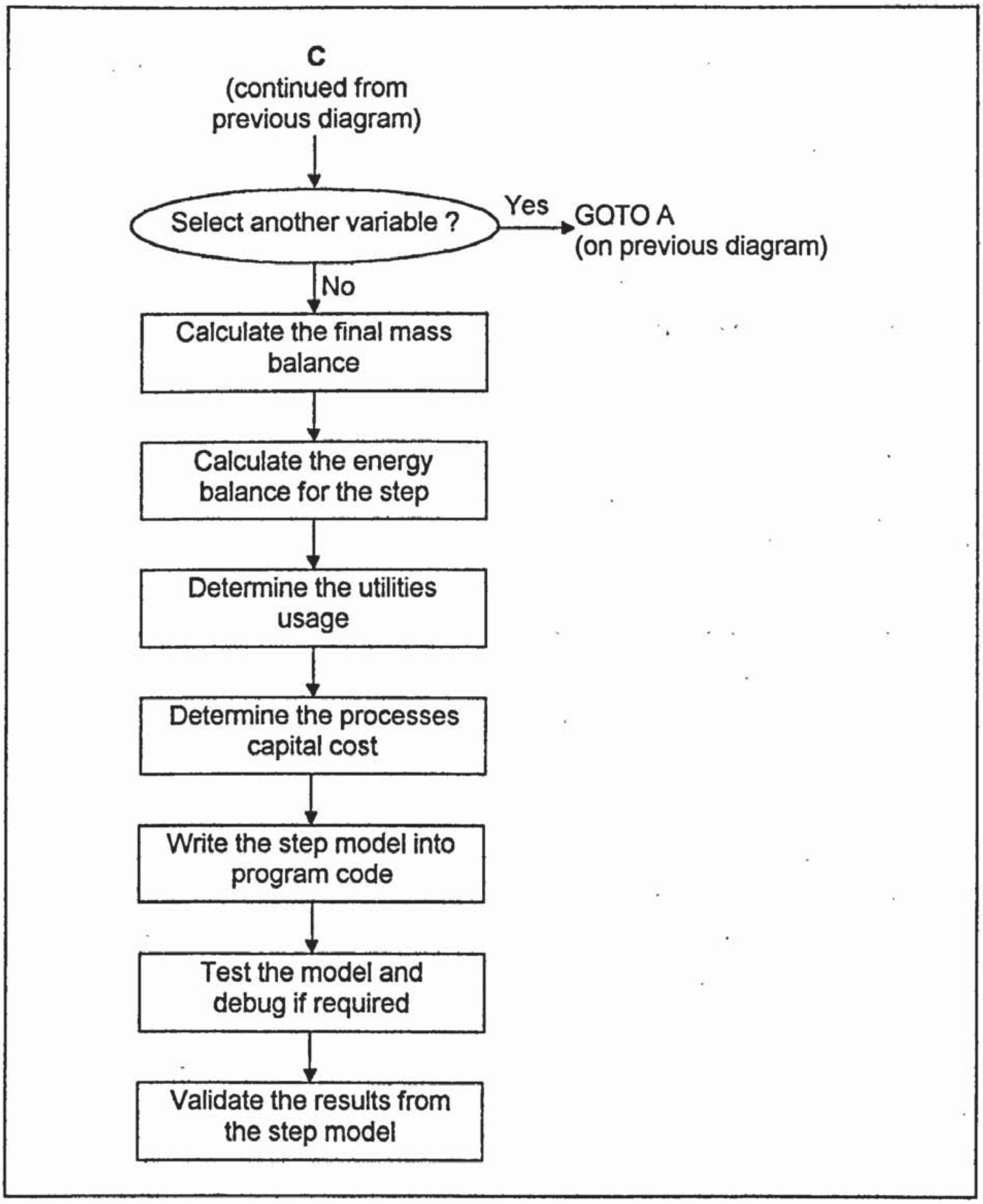


Figure 4.4(b). The stages in the design of a process step model.

5. If sufficient information is available on the process, algorithms are constructed to model the effect of the process variable and the mass balance is reworked. If there is insufficient information available then another known process variable is selected (Return to step 4).

6. The whole process of selecting and testing a process variable, (4) and (5), is repeated until all possible process variables have been considered and the final mass balance, consisting of algorithms, has been constructed.
7. The energy balance for the process is calculated for all the process streams in the mass balance.
8. The capital cost for the process is determined based on the information retrieved in (2) and the final mass balance calculated in (6).
9. The utility requirements and additional costs for the process are determined using the final mass balance developed in (6), the energy balance determined in (7) and the initial information gathered in the literature survey (1).
10. Having completed stages (1) through to (9) the step model is constructed and incorporated into the program where it is run to check that the results are consistent. If the results are inconsistent the step model is checked and debugged.
11. The final stage in the step model development is validation of the results. The results may be validated by comparison with published data or by the company or researcher who provided the original data.

4.5.3. The financial calculations in the DLL.

The economics of a process are usually determined using a cash flow table constructed from the capital cost, operating costs and the income derived from product sales. The cash flow table is utilised to calculate the profitability of the process as measured by payback time, net present worth, discounted cash flow rate of return and return on investment. However, the income derived from selling the liquid fuels produced in a biomass conversion process is hard to determine and so the production cost of the biomass derived product is calculated per tonne and per unit energy.

The production cost of a biomass derived liquid is composed of the cost of the feedstock, the labour costs, utilities cost, byproduct disposal costs, maintenance and overhead costs per year and the total capital cost amortised over the project life. The costs are derived using standard methods which have been demonstrated and validated in other computer simulation packages [131, 135].

4.6. The Graphical User Interface.

4.6.1. Introduction to the GUI.

BLUNT's Graphical User Interface (GUI) allows the program User to enter the process inputs required to run the DLL and view the results from a simulation run. The GUI allows the program User to select the feedstock, product, plant feed input and conversion route for a simulation run. It also allows the selected variables in four of step models (wood drying, Generic flash pyrolysis, hydrotreating and power production) to be altered by the User to customise the conversion process. The financial base case, utilities costs and the base currency unit can also be altered by the program User to customise the financial calculations.

Notice was taken of the limitations of AMBLE [28] when BLUNT was designed and the result was the development of the graphical interface allowing simple data entry and multiple simulation runs without quitting the program. Where AMBLE crashed if a feedstock / product combination was not permitted, BLUNT has been more strictly controlled and only conversion routes which are available in the program are actually visible in the GUI. If values are entered which are out of range, message boxes appear informing the User of the control limits and resetting the incorrect value to the default setting.

Visual Basic has some basic terminology which may not be familiar to the general program User and some of the more common phrases which are used to describe the controls on the screen of a Visual Basic application are given below and shown pictorially in Figure 4.5.

Command buttons	raised, rectangular buttons with text on the face. They are activated by clicking with the mouse or pressing the <RETURN> key when the focus is on the button. Command buttons are referred to by the text on the button e.g. *RUN*.
Radio buttons	are small round buttons with text on the right hand side. They are activated by clicking with the mouse or pressing the <RETURN> key when the focus is on the button. Radio buttons are referred to by the text next to the button e.g. *YES*.

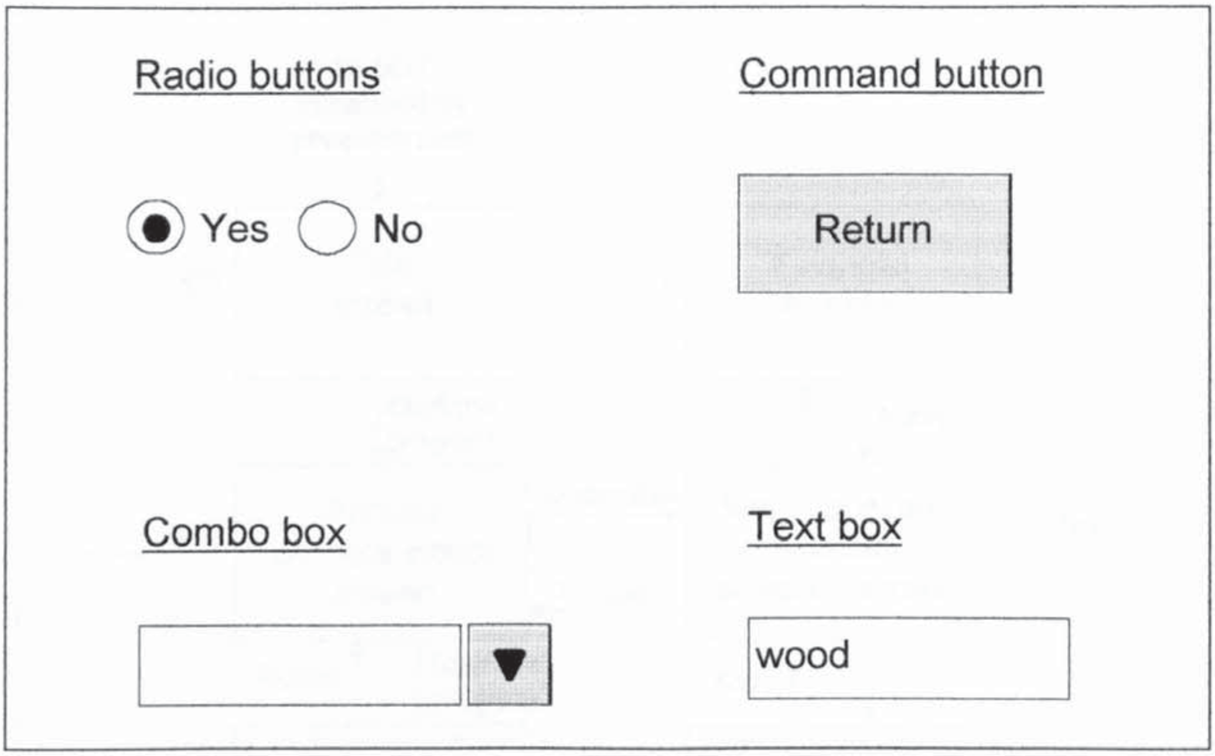


Figure 4.5. Common controls seen on a Visual Basic screen.

Text boxes	are rectangular boxes containing numbers or text. The contents of the text box may altered by the program User by clicking on, highlighting and overwriting. Text boxes are referred to by the text on the left of the box i.e. their titles for example, Plant feed input.
Combo boxes	are text boxes with an arrow on the right hand side. Clicking on the arrow or box activates a list within the combo box and allows the User to scroll down the list to the item to be selected. Combo boxes are referred to by the text on the left of the combo box i.e. their titles for example, Product.

The graphical interface consists of twenty screens which can be broken down into six sections as shown below. The interaction between the screens is shown in Figure 4.6.

Initialisation	consisting of the BLUNT title screen, the Primary Process Inputs (PPI) screen and the Customisation Selection screen (described in Section 4.6.2.

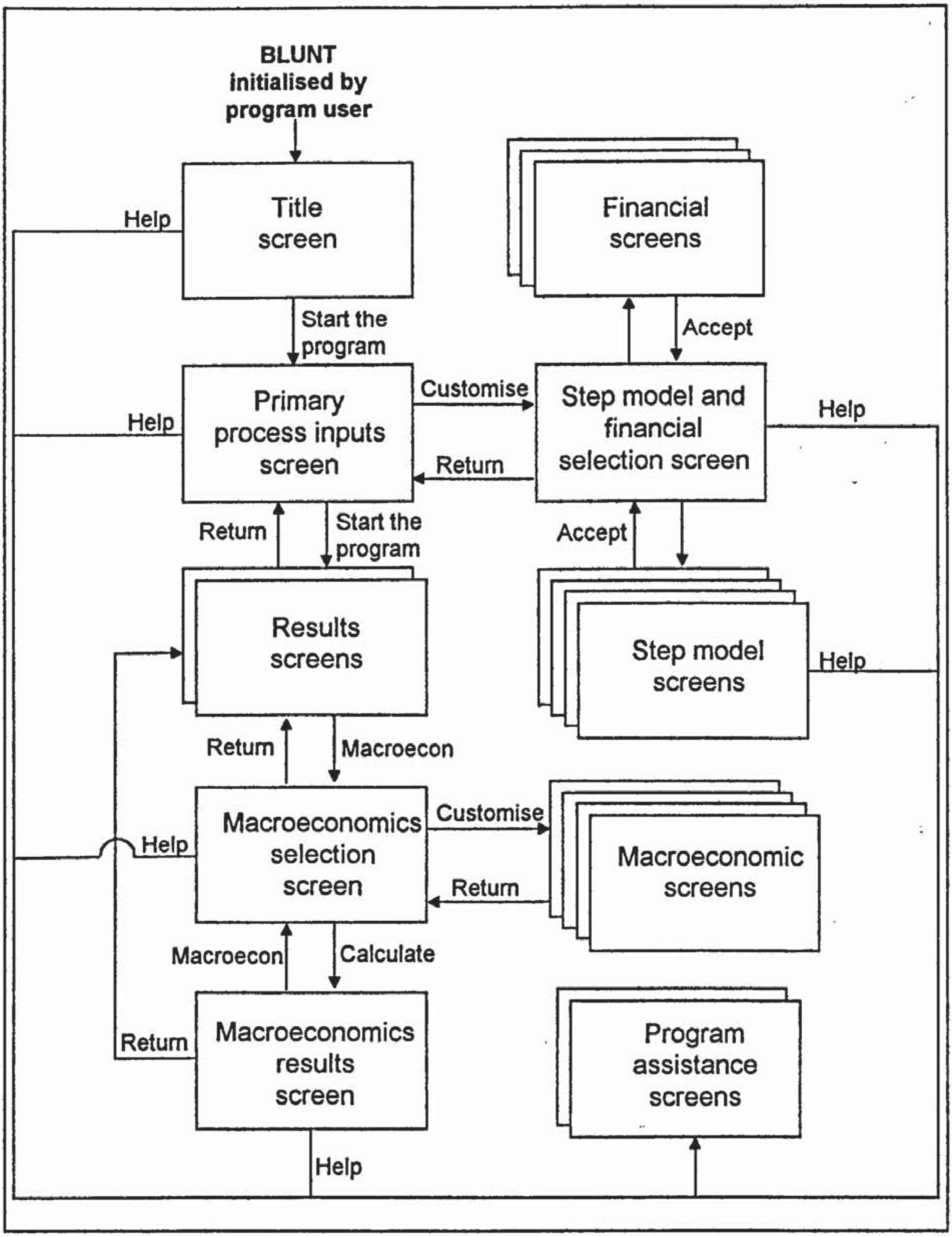


Figure 4.6. The interaction between the screens which make up the graphical interface.

Step models	consisting of four step model customisation screens (described in Section 4.6.3.);

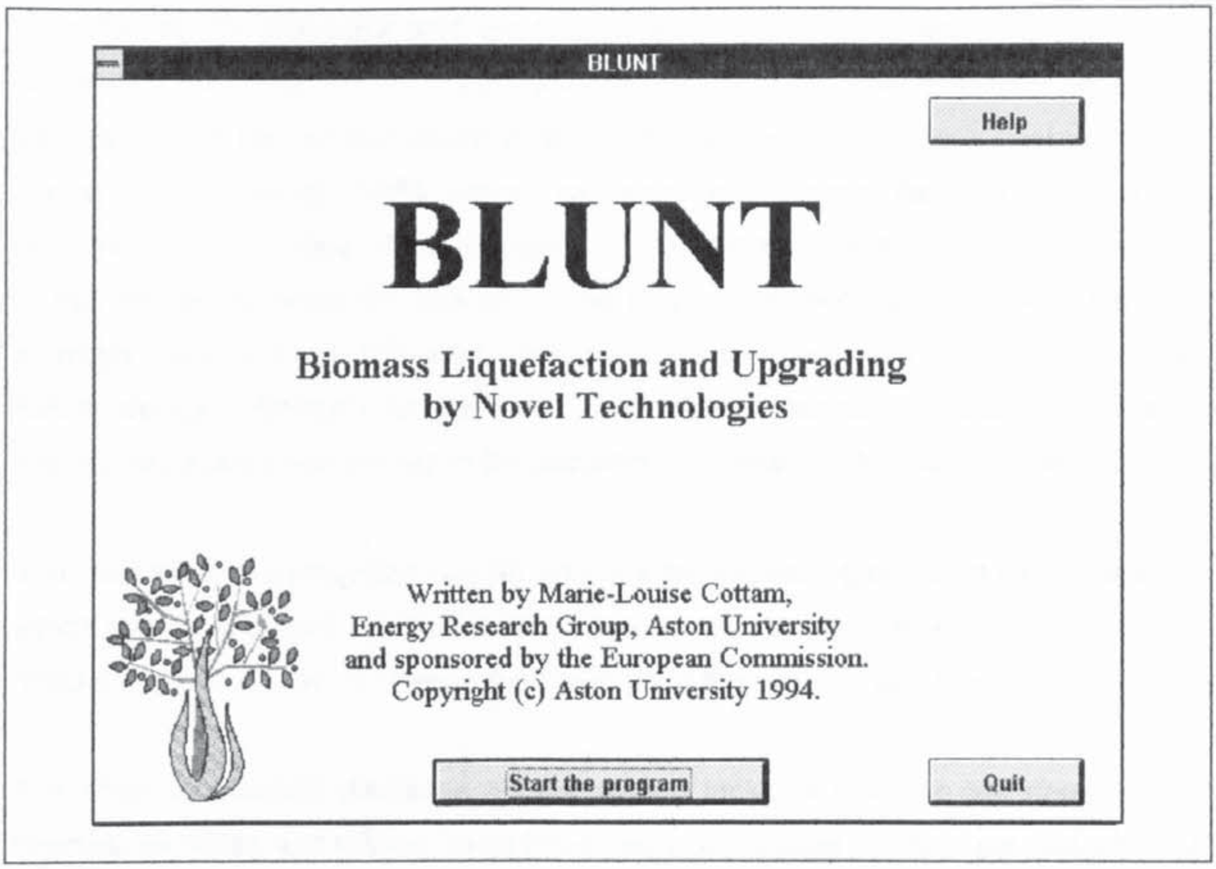
Financial	consisting of three financial customisation screens (described in Section 4.6.4.);
Results	consisting of the results summary and secondary display screens (described in Section 4.6.5.);
Macroeconomics	consisting of six screens dealing with the selection, customisation and results display for the macroeconomic factors (described in Section 4.6.6.); and
Help	consisting of two screens which display the program assistance for BLUNT (described in Section 4.6.7.).

The presence of a *HELP* button on any screen allows the program User to access the Program Assistance Screens which are described in Section 4.6.7. and *QUIT* always allows the User to leave the BLUNT and return to the Windows Program Manager.

4.6.2. The initialisation screens.

The initialisation screens consist of the BLUNT title screen, the primary process inputs screen and the customisation selection screen. When initialised, BLUNT activates the title screen as shown in Figure 4.7. The BLUNT title screen gives some basic information about the program and includes the program's logo, a burning tree symbolising biomass thermochemical conversion. The initialisation screen has three command buttons, *START THE PROGRAM, HELP* and *QUIT. HELP* and *QUIT* have been previously explained and *START THE PROGRAM* closes the title screen and opens the Primary Process Inputs (PPI) Screen for main data entry.

The PPI screen allows the User to select the essential information required for running BLUNT. The primary data consists of the feedstock, product, conversion route and the feed input to the plant. The PPI screen is shown in Figure 4.8. The feed, product and conversion route are selected from a list of items which appear when the mouse is clicked on the relevant combo box. The User is not permitted to type in extra items to the lists because only the items present in the combo boxes have been modelled by the program. The feeds and products which are modelled in the program are detailed in the program scope in Section 1.3. and the combinations of feed, products and conversion routes which are available are summarised in the tables in Section 10.1.



BLUNT

Biomass Liquefaction and Upgrading
by Novel Technologies

Written by Marie-Louise Cottam,
Energy Research Group, Aston University
and sponsored by the European Commission.
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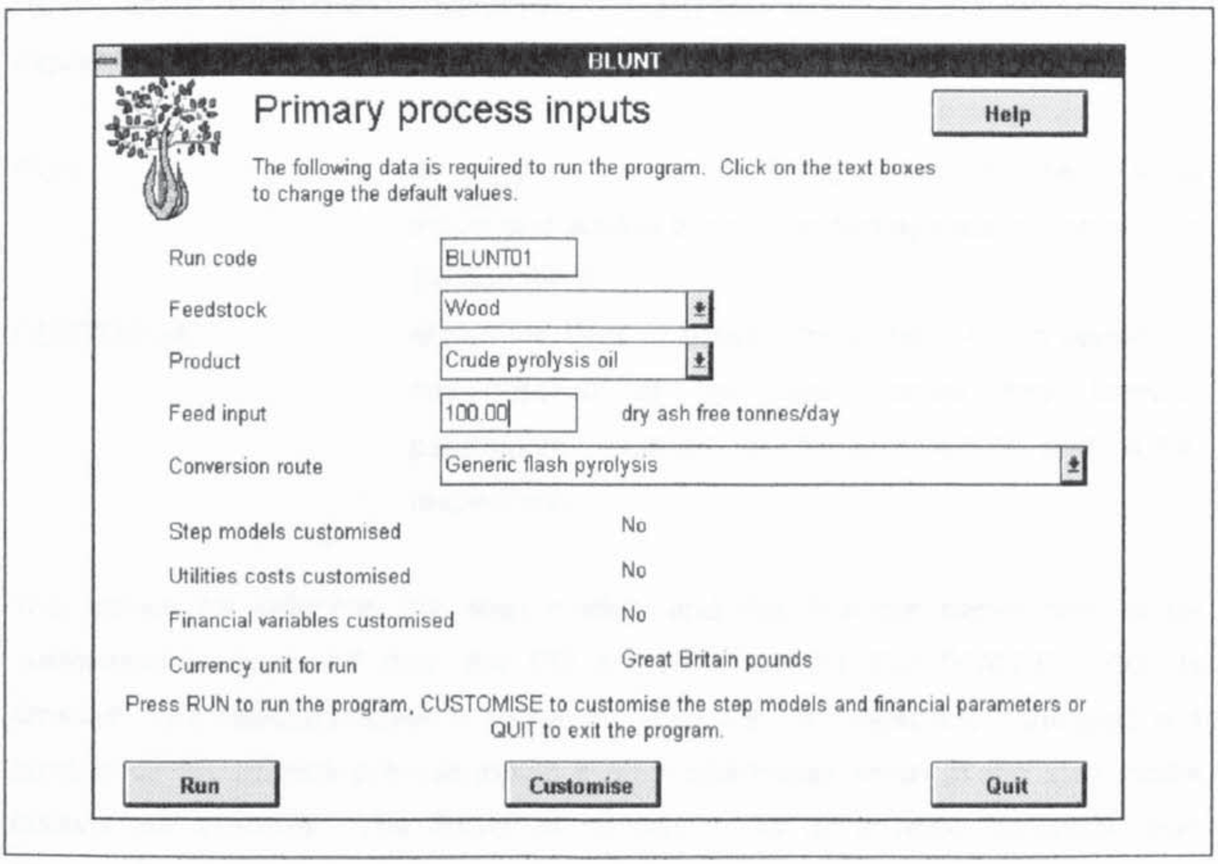
Figure 4.7. The BLUNT title screen.



Primary process inputs

The following data is required to run the program. Click on the text boxes to change the default values.

Run code	BLUNT01	
Feedstock	Wood	
Product	Crude pyrolysis oil	
Feed input	100.00	dry ash free tonnes/day
Conversion route	Generic flash pyrolysis	
Step models customised	No	
Utilities costs customised	No	
Financial variables customised	No	
Currency unit for run	Great Britain pounds	

Press RUN to run the program, CUSTOMISE to customise the step models and financial parameters or QUIT to exit the program.

Figure 4.8. The Primary Process Inputs Screen.

Selection of the feedstock and product determines the items which appear in the conversion route combo box. For example, selecting a wood feedstock and crude pyrolysis oil as the product would give conversion routes of Generic flash pyrolysis, Ensyn RTP pyrolysis, NREL vortex pyrolysis and Fenosa flash pyrolysis in the conversion route list box. If the feedstock is changed to straw the only item appearing in the conversion route list box would be Ensyn RTP pyrolysis as this is the only pyrolysis process in BLUNT which can take a straw feedstock (See Chapter 6 for further details). Defaults for the feed and product of wood and crude pyrolysis oil respectively have been pre-set in the program but these can be altered by the User.

The feed input or throughput has an arbitrary default setting of 100 d.a.f. tonnes/day which can be changed by the program User to accept any values between 25 and 1000 d.a.f. tonnes/day by altering the numbers in the relevant text box.

The other information displayed on the screen includes whether the step models, financial variables and utilities costs have been customised by the User and whether the base currency unit has been changed from Great Britain pounds to one of the other currencies available. The command buttons available on the PPI Screen are *HELP*, *QUIT*, *RUN* and *CUSTOMISE*. *QUIT* and *HELP* have been previously explained and the functions of the other buttons are:-

RUN	allows the User to run the program with the process inputs and access the results display screens, detailed in Section 4.6.5.
CUSTOMISE	allows the User to access the screens which permit the customisation of the step models and financial parameters, detailed in Section 4.6.3. and 4.6.4. respectively.

The screen for selecting the step models and the financial parameters to be customised is accessed from the PPI screen when the *CUSTOMISE* button is pressed. The selection screen is shown in Figure 4.9. The selection of the feed and product on the primary process inputs screen determines which of the step model buttons are available. The drying model only works for a wood feedstock (See Chapter 5 for details on the wood drying model) so selection of straw or sorghum bagasse as the feedstock deactivates the *DRYING* button. If generic flash pyrolysis is selected as the pyrolysis process from the list of Generic, Ensyn, Fenosa and NREL the *GENERIC FLASH PYROLYSIS* button will be activated. Selection of any of the other three pyrolysis processes deactivates the button.

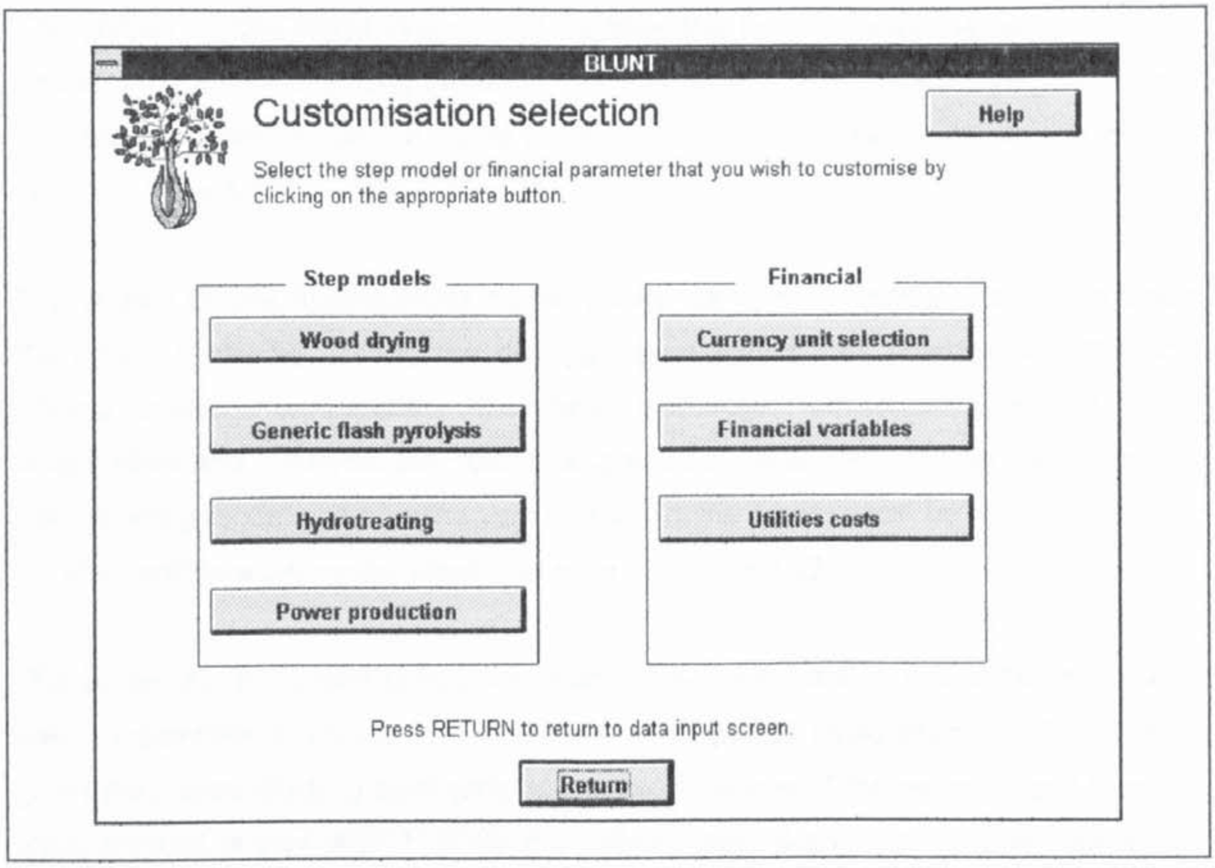
If either hydrotreated oil or diesel or power (when the processing route involves hydrotreating) is selected as the product on the primary process inputs screen the *HYDROTREATING* button will be activated. Selection of any other products or processing routes deactivates this button. The *POWER PRODUCTION* button is only activated when electricity is selected as a product on the PPI screen. Clicking on any of the step model buttons which are activated sends the User to the appropriate screen (detailed in Section 4.6.3). Clicking on the *RETURN* button at the bottom of the screen sends the User back to the PPI screen.

4.6.3. The step models screens.

The step models for generic flash pyrolysis, drying, hydrotreating and power production have variables which the program User can change to alter the ways in which the models behave. Each of the step model screens have two buttons - *DEFAULTS* and *ACCEPT*. The function of these buttons is always the same.

ACCEPT	accepts any changes made to the default variables on the screen and switches the User back to the selection screen.
DEFAULTS	changes the variables on the screen back to the default settings. *ACCEPT* must be pressed to return to the selection screen.

The screen for the generic flash pyrolysis model allows the User to customise the model by altering the pyrolysis reactor temperature, the char/off-gas combustion efficiency, the quench column energy balance, the source of the energy required for the pyrolysis process and the char filter efficiency. Further details on the Generic flash pyrolysis model and the variables which can be used to customise the model can be found in Chapter 6. The generic flash pyrolysis model screen is shown in Figure 4.10.

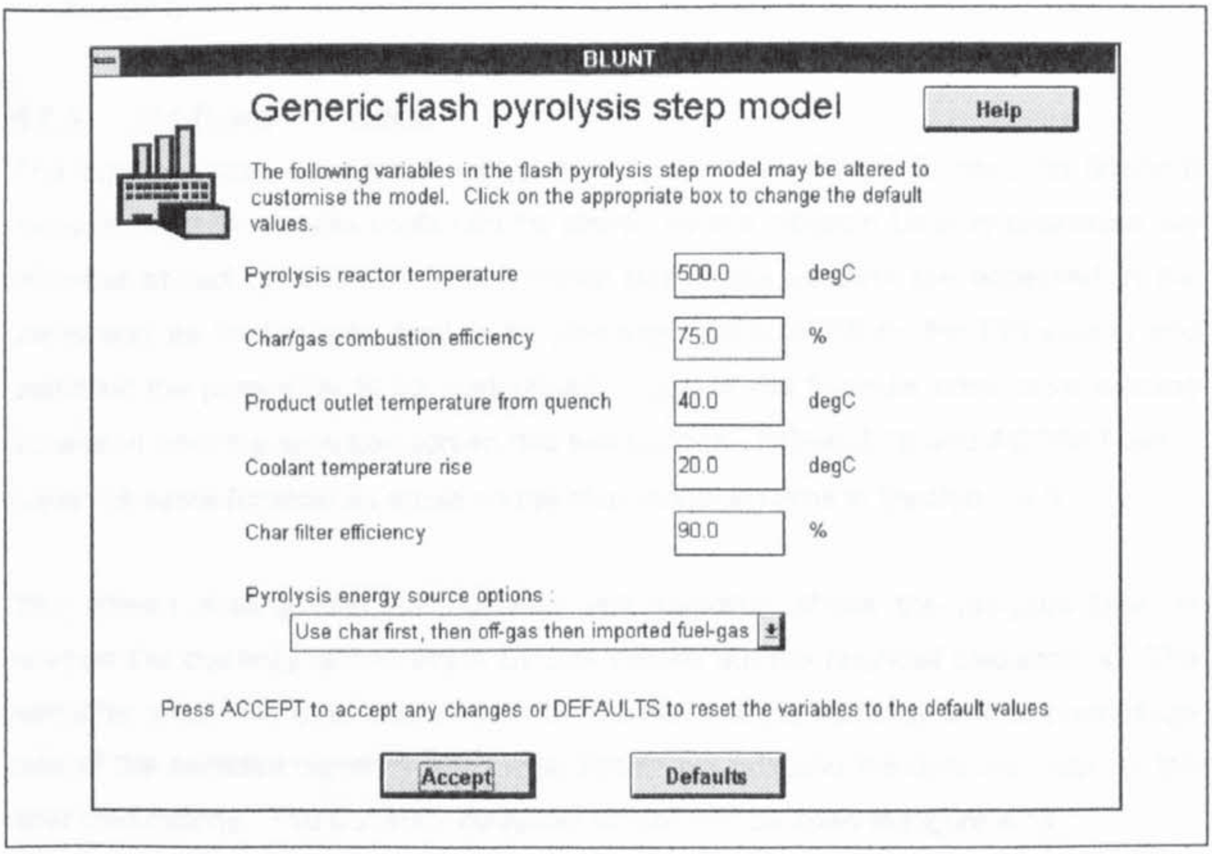


Customisation selection

Select the step model or financial parameter that you wish to customise by clicking on the appropriate button.

Press RETURN to return to data input screen.

Figure 4.9. The step models and financial parameters selection screen.



Generic flash pyrolysis step model

The following variables in the flash pyrolysis step model may be altered to customise the model. Click on the appropriate box to change the default values.

Pyrolysis reactor temperature	500.0	degC
Char/gas combustion efficiency	75.0	%
Product outlet temperature from quench	40.0	degC
Coolant temperature rise	20.0	degC
Char filter efficiency	90.0	%

Pyrolysis energy source options :
Use char first, then off-gas then imported fuel-gas

Press ACCEPT to accept any changes or DEFAULTS to reset the variables to the default values

Figure 4.10. The screen for the generic flash pyrolysis model.

The screen for the wood drying model allows the User to alter the wood moisture contents entering and exiting the dryer. Further details on the wood drying model and it's process variables can be found in Chapter 5. The drying model screen can be seen in Figure 4.11.

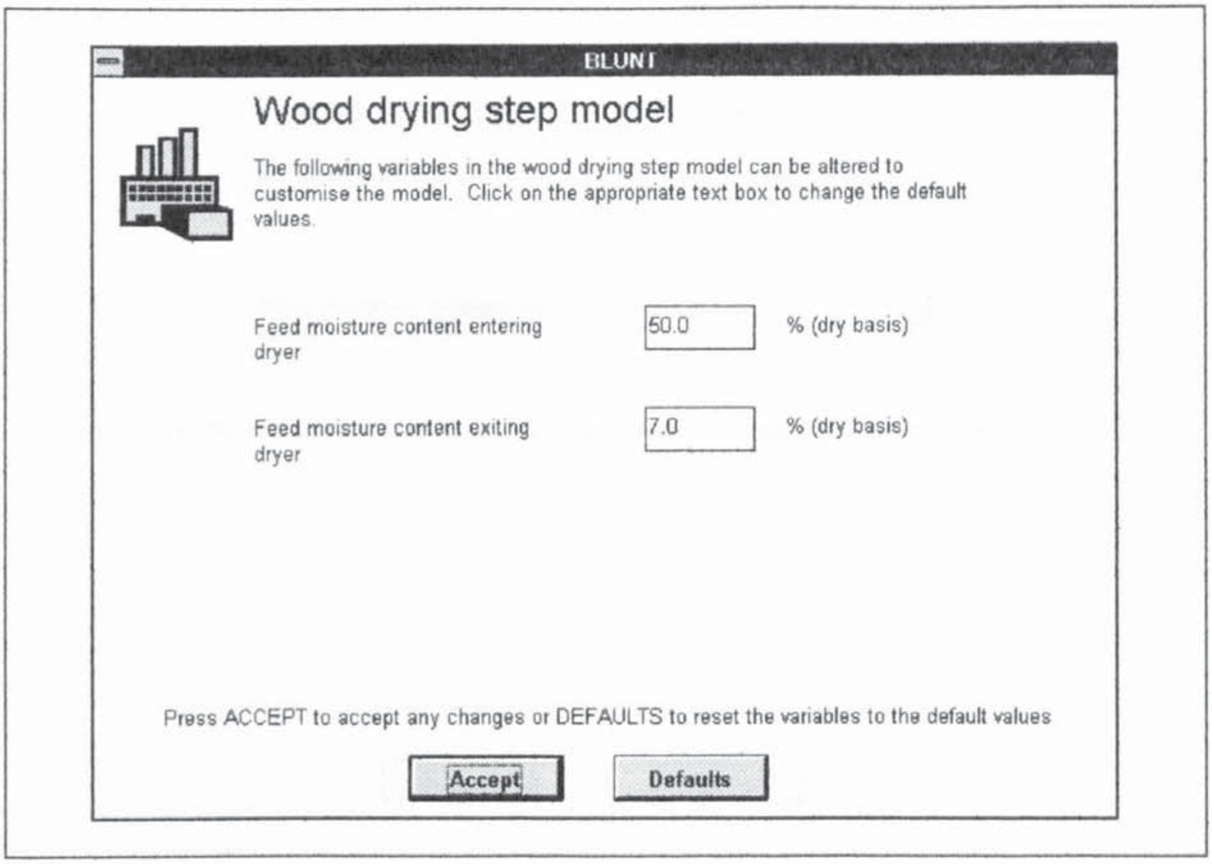
The screen for the hydrotreating model allows the User to change the way in which the model reacts by altering the deoxygenation extent, gas combustion efficiency, cooling medium temperature rise, heat exchanger efficiencies, product outlet temperature and the hydrogen source/off-gas disposal option. Further details on the hydrotreating model and how the variables affect the process can be found in Chapter 7. The hydrotreating model screen is shown in Figure 4.12.

The screen for the power production model allows the User to select the technology used to generate electricity from the crude or upgraded liquid product. The User is given the choice of using an engine or turbine. However, if the power output from the liquid product is less than 5 MWe the default technology is an engine and above 20MWe the default technology is a turbine. Figure 4.13. shows the screen for the power production model. Further details on the power production model can be found in Chapter 8.

4.6.4. The financial screens.

The financial base case for the program i.e. the currency unit in use, the financial variables and the utilities costs can be altered by the program User to customise the financial aspect of BLUNT. The financial base case screens are accessed in the same way as for the step models by pressing *CUSTOMISE* on the PPI screen and selecting the parameter to be customised. Each of the financial base case screens accessed from the selection screen has two buttons - *DEFAULTS* and *ACCEPT* which have the same function as those on the step model screens in Section 4.6.3.

The screen dealing with the currency unit selection allows the program User to change the currency unit in which BLUNT carries out the financial calculations. The variables which the User can alter on the screen are the currency unit, the exchange rate of the selected currency with Great Britain pounds and the financial year for the cost calculations. The currency selection screen can be seen in Figure 4.14.



Wood drying step model

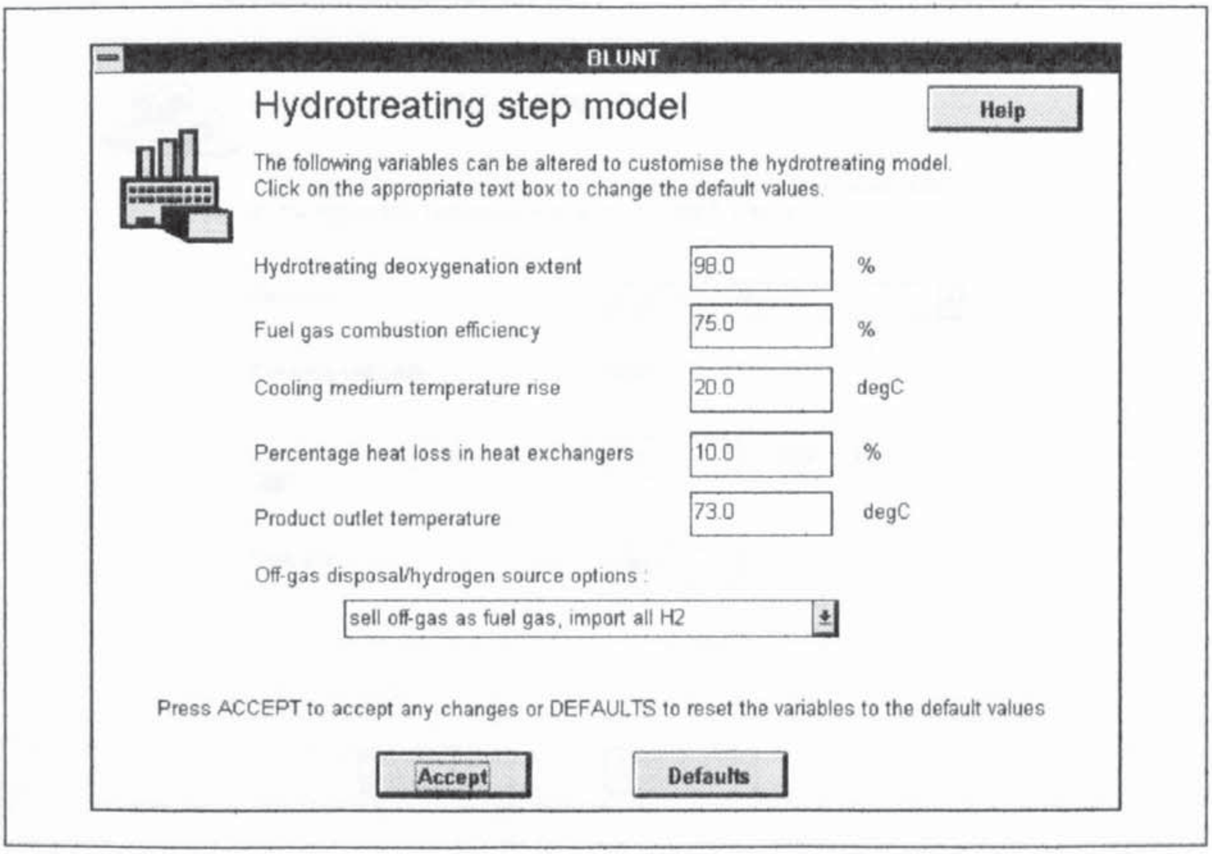
The following variables in the wood drying step model can be altered to customise the model. Click on the appropriate text box to change the default values.

Feed moisture content entering dryer 50.0 % (dry basis)

Feed moisture content exiting dryer 7.0 % (dry basis)

Press ACCEPT to accept any changes or DEFAULTS to reset the variables to the default values

Figure 4.11. The screen for the wood drying model.



Hydrotreating step model

The following variables can be altered to customise the hydrotreating model. Click on the appropriate text box to change the default values.

Hydrotreating deoxygenation extent 98.0 %

Fuel gas combustion efficiency 75.0 %

Cooling medium temperature rise 20.0 degC

Percentage heat loss in heat exchangers 10.0 %

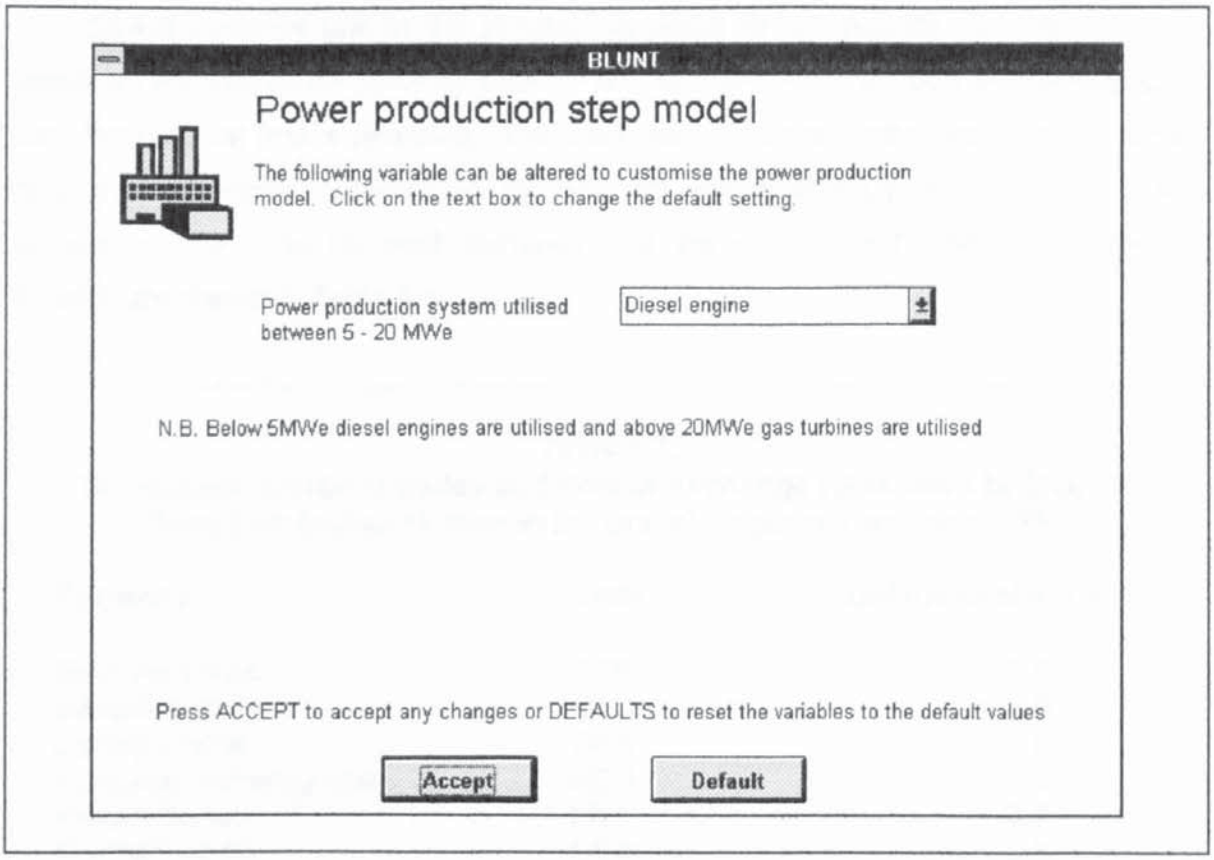
Product outlet temperature 73.0 degC

Off-gas disposal/hydrogen source options :

sell off-gas as fuel gas, import all H2

Press ACCEPT to accept any changes or DEFAULTS to reset the variables to the default values

Figure 4.12. The screen for the hydrotreating model.



Power production step model

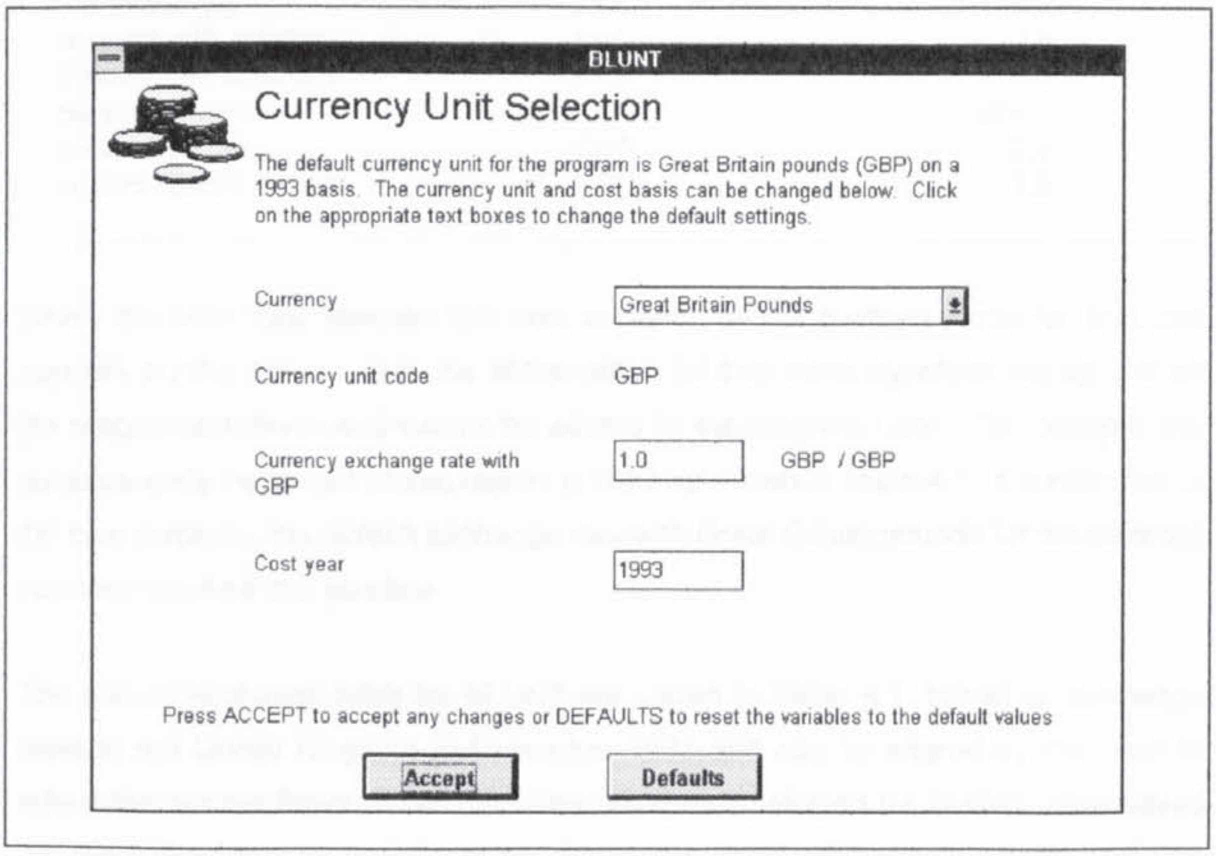
The following variable can be altered to customise the power production model. Click on the text box to change the default setting.

Power production system utilised between 5 - 20 MWe: Diesel engine

N.B. Below 5MWe diesel engines are utilised and above 20MWe gas turbines are utilised

Press ACCEPT to accept any changes or DEFAULTS to reset the variables to the default values

Figure 4.13. The screen for the power production model.



Currency Unit Selection

The default currency unit for the program is Great Britain pounds (GBP) on a 1993 basis. The currency unit and cost basis can be changed below. Click on the appropriate text boxes to change the default settings.

Currency: Great Britain Pounds

Currency unit code: GBP

Currency exchange rate with GBP: 1.0 GBP / GBP

Cost year: 1993

Press ACCEPT to accept any changes or DEFAULTS to reset the variables to the default values

Figure 4.14. The screen for currency unit selection.

The default currency unit in the program is Great Britain pounds but the User can select an alternative currency by clicking on the appropriate list box and selecting an item from the list that is revealed. The User cannot add any other items to the list as only the currencies included in the list are recognised by the program. The currencies, the code for each currency and the default exchange rates used by BLUNT are shown in Table 4.1.

Table 4.1.
Currencies, currency codes and default exchange rates used by BLUNT.
(Based on exchange rates in the United Kingdom, December 1993)

Currency	Code	Default exchange rate
Belgium francs	BEF	53.5
Canadian dollars	CAD	2.0
Danish kroner	DKK	10.0
European currency units	ECU	1.3
French francs	FRF	8.7
German marks	DEM	2.5
Greek drachma	GRD	359.3
Irish punts	IRP	1.1
Italian lira	ITL	2476.1
Japanese yen	JAY	158.6
Netherlands guilder	NLG	2.8
Portuguese escudos	PTE	256.7
Spanish pesetas	ESP	204.1
Swiss francs	CHF	2.2
United States dollars	USD	1.5

When the User has selected the new currency unit a currency code for that unit appears on the screen. It is the abbreviation for that currency which will appear on the program's screens and cannot be altered by the program User. For example, the currency code for United States dollars is USD as shown in Table 4.1. On selection of the new currency, the default exchange rate with Great Britain pounds for the selected currency appears in a text box.

The default exchange rates for BLUNT are shown in Table 4.1. based on exchange rates in the United Kingdom in December 1993 and can be altered by the User to reflect the current financial climate. The currencies selected for BLUNT were based on where the program is likely to be distributed which will be mainly in Europe and North America. The base financial year for the the financial calculations in BLUNT is

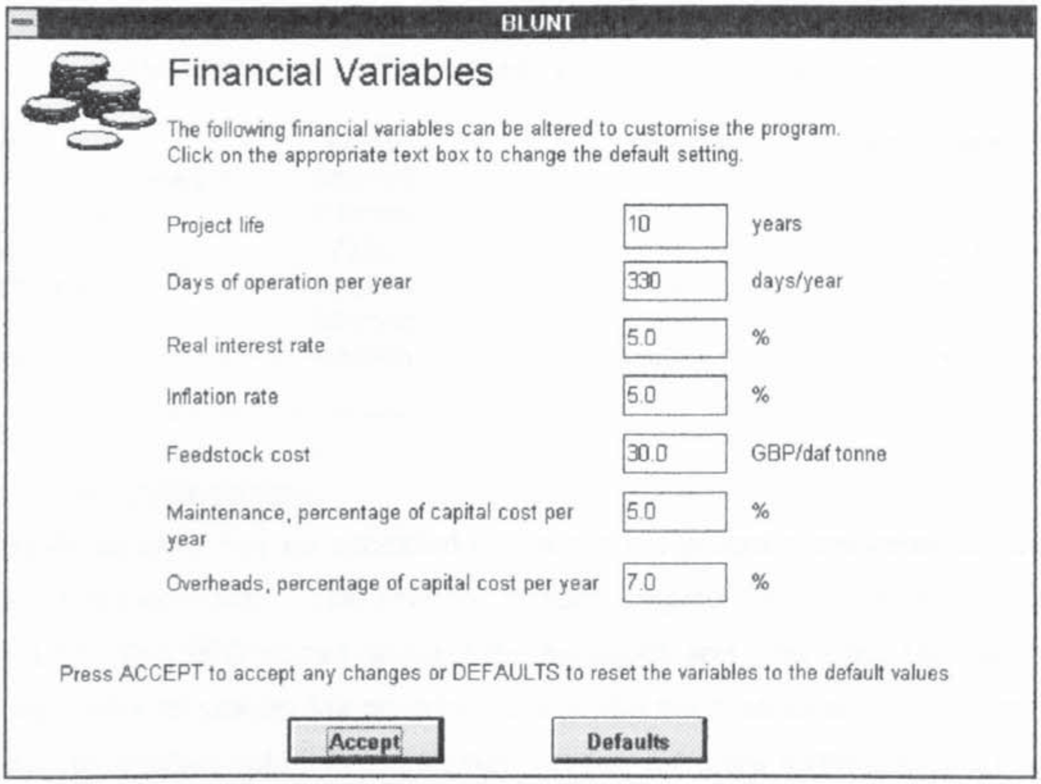
1993. However, the base year can be updated using a cost index to recalculate the capital costs for the various parts of the selected processing route.

The screen dealing with BLUNT's financial variables allows the User to customise the financial calculations. The variables which the User can alter include the project life, operational days per year, real interest rate, inflation rate, feedstock cost and the maintenance and overheads costs as a percentage of capital cost. The financial variables screen can be seen in Figure 4.15. The default settings for the financial variables are shown in Table 4.2. and are based on the settings used in AMBLE [28] with some slight modifications based on discussions with other researchers carrying out economic analyses. As before, the default settings can be altered by the program User to customise the BLUNT when it is being operated.

Table 4.2.
Default settings and permitted ranges for the financial variables

Item	Units	Default	Permitted range
Project life	years	10	1 - 50
Operational days	days/year	330	1 - 365
Real interest rate	%	5	1 - 100
Inflation rate	%	5	1 - 100
Feedstock cost	£/d.a.f. t	30	0 - 100
Maintenance cost	% Cap Cost	5	1 - 100
Overheads cost	% Cap Cost	7	1 - 100

The utilities cost screen allows the User to affect the viability of the process by changing the utilities costs and hence production cost of the product. The utilities cost screen is shown in Figure 4.16. The utilities used by the program are low pressure (5 bar) steam, cooling water, fuel gas, hydrogen, oxygen and power. The only limitation on the production cost of the utilities is that they cannot be zero. Table 4.3. gives the arbitrary default settings for the utilities costs, which are based on the prices used in AMBLE [28] with slight increases to allow for inflation. The hydrogen cost has been reduced from the 1500 £/tonne used in AMBLE (1990) to 360 £/tonne which is thought to be a more realistic cost. It must be stressed however, that the default costs are only included as a guide and can be altered by the program User when BLUNT is run.



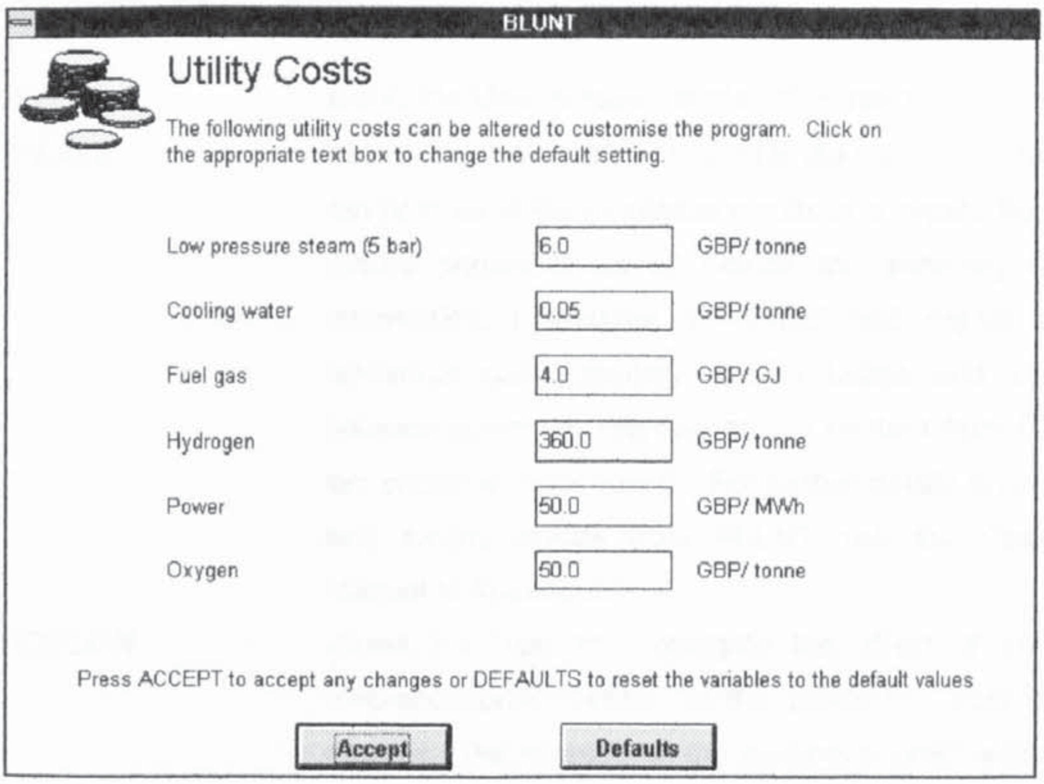
Financial Variables

The following financial variables can be altered to customise the program. Click on the appropriate text box to change the default setting.

Project life	10	years
Days of operation per year	330	days/year
Real interest rate	5.0	%
Inflation rate	5.0	%
Feedstock cost	30.0	GBP/daf tonne
Maintenance, percentage of capital cost per year	5.0	%
Overheads, percentage of capital cost per year	7.0	%

Press ACCEPT to accept any changes or DEFAULTS to reset the variables to the default values

Figure 4.15. The screen for customising the financial variables.



Utility Costs

The following utility costs can be altered to customise the program. Click on the appropriate text box to change the default setting.

Low pressure steam (5 bar)	6.0	GBP/ tonne
Cooling water	0.05	GBP/ tonne
Fuel gas	4.0	GBP/ GJ
Hydrogen	360.0	GBP/ tonne
Power	50.0	GBP/ MWh
Oxygen	50.0	GBP/ tonne

Press ACCEPT to accept any changes or DEFAULTS to reset the variables to the default values

Figure 4.16. The screen for customising the utilities costs.

Table 4.3.
Default settings and permitted ranges for the utilities costs.

Item	**Units**	**Default**	**Permitted range**
Low pressure steam	£/tonne	6	> 0
Cooling water	£/tonne	0.05	> 0
Fuel gas	£/GJ	4	> 0
Hydrogen	£/tonne	360	> 0
Oxygen	£/tonne	50	> 0
Power	£/MWh	50	> 0

4.6.5. The results screens.

The results screens may be accessed only when the program has been run with the selected process inputs. The Primary Results Display (PRD) screen is shown in Figure 4.17. The PRD screen displays the feedstock and feed input selected on the PPI screen prior to running the program. Below this the product and conversion route selected, the product output in tonnes/day, product moisture content, product heating value in GJ/tonne, the production cost in the selected currency per tonne and the capital cost in the selected currency are displayed. There are several buttons on the results screen - *RETURN, MORE, PRINT/SAVE, MACROECON* and *QUIT*. *QUIT* has been explained earlier and the function of the other buttons is given below:

RETURN	allows the User to return to the PPI screen.
PRINT/SAVE	allows the User to print out the results from the simulation run or to save the simulation results to a results file. The results printed or saved include the summary results information, breakdown of annual and capital costs, additional cost summary, utilities usage and a mass balance summary. Sample results printout from BLUNT are shown in Appendix B. For further details on printing and saving results from BLUNT see the Operating Manual in Appendix A.
MACROECON	allows the User to investigate the effect of selected macroeconomic factors on the production cost of the product. The screens for the macroeconomic factors are detailed in Section 4.6.7. For more information on the macroeconomic factors see Chapter 9.

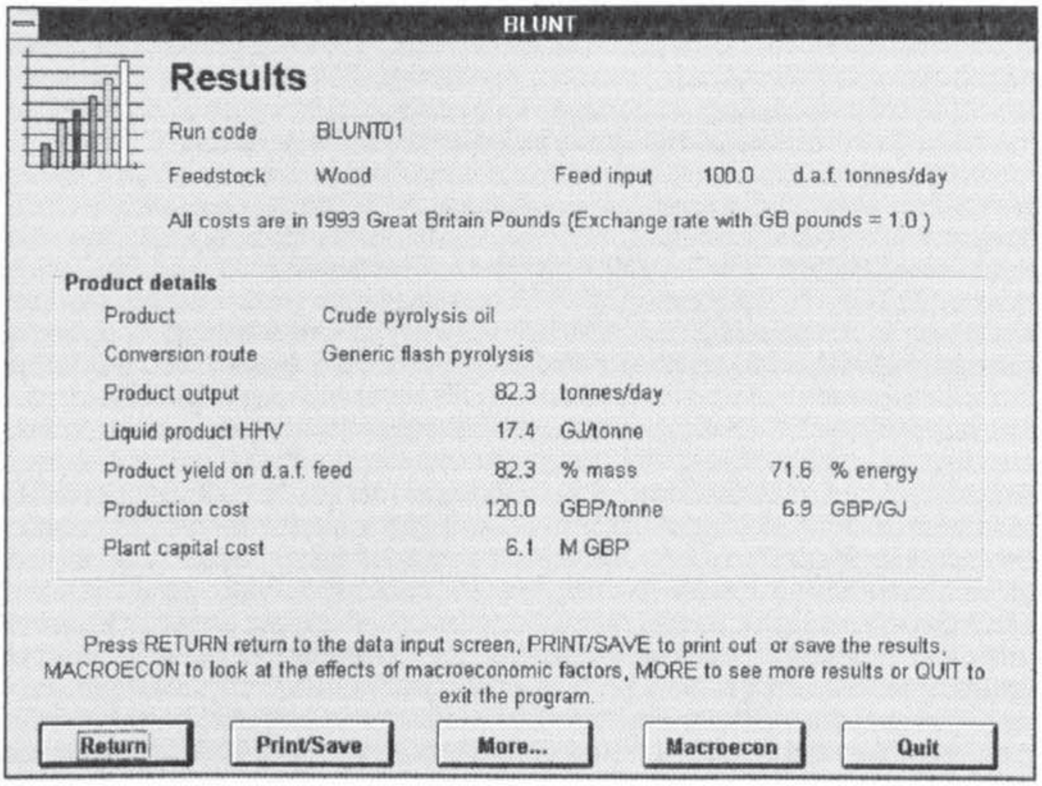
MORE	allows the User to view more detailed results from the simulation run (See below).

If the *MORE* button on the PRD screen is pressed the Secondary Results Display (SRD) screen is opened. The SRD screen allows the User to display the capital, additional and annual cost breakdowns and the utility requirements for the simulation run. The SRD screen has five buttons - *CAPITAL, ADDITION, RETURN, UTILITIES* and *ANNUAL.* The function of each button is described below.

RETURN	returns the program User to the PRD screen.
ANNUAL	displays the breakdown of the annual costs (feedstock, capital amortisation, overheads, utilities, maintenance, labour, byproduct disposal and additional costs) for the simulation run. The display of this information is illustrated in Figure 4.18.
CAPITAL	displays the breakdown of the capital cost for each process step in the conversion route. The display of this information is illustrated in Figure 4.19.
UTILITIES	displays the breakdown of the utility (low pressure steam, cooling water, fuel gas, hydrogen, oxygen and power) requirements for each process step in the conversion route. The display of this information is illustrated in Figure 4.20.
ADDITION	displays the breakdown of the byproducts, waste streams and additional costs incurred by the processes in the conversion route. The display of this information is illustrated in Figure 4.21.

4.6.6. The macroeconomic screens.

The macroeconomic section of BLUNT overlay the impact of selected macroeconomic issues on the technical criteria employed by the program to assess the financial viability of a conversion process. The factors considered to be significant and which are modelled in BLUNT are: measures to control the use of high sulphur fossil fuels; carbon taxes on fossil fuels; water resources; and specific capital subsidies or credits. Further details on the macroeconomic factors can be found in Chapter 9.



Results

Run code BLUNT01

Feedstock Wood Feed input 100.0 d.a.f. tonnes/day

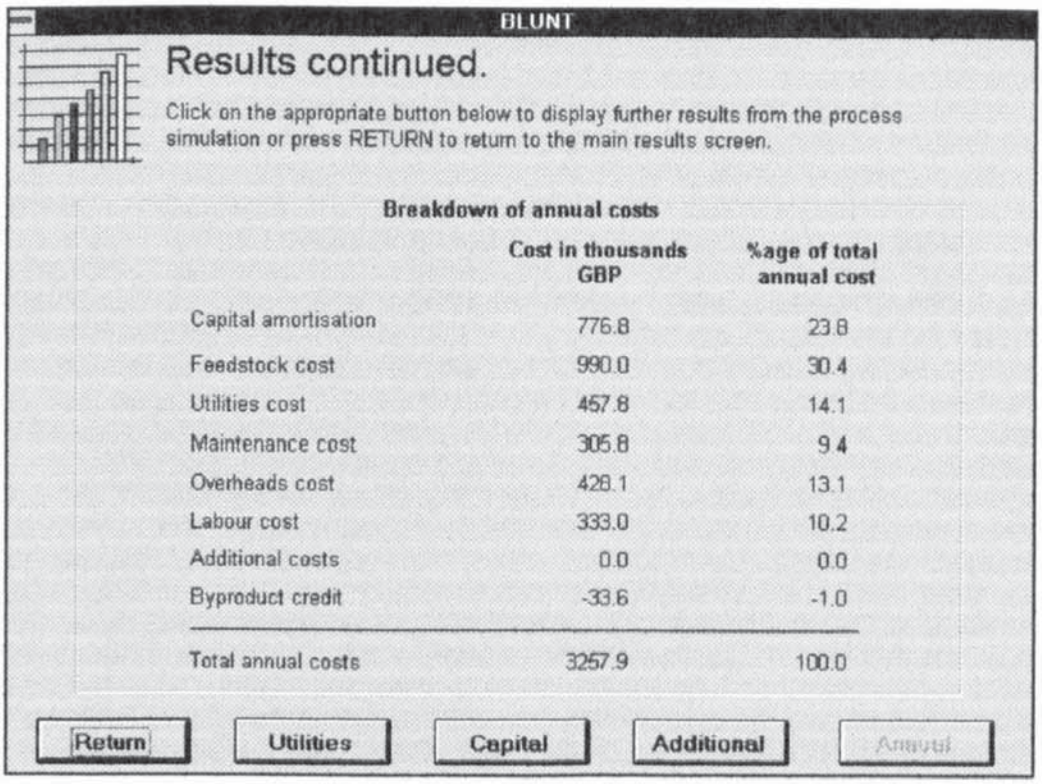
All costs are in 1993 Great Britain Pounds (Exchange rate with GB pounds = 1.0)

Product details

Product	Crude pyrolysis oil			
Conversion route	Generic flash pyrolysis			
Product output	82.3	tonnes/day		
Liquid product HHV	17.4	GJ/tonne		
Product yield on d.a.f. feed	82.3	% mass	71.6	% energy
Production cost	120.0	GBP/tonne	6.9	GBP/GJ
Plant capital cost	6.1	M GBP		

Press RETURN return to the data input screen, PRINT/SAVE to print out or save the results, MACROECON to look at the effects of macroeconomic factors, MORE to see more results or QUIT to exit the program.

Figure 4.17. The Primary Results Display Screen.



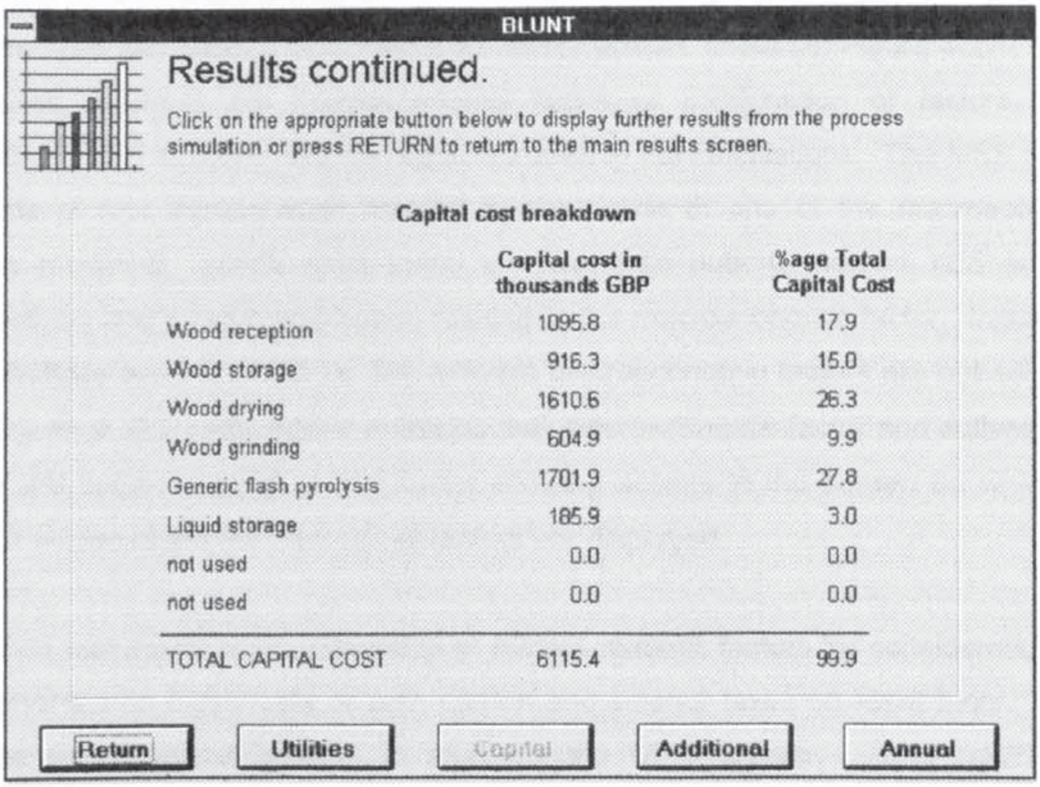
Results continued.

Click on the appropriate button below to display further results from the process simulation or press RETURN to return to the main results screen.

Breakdown of annual costs

	Cost in thousands GBP	%age of total annual cost
Capital amortisation	776.8	23.8
Feedstock cost	990.0	30.4
Utilities cost	457.8	14.1
Maintenance cost	305.8	9.4
Overheads cost	428.1	13.1
Labour cost	333.0	10.2
Additional costs	0.0	0.0
Byproduct credit	-33.6	-1.0
Total annual costs	3257.9	100.0

Figure 4.18. The SRD screen displaying the breakdown of annual costs.



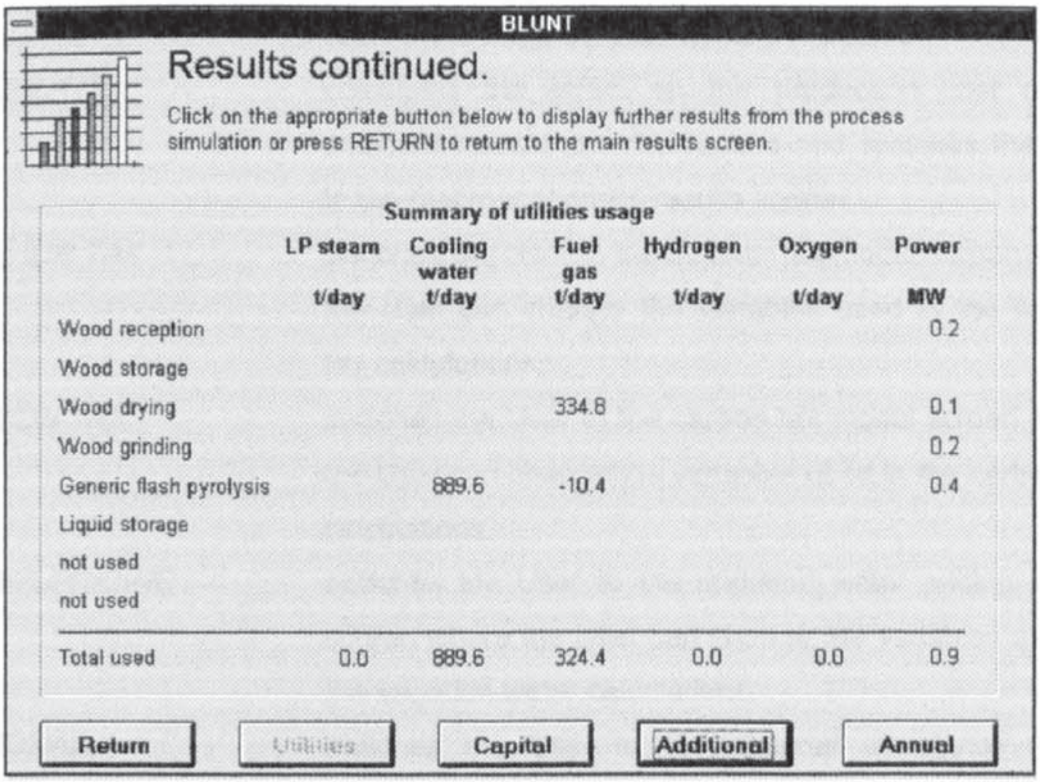
Results continued.

Click on the appropriate button below to display further results from the process simulation or press RETURN to return to the main results screen.

Capital cost breakdown

	Capital cost in thousands GBP	%age Total Capital Cost
Wood reception	1095.8	17.9
Wood storage	916.3	15.0
Wood drying	1610.6	26.3
Wood grinding	604.9	9.9
Generic flash pyrolysis	1701.9	27.8
Liquid storage	185.9	3.0
not used	0.0	0.0
not used	0.0	0.0
TOTAL CAPITAL COST	6115.4	99.9

Figure 4.19. The SRD screen displaying the breakdown of capital costs.



Results continued.

Click on the appropriate button below to display further results from the process simulation or press RETURN to return to the main results screen.

Summary of utilities usage

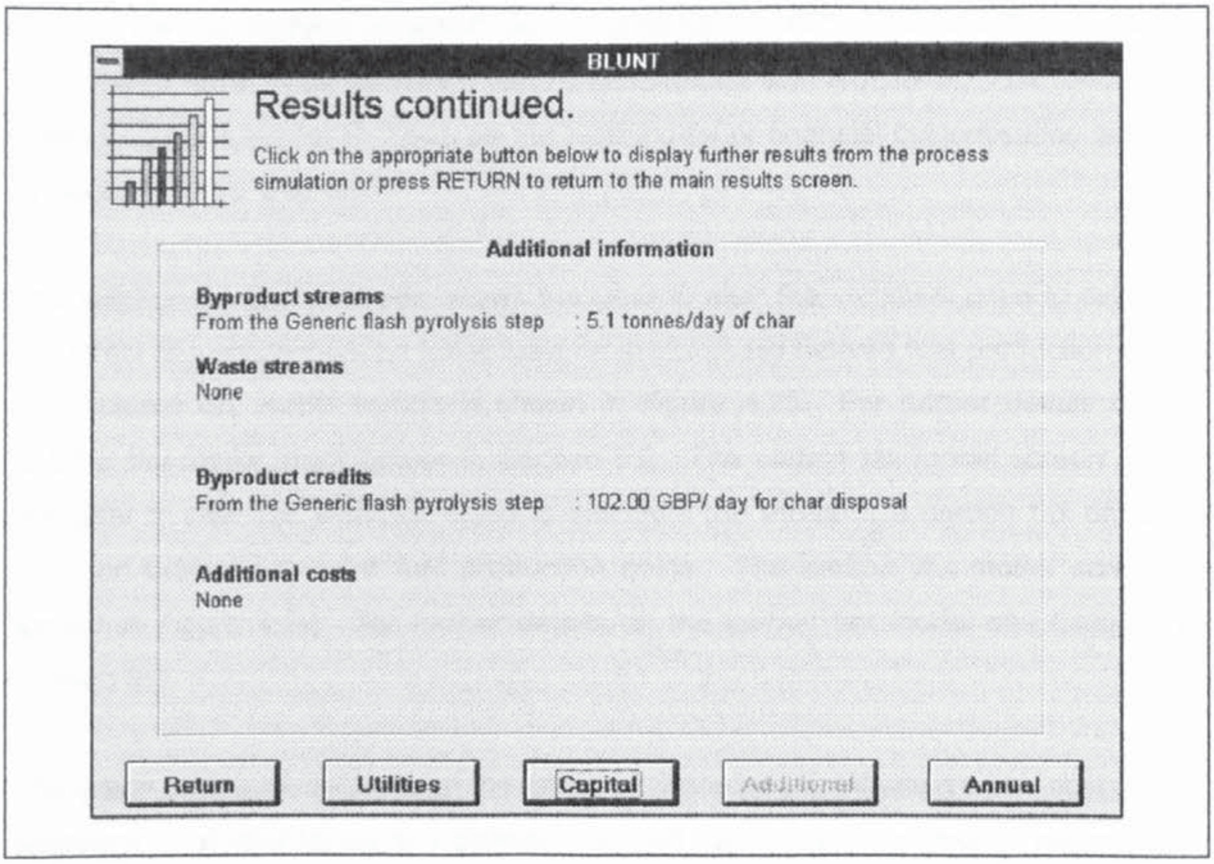
	LP steam t/day	Cooling water t/day	Fuel gas t/day	Hydrogen t/day	Oxygen t/day	Power MW
Wood reception						0.2
Wood storage						
Wood drying			334.8			0.1
Wood grinding						0.2
Generic flash pyrolysis		889.6	-10.4			0.4
Liquid storage						
not used						
not used						
Total used	0.0	889.6	324.4	0.0	0.0	0.9

Figure 4.20. The SRD screen displaying the breakdown of utilities requirements.

The macroeconomics screens are accessed by pressing *MACROECON* on the PRD screen. The macroeconomics selection (MES) screen, shown in Figure 4.22., allows the User to select the macroeconomic factor or combination of factors to be considered and to customise the variables used in the calculations. The MES screen consists of four frames each headed by the name of one of the macroeconomic factors modelled. Inside each frame are two radio buttons marked *YES* and *NO* respectively and a command button relating to the macroeconomic factor. Initially the radio buttons are set to *NO* i.e. the relevant macroeconomic factors are not selected. Clicking on a *YES* radio button activates that macroeconomic factor and activates the command button relating to that factor allowing access to the screen on which the default values of the macroeonomic factors are displayed.

The User can select any combination of macroeconomic factors for consideration e.g. a sulphur tax on fossil fuels or both carbon and sulphur taxes on fossil fuels. There are seven command buttons in total on the MES screen - *HELP*, *RETURN*, *CALCULATE*, *SULPHUR TAX*, *CARBON TAX*, *WATER PRICING* and *REGIONAL*. *HELP* has been explained earlier and the other buttons are described below:

RETURN	unloads the macroeconomics screens from memory and returns the program User to the PRD screen.
CALCULATE	calculates the effect on the production cost of the selected macroeconomic factors and switches the User to the macroeconomics results screen.
SULPHUR TAX	switches the User to the sulphur tax model screen where the User can change the variables used in the sulphur tax calculations.
CARBON TAX	switches the User to the carbon tax model screen where the User can change the variables used in the carbon tax calculations.
WATER PRICING	switches the User to the irrigation water pricing model screen where the User can change the variables used in the irrigation water calculations.
REGIONAL	switches the User to the regional assistance model screen where the User can change the variables used in the regional assistance calculations.



Results continued.

Click on the appropriate button below to display further results from the process simulation or press RETURN to return to the main results screen.

Additional information

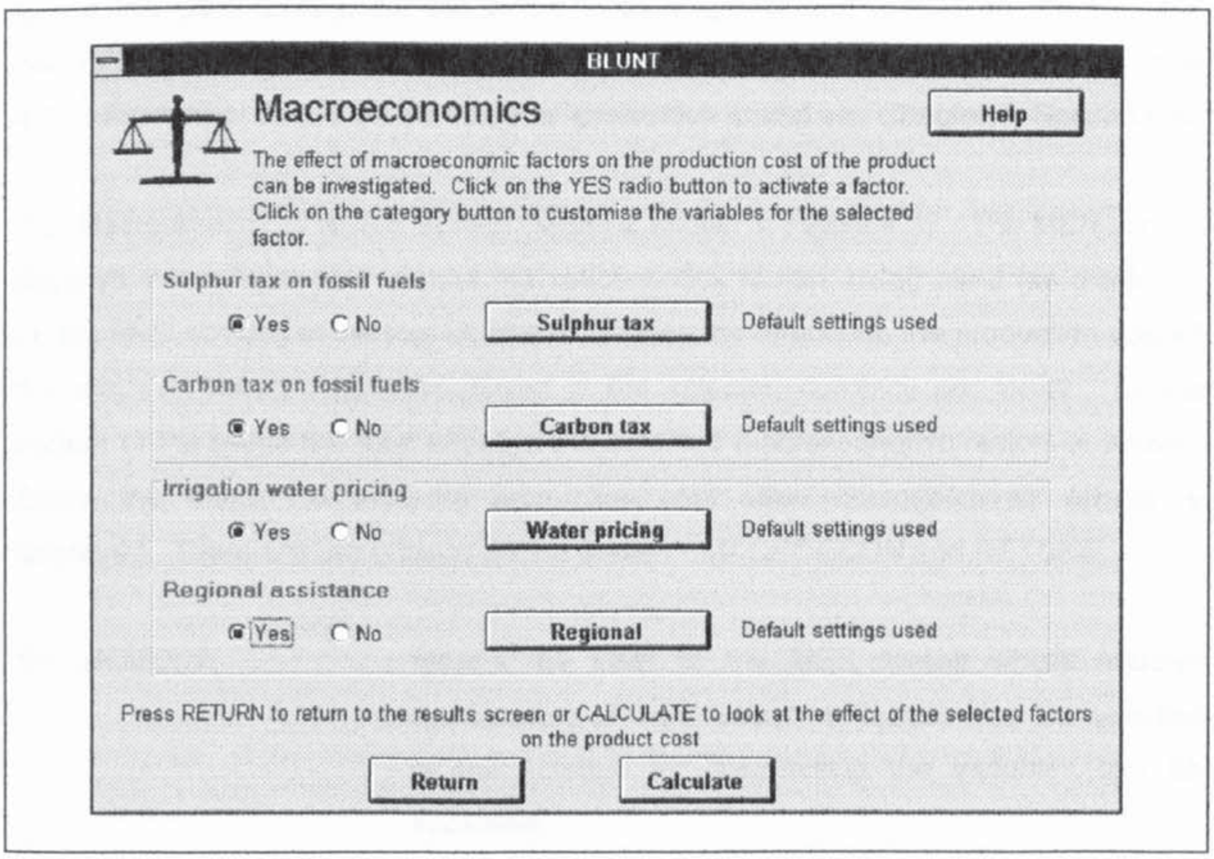
Byproduct streams
From the Generic flash pyrolysis step : 5.1 tonnes/day of char

Waste streams
None

Byproduct credits
From the Generic flash pyrolysis step : 102.00 GBP/ day for char disposal

Additional costs
None

Figure 4.21. The SRD screen displaying the breakdown of additional costs.



Macroeconomics

The effect of macroeconomic factors on the production cost of the product can be investigated. Click on the YES radio button to activate a factor. Click on the category button to customise the variables for the selected factor.

Sulphur tax on fossil fuels
● Yes ○ No Default settings used

Carbon tax on fossil fuels
● Yes ○ No Default settings used

Irrigation water pricing
● Yes ○ No Default settings used

Regional assistance
● Yes ○ No Default settings used

Press RETURN to return to the results screen or CALCULATE to look at the effect of the selected factors on the product cost

Figure 4.22. The macroeconomic selection screen.

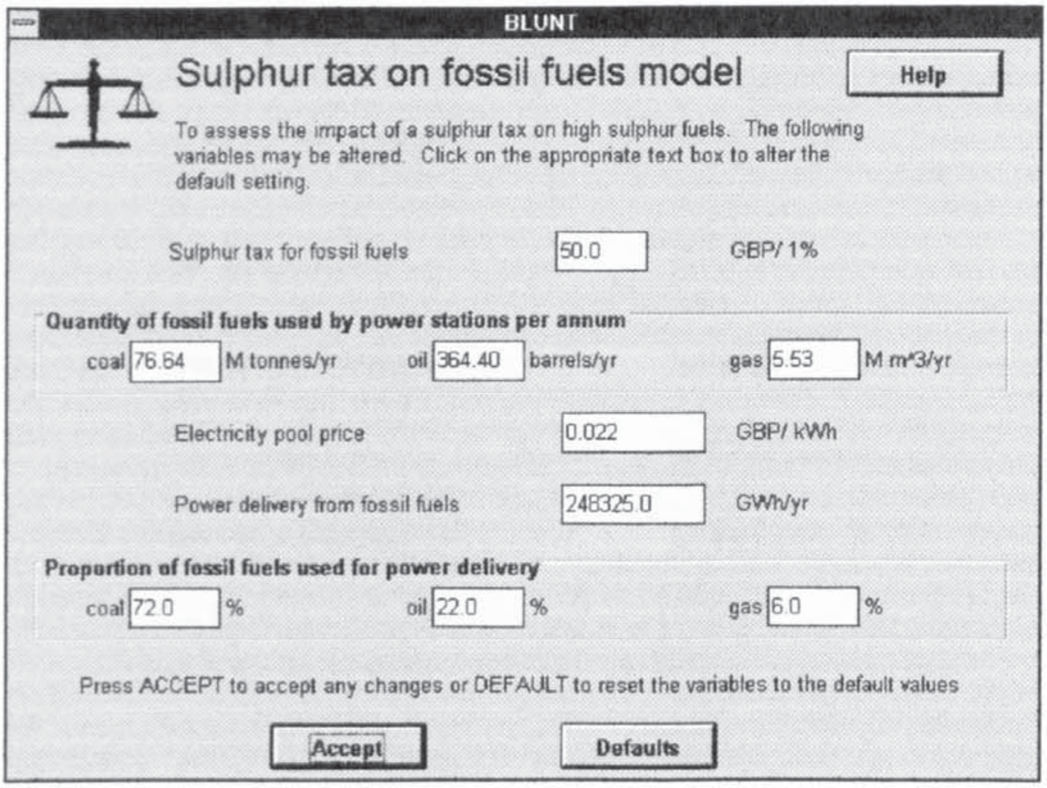
Each of the four screens on which the variables for the macroeconomic factors can be customised have three buttons - *HELP*, *DEFAULTS* and *ACCEPT*. The function of these buttons is similar to those on the step model or financial customisation screens in Sections 4.6.3. and 4.6.4.

The sulphur tax model screen allows the User to alter the variables used to calculate the effect of a sulphur tax on fossil fuels on the biomass derived fuel production costs. The sulphur tax model screen is shown in Figure 4.23. For further details on the sulphur tax model see Chapter 9, Section 9.2. The carbon tax model screen allows the User to alter the variables used to calculate the effect of a carbon tax on fossil fuels on biomass derived fuel production costs. The carbon tax model screen is shown in Figure 4.24. For further details on the carbon tax model see Chapter 9, Section 9.3.

The water pricing model looks at the effect of changing the irrigation water cost on the feedstock cost and hence the biomass derived fuel production cost. The irrigation water pricing screen is shown in Figure 4.25. For further details on the irrigation water pricing model see Chapter 9, Section 9.4. The regional assistance model screen allows the User to consider the effect of local grants and credits on the biomass conversion process viability. The regional assistance model screen is shown in Figure 4.26. For further details on the regional assistance model see Chapter 9, Section 9.5.

The Macroeconomics results screen (MER) is shown in Figure 4.27. The MER screen displays the product cost without macroeconomics factors being used (as displayed on the PRD screen) at the top. A frame contains the effects on the production cost of the macroeconomic factors employed in the selected currency per tonne. At the bottom of the frame the total effect of the selected macroeconomic factors is shown. Below the frame the adjusted production cost after macroeconomic effects is displayed. There are two buttons on the screen - *MACROECON* and *RETURN*.

MACROECON	returns the User to the MES screen where another combination of macroeconomic factors may be selected or the screens for customising the variables can be accessed.
RETURN	returns the User to the PRD screen.



Sulphur tax on fossil fuels model

To assess the impact of a sulphur tax on high sulphur fuels. The following variables may be altered. Click on the appropriate text box to alter the default setting.

Sulphur tax for fossil fuels: 50.0 GBP/ 1%

Quantity of fossil fuels used by power stations per annum

coal 76.64 M tonnes/yr; oil 364.40 barrels/yr; gas 5.53 M m^3/yr

Electricity pool price: 0.022 GBP/ kWh

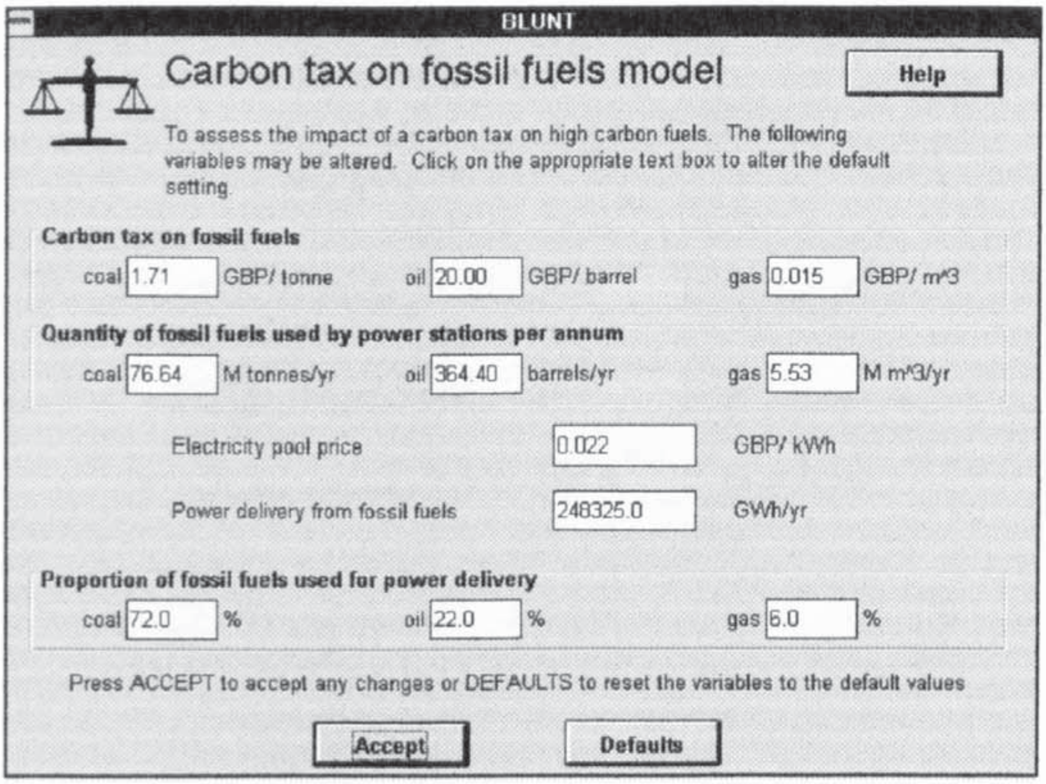
Power delivery from fossil fuels: 248325.0 GWh/yr

Proportion of fossil fuels used for power delivery

coal 72.0 %; oil 22.0 %; gas 6.0 %

Press ACCEPT to accept any changes or DEFAULT to reset the variables to the default values

Figure 4.23. The sulphur tax model screen.



Carbon tax on fossil fuels model

To assess the impact of a carbon tax on high carbon fuels. The following variables may be altered. Click on the appropriate text box to alter the default setting.

Carbon tax on fossil fuels

coal 1.71 GBP/ tonne; oil 20.00 GBP/ barrel; gas 0.015 GBP/ m^3

Quantity of fossil fuels used by power stations per annum

coal 76.64 M tonnes/yr; oil 364.40 barrels/yr; gas 5.53 M m^3/yr

Electricity pool price: 0.022 GBP/ kWh

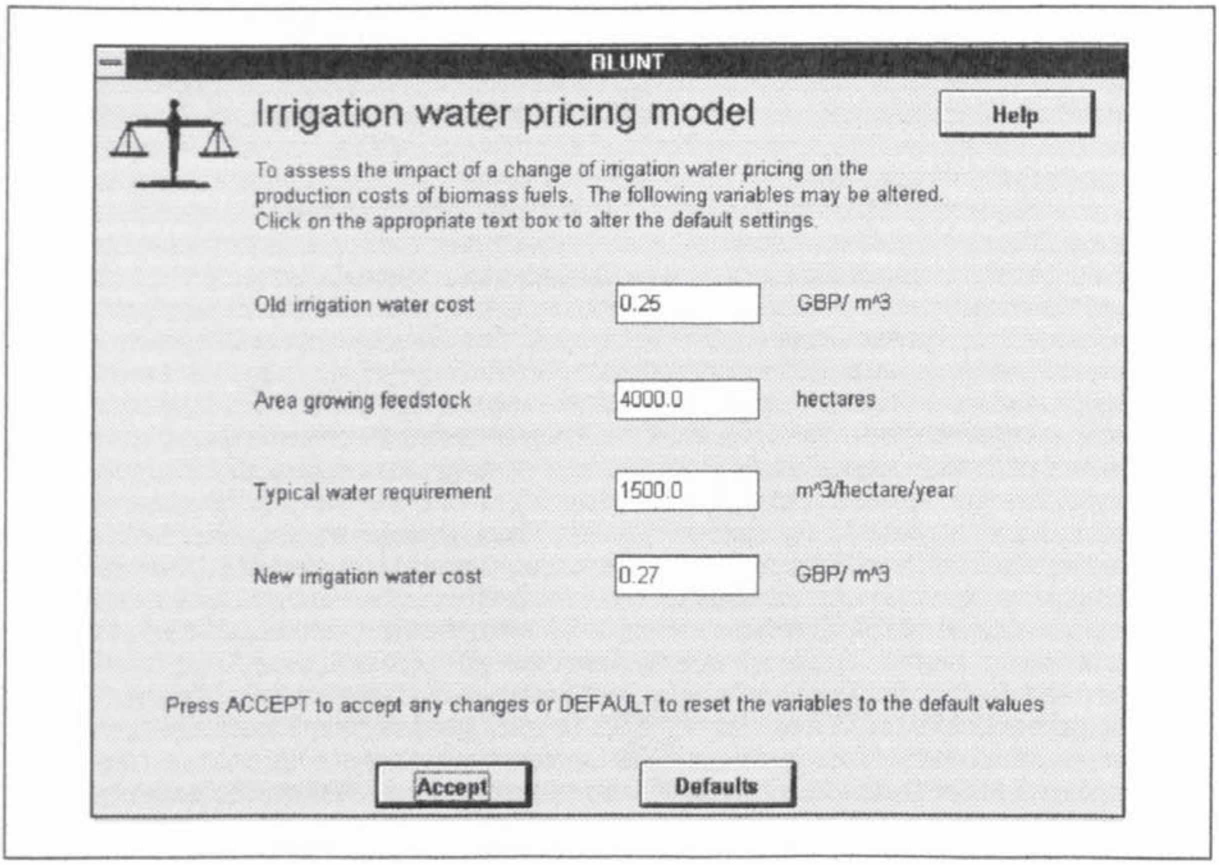
Power delivery from fossil fuels: 248325.0 GWh/yr

Proportion of fossil fuels used for power delivery

coal 72.0 %; oil 22.0 %; gas 6.0 %

Press ACCEPT to accept any changes or DEFAULTS to reset the variables to the default values

Figure 4.24. The carbon tax model screen.



Irrigation water pricing model

To assess the impact of a change of irrigation water pricing on the production costs of biomass fuels. The following variables may be altered. Click on the appropriate text box to alter the default settings.

Old irrigation water cost: 0.25 GBP/ m^3

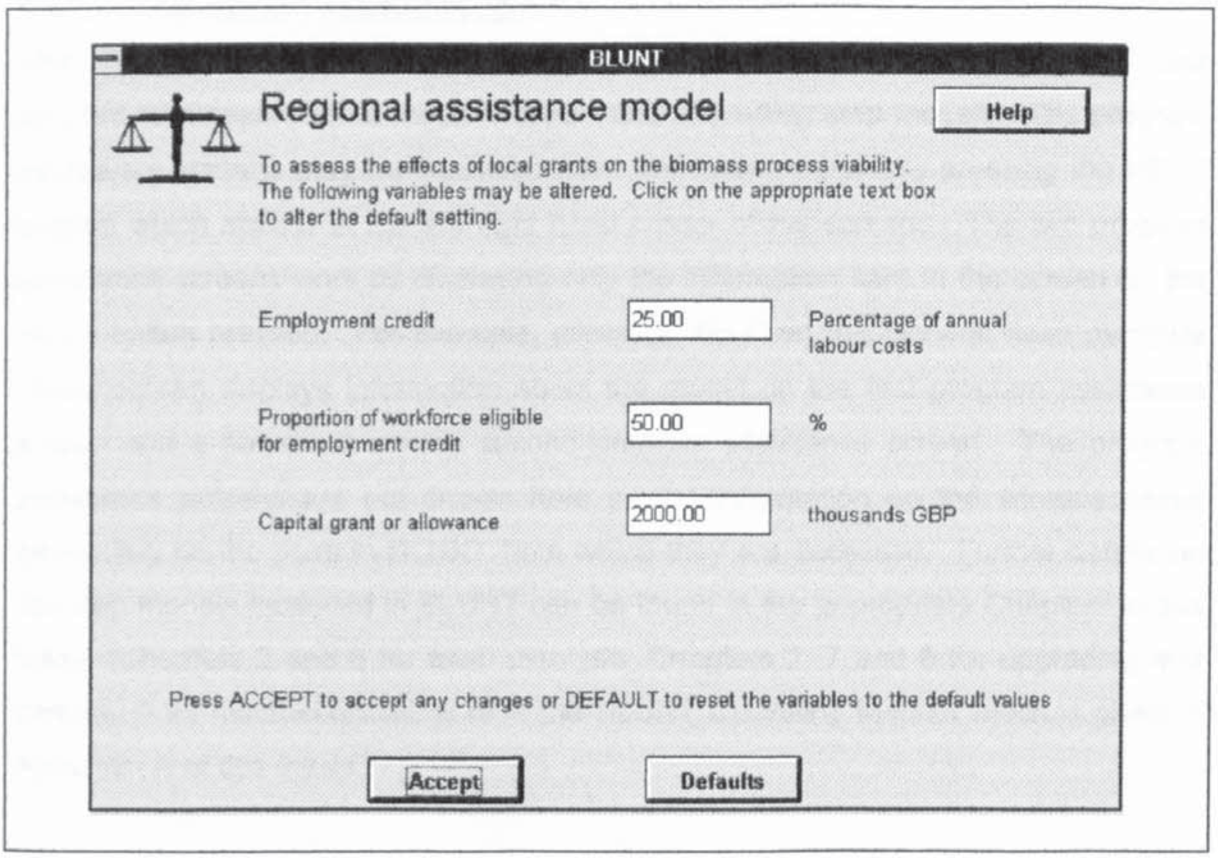
Area growing feedstock: 4000.0 hectares

Typical water requirement: 1500.0 m^3/hectare/year

New irrigation water cost: 0.27 GBP/ m^3

Press ACCEPT to accept any changes or DEFAULT to reset the variables to the default values

Figure 4.25. The water pricing model screen.



Regional assistance model

To assess the effects of local grants on the biomass process viability. The following variables may be altered. Click on the appropriate text box to alter the default setting.

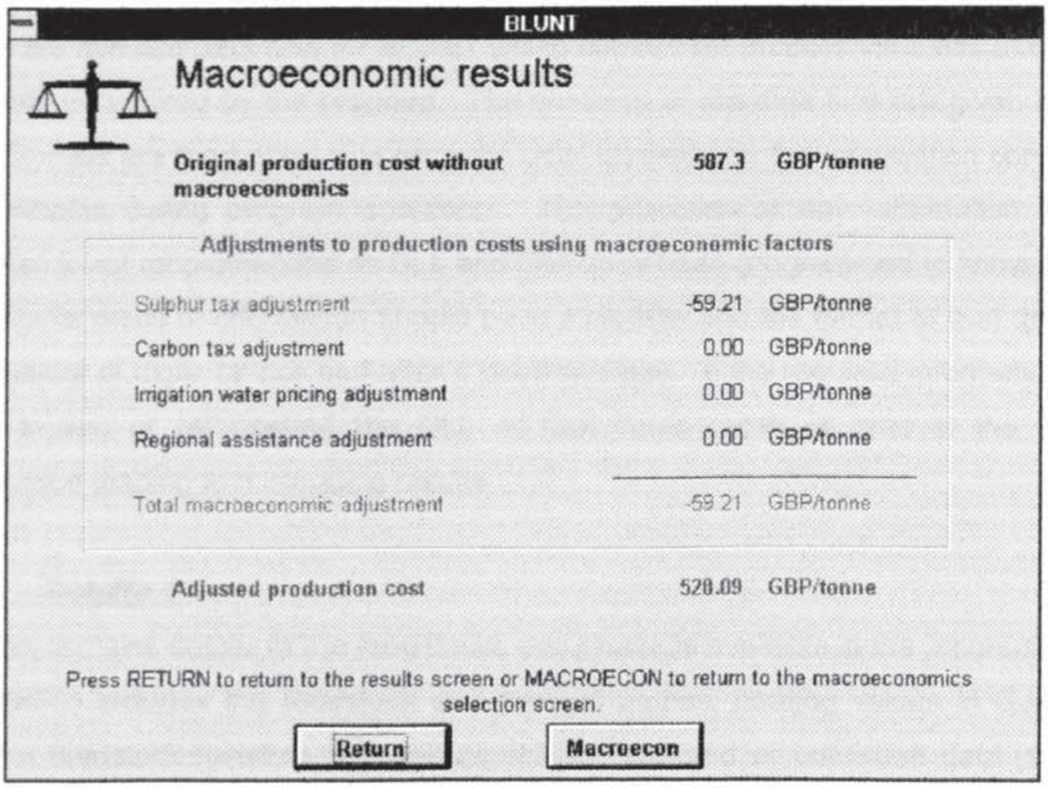
Employment credit: 25.00 Percentage of annual labour costs

Proportion of workforce eligible for employment credit: 50.00 %

Capital grant or allowance: 2000.00 thousands GBP

Press ACCEPT to accept any changes or DEFAULT to reset the variables to the default values

Figure 4.26. The regional assistance model screen.



BLUNT

Macroeconomic results

Original production cost without macroeconomics	**587.3**	**GBP/tonne**

Adjustments to production costs using macroeconomic factors

Sulphur tax adjustment	-59.21	GBP/tonne
Carbon tax adjustment	0.00	GBP/tonne
Irrigation water pricing adjustment	0.00	GBP/tonne
Regional assistance adjustment	0.00	GBP/tonne
Total macroeconomic adjustment	-59.21	GBP/tonne
Adjusted production cost	**528.09**	**GBP/tonne**

Press RETURN to return to the results screen or MACROECON to return to the macroeconomics selection screen.

Figure 4.27. The macroeconomics results screen.

4.6.7. The program assistance screens.

The program assistance screens display general information about BLUNT and detailed information about the conversion and upgrading step models. The program assistance screens may be accessed from the main program by pressing the *HELP* buttons which appear at the top right hand corner of the screens. The two program assistance screens work by displaying only the information sent to the screen by the *HELP* button pressed. For example, pressing *HELP* on the Generic flash pyrolysis model screen displays information about the model on the first program assistance screen and a flowsheet on the second program assistance screen. The program assistance screens are not shown here as the information on the screens varies depending on the point in BLUNT from which they are accessed. Further details on the step models modelled in BLUNT can be found in the appropriate Chapters in this thesis (Chapters 2 and 6 for flash pyrolysis, Chapters 2, 7 and 8 for upgrading and Chapter 9 for macroeconomics) or in the BLUNT Operating Manual which is given in Appendix A of this thesis.

4.7. The text datafiles.

There are five text data files for BLUNT which contain set process variables and other information required by the program. The contents of the data files are given below. The datafiles are read only i.e. a program User cannot alter the information contained in a datafile during program operation. The alteration of any information in the datafiles is not recommended as DLL and GUI have been programmed to know where a particular piece of information should be in a datafile and the format of that data i.e. the number of digits before and after a decimal place. If the required information has been moved or reformatted the DLL or GUI could crash or read in the wrong information leading to misleading results.

4.7.1. Datafile 1

This file contains details of the feedstocks and products available in the program. The information includes the feedstock and product names, heating values in GJ/tonne and the feedstock moisture and ash contents. Standard or published data [136] is utilised where it is known and the rest are estimates. The heating values of the crude pyrolysis oil and hydrotreated oil depend on the processing conditions and algorithms have been developed which are given in the relevant Chapters - Chapter 6 for the flash pyrolysis value and Chapter 7 for the hydrotreating value.

4.7.2. Datafile 2

This contains details of the process steps modelled in the program. The information includes a step number unique to a particular step model and the name of the process step model. The step number is a three digit number by which BLUNT recognises a particular step and is able to call and execute that step during program operation. The step numbers alter according to where they are in a processing route. For example, the step number for wood reception is 101 as this is the first step in any wood conversion route. The step numbers are utilised only by the program and are not on any of the screens visible to the program User.

4.7.3. Datafile 3

This contains details of the process routes modelled in the program. The information includes the combination of feedstocks and products that are permitted in the program and the number of potential processing routes available for any feed/product combination. For each processing route available in the program a route name and process step sequence composed of the numbers of the process steps from which a conversion route is composed are given. (For further details on the possible feed, product and conversion route combinations available in BLUNT see Section 10.1). For example, for the flash pyrolysis of straw to produce crude pyrolysis oil the entry in the data file and the interpretation made by the program are shown in Table 4.4.

Table 4.4.
Data read from text file and the interpretation of the data made by BLUNT.

Data read from text file	Interpretation
2 1 01	Straw to crude pyrolysis oil, one conversion route possible
837 Ensyn RTP pyrolysis	Conversion route number 837 using the Ensyn RTP pyrolysis process only
102 112 202 222 303 701 000 000 000 000 end	Process steps required: straw reception, straw storage, straw size reduction, straw fine chopping, Ensyn RTP pyrolysis and liquid product storage.

4.7.4. Datafile 4

This contains the names of all the currencies available for use in the program and information about each currency. The information contained in the datafile includes a three letter code of abbreviation and the exchange rate of the currency against Great Britain pounds. See Section 4.6.4. for details on the currencies available in BLUNT, their currency codes and the default values.

4.7.5. Datafile 5

This contains details of all the chemical components utilised in the program. The information includes component heating values in GJ/tonne, specific heat capacities in GJ/tonne K, enthalpies of condensation in GJ/t, molecular weights and enthalpies of combustion in GJ/tonne. All of the values for standard chemicals have been taken from published sources [136] and those for the non-standard chemicals e.g. crude pyrolysis oil have been taken from similar chemicals (See Chapter 6 concerning estimation of specific heat capacities for flash pyrolysis oil). Where the specific heat capacity of a chemical is related to temperature e.g. carbon dioxide, procedures have been written and incorporated into BLUNT incorporating formulae from published sources [136].

4.8. Operating BLUNT.

A copy of the BLUNT Operating Manual is available in Appendix A and a copy of BLUNT is included at the end of this thesis. It is not intended to detail the necessary steps in the installation and operation of BLUNT in this Section as this is included in the Operating Manual in Appendix A. The Operating Manual gives full details on installation and operation of BLUNT, including taking the program User through simulations for simple program operation, customising step model variables, customising the financial parameters and using macroeconomic factors.

4.9. Summary

BLUNT is a unique computer simulation package for biomass flash pyrolysis and upgrading which has been written as part of this project. It incorporates the latest information on the conversion processes, full results display and printouts and allows the User to consider the effects of macroeconomic factors on the selected conversion process. Program assistance screens have been written giving details on the program and include summaries of the conversion processes modelled and simplified flowsheets. BLUNT runs in the Windows environment and operates like a true Windows program allowing the use of a mouse or pointer to switch on buttons, open list boxes and move the focus to text boxes. The screens and controls have been kept as simple as possible to allow Users of all levels to use the program including those not too familiar with computer packages. An Operating Manual for BLUNT is included in Appendix A and a copy of BLUNT is included at the end of this thesis.

CHAPTER 5.
THE PRETREATMENT MODELS.

5.1. Introduction.

The flash pyrolysis processes detailed in Chapter 2 do not utilise a feedstock in it's raw condition for example, a whole tree which has been freshly felled. There are certain stages preceding the pyrolysis process and which could include a reduction in size to enable the wood to be moved, some sort of seasoning (e.g. storage and drying) and secondary size reduction (e.g. chipping). Most of these steps take place before the wood is delivered to a biomass conversion plant and will be not be covered here. This Chapter is concerned with the steps required to prepare or "pretreat" the biomass feedstock once it has been delivered to a conversion plant to enable it to be utilised in a flash pyrolysis process.

In Chapter 2 factors which affect the flash pyrolysis process and the derived products were identified and discussed. In particular the moisture content and particle size of the feedstock affect the energy demands of the pyrolysis process, the rate of heat transfer and the water content of the bio-oil product in addition to several other aspects of the flash pyrolysis process (See Section 2.3.1). Consequently, the pretreatment steps considered in this Chapter are designed to reduce the moisture content and particle size of a feedstock to enable it to be utilised in a pyrolysis plant.

Section 5.2. specifies the feedstocks utilised in BLUNT and for which the feed pretreatment models are written. Sections 5.3. and 5.4. look in detail at the two pretreatment routes modelled in BLUNT and Section 5.5. summarises the key information arising from this Chapter.

5.2. The biomass feedstocks.

5.2.1. The feedstocks modelled.

The feedstocks considered and modelled in BLUNT are wood, straw and sorghum bagasse. Wood has been included because it has been widely used in biomass conversion research [56, 57, 77, 116]. There has however, been growing interest lately in the utilisation of residues or wastes (hence low cost feedstocks) from farming, pulping and sugar extraction. Straw is increasingly becoming popular as a biomass feedstock [137], as it is renewable and widely available as a farming waste product. Sorghum [138] is a sugar containing crop grown in Southern Europe and the bagasse (the fibrous residue from sorghum after sugar extraction) is being investigated as another potential biomass feedstock. As sorghum bagasse has only recently been recognised as a possible biomass feedstock there is little information available in the literature on it's pretreatment or physical properties of sorghum bagasse or on those of sugar cane bagasse. However, from observation sorghum bagasse is similar to straw and hence the pretreatment steps required for straw and sorghum bagasse are considered to be the same.

5.2.2. Feed specification.

The feed specification includes the characteristics of the feed arriving at the biomass conversion facility and the constraints imposed by the flash pyrolysis conversion process. Both are considered in this Section.

The form in which the required feedstock arrives at the conversion facility will determine the pretreatment stages required before it can be pyrolysed. For example, for wood it is assumed that the feed comes prechipped to the pretreatment plant and that it had been stored previously in such a way as to be air dried. Table 5.1. gives the supplied specifications of the wood and straw feedstocks which are assumed in the modelling.

Table 5.1.
Required feedstock specifications for the pretreatment processes (raw feedstock to the pretreatment plant).

	Wood	**Straw**
Size range supplied	Chips	0.5 tonne bales
Moisture content (wt%, wet basis)	25 - 50	12.5
Ash content (wt%)	0.4	4.2

There are several physical properties of straw which have historically dictated the way in which it is stored and hence the conditions in which it is available for use as a process feedstock. Straw is traditionally stored in bales of which there are several different standard sizes of bales in use. After consultation with firms specialising in the supply and handling of straw, it was decided to concentrate on 0.5 tonne or

Hesston bales as these are the largest practicable bale size and are used most consistently for large capacity systems [139].

The final "product" from the pretreatment processes is the "feed" for the subsequent flash pyrolysis conversion process. The specifications for the flash pyrolysis processes as regards the physical size and moisture content of the feedstocks are given in Table 5.2. The specified moisture content of straw for the flash pyrolysis process in Table 5.2. is < 15wt% (wet basis) and since the raw moisture content of the straw as given in Table 5.1. falls within this range a drying step is not required for straw. This is discussed further in Section 5.4.

Table 5.2.
Required feedstock specifications for the flash pyrolysis processes, post pretreatment to reduce particle size and moisture content.

	Wood	**Straw**
Size range supplied (average) to 7mm	powdered or chipped 30 - 10 mm	chopped to
Moisture content (wt%, wet basis)	< 10	< 15

5.3. Wood pretreatment.

The design and specification of the wood pretreatment models is derived from previous work (AMBLE) [28, 131] as this was the best available source of information on the processes specified and there was sufficient information to allow adaptation of the processes to meet the specifications of the present project. Several modifications and additions were made to the wood pretreatment models before they were included in BLUNT which are given below.

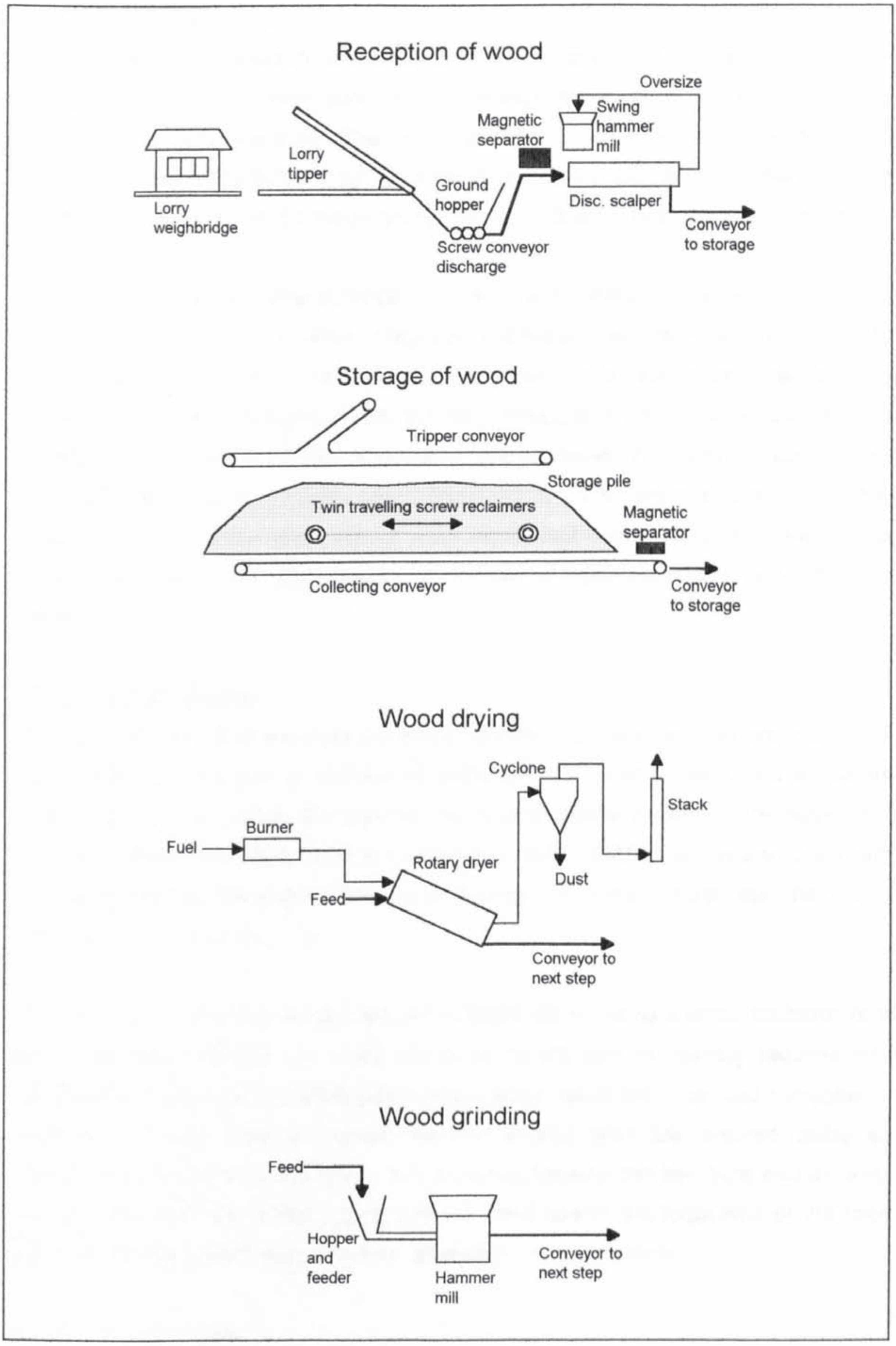
Alteration of the plant size limits and hence equipment specifications to allow 25-1000 d.a.f. tonnes/day of feedstock to be processed (from the 200-2000 d.a.f. tonnes/day utilised in AMBLE [28]). The pretreatment steps modelled in AMBLE included plant sizes from 200-2000 d.a.f. tonnes/day split into sections to allow for alterations in equipment specifications and numbers. For example, the storage step is split into three sections: less than 300 d.a.f. tonnes/day, 300-600 d.a.f. tonnes/day and 600-2000 d.a.f. tonnes/day allowing either multiple units of equipment to be used or alternative equipment to be specified. The wood pretreatment models in BLUNT work in a similar way although the plant size limits have been lowered to allow conversion facilities of 25 d.a.f. tonnes/day.

- An update of the equipment and hence capital costs to a 1993 basis. The equipment costs for the pretreatment steps in AMBLE were from 1990, these have been scaled using a cost index to give equivalent costs in 1993, the base year for the BLUNT program.

- A deletion of all units designed but not actually used in AMBLE for wood pretreatment e.g. screening. Although a screening step was specified in AMBLE for wood pretreatment it is not actually utilised as part of a flash pyrolysis conversion route and so was not included in BLUNT's wood pretreatment modelling.

- The design of a new drying model incorporating the feed inlet and outlet moisture contents as variables. AMBLE had two drying models drying to 7 and 15% moisture content (wet basis) respectively. However, drying has been highlighted as one of the most important stages in a pretreatment process [30], due to the effects of feed moisture content on the pyrolysis energy requirements (See Section 2.3.1. and Chapter 6) and it was decided to design a new drying model allowing variable feed moisture leaving the dryer. The new feed drying model is detailed in Section 5.3.3.

The main steps which have previously been identified (Section 5.1.) as important in the preparation of a wood feed for flash pyrolysis are drying and grinding to reduce the feed moisture content and size to the limits specified by the flash pyrolysis technology. In addition, steps are required to receive the wood into the conversion facility and store it until required for processing. These steps are designated as reception and storage and are described in Sections 5.3.1. and 5.3.2. respectively. The drying and grinding steps are described in Sections 5.3.3. and 5.3.4. respectively. Figure 5.1. gives an overview of the pretreatment processes and equipment specified in BLUNT for a wood feedstock. The reception, storage and grinding steps are just summarised as these have been detailed in the previous work [28]. All of the steps specified include conveying of the wood feedstock to the following step.



Reception of wood

Storage of wood

Wood drying

Wood grinding

Figure 5.1. Wood pretreatment processes.

5.3.1. Wood reception.

The first step in a pretreatment process is the reception of the feedstock at the conversion facility. The reception step is as specified previously [28] as this was the best source of data available. The equipment costs and power requirements for the model included in BLUNT are as determined previously with the exception that the capital cost calculated by the wood reception step has been updated to a 1993 basis.

The wood pretreatment step consists of several units: weighbridge, tipper, receiving hopper, conveyor, disc scalper, magnetic separator, rechipper and conveyor to storage (See Figure 5.1.). The number of equipment units used in the reception step alters with the feed throughput with the feed throughput range being split into two categories: less than 1020 wet tonnes/day (approximately 25 - 500 dry tonnes/day) and 1020 - 2040 wet tonnes/day (approximately 500 - 1000 dry tonnes/day). For the < 1020 wet tonnes/day range one of each equipment unit is specified while for the 1020 - 2040 wet tonnes/day range the number of each equipment unit utilised is doubled.

5.3.2. Wood storage.

The wood storage step provides a method of storing the feedstock in order to cover any disruptions in supply or changes in demand. The wood storage step is also as specified previously [28] as this was the best source of data available. The equipment costs and power requirements for the model included in BLUNT are as previously with the exception that the capital cost calculated by the wood storage step has been updated to a 1993 basis.

The feedstock is stored in large piles and is reclaimed either by a screw reclaimer or a shovel equipped vehicle. A wood pile of up to 612 wet tonnes/day requires one conical pile, using one front-end loader and a screw reclaimer. For feed throughputs between 612 and 1224 wet tonnes/day two conical piles are required, using an additional conveyor from reception, two screw reclaimers and two front end loaders. Above 1224 wet tonnes/day a parascrew is used due to the large size of the feed storage pile, associated with magnetic separator and a conveyor.

5.3.3. Wood drying.

The key parameter in the thermochemical conversion of wood is it's moisture content (See Chapter 2) as a lower moisture content generally means a lower energy demand on the pyrolysis unit and a lower moisture content, higher HHV liquid product (See Chapter 6 for further details).

Of the many types of commercial dryers available only a few are suitable for use in this application due to the nature of the feedstock being dried. Wood represents a considerable fire risk if over heated in the presence of excess oxygen and thus requires reasonable temperature control and a method of restricting the oxygen supply. In addition the large feed throughputs passing through the drier may cause uneven drying conditions. Rotary dryers are the most common solution to this problem as they offer good mixing and hence higher possibility of even drying, even at high temperatures. Exhaust gases may be recycled through the dryer, reducing the oxygen content of the atmosphere inside the dryer and reducing the risk of fires, as well as increasing the fuel economy.

The drying models included in the previous work [28, 131] had limitations - they did not permit a variable feed moisture content into the dryer to allow for fluctuations in feed quality or permit variable feed moisture contents exiting the dryer to meet the constraints on the flash pyrolysis models. Additionally, the original drying model [28] had only one size dryer specified which meant that for smaller capacity plants the drying unit was overspecified and the capital expenditure was increased unnecessarily.

It was decided to design a new drying model allowing the original wood moisture content and the wood moisture content ex-drying as variables, permitting the dryer size and duty, the subsequent duty on the pyrolyser unit and the heating value of the liquid product to be determined by the User of BLUNT giving a more flexible programming and design tool. The new model has two drying size ranges - one for feed throughputs between 25-100 d.a.f. tonnes/day and the other for 100 - 1000 d.a.f. tonnes/day. This allows the smaller capacity plant with feed input between 25-100 d.a.f. tonnes/day to utilise multiple rotary dryers of 25 tonnes/day capacity and reduces the over specification and hence increased capital expenditure of using larger units. Likewise, larger capacity plants in the 100-1000 d.a.f. tonnes/day range utilise multiple rotary dryers of 100 tonnes/day capacity.

The actual number of dryers required for any given capacity is dependent on the feed intake, the feed inlet moisture content (determined by the program User) and the required feed outlet moisture content (determined by the program User) as shown by equation {1}. Equation {1} was generated by from analysing similar equations in the models for drying to 7% and 15% [28].

No. of Dryers
$$TRUNC\left[\frac{Db}{\left(\frac{m_{out}}{m_{in}}\right)\times 100\times a}\right]+1 \qquad \{1\}$$

where Db is the dry feed throughput (d.a.f. tonnes/day);

m_{out} is the feed moisture content leaving the dryer (wt% dry basis);

m_{in} is the feed moisture content entering the dryer (wt% wet basis);

$TRUNC$ is a function which gives the truncated value of the answer i.e. it rounds the real number answer of the calculation to the nearest integer;

a is an empirical unit size variable equal to 2 for the 25-100 d.a.f. tonnes/day range and 12 for the 100-1000 d.a.f. tonnes/day range.

The capital cost of the drying unit in 1993 Great Britain pounds is determined by equation {2}.

Capital cost (wood drying)
$$= 3 \times Dc \times Nd \qquad \{2\}$$

where Nd is the number of dryer units specified by equation {1}.

Dc is the cost of a dryer unit on a 1993 basis);

The utilities required for the drying step are power and fuel gas - the fuel gas requirements are determined by the amount of water removed by the dryers as per the previous work [28]. The power requirements are determined by the number of dryer units utilised and the power requirements of each unit as given in equation {3}.

Power requirement (MW)
$$= \frac{Dp \times Nd}{1000} \qquad \{4\}$$

where Dp is the power required by a dryer unit

5.3.4. Wood grinding.

The flash pyrolysis conversion process requires a small feed particle size which can generally only be achieved by grinding. The wood grinding step is as specified in previously [28, 131] as this was the best available source of data. The equipment costs and power requirements for the model included in BLUNT are as determined previously with the exception that the capital cost calculated by the wood grinding step has been updated to a 1993 basis.

The equipment specified in the grinding model consists of a feed bin and feeder, a grinder and a conveyor and is based on grinding down to a 3mm particle size with multiple units for high feed throughputs. This quickly imposes capital cost and energy usage penalties as there is no economy of scale incorporated in the calculation. The number of grinders specified for a particular feed throughput is calculated by the algorithm given in equation {4} below.

$$\text{Number of grinders} = TRUNC\left[\frac{Db}{Gs}\right]+1 \qquad \{4\}$$

where Db is the dry feed throughput (d.a.f. tonnes/day);

Gs is the grinding unit size specified which has a capacity of 120 d.a.f. tonnes/day of feedstock.

5.3.5. Comments.

The wood pretreatment processes required in a biomass conversion facility are reception, storage, drying and grinding. The reception, storage and grinding models included in BLUNT are based on the previous work [28] with the equipment costs updated to a 1993 basis. The lower limits on each model were lowered from the 200 tonnes/day used previously to the 25 tonnes/day required for BLUNT. However, the equipment specifications for the lower limits was not adjusted to allow for minimal equipment and high labour utilised at small conversion plants due to lack of data. This confers considerable uncertainty on the equipment specifications and hence capital costs of these steps at low feed inputs where the costs are undoubtedly too high. The wood drying model, being a new model, does allow for low feed inputs and the resultant capital costs are probably more robust.

5.4. Straw / bagasse pretreatment.

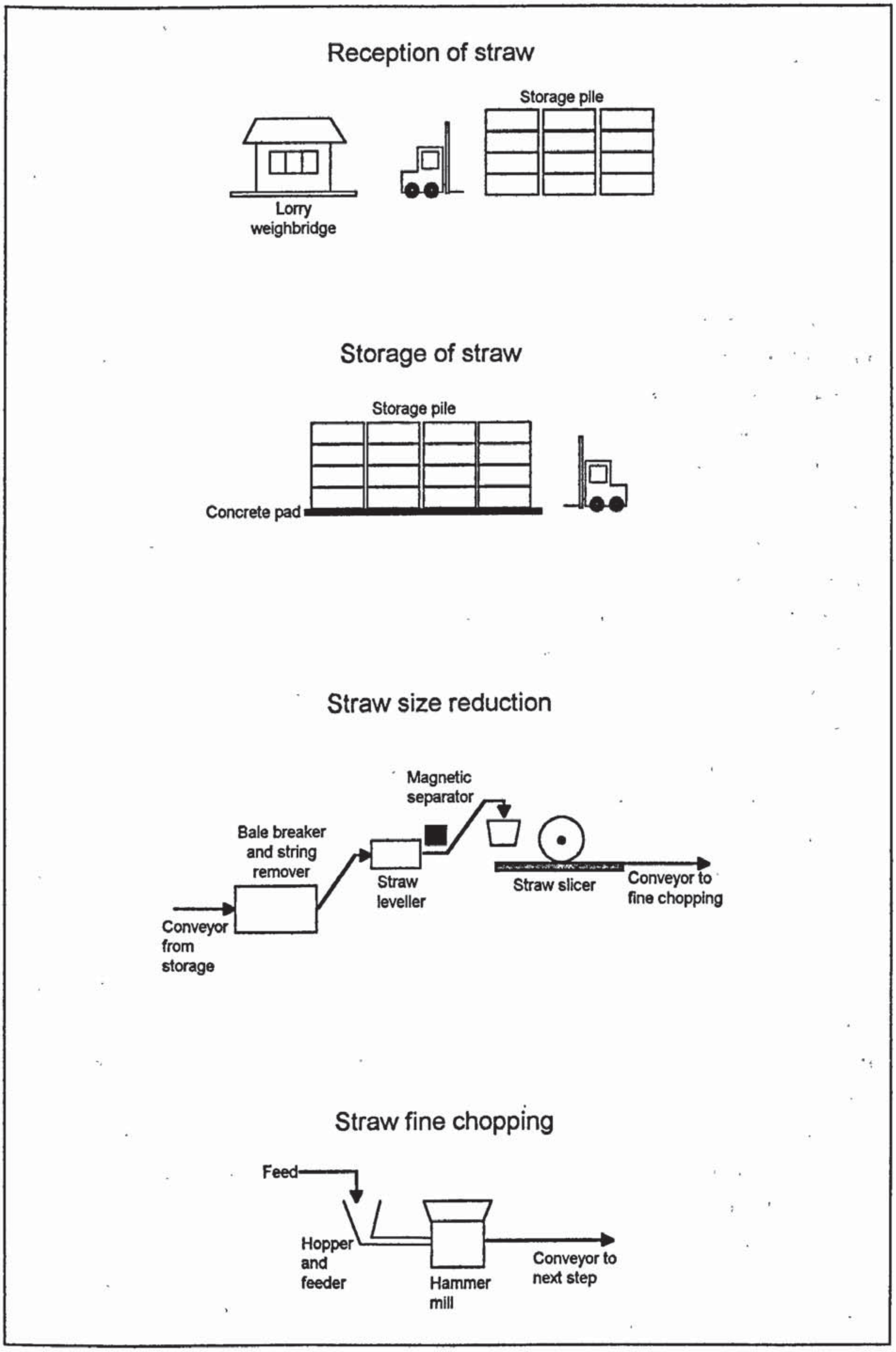
The pretreatment models and equipment described in this Section were designed for the pretreatment of straw. However, as mentioned in Section 5.2., since there is no data available on the pretreatment of sorghum bagasse for flash pyrolysis processes these models also represent sorghum bagasse pretreatment.

One important difference in the pretreatment of straw to that of wood (apart from those dealing with the difference in size / shape of the feedstock) is that due to the properties of straw no attempt is made to dry it. The low bulk density of straw makes the use of rotary dryers difficult due to straw being blown from the dryer. The same characteristic produces a very large volumetric throughput even at relatively small mass throughputs. Straw also represents a considerable fire risk (more so than wood) due to the presence of fine particles and thus any drying step would have to be carefully controlled to prevent hot spots forming and subsequent ignition leading to potentially very damaging fires. As straw is supplied with a relatively low moisture content (See Table 5.1.) and as the cost of drying is very high relative to the rest of the process [30], it was decided not to specify the need for drying. The savings made here should be more than sufficient to offset any increase in pyrolyser cost incurred due to the higher moisture content.

The straw pretreatment steps required and described in the following subsections are reception of straw at the conversion plant (Section 5.4.1.), storage of straw until required (Section 5.4.2.), primary size reduction (5.4.3.) and fine chopping (5.4.4.). The models have been derived using information provided from research institutes dealing with straw [139 , 140] and the manufacturers of straw handling equipment [141 , 142]. The straw pretreatment steps modelled are shown in Figure 5.2.

5.4.1. Straw reception.

The straw reception step receives and weighs the baled straw and unloads it from the transport vehicles. The straw is delivered in lorry loads of 39 bales, approximating to 20 tonnes. The lorries are received through a weighbridge before being unloaded using rough terrain telescopic handling machines. Unloading three bales at a time, one machine can unload a lorry in ten minutes. The unloaded straw is stacked for immediate collection by telescopic handling machines from the storage step.


Reception of straw

Storage of straw

Straw size reduction

Straw fine chopping

Figure 5.2. Straw pretreatment processes.

The reception model splits the plant feed limits of 25-1000 d.a.f. tonnes/day into two capacity ranges, 25-500 d.a.f. tonnes/day and 500-1000 d.a.f. tonnes/day. The equipment specified is the same for each capacity range with the number of equipment units doubled for the 500-1000 d.a.f. tonnes/day range. The equipment costs for each unit utilised are summed and multiplied by a factor to give the total plant cost.

5.4.2. Straw storage.

The straw storage step is designed to store sufficient quantities of straw to overcome any short term supply difficulties and to supply this straw to the subsequent pretreatment steps when necessary. Although straw represents a significant fire risk if stored in any quantity, the nature of the pretreatment and conversion processes modelled demand that a supply is maintained in storage. The model is based on an storing 2.5 days worth of straw - the time period was selected to minimise the potential loss should a fire occur and to prevent degradation of the straw during extended storage. The straw is stored on a concrete hard standing [139] which is the safest storage method since a straw fire cannot be extinguished and must be left to burn itself out. If the straw is stored under cover and ignites, the cover or building would be lost in the fire. The straw is stored in two stacks of sufficient size to each represent one and a quarter days processing supply. The use of such large stacks minimises the amount of straw lost due to inclement weather as experience has demonstrated that only the outer layer of straw (of depth roughly equivalent to the depth of rainfall) will be degraded to below the required specifications [139]. The straw is stacked and reclaimed by rough terrain telescopic loading machines as for the reception step. Two forklifts are specified for the straw storage step and a concrete standing.

5.4.3. Straw size reduction.

The size reduction step is utilised to chop the straw into lengths of approximately 30mm for subsequent reduction to smaller size particles. Before being chopped, the straw must first be teased from the bales and any impurities e.g. stones are extracted. There are several possible technologies available for chopping the straw, each with it's own merits. The properties of straw make chopping it an energy intensive operation and thus expensive to run, a factor which must be considered when specifying the equipment for this duty. The model has been specified with a Cormall straw line [143 , 144] which includes all the equipment for separating the bales and screening the product. The chopping operation is carried out by a rotary slicer which offers good slicer performance and low maintenance costs coupled with low energy consumption. The technology is not necessarily the cheapest available, however it is intended to represent the type of technology which is currently starting to become available offering significant advantages over more traditional methods of processing [143]. Each straw line can process up to 10 tonnes/hour and above this capacity multiple units are specified. The equipment specified for the size reduction model equates to an intake conveyor, a Hesston bale breaker, a string remover, a buffer section, an inclined conveyor, a straw levelling wheel, a magnet, a feed conveyor, a feeding device, a straw slicer, a straw blower and sundry pipe work. The number of equipment units required is given by equation {5} which is based on a 240 d.a.f. tonnes/day plant.

$$\text{Number of units} \quad = \quad TRUNC\left[\frac{Db}{240}\right]+1 \qquad \{5\}$$

where Db is the dry feed throughput (d.a.f. tonnes/day);

5.4.4. Straw fine chopping.

The fine chopping step is utilised to chop the 30mm straw lengths to the 10mm required by the flash pyrolysis processes. Machines specifically designed to fine-chop straw by various methods are available, with very little to choose between them. However, all the machinery suitable for industrial capacities require the straw to be pre-chopped to approximately 50mm lengths or smaller, for acceptable performance. The equipment specified for the fine chopping model is a hammer mill, typical of the range of equipment currently available. However, a current cost for the equipment is not available and the cost for the equipment is based on the cost of comparable technology specified in the earlier work [28]. Each machine has a capacity of up to 10 tonnes/hour with multiple units specified for higher throughputs. The total equipment specification for fine chopping is conveyors, a feed bin, a feeder, a hammer mill, a conveyor, a straw blower and pipe work. The number of equipment units required is given by equation {6} based on 120 d.a.f. tonnes/day plant.

$$\text{Number of units} \quad = \quad TRUNC\left[\frac{Db}{120}\right]+1 \qquad \{6\}$$

where Db is the dry feed throughput (d.a.f. tonnes/day);

5.4.5. Comments.

The straw pretreatment steps include reception, storage, size reduction and fine chopping. The models are based on conventional straw processing e.g. for agriculture with high feed throughputs as there is a sparcity of data concerning the pretreatment of straw for thermochemical conversion. Each model was based on a single equipment specification (for example, the straw fine chopping model was based on a 120 d.a.f. tonnes/day process) and the plant feed limits for BLUNT were met by the use of multiple units. Consequently the uncertainty attached to the straw pretreatment models is quite high particularly at the lower end of the plant limits where equipment is almost certainly overspecified. However, the models do give a guide to the processing required and allow the comparison of straw with wood as a biomass feedstock.

5.5. Summary.

The models described in this Chapter prepare the biomass feedstock for the subsequent flash pyrolysis process. This preparation takes the form of drying and grinding to reduce the moisture content and particle size respectively of a wood feedstock to meet the specifications in Table 5.2. The drying model allows the feed moisture content ex-drying to be varied which will enable the specific effects of feed moisture content on the pyrolysis energy balance and bio-oil water content to be determined (See Chapter 6 for further details). The required pretreatment deemed necessary for a straw or bagasse feedstock is size reduction and fine chopping to meet the specifications in Table 5.2. Drying is not considered advisable for straw and so the effects of feed moisture content cannot be determined for straw.

A comparison of the costs involved in the pretreatment processes for wood, straw and sorghum bagasse has not been carried out here as the steps involved are different. However, the choice of feedstock (and hence pretreatment process) related to the production costs of crude pyrolysis oil has been included in Chapter 10

CHAPTER 6.

FLASH PYROLYSIS MODELLING.

6.1. Introduction.

In Chapter 2 the products and main process variables for the flash pyrolysis of biomass were identified and three flash pyrolysis processes which are at either demonstration or commercial level were detailed (Sections 2.4. - 2.6.). The process variables identified were the vapour residence time, pyrolysis temperature, feed water content, feed composition (CHO) and feed particle size and this Chapter is concerned with the modelling the effects of these variables on flash pyrolysis processes.

There are four flash pyrolysis models in BLUNT representing a "generic" process and the Ensyn [19] (detailed in Section 2.4.), NREL [17] (detailed in Section 2.6.), and Union Fenosa [73] (detailed in Section 2.5.), processes. The "generic" model is based on a fluidised bed flash pyrolysis system and has been included because it demonstrates the effects of pyrolysis temperature, residence time and feed moisture content on the product yields and energy balance. The models described in this Chapter are not detailed chemical or mechanical engineering designs but are of the mass and energy balances from which are derived the product yields and utilities usage.

The background and process description for the generic flash pyrolysis model is given in Section 6.2. Section 6.3. details the development of the mass balance for the model and Section 6.4. the energy balance. The modelling of feed composition and vapour residence time are discussed in Section 6.5. and the derivation of the capital cost and utility requirements are given in Section 6.6. Section 6.7. describes the modelling of the Ensyn, NREL and Union Fenosa flash pyrolysis process. The main points arising from this Chapter are summarised in Section 6.8.

6.2. The generic flash pyrolysis model.

6.2.1. Background.

The other flash pyrolysis processes which have been included in BLUNT (the Ensyn RTP, NREL vortex and Union Fenosa processes) are based on existing flash pyrolysis processes in which the process conditions have been optimised to increase liquid product yields. The generic flash pyrolysis model has been designed to demonstrate the effects of the pyrolysis process variables highlighted in Section 2.3. on product yields, product quality (e.g. heating value and moisture content) and energy requirements. Although the model utilises data from the University of Waterloo fluidised bed system [6, 14] it does not represent this process but is a "hybrid" model specifically designed to give a greater understanding of the flash pyrolysis process.

The generic flash pyrolysis model uses empirical relationships [35] to determine the pyrolysis product yields and hence the mass balance. The empirical relationships are based on published results from the University of Waterloo [6, 41, 57, 58, 116] because their process is representative of flash pyrolysis and they have published consistent data. In addition, the fluidised bed technology used in the Waterloo process [14, 58, 145] is known to be scaleable, to give results comparable to other systems and is a reasonable starting point for the generic flash pyrolysis model. The generic flash pyrolysis model works within the limits given below:

- Pyrolysis temperatures between 400 and 650°C. The pyrolysis temperature is a process and is defined by the program User between the set limits.

- The yields of organics, reaction water, char and off-gas are determined by empirical relationships based on the pyrolysis temperature, the feed moisture content and the feed input to the plant. The yields are model variables in that they cannot be directly altered by the program User but may be altered indirectly when the pyrolysis temperature, the feed moisture content and the feed input to the plant are defined by the User.

- The quantity of water in the final product is determined by the yield of reaction water and the moisture content of the feedstock. The water content of the bio-oil product is a model variable as it is determined only indirectly by the User when the pyrolysis temperature and the feed moisture content are defined.

- The quantity of ash/char remaining in the bio-oil product is determined by the char filter efficiency. The char filter efficiency is a process variable and can be defined by the program User. The default value for the filter efficiency is 90% i.e. 10% of the total char produced remains in the liquid product.

- The energy requirement of the pyrolyser is met by burning a combination of char, off-gas and imported fuel gas as explained in Section 6.2.4. This can be directly influenced by the program User who can select the fuel source or combination of sources to be used and hence is a process variable.

- The product oil and off-gas are cooled by a direct quench system, as favoured by Ensyn and Egemin [25] with cooled stored product oil from previous runs.

A description of the process around which the generic model is based is given in Section 6.2.2. The process starts with a prepared wood feedstock which is within the limits given in Table 5.2. in Chapter 5 and ends with either cooled, condensed pyrolysis oil for storage, power production or hydrotreatment or pyrolysis vapours for upgrading by zeolite cracking.

6.2.2. Process description.

In the generic process the prepared wood feedstock passes from a hopper and through a feed screw into the fluidised sand bed reactor. The pyrolysis vapour products leave the reactor and pass through a cyclone where the char is removed for storage. At this point there are two possible directions for the vapour products to go: direct to a zeolite cracking process for upgrading (See Section 3.2.) or to be cooled and condensed to give crude pyrolysis oil which may be stored for direct use, power production or upgrading by hydrotreatment (See Section 3.3.).

If the product is to be crude pyrolysis oil the hot vapour stream passes from the cyclone into a quench column where it is cooled using stored product from previous runs. Trials have demonstrated that a system of multiple quench units are required for the condensation of crude pyrolysis oil [74] and the single unit shown in Figure 6.1. should be taken as a representation of the quench system as a whole. The condensed pyrolysis oil product is removed for intermittent storage. Some of the recovered off-gas is preheated in a furnace before being utilised as the fluidising gas in the pyrolyis reactor. The furnace is fuelled by some combination of char, off-gas or imported fuel gas. Any surplus char is exported as a byproduct and any surplus off-gas is exported as fuel gas to the utilities. The flowsheet for the generic flash pyrolysis model incorporating a condensation process is shown in Figure 6.1.

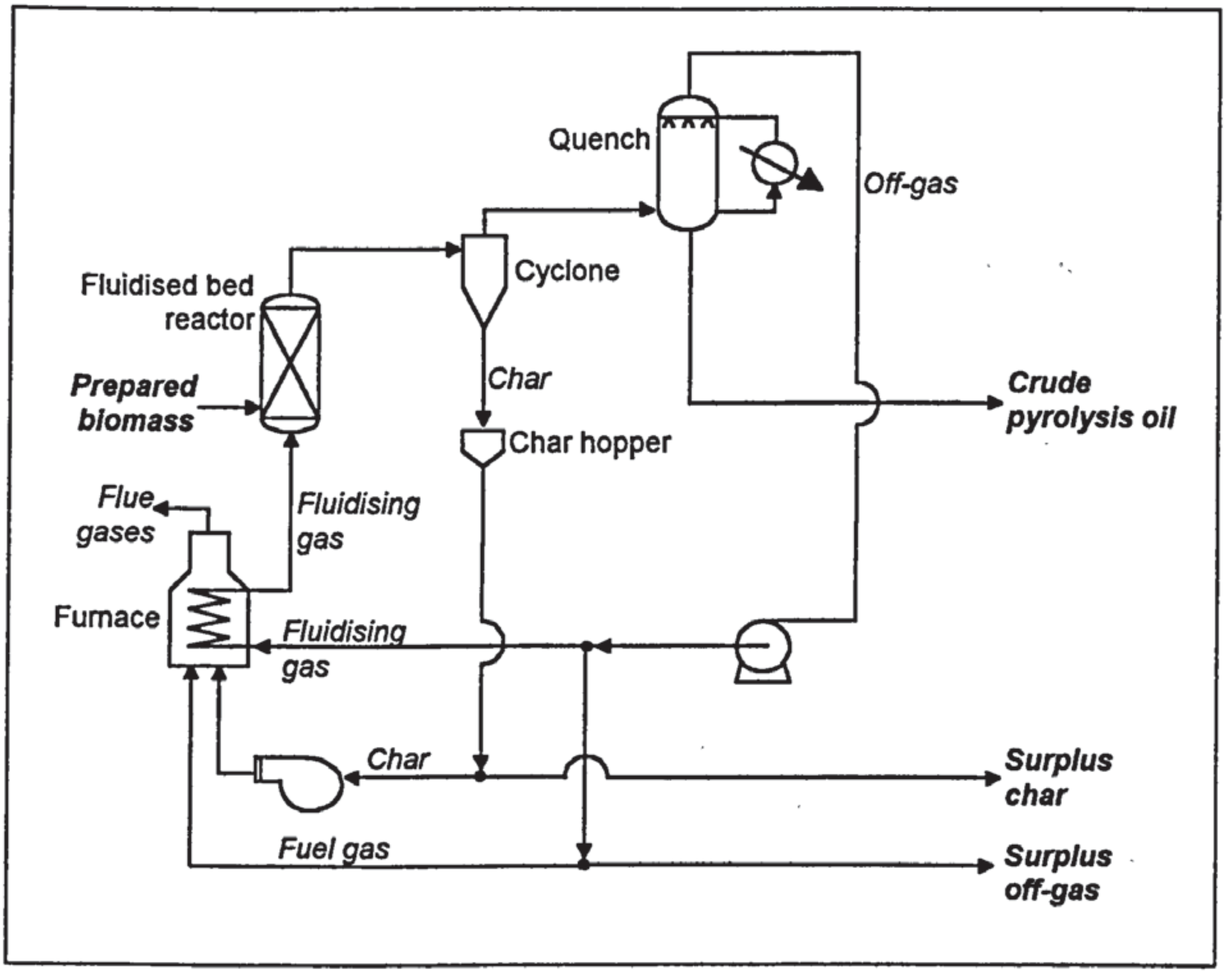


Figure 6.1. Flowsheet for the generic flash pyrolysis model incorporating a liquid product condensation and recovery step.

If a condensation process is not to utilised and zeolite cracking is the next stage in the overall conversion route, the primary vapours exit the char cyclone and pass directly to a cracking reactor. See Chapter 8 for further details on the zeolite cracking of crude pyrolysis vapours. In this case off-gas is not produced and hence not exported as fuel gas and the energy demand for the process must be met by a combination of char and imported fuel gas. A flowsheet for this process is not included as it is essentially the same as Figure 6.1. but without the condensation step.

6.3. The mass balance for the generic flash pyrolysis model.

6.3.1. Introduction.

As mentioned earlier the mass balance for the generic flash pyrolysis model is based on empirical relationships representing the organic, reaction water, char and off-gas yields. The empirical relationships were derived from the Waterloo data [58] shown in Table 6.1. and illustrated in Figure 2.3. (shown in Section 2.3.4.). Additional sets of data were not included because of the differences in feed type, moisture content and processing conditions.

Table 6.1.
Waterloo raw data used to derive the empirical relationships [58].
(Pilot plant results on IEA standard Poplar-Aspen, 6.25% moisture content)

Temperature (°C)	**425**	**465**	**500**	**500**	**541**	**541**	**625**
Yields (% wood mf)							
Organic liquids	55.88	67.17	65.76	71.15	63.65	62.05	40.27
Reaction water	3.8	5.5	9.3	6.6	7.4	7.7	4.1
Char	30.51	18.88	12.15	11.2	8.99	9.72	7.81
Gas	5.95	8.53	12.45	12.07	21.22	19.11	36.65

Although the data shown in Table 6.1. shows all the correct temperature versus yield trends as described in Chapter 2 the water yields have subsequently been shown to be low. As a rule of thumb the reaction water yield at 500°C should be in the region of 11% [59] whereas the data in Table 6.1. gives a yield between 6.6 and 9.3%. The low reaction water yield could have been caused by inaccurate determination of the bio-oil's water content by Karl-Fischer analysis [59]. Apart from this the data shown in Table 6.1. is consistent with the product yields obtained from the Waterloo process and can be utilised as a representation of the process [59].

6.3.2. Development of the empirical relationships.

The Waterloo data given in Table 6.1. was plotted using a graphical package and the plots were curve fitted using the same graphical package. The equations resulting from the curve fits give the empirical relations for the reaction water, char, organic and off-gas yields based on the reaction temperature and are shown below in equations {1}, {2}, {3} and {4}. The range of accuracy for the curve fits is for temperatures between 400 and 650°C only as the original data in Table 6.1. gave product yields for temperatures between 425 and 625°C [58].

$$Y_{org} = 1.790\, T_{pyr} - 0.00179\, T_{pyr}^2 - 380.156 \qquad r^2 = 0.953 \qquad \{1\}$$

$$Y_{gas} = 0.112 \times 10^{0.00409\, Tpyr} \qquad r^2 = 0.983 \qquad \{2\}$$

$$Y_{char} = 409.844 \times 10^{-0.00294\, Tpyr} \qquad r^2 = 0.817 \qquad \{3\}$$

$$Y_{wat} = 0.4186\, T_{pyr} - 0.0004\, T_{pyr}^{2} - 102.679 \qquad r^2 = 0.802 \qquad \{4\}$$

where	T_{pyr}	is the pyrolysis reactor temperature in °C
	Y_{org}	is the organic yield, dry basis (% wood feed)
	Y_{gas}	is the off-gas yield (% wood feed)
	Y_{char}	is the char/ash yield (% wood feed)
	Y_{wat}	is the reaction water yield (% wood feed)
	r^2	gives an indication of the accuracy of the curve fit to the original data. The higher the value of r^2 the more accurate is the curve fit i.e. a value of 1 is the maximum curve fit and represents a complete fit of the generated curve to the original data. A value of r^2 greater than 0.8 is considered acceptable.

Due to the low reaction water yield in Table 6.1. adjustments were required in the reaction water yields as given in equations {4} to increase the reaction water yield to approximately 11% at a pyrolysis temperature of 500°C [59]. This was done by assuming that approximately 8% of the recorded organic yield in Table 6.1. was actually water and increasing the reaction water yields accordingly.

The total water yield from flash pyrolysis is determined by the reaction water yield and the moisture content of the prepared feedstock entering the pyrolysis process. Likewise, the total char/ash yield is given by the yield given in Table 6.1.(on dry, ash free feedstock) and the char/ash content of the feedstock. A further adjustment is also required to the pyrolysis yields given in equations {1} - {4} to normalise the yields to give a 100% pyrolysis yield on the feedstock. The resulting relationships for the pyrolysis yields incorporating all of the adjustments above are given in equations {5} - {8}, all derived yields are tonnes/day of product.

$$\text{Organic liquid (dry) yield} \quad = Y_{org} \times \left(\frac{F_{in}}{Y_{norm}}\right) \times 0.92 \qquad \{5\}$$

$$\text{Total water yield} \quad = (0.08 Y_{org} + Y_{wat}) \times \left(\frac{F_{in}}{Y_{norm}}\right) + F_{wat} \qquad \{6\}$$

$$\text{Char/ash yield} \qquad = Y_{char} \times \left(\frac{F_{in}}{Y_{norm}} \right) + F_{ash} \qquad \{7\}$$

$$\text{Off-gas yield} \qquad = Y_{gas} \times \left(\frac{F_{in}}{Y_{norm}} \right) \qquad \{8\}$$

$$\text{and } Y_{norm} \qquad = Y_{org} + Y_{wat} + Y_{gas} + Y_{char} \qquad \{9\}$$

where	T_{pyr}	is the pyrolysis reactor temperature in °C
	F_{in}	is the d.a.f. feed input to the pyrolysis process (tonnes/day)
	F_{wat}	is the moisture content of the feedstock entering the pyrolysis process (wt% wet basis)
	F_{ash}	is the char/ash content of the feedstock entering the pyrolysis process (wt%)
	Y_{norm}	is the total yield of the pyrolysis process before normalisation and is given by equation (wt%) {9}

6.3.3. Product yields using the empirical relationships.

The resultant yields for the pyrolysis products given by equations {5} - {8} are given in Table 6.2. for the case of 100 d.a.f. tonnes/day wood feed with a feed moisture content of 6.5% wet basis (to allow comparison with the data in Table 6.1.) at a range of reaction temperatures.

From the results shown in Table 6.2. it can be seen that at a pyrolysis temperature of 500°C the yield of crude, wet pyrolysis oil obtained is 80.4% (dry feed basis). The resulting moisture content of the crude pyrolysis oil produced at 500°C is 23.3% (wet basis). The results shown in Table 6.2. and their derivation, including the adjustment of the organic and water yields, have been validated by Waterloo [59] as being reasonable and representative.

Figures 6.2. -6.5. illustrate the original Waterloo data as shown in Table 6.1. and the curve fits on the data in comparison with the product yields derived from the empirical relationships and shown in Table 6.2. Figure 6.2. shows the dry organic yields, Figure 6.3. the reaction water yields, Figure 6.4. the char yields and Figure 6.5. the off-gas yields.

Table 6.2.
Derived product yields from the flash pyrolysis of 100 d.a.f. tonnes/day of wood feed.
(Basis: 6.5% moisture content feedstock (wet basis))

Reactor temperature, °C	**400**	**450**	**500**	**550**	**600**	**650**
Derived product yields, (t/day)						
Organic liquid, dry	55.1	60.9	61.7	58.2	48.9	29.7
Char	33.1	20.5	13.8	10.0	7.6	6.0
Gas	5.9	8.2	12.4	20.0	34.2	60.9
Total water	12.4	16.9	18.7	18.4	15.8	9.9
Total water yield composed of :						
Reaction water (t/day)	5.9	10.4	12.2	11.9	9.3	3.4
Feed water (t/day)	6.5	6.5	6.5	6.5	6.5	6.5
Moisture content of crude bio-oil (% wet basis)	18.4	21.8	23.3	24.0	24.4	25.0

The derived char and off-gas yields show good correlations to the curve-fitted Waterloo data between 450 and 600°C which is not surprising since the Waterloo data was for pyrolysis temperatures between 425 and 625°C. Outside the 450 - 600°C temperature range the derived data is less accurate and cannot be relied on. The dry organic and reaction water yields cannot be directly compared with the Waterloo base data as there have been adjustments to increase the reaction water yields. However, from Figure 6.3. it can seen that the reaction water yield at 500°C, 12.2wt% on dry feed, is much nearer the 11wt% proposed by Waterloo [59].

6.3.4. Feed moisture content.

In addition to the effect of pyrolysis temperature, the other variable which is used in the generic pyrolysis model and which affects the bio-oil yield is the feed moisture content. The total water yield from pyrolysis is derived from the moisture content of the feedstock and the reaction water yield as stated earlier.

The effect of feed moisture content on the bio-oil product moisture content is shown in Figure 6.6. for various pyrolysis temperatures and using the normalised correlations in equations {5} to {8}. The correlation for a particular pyrolysis temperature is a straight line as the reaction water yield is a function of temperature and is therefore constant.

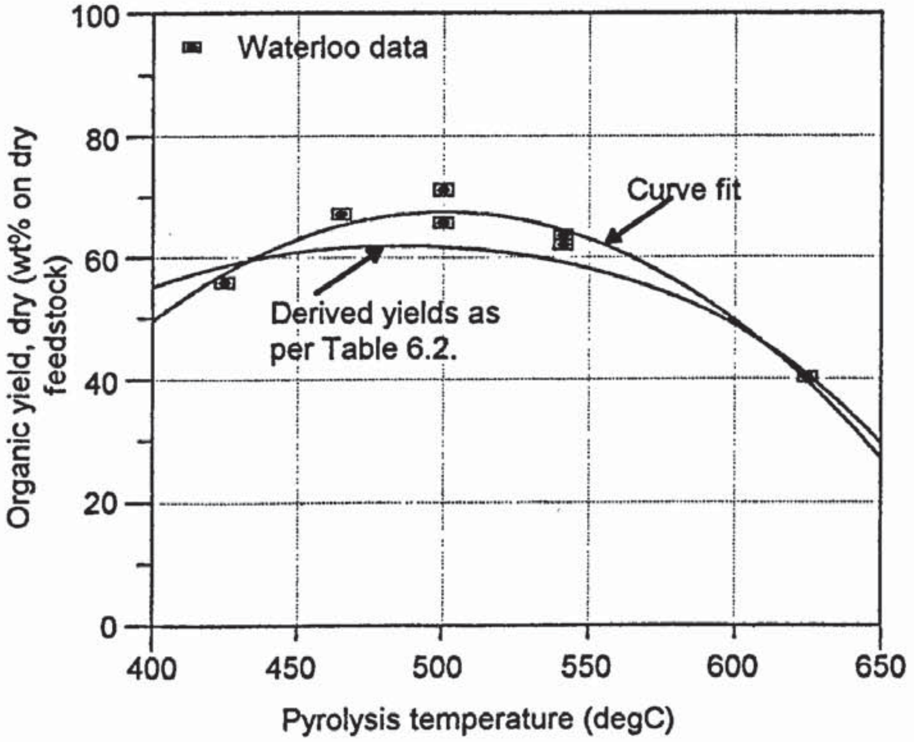


Figure 6.2. Organic yields from flash pyrolysis as a function of pyrolysis temperature.

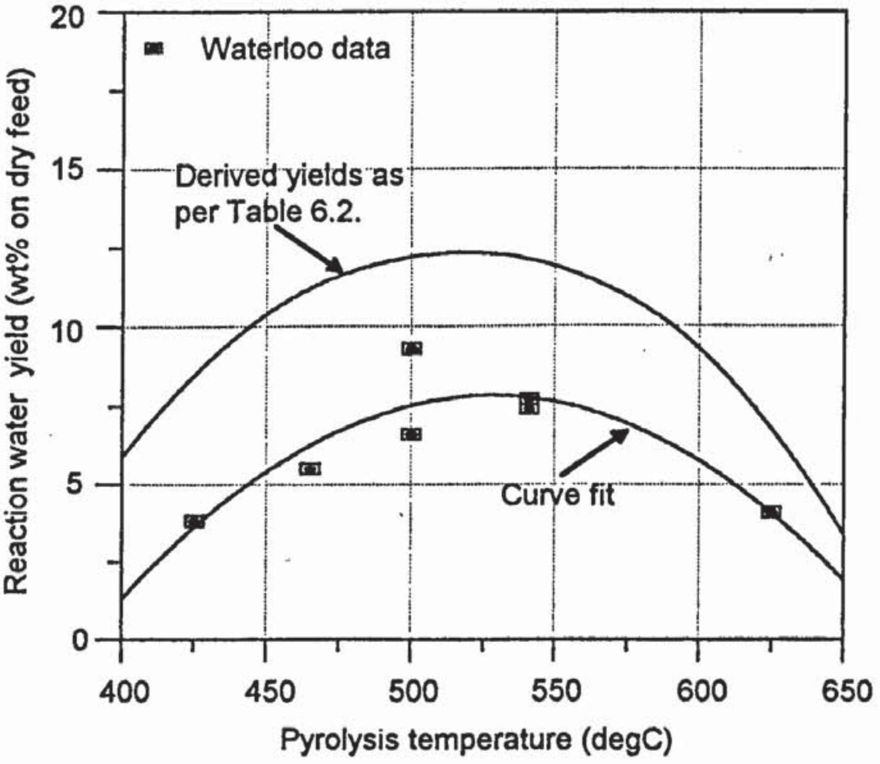


Figure 6.3. Reaction water yields from flash pyrolysis as a function of pyrolysis temperature.

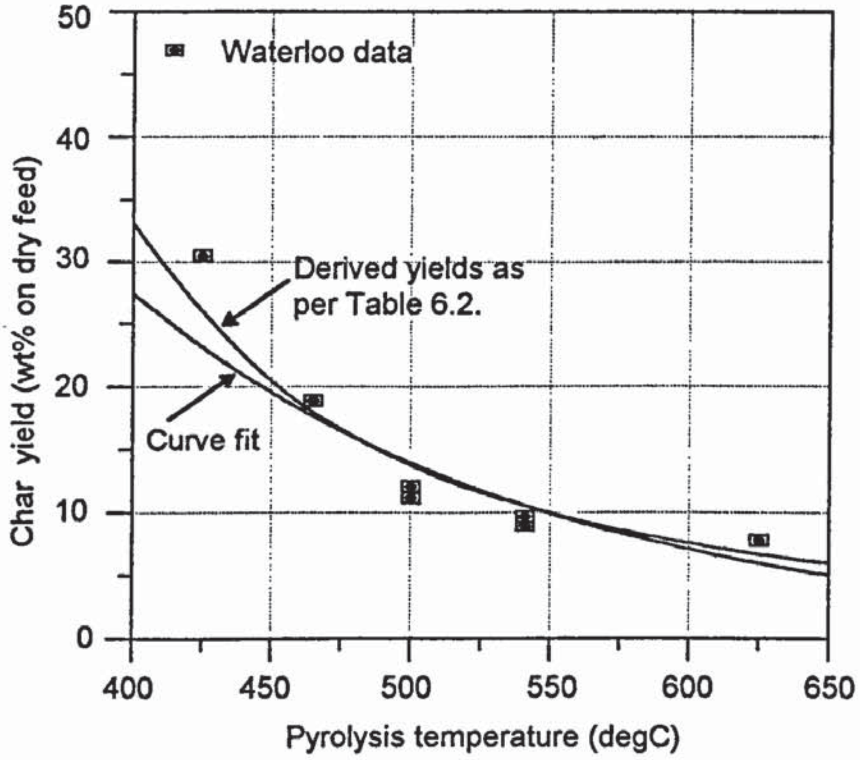


Figure 6.4. Char yields from flash pyrolysis as a function of pyrolysis temperature.

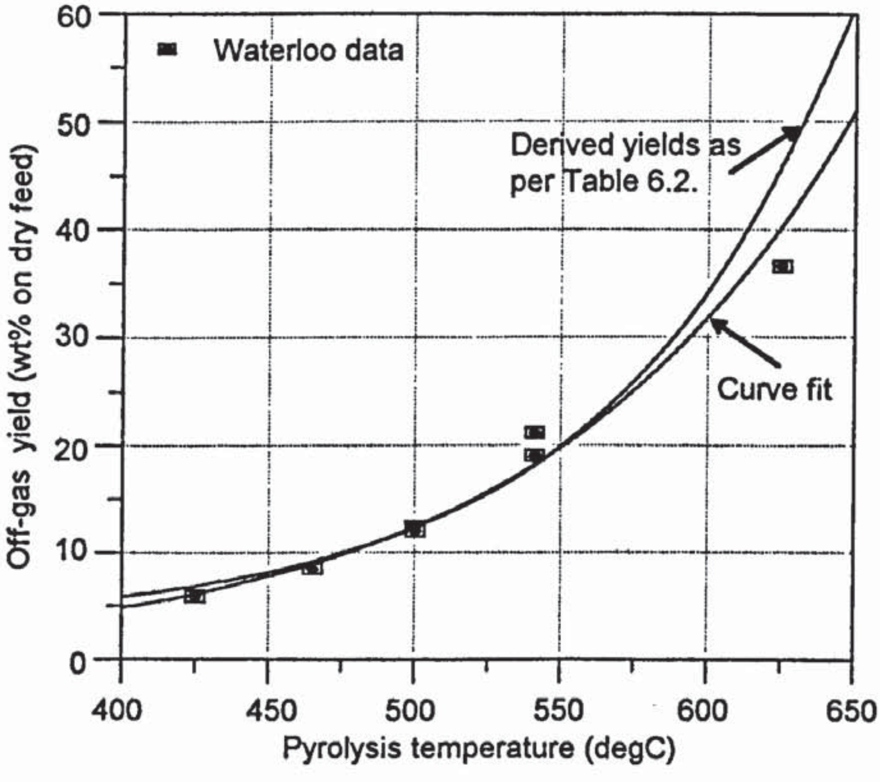


Figure 6.5. Off-gas yields from flash pyrolysis as a function of pyrolysis temperature.

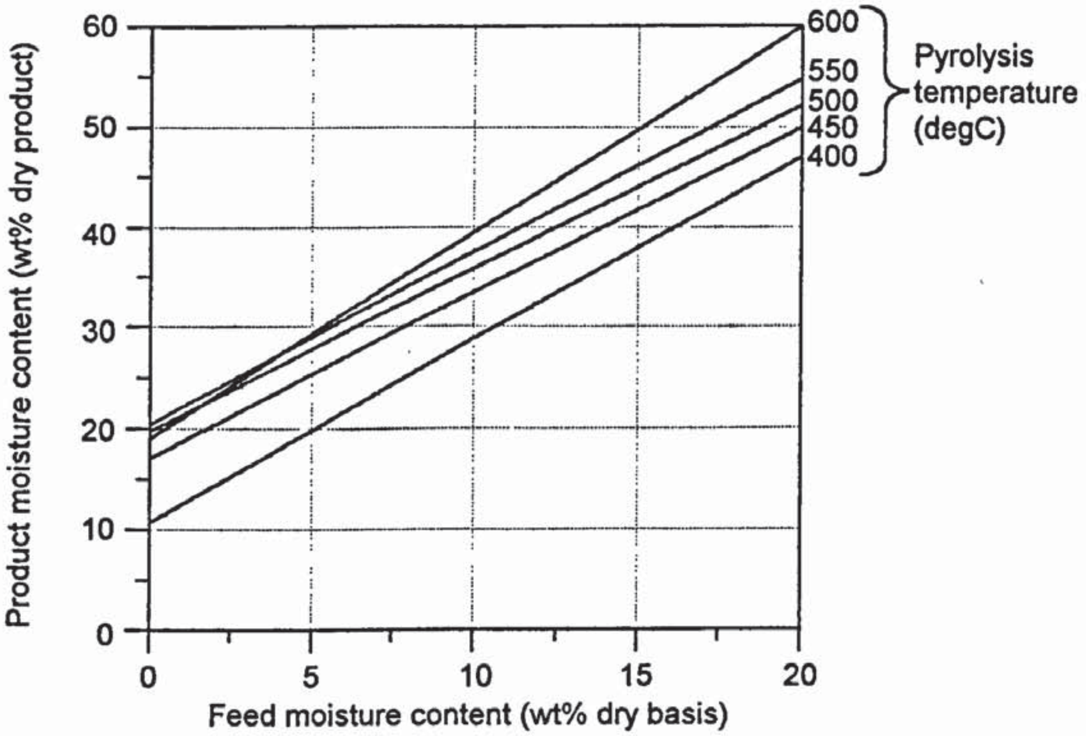


Figure 6.6. The effect of feed moisture content on the bio-oil moisture content.

However, the feed moisture content versus pyrolysis temperature correlations are not parallel due to the affect of temperature on reaction water yields as demonstrated in Figure 6.6. The line for a pyrolysis temperature of 600°C shown on Figure 6.6. actually crosses over the lines for temperatures of 550°C and 400°C when the feed moisture content is below 5wt% (dry basis). This apparent anomaly can be explained by the decrease in reaction water yields at increasing temperatures as shown in Figure 6.2.

Figure 6.7. shows the data in Figure 6.6. with the pyrolysis reactor temperature as the main variable and the resultant correlations are no longer linear. The correlation for a feed moisture content 5% has a domed shape similar to that of the reactant water curve given in Figure 6.3. as the amount of water entering with the feed is insignificant in comparison with the amount of water produced by the pyrolysis process. However, as the feed moisture content increases the shape of the curve alters as the feed moisture content becomes the most significant contributor to the product moisture content.

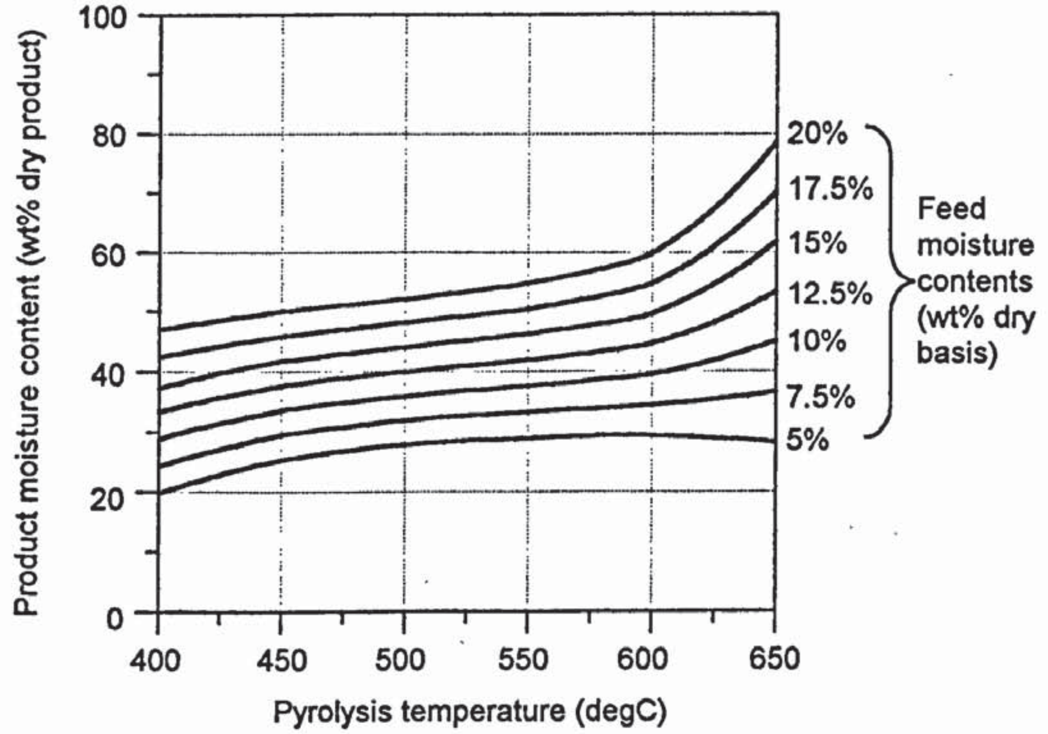


Figure 6.7. The effect of reaction temperature on the bio-oil moisture content.

6.4. Derivation of the energy balance.

6.4.1. The energy requirements.

The energy requirements of the pyrolysis process is an integral part of the generic flash pyrolysis model. The feed coming into the reaction must be heated to the pyrolysis reaction temperature before the pyrolysis reaction can commence. There are five components in the pyrolyser energy requirement: the energy required to raise the dry feed temperature from the inlet temperature to the reaction temperature, the energy required to pyrolyse the dry wood feed at the reaction temperature, the energy required to raise the temperature of the feed moisture content from the inlet temperature to 100°C (at which temperature water vaporises), the energy required to vaporise the feed moisture content at 100°C and the energy required to raise the water vapour from 100°C to the pyrolyser temperature. It is assumed that the amount of ash in the feedstock is negligible and will not affect the pyrolyser energy requirements. The five requirements can be put into two categories - those pertaining to the dry feed and those pertaining to the feed's moisture content. The energy requirements of the dry feed and the moisture content of the feed are summarised in equations {10} and {11}.

$$E_{wood} = M_{wood} \times ((C_{wood} \times (T_{pyr} - T_{in})) + \Delta Hf_{organics}) \qquad \{10\}$$

$$E_{water} = M_{water} \times ((C_{water} \times (100 - T_{in})) + \Delta Hv_{water} + (C_{vap} \times (T_{pyr} - 100))) \qquad \{11\}$$

where	E_{wood}	is the energy required by the dry wood (MJ)
	E_{water}	is the energy required by the feed moisture content (MJ)
	T_{pyr}	is the pyrolysis reactor temperature (°C)
	M_{wood}	is the amount of dry wood entering the reactor (tonnes)
	C_{wood}	is the average heat capacity of dry wood at constant pressure (MJ/tonne °C)
	T_{in}	is the feed inlet temperature (°C)
	C_{water}	is the average heat capacity of water at constant pressure (MJ/tonne °C)
	C_{vap}	is the average heat capacity of water vapour at constant pressure (MJ/tonne °C)
	$\Delta Hf_{organics}$	is the specific enthalpy of formation of dry organics (MJ/tonne)
	ΔHv_{water}	is the specific enthalpy of vaporisation of water (MJ/tonne)
	M_{water}	is the amount of water entering the reactor with the feedstock (tonnes)

Table 6.3. shows the energy requirements for the generic pyrolysis process for the case of 100 tonnes/day of d.a.f. wood feed with a moisture content of 7wt% (wet basis) based on equations {10} and {11}. The feed temperature on entering the pyrolyser is taken to be 20°C and the specific heat capacity of the wood is assumed to be 2.8MJ/tonne K [136], all standard values. The enthalpy of pyrolysis is taken as 40.0 MJ/tonne, a value which has been used by other researchers [146] although some researchers say that it's value could be zero. However, the energy required to pyrolyse the dry feed is negligible compared with the total energy requirements of the pyrolysis system and so the value makes little difference to the overall energy balance.

Table 6.3.
Predicted energy requirements for flash pyrolysis.
(Basis: 100 d.a.f. tonnes/day of wood, moisture content of 7wt% (wet basis) and all values are in GJ/tonne d.a.f. feed)

Reactor temperature (°C)	**400**	**450**	**500**	**550**	**600**	**650**
Energy required to heat and pyrolyse the dry wood	-1.09	-1.23	-1.37	-1.51	-1.65	-1.79
Energy required to heat and vaporised the feed's moisture content	-0.21	-0.22	-0.23	-0.23	-0.24	-0.25
TOTAL ENERGY REQUIRED	-1.30	-1.45	-1.60	-1.74	-1.89	-2.04

As could be expected, the energy requirements for the dry wood feed increase with increasing reactor temperature. The amount of energy required by the dry wood is considerably larger than that required by the feed's moisture content. This is not surprising considering the difference in quantities of wood and water fed to the pyrolysis reactor.

6.4.2. Fuel sources.

One of the objectives of the generic flash pyrolysis model was that it should be as energy self-sufficient as possible. To this end the char and gas by-products are utilised to satisfy the pyrolyser energy requirements where possible. Surplus off-gas is exported from the pyrolyser model as fuel gas for use in the feed drying step. The recoverable energy values for the off-gas assume that it is at 60°C to utilise the sensible heat of the gas and typical energy values of 0.0015 GJ m^{-3} and 30 GJ/tonne have been assumed for the off-gas and char respectively [35]. There are five options for supplying the energy requirements of the pyrolyser which are:-

- char followed by imported fuel gas to make up the deficit, all the off-gas and any surplus char are exported as by-products;
- char followed by the off-gas, any surplus char or off-gas is exported as by-product;
- off-gas followed by imported fuel gas to make up the deficit, all char is exported as a by-product;
- off-gas followed by char, any surplus char is exported as by-product; and

- imported fuel gas only, all char and off-gas are exported as by-products.

These options vary the sequence in which the char, off-gas and imported fuel gas are utilised to satisfy the pyrolyser energy requirement and have a direct impact on the utilities requirements of the generic pyrolysis process, the by-product exports and hence on the production cost of the bio-oil product. For example, if imported fuel gas only is used, there will be a high fuel gas demand on the utilities and hence a high utilities charge. All of the char is exported as by-product resulting in a by-product credit and the off-gas is exported as fuel gas. The five options can only really be evaluated quantitatively by using the production cost of the bio-oil product and this has been included in the results analysis in Chapter 11

Table 6.4. utilises the predicted energy requirements for the generic flash pyrolysis model as given in Table 6.3. in comparison with the energy recoverable from the char and off-gas assuming a 75% combustion efficiency. Based on 100 tonnes/day of d.a.f. wood feed with a moisture content of 7wt% (wet basis).

Table 6.4.
Predicted energy requirements for the flash pyrolysis of wood and the energy recoverable from the char and off-gas at 75% combustion efficiency.
(Basis: 100 d.a.f. tonnes/day of wood, moisture content of 7wt% (wet basis) and all values are in GJ/tonne d.a.f. feed)

Reactor temperature (°C)	**400**	**450**	**500**	**550**	**600**	**650**
Total energy required to pyrolyse the feedstock (from Table 6.3.)	-1.30	-1.45	-1.60	-1.74	-1.89	-2.04
Energy supplied by the char	7.23	4.42	2.93	2.08	1.58	1.27
Energy supplied by the off-gas	0.05	0.06	0.10	0.15	0.26	0.47
TOTAL ENERGY SURPLUS OR DEFICIT	5.98	3.03	1.43	0.49	-0.05	-0.30

Looking at the data in the table it is evident that the energy provided by combusting the char is sufficient to satisfy the energy requirements for pyrolysis temperatures less than 550°C, for a 7% moisture content feed. Also, the char contributes more to the energy balance than the off-gas. This information is significant if the char is not used as fuel for example, in the Ensyn, NREL and Union Fenosa step models (See Section

6.7.) as there is insufficient energy provided by the off-gas to satisfy the energy requirement and fuel gas would have to be imported from the utilities.

Figure 6.8. shows some energy balances over the generic pyrolyser for feed moisture contents between 5 and 25 % (wet basis) for temperatures between 400 and 650 °C. The balances were derived using the energy requirements of the pyrolyser and the energy recoverable from utilising the both the char and off-gas together at a combustion efficiency of 75%. All of the values shown are in GJ/tonne d.a.f. feed and are based on the product yields shown in Table 6.2. Figure 6.8. demonstrates quite clearly the temperatures below which the pyrolysis model is total energy self-sufficient i.e. no fuel gas is imported from the utilities.

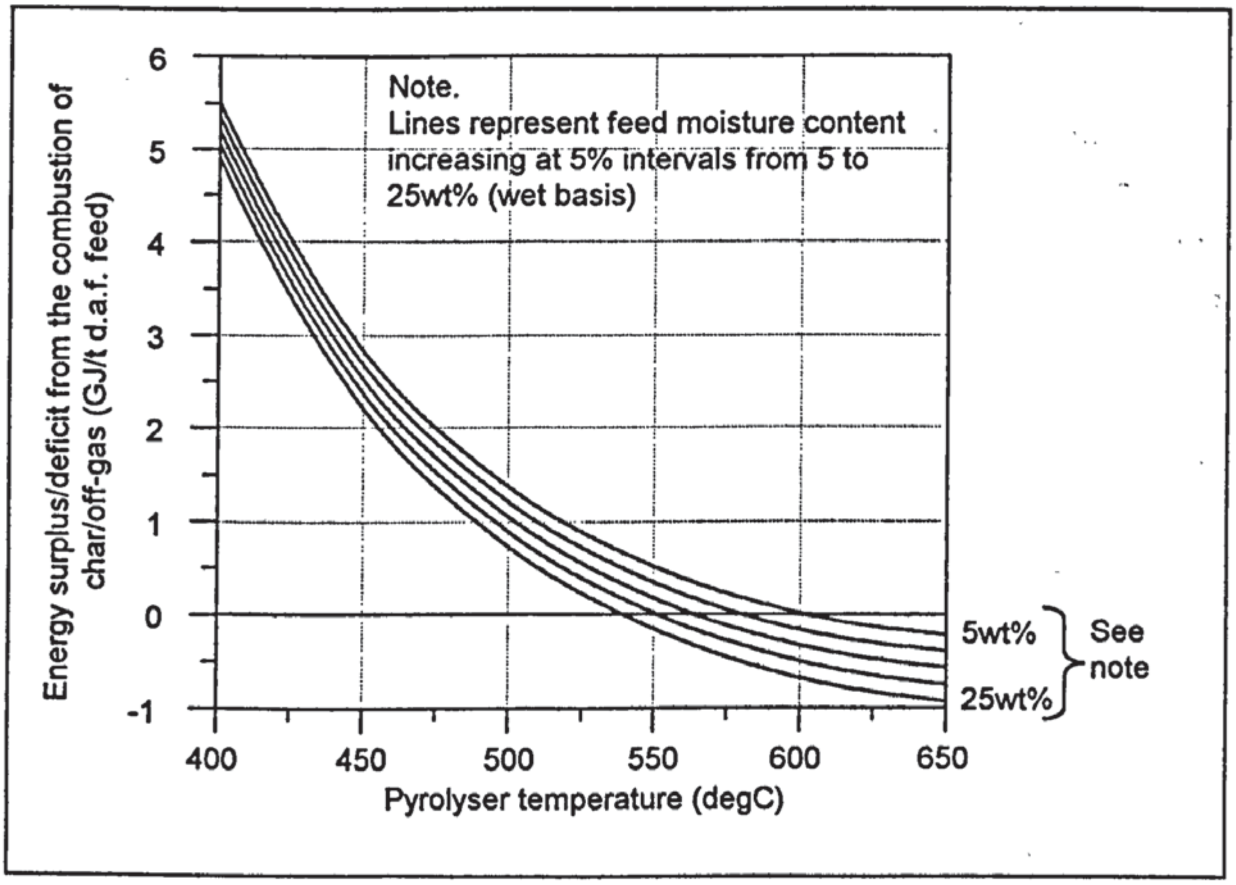


Figure 6.8. Energy balances for the generic flash pyrolysis model for a char/off-gas combustion efficiency of 75% over a range of pyrolysis temperatures and feed moisture contents.

For example, for a 10% moisture content feedstock (wet basis) the utilisation of the char and off-gas will be sufficient to satisfy the energy demand for temperature below 600°C. For a very wet feedstock with a moisture content of 25% (wet basis), the energy demand can be satisfied at temperatures below 580°C. In all cases where the energy surplus is negative fuel gas must be imported from the utilities to satisfy the energy demand. The energy balances follow an expected trend - there is an energy surplus in the lower temperature region where the char yields are relatively high (as demonstrated in Table 6.2.) and a deficit in the high temperature region where the char yields are low. Although the off-gas yields increase with an increase in temperature (as demonstrated in Table 6.2.), the off-gas has a lower energy value than the char and contributes less to the overall energy balance.

Figure 6.9. shows the maximum pyrolyser temperatures possible for energy self-sufficiency for char and off-gas combustion efficiencies between 20 and 100% and feed moisture contents between 5 and 25% (wet basis). The graph was derived using the energy requirements of the pyrolyser and the energy recoverable from the char and off-gas at a range of pyrolyser temperatures, char/gas combustion efficiencies and feedstock moisture contents. Char/off-gas combustion efficiencies below 50% could represent a system with inefficient combustion or indirect heat transfer for example, a system heated indirectly with external fuel combustion. Although a 100% combustion efficiency is perhaps unrealistic, a figure close to it would be possible with in-situ "combustion" where the heat losses are minimal for example, the recirculating fluidised bed system designed by CRES [21, 22].

For example, using a 5% moisture content (wet basis) feedstock at a char/off-gas combustion efficiencies of 70% the maximum temperature possible for energy self-sufficiency is 580°C. By increasing char/gas combustion efficiencies to 80% the maximum temperature shifts to 620°C. For a 25% moisture content (wet basis) feedstock at char/gas combustion efficiencies of 80% the maximum possible temperature for energy self-sufficiency is 550°C. This can be increased to 565°C if a 20% moisture content (wet basis) feedstock is utilised. The maximum pyrolysis temperature possible for energy self-sufficiency can be increased by improving the char/off-gas combustion efficiency or decreasing the feedstock moisture content, as shown in Figure 6.9.

5.4.3. Cooling duties.

When the required product from the flash pyrolysis process is crude pyrolysis which is to be utilised directly or for hydrotreatment or power production a cooling duty included in the energy balance.

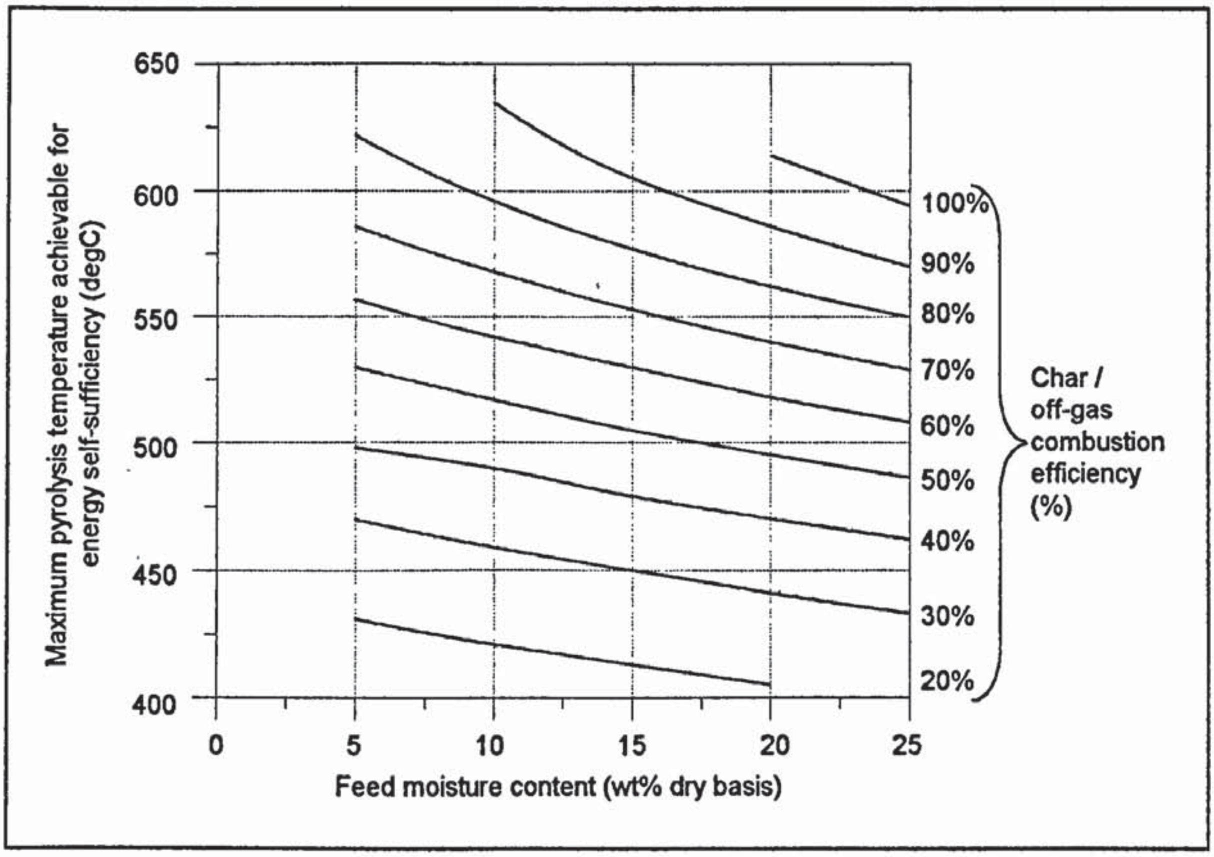


Figure 6.9. The effect of char/off-gas combustion efficiency on the maximum pyrolysis temperature achievable for energy self-sufficiency for variable feed moisture contents.

The duty on the product cooling and condensing system and the amount of cooling medium required are determined under the energy balance for the generic flash pyrolysis model. The duty on the cooling/condensing system is composed of the energy required to cool the pyrolyser off-gas (composed primarily of hydrogen, methane, carbon dioxide and carbon monoxide) from the pyrolysis reaction temperature to the condenser outlet temperature and the energy required to cool and condense the crude, wet bio-oil product and the small quantity of carried over char.

There is little information available on the physical properties of crude pyrolysis oil and so a model compound has to be utilised to substitute for the bio-oil. After consultation with other researchers it was decided to utilise the properties of an organic, oxygen containing compound with a molecular weight around 100. Table 6.5. shows the predicted cooling duties on the pyrolysis quench system based on a 100 d.a.f. tonnes/day wood feed plant. Three different organic compounds with molecular weights around 100 were utilised to demonstrate the effect that the choice of organic compound has on the cooling duty. The compounds were selected arbitrarily and were furfural, isovaleric acid and methyl isobutyl ketone, the properties of which are included in Table 6.5. The cooling duties were determined for pyrolysis temperatures between 400 - 650°C, a quench outlet temperature of 60°C.

Table 6.5.
Predicted cooling duties for the products from the flash pyrolysis of wood.
(Basis : 100 d.a.f. tonnes/day of wood, 7wt% moisture content (wet basis), all values are in GJ/tonne d.a.f. feed)

Reactor temperature (°C)	**400**	**450**	**500**	**550**	**600**	**650**
Cooling duty of wet organics						
using organic A	0.753	0.970	1.090	1.119	1.013	0.667
using organic B	0.827	1.074	1.218	1.260	1.150	0.761
using organic C	0.717	0.936	1.061	1.097	0.999	0.661
Cooling duty for off-gases	0.013	0.022	0.040	0.076	0.150	0.302
TOTAL DUTY ORGANIC A	0.766	0.992	1.130	1.195	1.163	0.969
TOTAL DUTY ORGANIC B	0.840	1.096	1.258	1.336	1.299	1.063
TOTAL DUTY ORGANIC C	0.730	0.958	1.101	1.173	1.148	0.963

NOTES.
Physical properties of the chosen organics:
A = furfural ($C_5H_4O_2$) MW = 96 Cp = 1.74 Lh = 450.04
B = isovaleric acid ($C_5H_{10}O_2$) MW = 102 Cp = 2.47 Lh = 423.00
C = methyl isobutyl ketone ($C_6H_{12}O$) MW = 100 Cp = 1.92 Lh = 345.01

where MW is the molecular weight, Cp is the specific heat capacity (MJ/t K) and Lh is the enthalpy of condensation (MJ/t) [136]

The largest portion of the cooling duty on the condenser is composed of cooling and condensing the wet organics to form the crude bio-oil. This duty follows the same trend as the yields of dry organics and water from flash pyrolysis i.e. a rise up to a temperature around 500°C and then a decline as the temperature rises further. The cooling duty from the pyrolyser off-gas is small in relation to that of the wet organics and it increases with temperature due to increasing gas yields (and increasing temperature). However, the contribution of the off-gas to the cooling duty is not significant enough to affect the trend set by the wet organics. As can be seen from the table, the choice of the model organic compound does not appear to have an impact on the cooling duty.

6.4.4. Determination of the heating value of the crude bio-oil.

As described in Chapter 2 the water content of the crude bio-oil product alters it's heating value (HHV) [3]. The drying model described in Chapter 5 allows the User to determine the feed moisture content entering the pyrolysis processes and hence the water content of the final product. A simple algorithm was derived using published data [3] relating the HHV of a crude bio-oil to it's water content and is shown in equation {10}. The algorithm was derived by taking the HHV of a dry bio-oil and that of a bio-oil with a known moisture content and taking a straight line correlation between the two.

$$\text{Bio-oil HHV (GJ/tonne)} \quad = 23+\left(\frac{M_{org}}{\left(M_{org}+M_{water}\right)\times 100}\right)\times -0.23 \qquad \{10\}$$

where M_{org} is the amount of dry organics produced (tonnes/day)

M_{water} the amount of water produced, reaction water plus feed water (tonnes/day)

6.5. The modelling of vapour residence time and feed composition.

6.5.1. Vapour residence time.

The yield and quality of the products from flash pyrolysis are both affected by the vapour residence time and the reaction temperature as discussed in Chapter 2. The Waterloo data utilised to determine the product yields includes a residence time effect but there is little information on the interaction between residence time and temperature for the range of temperatures modelled here. Some data is available but it is not directly applicable as it is for high temperature gas production [48] as described in Section 2.3.3. At the present time there is insufficient data for the temperature range utilised in the model for residence time to be included as a separate variable and possible effects of vapour residence time on product yields have been discussed in Section 2.3.3.

6.5.2. Feed composition.

It is necessary to be able to relate the products from flash pyrolysis to the feed specifications for the hydrotreating model if the models are to be used in a processing route. The methodology used is to determine the CHO analysis of the flash pyrolysis product and to use this in conjunction with the severity of the hydrotreating deoxygenation process to determine the oxygen content of the hydrotreated product. The extent of deoxygenation also determines the amount of separation of the hydrotreated product from the reaction water and the quantity of reaction water produced. The CHO of the flash pyrolysis product can be related to that of the respective feedstock using algorithms derived from data collated from published sources. The provisional correlations which have been derived for C/H, C/O and C/N are shown in equations {11} - {13} and in Figures 6.10. and 6.11.

Pyrolysis oil C/H ratio $= -9.7950 + 2.0099 \times C/H_{feed}$ $r^2 = 0.912$ {11}

Pyrolysis oil C/O ratio $= -0.65441 + 1.6146 \times C/O_{feed}$ $r^2 = 0.945$ {12}

Pyrolysis oil C/N ratio $= -283.47 + 9.0906 \times C/N_{feed}$ $r^2 = 0.9$ {13}

where C/H_{feed} is the feedstock C/H ratio
C/O_{feed} is the feedstock C/O ratio
C/N_{feed} is the feedstock C/N ratio

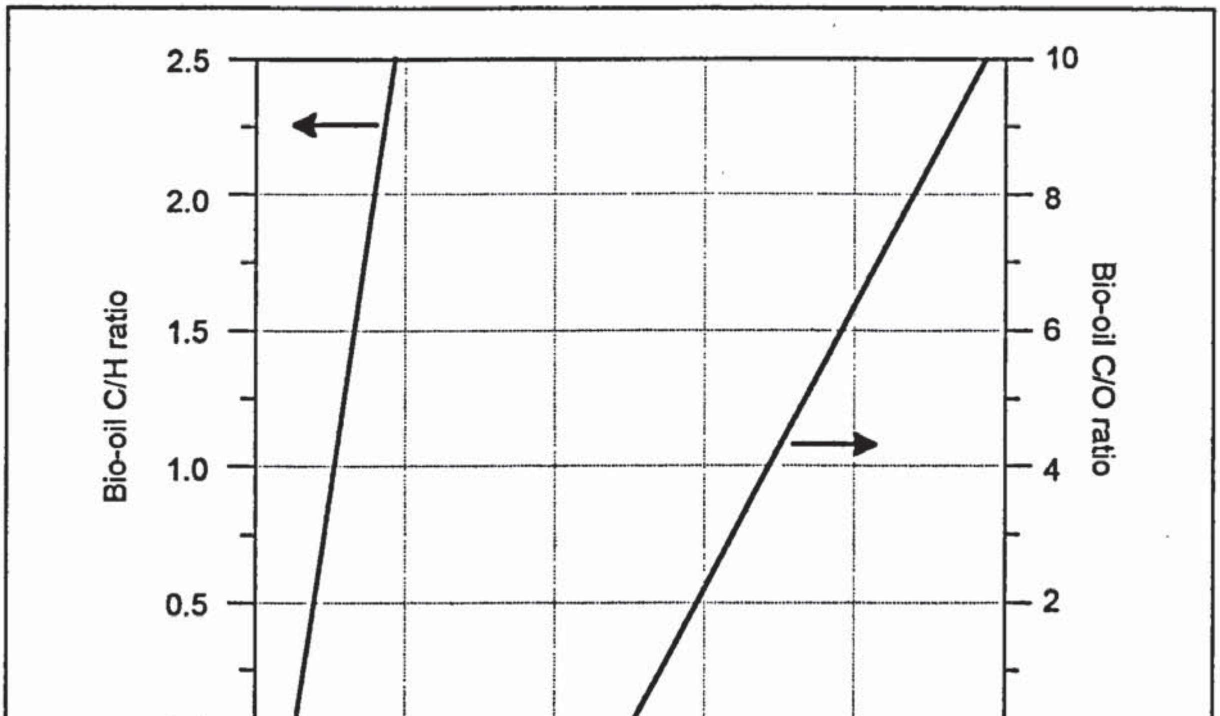


Figure 6.10. Correlations for the relations between feed and bio-oil C/H and C/O ratios.

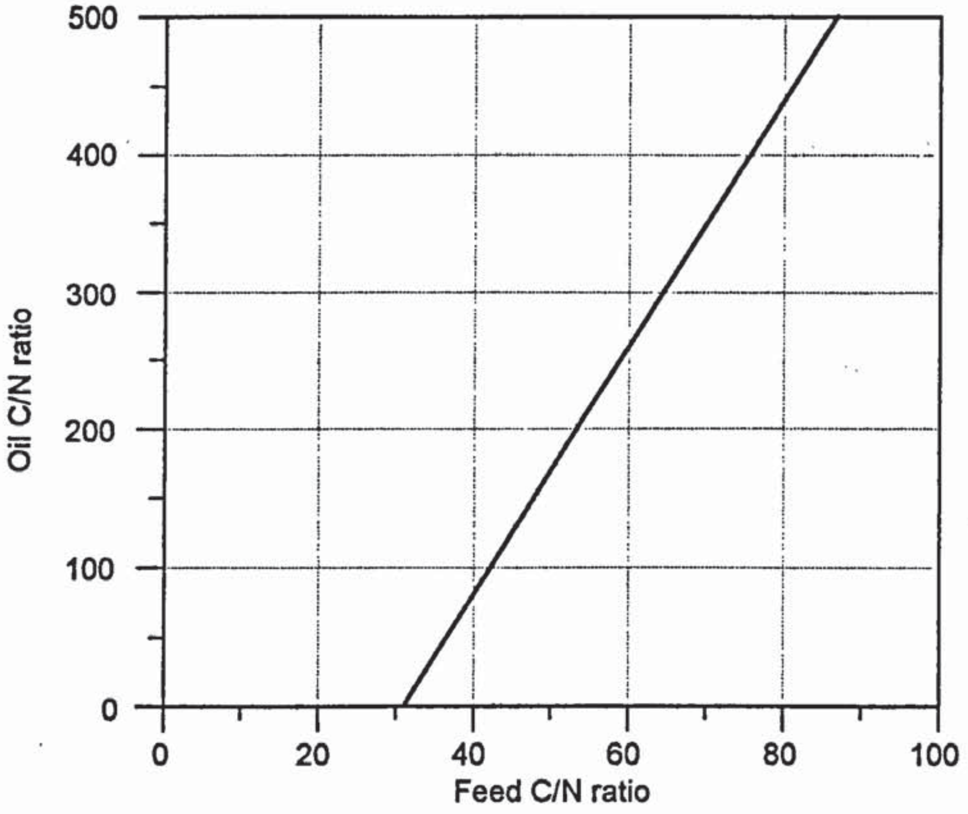


Figure 6.11. Correlation for the relationship between feed and bio-oil C/N ratios.

6.5. Capital cost and utility requirements.

The capital cost for the generic flash pyrolysis model is based on the algorithm in equation {14}. The basis for the algorithm is the installed cost of 1 tonne/hour flash pyrolysis plant [147] (pretreatment of the feedstock excluded).

Capital cost (1993 £) $= 673100 \times \left(\frac{F_{in}}{24}\right)^{0.65}$ {14}

where F_{in} is the d.a.f. feed input to the pyrolysis process (tonnes/day)

The possible utilities used by the generic flash pyrolysis model are power, cooling water and possibly fuel gas. The power requirement for the model is given by equation {15} and is taken from the previous work [28] as there was no other data available.

Power requirement $= 0.00363 \times F_{in}$ {15}

The quantity of cooling water required is dependent on the pyrolyser energy balance and is determined by the cooling duty calculated. Cooling water is not utilised to directly cool and condense the flash pyrolysis products but is utilised to cool the quench medium (recycled condensed pyrolysis oil). The permitted rise in the cooling water temperature is determined by the User and a 5% heat loss is assumed in the condenser to allow for heat transfer problems and fouling.

The fuel gas requirement (if any) is dependent on the pyrolyser temperature, the feed moisture content the char/off-gas combustion efficiency selected and the pyrolyser energy supply option selected. The energy requirements of the process are determined for the pyrolysis temperature and feed moisture content as described in Section 6.4.1. For example, for the data in Table 6.4. (7% moisture content feedstock, char and off-gas combustion efficiencies of 75%) at 500°C there is sufficient energy available from the char to satisfy the pyrolysis energy requirements and fuel gas will not be imported from the utilities. However, if the pyrolysis temperature is increased to 600°C there is insufficient energy provided by combustion of the char and off-gas and fuel gas would have to be imported from the utilities to make up the deficit.

6.7. The customised pyrolysis models.

6.7.1. Introduction.

The customised flash pyrolysis step models for the Ensyn, Fenosa and NREL processes are based on the yields shown in Tables 2.1., 2.3. and 2.4. in Chapter 2. Process descriptions and flowsheets for the processes are given in Chapter 2, Sections 2.4. - 2.6. The models are fixed models i.e. the process variables used in the step models have been fixed to accurately represent the technologies modelled as accurately as possible from the supplied data. The only differences between the generic and the customised flash pyrolysis processes is that the User determined variables in the Generic model for example, pyrolysis temperature have been preset in the customised models. The product yields in the customised models are not determined by empirical relationships but are designed to vary with feed moisture content and feed throughput only. The energy balances developed for the generic model are also utilised for the customised models.

6.7.2. The Ensyn flash pyrolysis model.

The Ensyn flash pyrolysis model is based on the Ensyn RTP process described in Section 2.4. Table 6.6. gives the process conditions used to develop the mass and energy balances and derive the capital cost and utilities requirements. The Ensyn pyrolysis model uses the off-gas produced by the process to provide the process heat at a combustion efficiency of 75% with the deficit being made up from fuel gas imported from the utilities. All of the char produced in the processes is exported from the model as a by-product and is credited to the model. The capital cost utilised in the Ensyn flash pyrolysis model is the same cost algorithm as developed for the generic flash pyrolysis process. The utilities requirements are developed in a similar way to those for the Generic flash pyrolysis model.

Table 6.6.
Process conditions and variables used in the Ensyn RTP model.

Item	Units	Default
Pyrolysis temperature	°C	510.0
Off-gas combustion efficiency	%	75.0
Char filter efficiency	%	90.0
Product temperature ex condenser	°C	40.0

The Ensyn flash pyrolysis model is the only pyrolysis model included in BLUNT which can accept straw and sorghum bagasse feedstocks in addition to wood as this is the only process for which there is data on straw and sorghum bagasse pyrolysis. Product yields for the three feedstocks permitted in the Ensyn model are shown in Table 6.7. The data utilised in the Ensyn model for the pyrolysis of straw and sorghum bagasse is based on actual processing of these feedstocks in the Ensyn RTP system [148 , 149].

Table 6.7.
Product yields for various feedstocks processed in the Ensyn RTP system.
(Basis: 100 tonnes/day d.a.f. feedstock)

Feedstock	Crude bio-oil (wet)	Off-gas	Char
Wood (10% m.c. dry basis)	82.5	16.5	11.4
Straw	71.3	11.4	17.3
Sorghum bagasse	69.5	13.2	17.3

6.7.3. The Fenosa flash pyrolysis model.

The Fenosa flash pyrolysis model is based on the Union Fenosa process described in Section 2.5. Table 6.8. gives the process conditions used to develop the mass and energy balances and derive the capital cost and utilities requirements. The Fenosa pyrolysis model uses the off-gas produced by the process to provide the process heat at a combustion efficiency of 75% with the deficit being made up from fuel gas imported from the utilities. All of the char produced in the processes is exported from the model as a by-product and is credited to the model. The capital cost utilised in the Fenosa flash pyrolysis model is the same cost algorithm as developed for the generic flash pyrolysis process. The utilities requirements are developed in a similar way to those for the Generic flash pyrolysis model.

Table 6.8.
Process conditions and variables used in the Fenosa flash pyrolysis model.

Item	Units	Default
Pyrolysis temperature	°C	520.0
Off-gas combustion efficiency	%	75.0
Char filter efficiency	%	90.0
Product temperature ex condenser	°C	40.0

6.7.4. The NREL vortex pyrolysis model.

The NREL pyrolysis model is based on the NREL vortex pyrolysis process described in Section 2. Table 6.9. gives the process conditions used to develop the mass and energy balances and derive the capital cost and utilities requirements.

Table 6.9.
Process conditions and variables used in the NREL vortex pyrolysis model.

Item Units	Default	
Pyrolysis temperature	°C	500.0
Off-gas combustion efficiency	%	75.0
Char filter efficiency	%	90.0
Product temperature ex condenser	°C	40.0

The NREL pyrolysis model uses the off-gas produced by the process to provide the process heat at a combustion efficiency of 75% with the deficit being made up from fuel gas imported from the utilities. All of the char produced in the processes is exported from the model as a by-product and is credited to the model. The capital cost utilised in the NREL vortex pyrolysis model is the same cost algorithm as developed for the generic flash pyrolysis process. The utilities requirements are developed in a similar way to those for the Generic flash pyrolysis model.

6.8. Summary

The four processes modelled in this Chapter demonstrate the different types of flash pyrolysis systems available. The pyrolysis processes modelled are the Ensyn RTP, Union Fenosa and NREL detailed in Chapter 2 and a user defined or generic model. A comparison of the four pyrolysis processes modelled with respect to the production costs of crude pyrolysis oil has been included in Chapter 10.

The generic model was developed to demonstrate the effect pyrolysis reactor temperature and feed moisture content have on the product yields (See Section 6.3.) from flash pyrolysis. The mass balance for the model is based on empirical relationships incorporating the pyrolysis temperature, feed moisture content and plant feed input. The model has been designed for energy self-sufficiency with the char and off-gas being available for combustion as fuel. The reaction temperature and feed moisture content have been shown to affect the energy self-sufficiency of the generic process and hence the fuel gas requirements from the utilities.

CHAPTER 7.

THE HYDROTREATING MODEL.

7.1. Introduction.

As described previously in Chapter 3, crude pyrolysis oil may be upgraded by hydrotreatment either partially to produce a more stable product or fully to produce a more stable hydrocarbon product which has properties similar to light distillate or kerosene [150]. The fully upgraded product may be introduced to a refinery as feedstock and utilised to produce a transportation fuel. Chapter 3 identified the main process variables for a bio-oil hydrotreating process as the pyrolysis oil oxygen content and the deoxygenation extent or severity of the hydrotreating process. Since, upgrading of bio-oil's reduces their affinity for water [3] any bio-oil upgrading process will have a waste water stream and hence the water content of the pyrolysis oil feed (See Chapter 6) to a hydrotreating process will affect the amount of waste water produced.

This Chapter describes the development of a process simulation model for a pyrolysis oil hydrotreating process. The model simulates the effects of deoxygenation extent and pyrolysis oil composition on the hydrotreating process and products. The purpose of the work detailed in this Chapter is not to design, evaluate or review hydrotreating but to utilise the best available information to construct a process simulation of hydrotreating using the identified process variables.

The hydrotreating models incorporated in the program are based on a demonstration plant design [130] commissioned for this work and developed by Battelle PNL who have considerable experience in bio-oil hydrotreating [109, 111, 128, 129]. The process description for the Battelle design was detailed in Section 3.3. The design included a mass balance, equipment specification and cost estimation for a hydrotreating pilot plant deoxygenating crude pyrolysis oil by 98%.

From the provided process design [130], published information on hydrotreating and information from Veba Oil [106] (See Section 3.3.) a mass balance incorporating algorithms allowing variable deoxygenation extent was developed which in turn led to the derivation of the hydrotreating model described in this Chapter.

Section 7.2. describes the scope of the derived hydrotreating model. Section 7.3. describes the design and development of the algorithms which form the basis of the model and Section 7.4. describes the determination of the energy balances and utilities requirements for the model. Section 7.5. details the two off-gas disposal / hydrogen source options available. A summary of the key points from this Chapter is in Section 7.6.

7.2. Scope of the hydrotreating model.

7.2.1. Background.

The hydrotreating model starts with crude, condensed pyrolysis oil fed from one of the flash pyrolysis processes described in Chapter 6. The process utilises a two-stage up-flow reactor, the first stage for stabilisation and the second for upgrading (See Section 3.3.) which has been shown to reduce coking and catalyst deactivation [127].

The main variable in the model is the deoxygenation extent of the crude pyrolysis oil feedstock which controls:

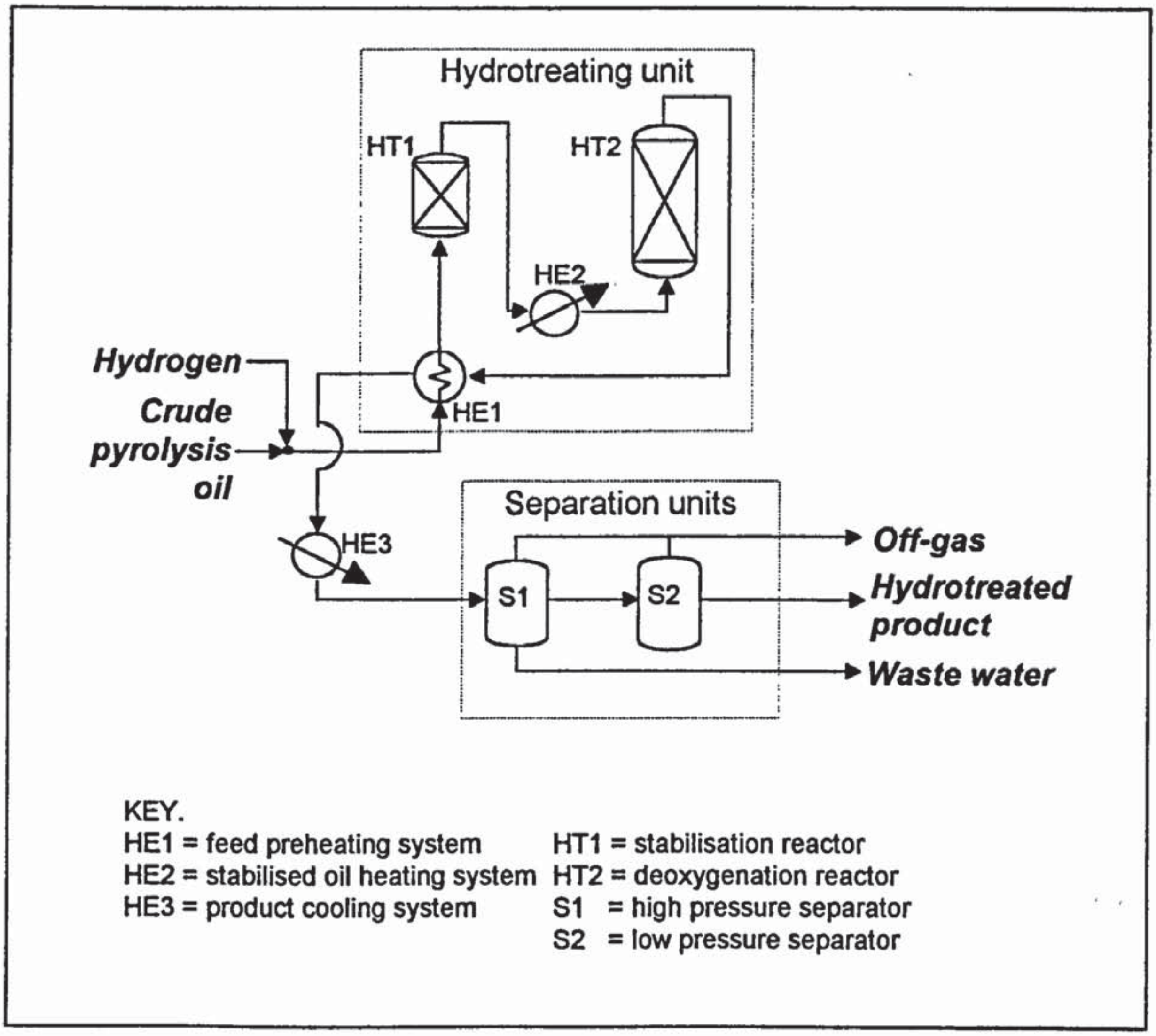
- the hydrogen requirement for the process and hence the operating cost (See Section 7.3.)
- the liquid hourly space velocity and hence the reactor size and capital cost (See Section 7.5.).
- oxygen content of the hydrotreated product and hence the quantities of all products produced, separation of the upgraded oil and water and the quality of the final product (See Section 7.3.).

The model includes two hydrogen sources and off-gas disposal options which may be selected by the User and are :

- the hydrogen rich off-gas is sold as fuel gas to the utilities and all of the required hydrogen is imported from the utilities (See Section 7.5.).
- hydrogen is produced by separation of the hydrogen rich off-gas using for example a membrane and the remaining off-gas is sold as fuel gas to the utilities (See Section 7.5.).

7.2.2. Process description.

The hydrotreating process modelled is based on the Battelle pilot plant design [130] described in Section 3.3. However, the description and a flowsheet are included here for clarity. The flowsheet for the hydrotreating process is shown in Figure 7.1.



KEY.
HE1 = feed preheating system
HE2 = stabilised oil heating system
HE3 = product cooling system
HT1 = stabilisation reactor
HT2 = deoxygenation reactor
S1 = high pressure separator
S2 = low pressure separator

Figure 7.1. Flowsheet for the hydrotreating model.

The process involves the use of two hydrotreatment reactors in series. The crude pyrolyis oil feedstock is mixed with the hydrogen gas stream and preheated over catalyst (to prevent polymerisation and coking) to 250°C before being catalytically stabilised in the first hydrotreating reactor which is maintained between 250-300°C. The stabilised product is preheated before entering a second stage reactor which uses temperatures around 350°C to deoxygenated the stabilised oil.

Operating pressures for the system in the 170-190 bar range are required to keep the products liquid and to maintain the hydrogen partial pressure. The upgraded product from the second stage reactor is cooled to 70-80°C to condense most of the water present. A two-stage separation process separates the off-gas, waste water and upgraded oil products.

The off-gas produced may be sold directly to the utilities as fuel gas or may be fed to a membrane separation and hydrogen recovery unit. These options are discussed further in Section 7.5.

7.3. Design and development of the hydrotreating model.

The development of the hydrotreating model for incorporation into the program consisted of several stages as listed below and detailed in the following subsections:-

1. determining the hydrogen required for deoxygenation of the crude pyrolysis oil (Section 7.3.1.);
2. developing the product yields based on variable deoxygenation of the crude pyrolysis oil (Section 7.3.2.);
3. determining the amount of contamination of the waste water by organics compounds (Section 7.3.3.);
4. calculating the split of the products between the off-gas, product and waste water streams (Section 7.3.4. and Section 7.3.5.); and
5. calculating the second stage reactor size based on the deoxygenation extent (Section 7.3.6.).

7.3.1. Determination of the process hydrogen requirements.

The hydrogen required for the process is calculated from the amount of oxygen present in the crude pyrolysis oil which has to be removed and the split of the oxygen between carbon dioxide and water production. From the Battelle design [130] it was apparent that more water was produced from the hydrotreating process than carbon dioxide, even allowing for the water content of the pyrolysis oil feed. It was apparent from the design mass balance that approximately 85% of the oxygen being removed goes to water and the remaining 15% to carbon dioxide. This was utilised to determine the hydrogen requirement for hydrotreating as shown in equation {1} and the carbon dioxide yield (See Section 7.3.2.).

$$H_{req} = \left(\frac{D_e}{100}\right) \times \left(\frac{P_{O2} \times P_{oil}}{100 \times 2 \times MW_O}\right) \times \left(\frac{W_O \times 8 \times MW_H}{100}\right) \quad \{1\}$$

where		
	D_e	is the deoxygenation extent, (%)
	H_{req}	is the hydrogen required for hydrotreating (tonnes/day)
	MW_H	molecular weight of hydrogen
	W_O	proportion of pyrolysis oil's oxygen extracted as water (%)
	MW_O	molecular weight of oxygen
	P_{oil}	amount of dry pyrolysis oil fed to hydrotreating (tonnes/day)
	P_{O2}	oxygen content of dry pyrolysis oil (wt% dry basis)

When the amount of hydrogen required has been calculated a surplus of 1.5 times [130] the actual amount required is added to enable sufficient hydrogen partial pressure to be maintained in the hydrotreating reactors.

7.3.2. Determination of the yields of products for variable deoxygenation extent.

The yields of the components in the product stream from the second stage reactor (prior to condensation and separation of the products) are based on the yields estimated by Battelle PNL for 98% deoxygenation of crude pyrolysis oil [130]. The composition of the bio-oil feed entering i.e. dry organics and water are utilised to give the "product" composition at a deoxygenation extent of 0%. The composition of the bio-oil feed varies according to the pyrolysis process (See Chapter 6), pyrolysis temperature and feed moisture content (See Chapter 6). The yields of the hydrotreating products at different deoxygenation extents are determined by interpolating between the bio-oil feed composition and the product distribution determined by Battelle. The yields are linearly interpolated due to the lack of data on product yields for lower deoxygenation extents available to establish other trends.

For example, Figures 7.2. and 7.3. show the product yield distributions from a crude bio-oil composed of 82% dry organics and 18% water. As can be seen from Figure 7.2. the yield of C_{5+} organics (represented by the crude pyrolysis oil at 0% deoxygenation) decreases as the deoxygenation extent increases due to cracking into

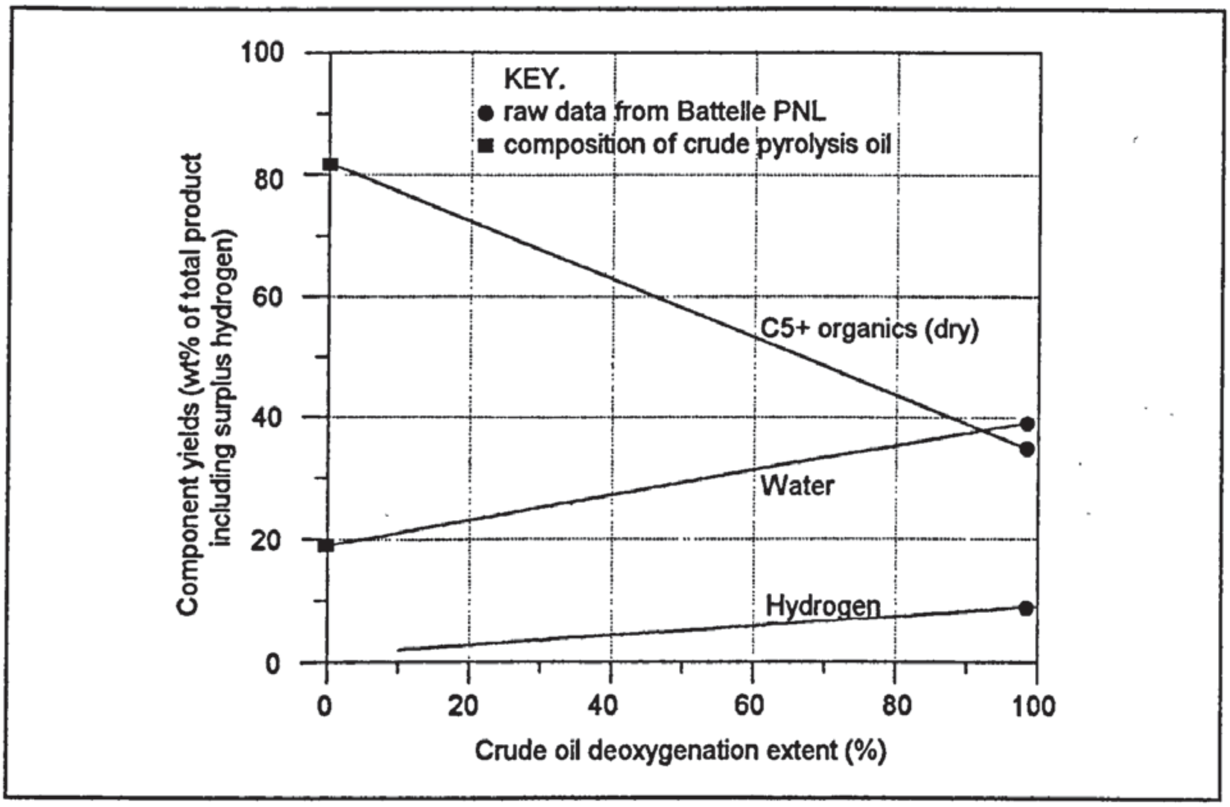


Figure 7.2. The yields of water, C_{5+} organics and hydrogen from the hydrotreating of crude pyrolysis oil (See Figure 7.3. for remaining yields)

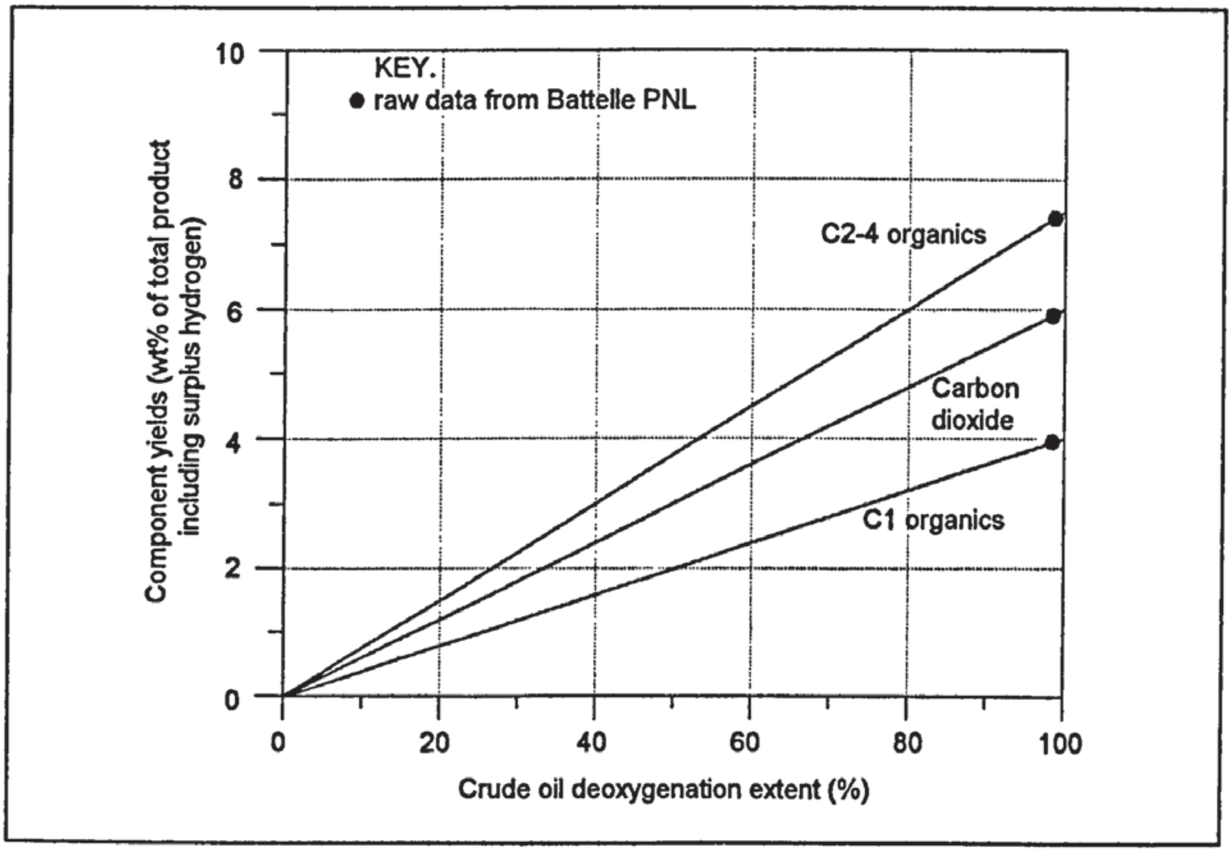


Figure 7.3. The yields of C_1, C_{2-4} organics and CO_2 from the hydrotreating of crude pyrolysis oil (See Figure 7.2. for remaining yields)

smaller chain organics and the formation of carbon dioxide. The yield of water increases from the pyrolysis oil water content shown at 0% deoxygenation as more oxygen is removed from the crude oil and water is formed as a result. The yields of carbon dioxide, C_1 and C_{2-4} organics all increase as the deoxygenation extent increases due to elimination of the oxygen in the dry bio-oil as carbon dioxide and the increased cracking of the longer chain organics respectively. This is demonstrated in Figure 7.3. There is one apparent anomaly in that the amount of surplus hydrogen present in the product stream increases with increasing deoxygenation. However, this can be explained by the hydrogen surplus supplied to the process which is dependent on the deoxygenation extent of the crude bio-oil as explained in Section 7.3.2. The correlations derived for the product yields are given in equations {2} to {7}

$$U_{CO2} \qquad = \left(\frac{D_e}{100}\right) \times \left(\frac{P_{O2} \times P_{oil}}{100 \times MW_O \times 2}\right) \times \left(\frac{C_O \times MW_{CO2}}{100}\right) \qquad \{2\}$$

$$\text{Total Water} \qquad = U_{water} + P_{wat} \qquad \{3\}$$

$$U_{water} \qquad = \left(\frac{D_e}{100}\right) \times \left(\frac{P_{O2} \times P_{oil}}{100 \times MW_O \times 2}\right) \times \left(\frac{W_O \times 2 \times MW_{H2O}}{100}\right) \qquad \{4\}$$

$$U_{C1} \qquad = U_{CO2}/1.45 \qquad \{5\}$$

$$U_{C2\text{-}4} \qquad = 1.283 \times U_{CO2} \qquad \{6\}$$

For $D_e > 0$,

$$U_{C5+} \qquad = P_{oil} - (U_{C2\text{-}4} + U_{C1} + U_{CO2} + U_{water}) \qquad \{7\}$$

where	D_e	is the deoxygenation extent, (%)
	H_{req}	is the hydrogen required for hydrotreating (tonnes/day)
	MW_H	molecular weight of hydrogen
	W_O	proportion of pyrolysis oil's oxygen extracted as water (%)
	MW_O	molecular weight of oxygen

P_{oil}	amount of dry pyrolysis oil fed to hydrotreating (tonnes/day)
P_{O2}	oxygen content of dry pyrolysis oil (wt% dry basis)
P_{wat}	amount of water fed to hydrotreating with the pyrolysis oil (tonnes/day)
U_{water}	reaction water formed from hydrotreating (t/day)
U_{CO2}	carbon dioxide yield from hydrotreating (t/day)
U_{C1}	C_1 organics yield from hydrotreating (t/day)
$U_{C2\text{-}4}$	$C_{2\text{-}4}$ organics yield from hydrotreating (t/day)
U_{C5+}	C_{5+} organics yield from hydrotreating (t/day)
MW_{H2O}	molecular weight of water
MW_{CO2}	molecular weight of carbon dioxide
C_O	proportion of pyrolysis oil's oxygen extracted as carbon dioxide (%)

As there are no published research results for hydrotreating at low deoxygenation extents a comparison of the results obtained using the above correlations cannot be made. It is believed however, that the correlations reasonably represent the effect of deoxygenation on product yields.

7.3.3. Determination of the severity of organic contamination of the waste water stream.

The contamination of the waste water stream with organics and the amount of treatment required on the stream to remove the organic contamination depends on the deoxygenation extent of the crude pyrolysis oil. The more severe the deoxygenation (or complete the hydrotreating) the less miscible the upgraded product is with water (the higher is it's quality as measured by water content) and hence the amount of organics in the waste water is reduced. Veba Oil [106] had carried out trials on bio-oil hydrotreatment and the results indicated that above 88% deoxygenation the upgraded organics and water components are no longer miscible and two discrete streams are formed. At 0% deoxygenation there is no waste water stream because the bio-oil feed is unchanged and is completely miscible with the water present. There is no data available on the amount of water produced or the amount of organics present in the waste water stream for deoxygenation extents between 0 and 88%. However, it is not believed that a waste water stream will be produced at very low deoxygenation extents (See Section 7.3.5.). Veba had generated data for the organic content of the waste water stream for deoxygenation extents between 88 and 98 [106] which was extrapolated to give equation the percentage of the C_{5+} organics that are removed in the waste water stream as shown in equation {8}.

For deoxygenation extents between 10 and 88%

$$O_w = (-4.303\times10^{-6}) - (2.38\times10^{-6}\,D_e) + (3.77\times10^{-6}\,D_e^2) - (4\times10^{-9}\,D_e^3) \qquad \{8\}$$

where D_e is the deoxygenation extent, (%)

O_w is the quantity of C_{5+} organics present in the waste water stream (percentage C_{5+} organics remaining in product stream after the off-gas is removed).

7.3.4. Determination of the hydrotreating off-gas composition.

The final off-gas composition depends on the separation stages (as shown in Figure 7.1.) and whether any hydrogen recycle is utilised. The gas streams from each separation stage are combined to form the final gas product. The proportion of components in each stream is taken from the Battelle design [130]. The gas stream from the first separation (high pressure [130]) stage is shown in Table 7.1. below. The proportions of the components are based on the vapour product stream from the hydrotreating reactor (the composition of which is given by the equations in Section 7.3.2.).

Table 7.1.
Composition of the off-gas from the first separation stage
(The compositions as given as proportions of the total yields from the hydrotreating process.

Component	Proportion of the total yield (%)
Hydrogen (% surplus hydrogen)	99.5
C5+ organics (% total C5+ produced)	4.0
Methane (% total methane produced)	98.0
C2-4 organics (% total C2-4 organics produced)	88.5
Carbon dioxide (% total CO2 produced)	96.0
Water (% total water produced)	0.5

If hydrogen recovery is used to recover some of the surplus hydrogen, it is performed on the product stream from the first separation stage before it is mixed with the off-gas stream from the second separation stage (See Section 7.5. for details on the hydrogen recovery options). The composition of the gas stream after hydrogen recovery will depend on the recovery process used for example, if membrane separation is utilised, the composition will be as in Table 7.1. but with most of the hydrogen removed [130].

The products not removed as off-gas in the first separation stage are fed into the second separator (low pressure [130]). The off-gas stream from the second separation stage contains 0.5% of the total surplus hydrogen, 87% of the remaining carbon dioxide (4% left after the first separation), 0.2% of the remaining C_{5+} organics (95% left after the first separation), 87.5% of the remaining methane (2% left after the first separation) and 51% of the remaining C_{2-4} organics (11.5% left after the first separation). This is shown as proportions of the total yields from hydrotreating, ex-reactor in Table 7.2.

Table 7.2.
Composition of the off-gas from the second separation stage.
(The compositions as given as proportions of the total yields from the hydrotreating process.

Component	**Proportion of the total yield (%)**
Hydrogen (% surplus hydrogen)	0.5
C_{5+} organics (% total C_{5+} produced)	19.0
Methane (% total methane produced)	1.75
C_{2-4} organics (% total C_{2-4} organics produced)	5.87
Carbon dioxide (% total CO_2 produced)	3.48

7.3.5. The composition of the upgraded oil product.

The composition of the product leaving the hydrotreater will vary with the deoxygenation extent. Removing oxygen from the oil during upgrading affects the miscibility of the product with water as well as the quantity of water produced. This is due to the fact that removing the oxygen decreases the polarity of the oil and hence it's hydrogen bonding with water. One important measure of the quality of a hydrotreated oil is, therefore, it's water content.

At deoxygenation extents of 88 - 100% it has been demonstrated that the product oil is no longer miscible with water [106]. There is no data available to suggest how the miscibility of the upgraded oil with water is affected by deoxygenation extents outside of this range. However, at low deoxygenation extents, possibly in the region of 0 - 10%, there will be saturation or "stabilisation" of the bio-oil by the hydrogen and consequently no water will be formed. These two cases fix the top and bottom of a water production versus deoxygenation curve. At some point in the central region of the deoxygenation range the water will start to separate out from the hydrotreated oil as it becomes less polar. The actual mechanisms and the deoxygenation extent at which this occurs are unknown as there have been no trials conducted in this central deoxygenation zone. A preliminary curve has been assumed for water content versus deoxygenation and is shown in Figure 7.4. For this curve the water loss from the bio-oil is assumed to be gradual as the upgrading process occurs and the oxygen remaining in the oil is reduced. An alternative scenario would be a curve in which the water content of the bio-oil remains constant until a certain deoxygenation extent is reached at which point there is a sudden loss of water. Both curves were discussed with Battelle and the gradual water loss curve as shown in Figure 7.4. was validated as being reasonable and more likely to occur based on their knowledge of bio-oil hydrotreatment [151].

It was not possible to accurately characterise the upgraded oil water content curve by a single equation to give a relation for water content and deoxygenation extent because of it's shape. This problem has been overcome by splitting the curve into five regions as shown and characterising the curve for each region. The relations derived are shown in equations {9} to {13} and have been incorporated into the model to give the amount of the total water present remaining in the upgraded product.

For deoxygenation extents between 0 and 10 %

$$W_r \quad = \quad 100 \qquad \{9\}$$

For deoxygenation extents between 10 and 40 %

$$W_r \quad = \quad 108.0 - 1.483D_e + 0.085D_e^2 - 0.0017D_e^3 \qquad \{10\}$$

For deoxygenation extents between 40 and 70 %

$$W_r \quad = \quad 570.0 - 21.767D_e + 0.290D_E^2 - 0.0013D_e^3 \qquad \{11\}$$

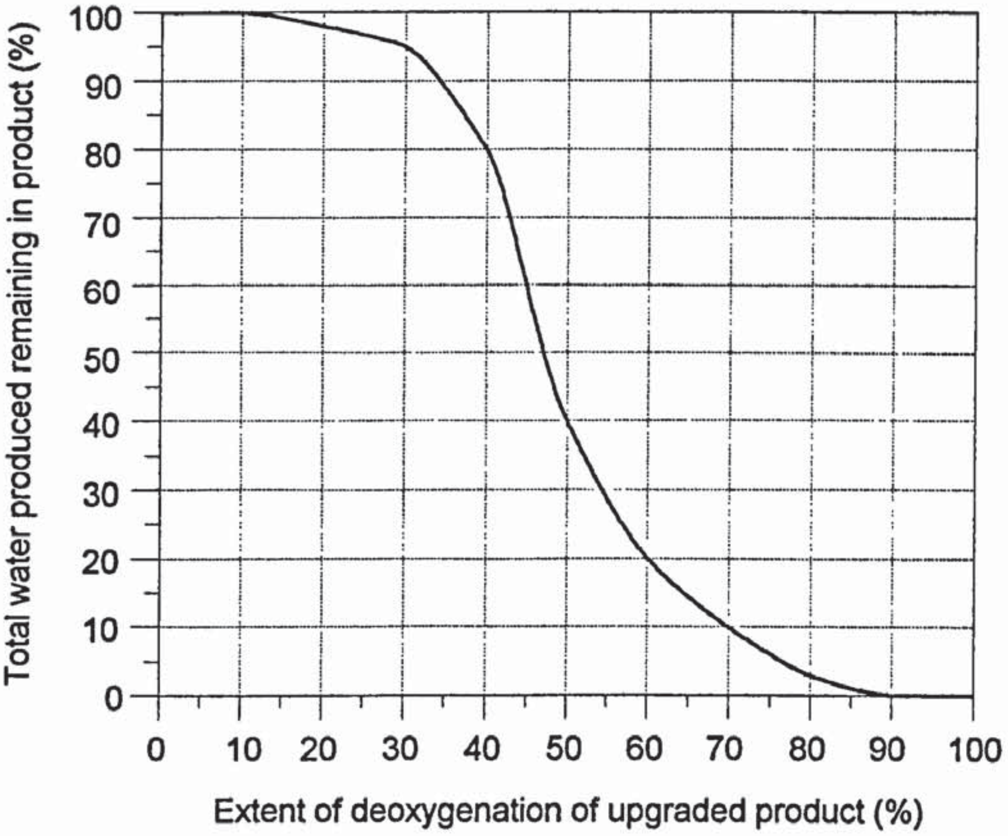


Figure 7.4. Hydrotreated oil water content as a function of deoxygenation extent.

For deoxygenation extents between 0 and 88 %

$$W_r = -127.482 + 6.601\, D_e - 0.096 D_e^2 + 0.0043 D_e^3 \qquad \{12\}$$

For deoxygenation extents between 88 and 100 %

$$W_r = 0 \qquad \{13\}$$

where D_e is the deoxygenation extent (%),

W_r is the proportion of the total water produced remaining in the upgraded product oil (%).

The quantity of the other components in the final upgraded product is determined by the two separation processes to remove the gases as detailed in Section 7.3.4. The quantity of C_{5+} organics depend on the deoxygenation extent which determines the organic contamination of the waste water stream (See Section 7.3.3.) and the quantity lost to the off-gas stream during the separation process. There will be no hydrogen present as that is all removed in the separation processes and is contained in the off-gas stream. The quantities of methane, carbon dioxide and C_{2-4} organics present in the product are 0.25, 0.5 and 5.5% of the total amount produced respectively [130].

7.3.6. Determination of the second stage reactor's size.

Battelle PNL provided some data relating the oxygen content of the upgraded product to the liquid hourly space velocity (LHSV) of the second stage reactor [130]. This data was plotted and characterised as shown in Figure 7.5. to give a relation between the product oxygen content and the LHSV of the upgraded products in the second stage reactor. The relation developed is shown in equation {14}.

$$H_O = 12.976 + 11.664\, Log(LHSV) \qquad r^2 = 0.992 \quad \{14\}$$

where		
	Ho	is the hydrotreated oil oxygen content (wt%)
	$LHSV$	is the liquid hourly space velocity (vol oil/vol cat) of the upgraded product

Veba Oel [106] also provided some oxygen content/LHSV data from trials which is shown in Figure 7.5. with that provided by Battelle [130]. The Veba data provides a similar shape curve to that from Battelle however the data gives lower oxygen contents for a similar space velocity. The Veba data is potentially very interesting as it indicates that smaller reactors could be used to achieve the same deoxygenation extents as in the Battelle system. However, it is not clear whether the Veba system is similar to that of Battelle (i.e. a two stage system with preheating of the crude pyrolysis oil) and so for consistency the Battelle data was used in the development of the model .

The quantity and oxygen content of the pyrolysis oil feed is known from the flash pyrolysis model and can be used with the deoxygenation extent to determine the hydrotreated oil oxygen content (wt%). The Liquid Hourly Space Velocity for the hydrotreating system is calculated using equation {14} and is then used to determine the volume of the second stage reactor by rearranging equation {15} [152].

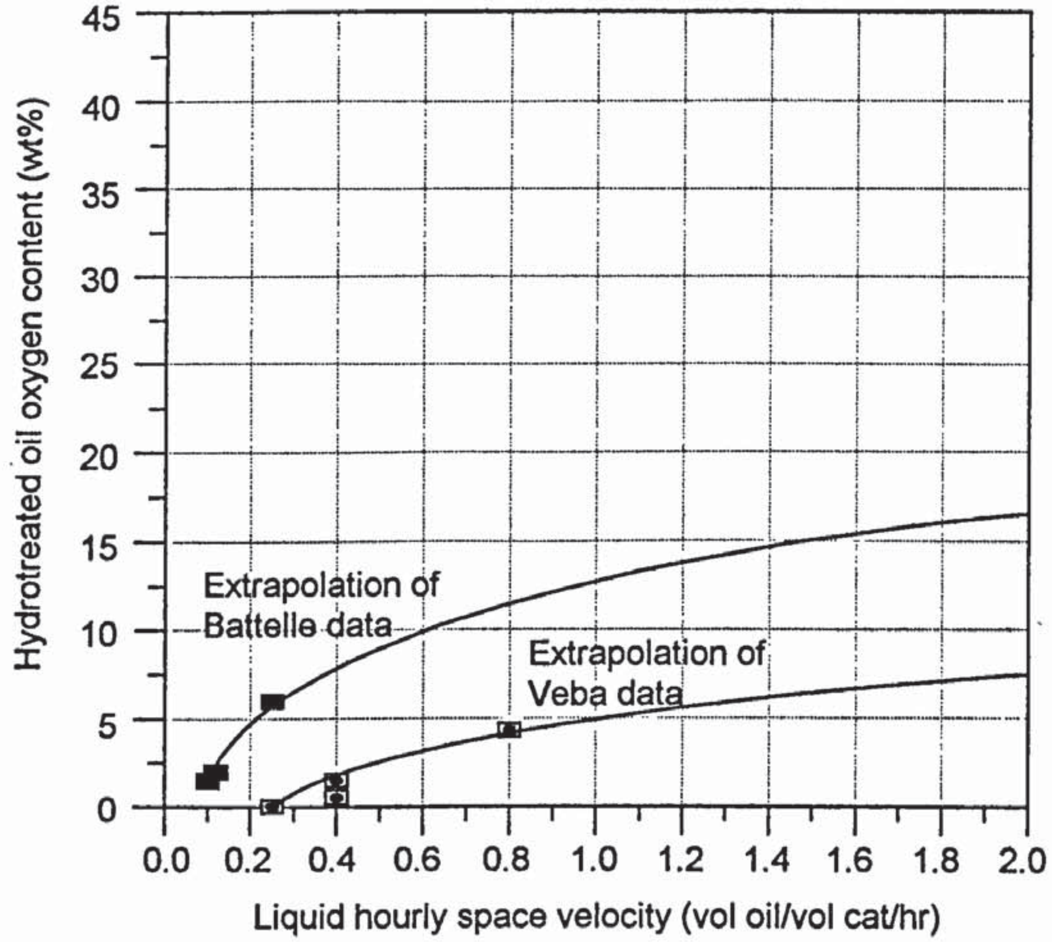


Figure 7.5. The hydrotreated oil oxygen content as a function of it's LHSV in the second stage reactor.

Liquid hourly space velocity (LHSV) $= \dfrac{\left(V/_{t}\right)}{R}$ {15}

where V is the volume of entering feed at specified process conditions (m^3) in time t

t is time (hours)

R is the void volume of the second stage reactor (m^3).

The second stage reactor is the most important item of equipment in the plant as it is where the upgrading process itself takes place. The size of this reactor is determined both by the deoxygenation extent of the process and the throughput of the plant. The reactor subsequently determines the size and cost of the downstream processing of the upgraded liquid product and off-gas. It was decided to utilise the size of the second stage reactor to determine the cost of the hydrotreating plant. The Battelle design [130] included the cost of a hydrotreating pilot plant and this cost is scaled to give a system capital cost for the step model using the size of the second stage reactor as per equation {16}.

$$Hcc = Pcc \times \left(\frac{R}{Rp}\right)^{0.7} \qquad \{16\}$$

where	Hcc	is the capital cost of the hydrotreating plant in thousands of Great Britain pounds
	Pcc	is the capital cost of the hydrotreating pilot plant designed by Battelle PNL (thousands Great Britain Pounds),
	R	is the void volume of the second stage reactor (m^3) in the hydrotreating model as determined by rearranging equation {15}
	Rp	is the void volume of the second stage reactor in the Battelle pilot plant design (m^3) [130].

The viability of using this method to determine the capital cost of the process is discussed further in Section 7.5.

7.4. Determination of the energy balances and utilities requirements.

There are three heat exchange systems in the hydrotreating model which contribute to the energy balance for the system. These are the feed preheater/primary product cooling system (HE1), the stabilised oil heating system (HE2) and the secondary product cooling system (HE3). The three heat exchange systems give the amounts of fuel gas and cooling water required by the hydrotreating model from the utilities. Some of the fuel gas requirement can be offset by the export of the off-gas to the utilities as fuel gas thus reducing the utilities charges. The heat exchange systems are only minor changes on the overall system and will not affect the process yield or capital cost. The amount of power required by the process is calculated by linearly scaling the power requirements of the Battelle plant based on the calculated plant size of the process in the model as per equation {17}.

$$H_p = P_p \times \left(\frac{R}{Rp}\right) \qquad \{17\}$$

where	H_p	is the power requirement of the hydrotreating plant in MW
	R	is the void volume of the second stage reactor (m^3) in the hydrotreating model as determined by rearranging equation {8}
	Rp	is the void volume of the second stage reactor in the Battelle pilot plant design (m^3) [130].
	P_p	is the power requirement of the hydrotreating pilot plant designed by Battelle PNL in kW

7.4.1. The feed preheating system, HE1.

HE1 consists of a heat exchanger and a furnace system as shown in Figure 7.6. The heat exchanger uses the upgraded product from the second stage hydrotreater to preheat the crude pyrolysis oil feed entering the first stage reactor from 50 to 250°C [130]. It also serves as the primary product cooler to reduce the temperature of the product leaving the second stage reactor before it enters the secondary product cooling system, HE3, from 400 to 110°C.

The heat requirement for the cold side to heat the crude feed from 50 to 250°C is calculated and compared with the heat provided by the hot side being cooled from 400 to 110°C. A heat transfer efficiency of 95% is assumed in the calculations to allow for heat losses. The heat deficit is provided by using the furnace burning fuel gas at 75% efficiency. The amount of fuel gas required to fuel the furnace is calculated and purchased from the utilities.

7.4.2. Stabilised oil heating system, HE2.

HE2 is a furnace between stages one and two of the hydrotreating reactor which is used to heat the stabilised oil leaving the first stage at 250°C to the 400°C required in the second stage [130]. The heat required to increase the temperature of the stabilised oil provided by burning fuel gas at 75% efficiency in the furnace. The amount of fuel gas required to meet this demand is calculated and purchased from the utilities. Figure 7.7. illustrates the heating system.

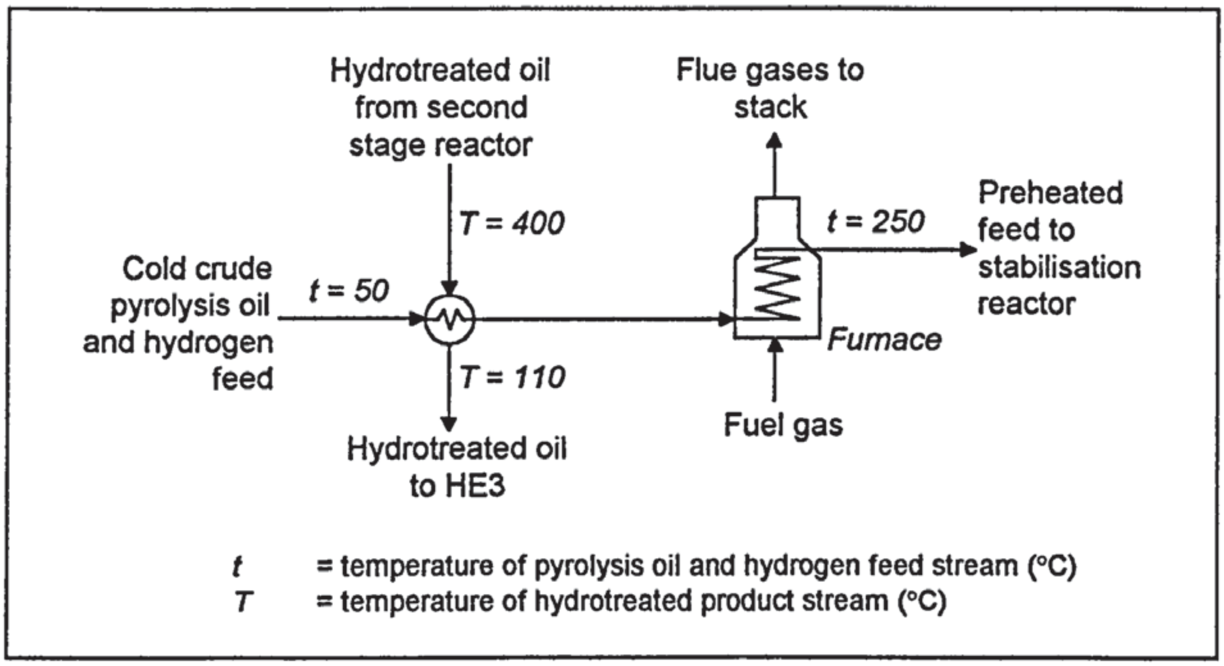


Figure 7.6. Schematic diagram of the feed preheating system, HE1.

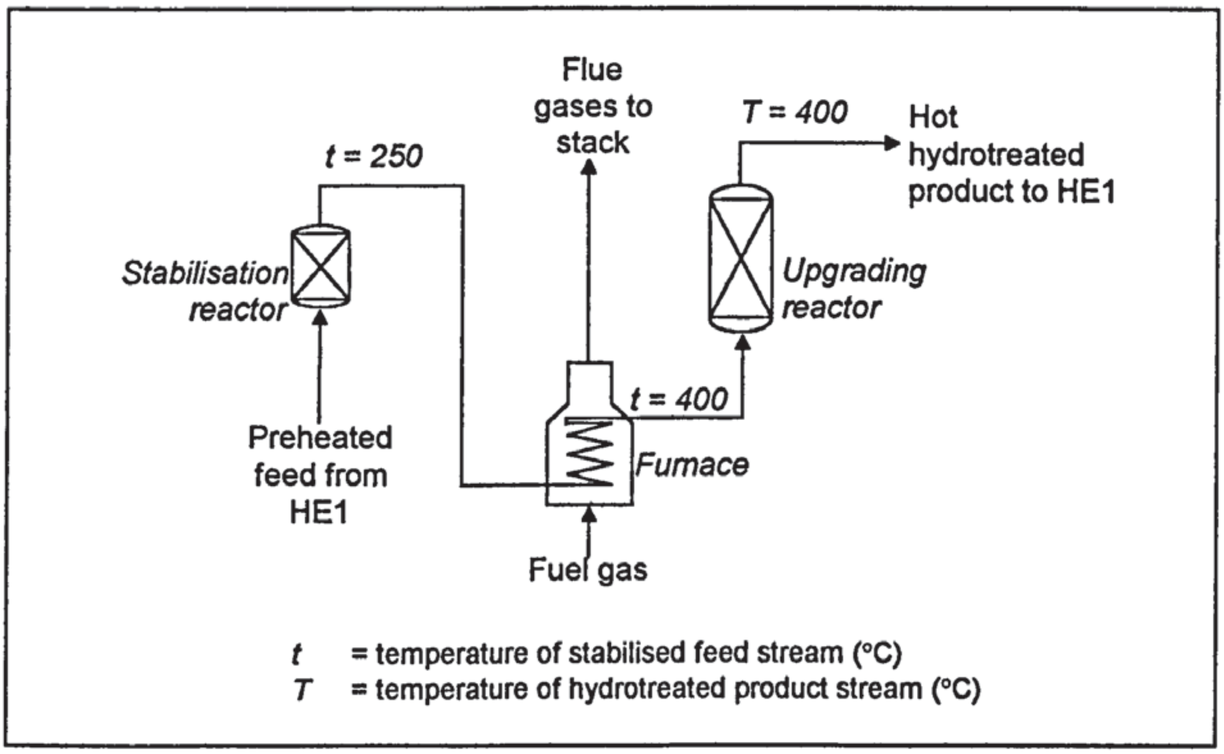


Figure 7.7. Schematic diagram of the stabilised oil heating system, HE2.

7.4.3. Secondary product cooling system, HE3.

HE3 is the second upgraded product cooling system consisting of a water cooled heat exchanger. It is used to cool and condense the upgraded product from the 110°C at which it leaves HE1 to the 73°C required for the separation stages [130]. The cooling duty required to lower the temperature of the upgraded oil is calculated and is met by using cooling water on the cold side of the exchanger with a 20°C temperature rise allowed. A 95% efficiency is assumed on the exchanger. Figure 7.8. illustrates the cooling system.

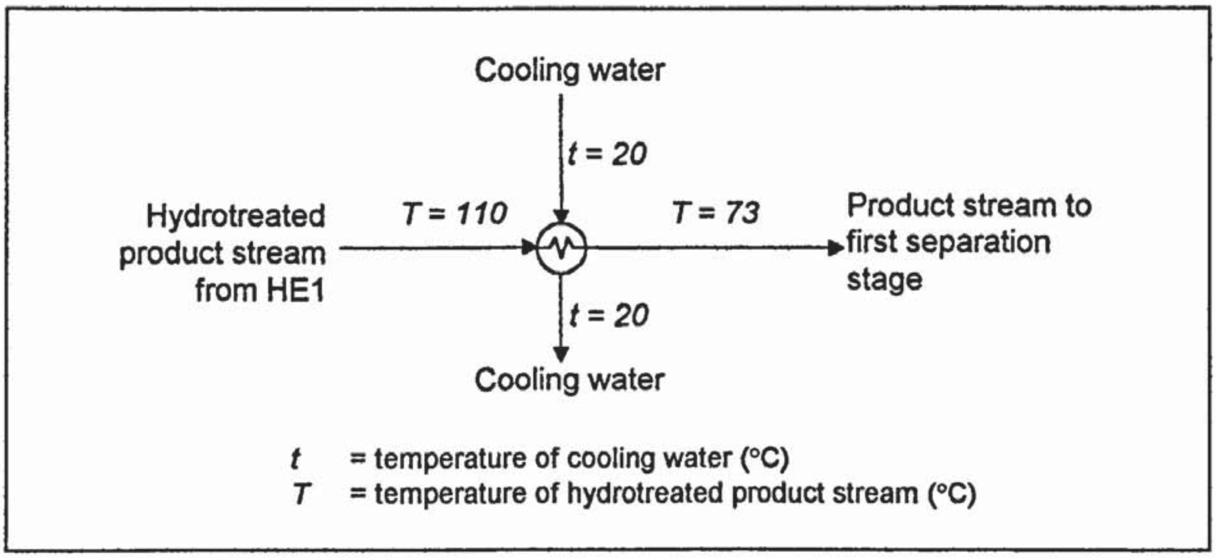


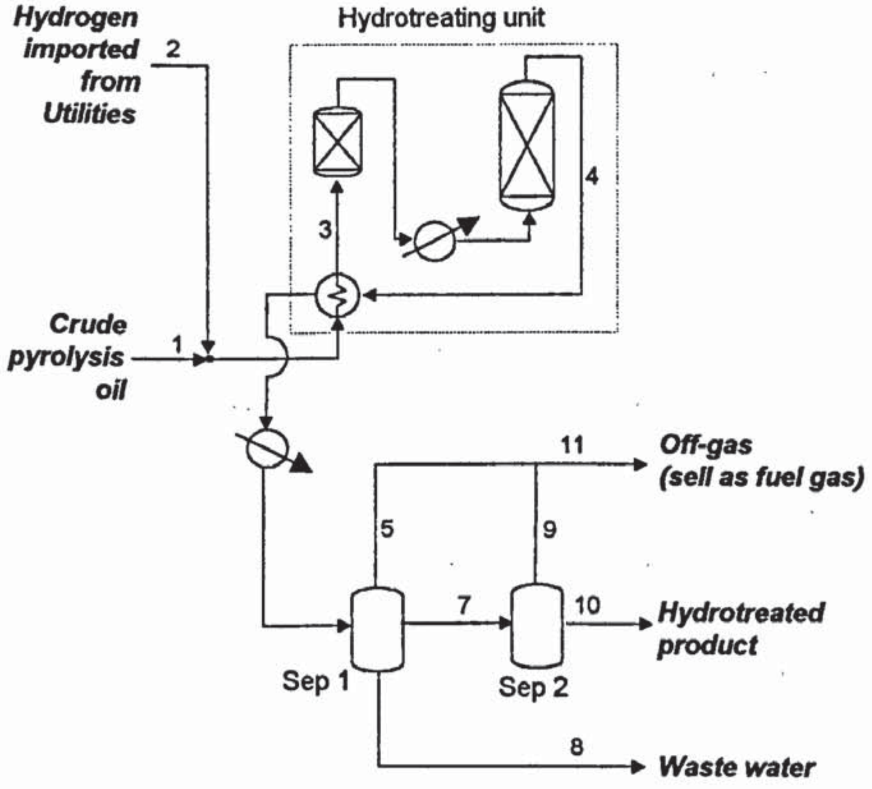
Figure 7.8. Schematic diagram of the secondary product cooling system, HE3.

7.5. Off-gas disposal/hydrogen source options.

There are two hydrogen source/off-gas disposal options that are modelled in the hydrotreating model. These are: selling the off-gas intact and importing all of the hydrogen requirements; and membrane recovery and recycling of the surplus hydrogen, importing the deficit from the utilities. The two options are described in the following subsections.

7.5.1. Hydrotreating option 1 - purchasing all required hydrogen from the utilities

This is the base case option of the program because of the two options modelled it requires the least equipment. The option is attractive because it has the lowest capital cost. It is the most likely option for the first hydrotreating plant built, possibly constructed inside an existing refinery and using cheap hydrogen produced in the refinery. After the off-gas has been cooled and separated out from the hydroteated liquid product and waste water it is sold intact to the utilities as fuel gas for the drying step. The fuel gas is rich in hydrogen due to the hydrogen surplus required by the process to maintain the correct hydrogen partial pressure and hence has a high energy value. The equipment cost for the process is lower than for the second option because gas separation equipment is not required. However, there is a high utilities cost incurred because all of the hydrogen required for the process has to be bought in from the utilities. This option could for example be used in a small scale plant where the low hydrogen recovery does not justify the cost of the off-gas separation equipment. Alternatively, the model would apply to a pyrolysis oil hydrotreater situated in a refinery with a fuel-value hydrogen surplus available. Figure 7.9 shows the schematic flowsheet for the process.



Stream numbers refer to stream numbers on associated mass balance

UNITS.

SEP1 = high pressure separation unit

SEP2 = low pressure separation unit

	STREAMS
1	crude pyrolysis oil feed
2	hydrogen
3	hydrotreater feed
4	hydrotreated product
5	vapour stream from high pressure separation
7	oil stream from high pressure separation
8	waste water stream
9	vapour stream from low pressure separation
10	hydrotreated oil product
11	off gas (Streams 5 & 9)

Figure 7.9. Schematic diagram of the base case hydrogen source / off-gas disposal option.

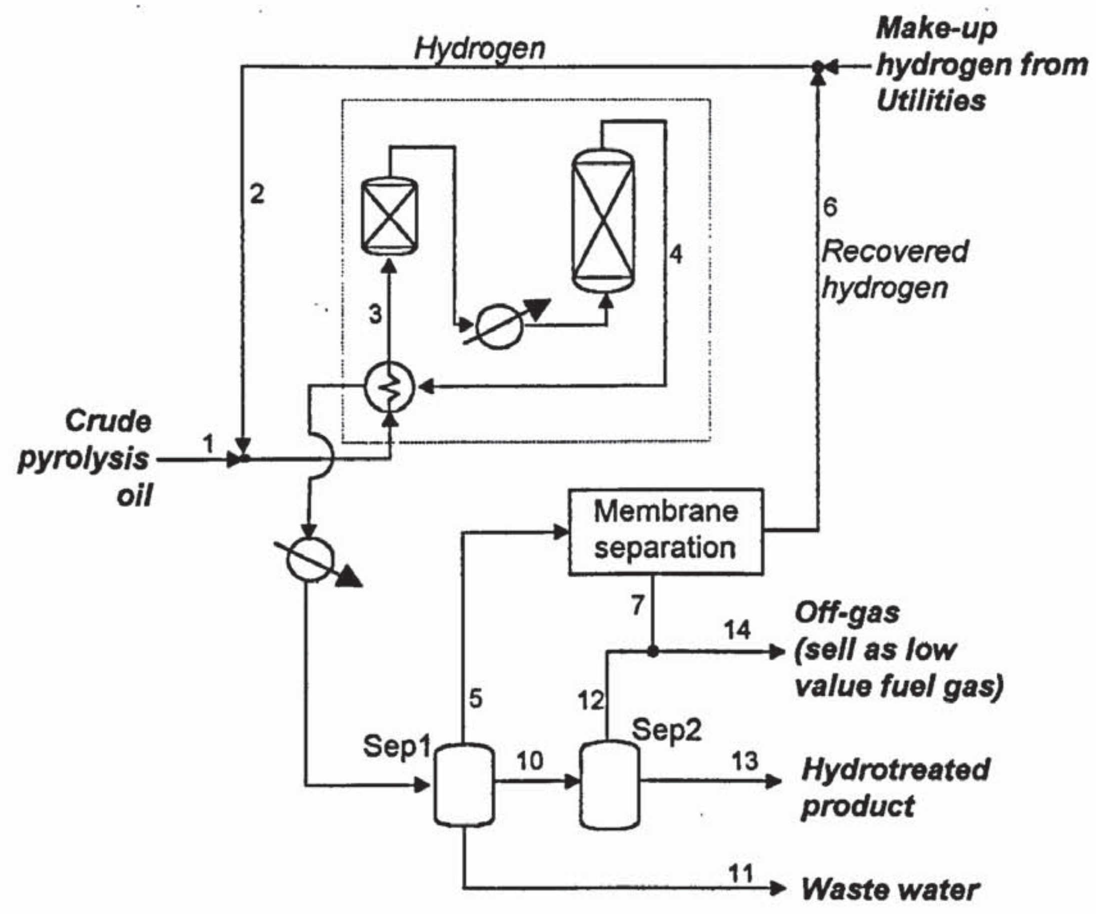
7.5.2. Hydrotreating option 2 - recovering surplus hydrogen from the off-gas by membrane separation.

This hydrotreating option is the case on which the Battelle design was based [130]. It is included in the model to allow a comparison with the base case option described in Section 7.5.1. and to demonstrate one possible hydrogen recovery system out of the possible options highlighted in Chapter 3. This hydrogen source case recovers some of the surplus hydrogen from the off-gas. After cooling the off-gas is separated from the liquid product and waste water as for the base case in Section 7.3.1. The off-gas is then subjected to a membrane separation process to recover the large amount of unused hydrogen in the off-gas. The excess hydrogen requirements for the process after hydrogen recovery and recycling are purchased from the utilities. The amount of additional hydrogen required from the utilities to make up the deficit depends on the efficiency of the membrane separation. The more efficient the membrane separation, the more hydrogen is recovered and the lower the utilities hydrogen requirement. The off-gas remaining after the separation process is sold to the utilities as a low value fuel gas. This membrane separation process involves a potential reduction in the utilities costs from the previous option because less hydrogen is imported from the utilities. However, there is a higher capital expenditure due to the membrane separation equipment. This option could be considered for a medium to large scale plant where the hydrogen demand justifies the purchase of additional equipment to recover the surplus hydrogen. Figure 7.10 shows the schematic diagram for the process.

7.5.3. Utility requirements and capital costs for the two hydrotreating options.

The utility requirements and capital costs for the two hydrotreating options modelled are given in Table 7.3. The basis for the utilities and capital costs are:

- a pyrolysis oil feed (wet) of 78.2 tonnes/day (from 100 d.a.f. tonnes/day wood feed to a generic flash pyrolysis process) to give 30.77 tonnes/day of hydrotreated product;
- the deoxygenation extent of the crude pyrolysis oil is 98% for a pyrolysis oil feed oxygen content of 37% (dry basis);
- a combustion efficiency of 75% is assumed for the fuel gas in the furnaces; and
- the cooling water is assumed to enter the system at 20°C and to leave at 40°C.



UNITS.

SEP1 = high pressure separation unit

SEP2 = low pressure separation unit

STREAMS

1 crude pyrolysis oil feed
2 hydrogen
3 hydrotreater feed
4 hydrotreated product
5 vapour stream from high pressure separation
6 hydrogen recycle
7 waste gas from membrane separator
8 make up hydrogen
10 liquid product from high pressure separator
11 waste water
12 gas stream from low pressure separator
13 hydrotreated oil product
14 off gas (11 & 7)

Figure 7.10. Schematic diagram of the hydrogen recovery by membrane separation option.

From Table 7.3. it can be seen that the power and cooling water requirements for the two hydrotreating processes are the same. This is due to the same algorithm and product cooling systems being utilised in the two options.

Table 7.3.
Utility requirements and capital costs for the two hydrotreating options.

Hydrotreating option :	1 Base Case	2 Membrane recovery
Capital cost		
Installed plant cost (k GB pounds)	8757.9	9094.2
Utilities requirements		
Fuel gas (t /t upgraded product)	- 28.22	- 12.59
Cooling water (t /t upgraded product)	75.66	75.65
Hydrogen (t /t upgraded product)	0.42	0.17
Power (MW)	0.5	0.5

However, there are significant difference in the fuel gas and hydrogen requirements. These are due to the different uses to which the off-gas from the hydrotreating process is put. The off-gas from the first option is exported whole to the utilities and has a high credit (hence the negative value) from the hydrogen content. The value of the off-gas from the second option is lower due to the hydrogen being recovered in the membrane separation unit and recycled. The hydrogen requirements are explained quite simply by the fact that the second option recovers the surplus hydrogen and the first option does not.

It is recognised that the method for determining the capital cost for the hydrotreating systems (i.e. based on the size of the second stage reactor) is unusual. It is also recognised that scaling the capital cost of a pilot plant could lead to an extremely high plant capital cost due to the high cost of pilot plants with respect to full scale plants. However, since hydrotreating plants for biomass pyrolysis oil have not yet gone beyond laboratory scale the cost of a full scale plant cannot be used for comparison. With these points in mind the capital cost of the plants for both options were re-calculated using equipment units as per the Timms method [153]. There are two equations for the Timms method which are shown in equations {18} and {19}. One only takes into account the number of functional units and the plant capacity (shown in equation {18}) while the other takes into account the materials of construction, temperature and pressure effects (shown in equation {19}). Both methods have been used in the derivation of the hydrotreating capital costs for comparison.

Timms (1) $$C = 8300\,N\,Q^{0.615} \qquad \{18\}$$

Timms (2) $$C = 3860\, N\, Q^{0.639}\, F_m\, (T_{max})^{0.066}\, (P_{max})^{-0.016} \qquad \{19\}$$

where	C	is the capital cost (Great Britain pounds, 1992 basis),
	N	is the number of functional units,
	Q	is the plant capacity (tonnes/year for the main product),
	F_m	is the materials of construction factor (which is 1.15 for low grade stainless steel),
	T_{max}	is the maximum process temperature (deg K), 450°C design temperature for the system modelled
	P_{max}	is the maximum process pressure (bar), 207 bar design pressure for the system modelled.

The equipment for hydrotreating option 1 (no recycling or recovery) equates into two units : hydrotreating reactor and hydrogen compressor. The equipment for hydrotreating option 2 (membrane separation and recovery) equates into four units : hydrotreating reactor, hydrogen recovery, hydrogen make-up compressor and the recycle hydrogen compressor. However, a fifth unit composed of a hydrogen cooler may be required so the actual number of units is taken to be 4.5. The results of the capital cost determinations using equations {11} and {12} as well as the derived capital costs are given in Table 7.4.

Table 7.4.
Capital costs for the hydrotreating plant.
(Basis : a pyrolysis oil feed of 78.2 t/day from 100 d.a.f. t/day wood feed through a generic flash pyrolysis process to give 30.77 t/day hydrotreated oil, 98% deoxygenation in the hydrotreating process plant, operating 330 days per year).

Method	**Installed cost in thousands 1993 Great Britain pounds**
<u>Option 1 (no hydrogen recovery)</u>	
Timms (1)	6,040.9
Timms (2)	5,716.3
Scaled pilot plant	8,757.9
<u>Option 2 (membrane separation and recovery)</u>	
Timms (1)	10,873.6
Timms (2)	10,289.4
Scaled pilot plant	9,094.2

As can be seen from Table 7.4 the calculated total capital cost using Timms Units Method are higher than the derived cost for the second option (membrane separation and recovery). This is probably due to the number of equipment units specified in the Timms calculations (4.5. specified) but illustrates that the scaling of the pilot plant cost by the method used is not unreasonable. For the first hydrotreating option (no hydrogen recovery or recycling) the scaled pilot plant cost is considerably higher than those derived from the Timms method and again the difference is probably due to the number of equipment units specified. However, since a bio-oil hydrotreating plant has not yet been constructed and the cost evaluated, the use of scaled pilot plant costs based on the size of the second stage reactor volume to calculate the total capital cost has been left in the hydrotreating model.

7.6. Summary.

The model described in this Chapter simulates the upgrading of biomass pyrolysis oil by hydrotreatment. The bio-oil composition and the deoxygenation extent were identified in Chapter 3 as process variables and are used as the basis for the hydrotreating model. The model uses hydrotreating deoxygenation extent, which can be varied between 0 and 100%, to determine the product yields and distribution, the hydrogen requirement and the size of the hdyrotreating reactor. The model is based on data from hydrotreating trials carried out by Veba Oil and pilot plant design commissioned from Battelle PNL. These were considered the best available sources of data on bio-oil hydrotreatment at the time of modelling. Two hydrogen source options are modelled. One represents membrane separation and recovery of hydrogen to minimise hydrogen consumption and the other represents a system in which all the required hydrogen is purchased.

CHAPTER 8.

ZEOLITE CRACKING, REFINING AND POWER PRODUCTION MODELLING.

8.1. Introduction.

This Chapter deals with the step models for upgrading by zeolite cracking, refining and power production which are included in BLUNT. The refining models take the products from fully hydrotreating or present case zeolite cracking of crude pyrolysis oil to produce diesel and gasoline respectively and are described in Section 8.4. The power production model, which simulates the generation of electricity from the crude or upgraded products from biomass conversion utilising both engines and turbines, is discussed in Section 8.5.

The Liquefaction Group of the IEA Biomass Agreement [30, 154] carried out technoeconomic assessments of the NREL biomass to gasoline process [77] as described in Chapter 3. The process was analysed as "present" and "potential" cases to illustrate the potential effects of further research and development. The present and potential cases catalytically upgrade the primary vapour stream from an NREL vortex pyrolyser (as described in Section 2.6.). The results of the IEA assessments were mass balances, equipment specifications and detailed cost analyses for the two cases [30].

The information derived by the IEA [30] was used to develop the zeolite cracking models detailed in Section 8.2. and 8.3. The mass balances developed by the IEA for the two cases were adapted to allow pyrolysis vapours from all of the pyrolysis processes modelled in Chapter 6 (i.e. generic, Ensyn RTP, NREL vortex and Union Fenosa). Both of the resulting zeolite cracking models are fixed, i.e. have no internal variables which can be altered by the program User, due to the sparcity of information on how changes in processing conditions would affect product yields. The individual reactions which occur in the various sections of the upgrading process have not been considered or evaluated here as the models are based solely on the systems designed by the IEA. The zeolite cracking models have been included in BLUNT to demonstrate the emergent zeolite technology and to allow comparison with the hydrotreatment of crude bio-oil. The capital costs for the models were derived by taking the equipment costs for the zeolite cracking sections from the IEA work [30].

8.2. The present case zeolite cracking model.

8.2.1. Introduction.

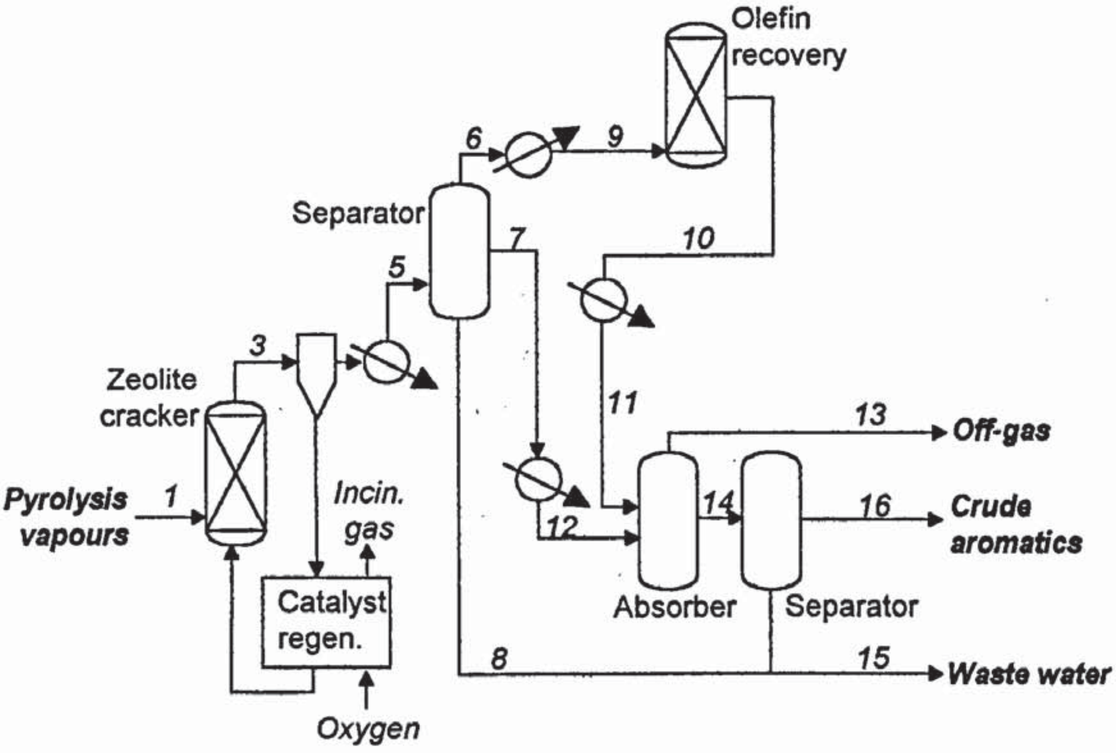
The zeolite cracking present case model represents a process based technology considered to be on commercially available [154]. The product from the process is an aromatic organic which is miscible with fossil fuels and may be utilised as a refinery feedstock to produce gasoline [30]. The yield of the aromatic product is based on experimental results [154]. The IEA process was modified to allow the vapours from any of the modelled pyrolysis processes in Chapter 6 to be utilised and the energy balance was optimised to make the system as energy self-sufficient as possible.

A process description for the present case model is given in Section 8.2.2. and a flowsheet for the process is given in Figure 8.1. Section 8.2.3. describes the derivation mass and energy balances for the model from the IEA data for the process. The derivation of the capital cost for the process is discussed in Section 8.2.4.

8.2.2. Process description.

The hot vapours and gases from the pyrolysis process (after char removal) enter the catalytic cracking reactor (a recirculating fluidised bed [130]) after being preheated with the product stream from the cracker. The cracker consists of a recirculating entrained bed with a 1 - 2 second [154] contact between the pyrolysis vapours and the zeolite catalyst for example, Mobil's HZSM-5. The temperature of the cracking reactor is between 500-525°C with a temperature rise of 20 to 50°C due to the exothermic reactions [154]. Attached to the cracking reactor is a fluidised bed catalyst regenerator in which the coke formed on the catalyst is burned off using air in an oxygen-starved environment. The flowsheet for the present case model is shown in Figure 8.1 [30]. In the reactor, the pyrolysis vapours are cracked and oxygen is removed in the form of water and carbon oxides to produce hydrocarbons.

After the process stream leaves the cracking reactor, it is cooled to condense the hydrocarbons heavier than benzene and any steam. The product stream is split into waste water, gases and condensed organics in the primary separator. The waste water is removed for treatment and disposal. The gas stream, consisting of permanent gases, gaseous olefins and a small amount of benzene vapour, is fed to the fixed-bed secondary catalytic reactor (olefin recovery) where the olefins alkylate to form cumene [154].



Stream Identification

1. Pyrolysis vapour feed
3. Reactor products
5. Cooled products
6. Separation gas stream
7. Separation waste water stream
8. Separation liquid stream
9. Olefin recovery feed
10. Olefin products
11. Olefin products to absorber
12. Liquid product to absorber
13. Absorber off-gas
14. Absorber liquid product
15. Waste water stream
16. Crude aromatic product

Figure 8.1. Flowsheet for the present case zeolite cracking model.

The organic liquids stream from the primary separator is cooled further before being recombined with the olefin stream in the absorber. The absorber product is a crude hydrocarbon mixture, rich in alkylated monocyclic aromatics but low in benzene [154]. It is separated from the waste water which is removed for treatment and disposal. The crude aromatics product is suitable for use in a refinery to produce gasoline. The off-gas stream is "sold" or credited to the utilities as fuel gas.

8.2.3. The mass and energy balances.

The mass balance was derived from one developed by the IEA [30] which has been modified to allow primary vapours to be fed from any flash pyrolysis process. In the IEA mass balance the pyrolysis vapour feedstock has a large amount of steam from the NREL vortex pyrolyser as steam is assumed as the fluidising gas [30] rather than nitrogen. The derived mass balance for the model does not have this steam input as the pyrolysis processes assume that the off-gas can be utilised for fluidisation. This difference in steam inputs to the zeolite cracker lowers the overall amount of waste water for treatment and disposal and alters the energy balance for the process.

The original design developed by the IEA ALPS group [30] was also modified to meet one of the prime objective of BLUNT - energy self-sufficiency. In the modified present case zeolite cracking process the energy balance is optimised by using hot product streams to preheat cold streams. An example of this is the product stream from the cracker, exiting the cyclone. The cracker product stream passes through the cracker feed preheater, the olefin recovery feed preheater and finally through a water cooled heat exchanger to reduce the stream temperature before it enters the primary separator. The off-gas produced from the process is sold as fuel gas to the utilities for use in the feed drying step.

8.2.4. Costs and utility requirements.

The capital cost for the present case step model was derived from the costings developed by the IEA [30]. The IEA developed the cost for a complete processing system pyrolysis from a wood feed to a crude aromatics product. The equipment required for the process was identified and approximately sized. The cost of each piece of equipment was estimated using known relationships between the size of equipment and cost [155] and updated using a CEP cost index. The total installed cost of the plant was based on the equipment costs with factors applied to account for erection, piping, instrumentation, electrics, civils, structures and buildings and installation to give the Direct Plant Cost (DPC). The DPC had further factors applied to account for the indirect costs of design, engineering. construction, overheads, contractor's fee, contingency and interest on capital during construction [30]. The sum of all the direct and indirect costs gave the Total Plant Cost.

The capital cost for the present case zeolite cracking model was developed by removing the pyrolysis and pretreatment sections from the IEA specification [30] to leave the cost of the zeolite cracker and ancillary equipment. The resultant cost, based on 1000 tonnes/day wood feed, was utilised in the expression given in equation {1} which allows the derived cost to be scaled according to the size of the system and gives a capital cost in 1993 Great Britain Pounds.

$$\text{Capital cost (1993 pounds)} \quad = 15.24 \times 10^6 \times \left(\frac{F_{in}}{1000}\right)^{0.7} \tag*{\{1\}}$$

where F_{in} is the d.a.f. feed input to the conversion plant i.e. the raw biomass feedstock (tonnes/day)

The utilities required by the present case zeolite cracking model include oxygen, cooling water, power and fuel gas. The oxygen requirement is based on the amount of oxygen required by the catalyst regeneration unit and is taken from the derived mass balance. The fuel gas requirement is based on the amount of fuel required by the cracker and the olefin feed preheaters. However, it is modified by the amount of off-gas which is exported to the utilities as fuel gas. The power requirement is based on the equipment specification from the IEA work [30] and is given by equation {2}.

$$\text{Power required (MW)} \quad = 4.991 \times \left(\frac{F_{in}}{1000}\right) \tag*{\{2\}}$$

where F_{in} is the d.a.f. feed input to the conversion plant i.e. the raw biomass feedstock (tonnes/day)

In addition to the utilities mentioned above there are charges incurred for waste water treatment and catalyst rejuvenation. However, since there was limited information available arbitrary values have been assigned to these charges. The catalyst replacement charge is given by equation {3}. It is based on the IEA design which specifies 0.7kg catalyst lost per tonne of upgraded product where the catalyst costs 8.81\$/kg [30]. The waste water treatment charge is given by equation {4} this is based on a value of £10/tonne of organics present in the waste water stream. The value of the charge is arbitrary since there was no other information available.

$$\text{Catalyst rejuvenation charge (pounds)} \quad = 3.63 \times Prod \tag*{\{3\}}$$

$$\text{Wastewater charge (pounds)} \quad = 10 \times W_{org} \tag*{\{4\}}$$

where	$Prod$	is the amount of aromatic product produced from the zeolite cracking present case model (tonnes/day)
	W_{org}	is the amount of organics which are present in the waste water stream (tonnes/day)

8.2.5. Design uncertainties and critical features.

The present case model has a relatively low degree of uncertainty as regards the technologies utilised. The reactor is based on a fluidised catalytic cracking reactor used in the petroleum industry to crack heavy hydrocarbons to gasoline products [30]. The product yields are based on experimental data in a small fixed bed reactor, rather than the fluidised bed specified here. Olefin recovery is being commercialised in the USA to lower the benzene content of gasolines and the capture of butenes and higher hydrocarbons from the fuel gases in the absorption column should be straightforward for the temperature and pressures specified.

8.3. The potential case zeolite cracking model.

8.3.1. Introduction

The zeolite cracking potential case model assumes progress in research and development to attain a modified process which produces gasoline directly without the need of a refining step [30]. A process description for the potential case model is given in Section 8.3.2. and a flowsheet for the process is given in Figure 8.2. Section 8.3.3. describes the mass and energy balances for the process. The IEA process was modified to allow the vapours from any of the modelled pyrolysis processes in Chapter 6 to be utilised and the energy balance was optimised to make the system as energy self-sufficient as possible. The derivation of the capital cost and utilities requirements for the process is discussed in Section 8.3.4.

8.3.2. Process description.

The pyrolysis vapours and gases enter the catalytic cracking reactor after passing through the char cyclones. The recirculating entrained bed reactor has a 1 - 2 second contact between the pyrolysis vapours and the zeolite catalyst. The temperature of the reactor is between 500 and 525°C with a temperature rise of 25 to 50°C from the exothermic reactions. The catalyst regenerator is as for the present case model. The flowsheet for the future case model is shown in Figure 8.2. [30]

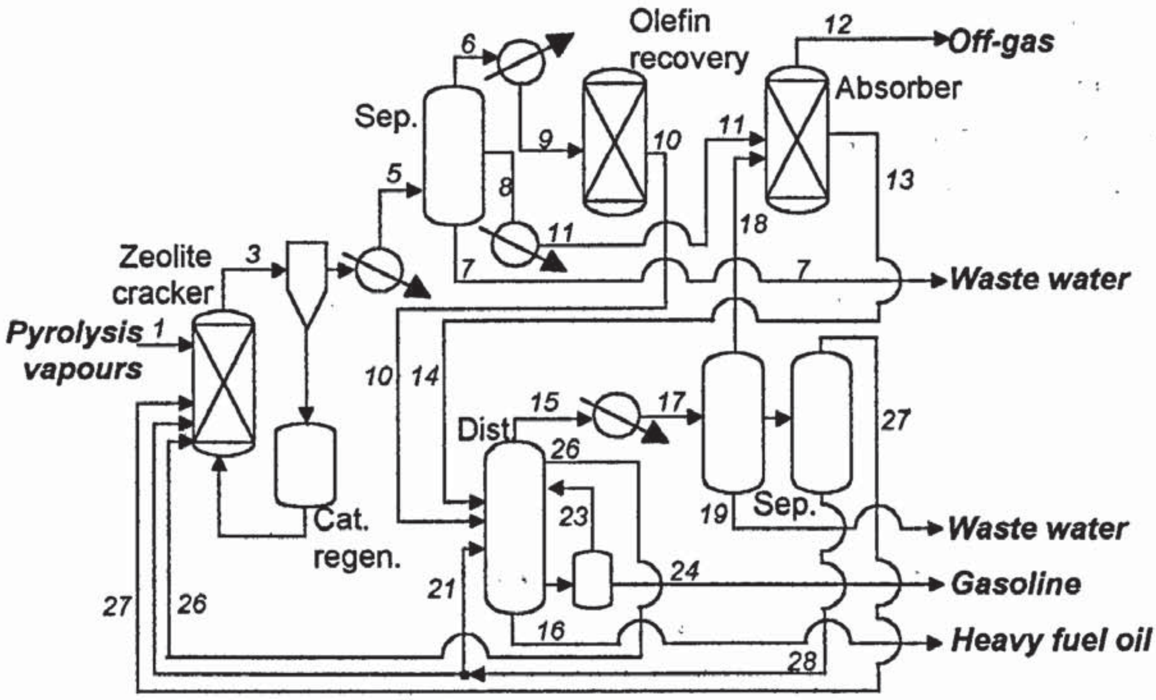
After the process stream leaves the reactor it is cooled to condense most of the steam and the hydrocarbons heavier than benzene. The permanent gases, gaseous olefins and benzene vapour are compresses to 100 psig before entering the secondary catalytic reactor (olefin recovery) [154]. In the fixed bed secondary reactor the olefins alkylate to form cumene. The hot vapours from the secondary reactor enter the distillation column where they supply the heat for distillation. Unreactive butanes and pentanes along with gas oil are recycled to the cracking unit to be transformed into reactive intermediates. A small amount of polycyclic aromatic oil may be drawn off the bottom of the distillation tower and can be burned for fuel, hydrocracked to gasoline in a refinery or sold as fuel oil [30]. The volatile hydrocarbons are recovered by absorption or condensation from the noncondensible gases in the process stream. The hydrocarbon product condensed at atmospheric pressure is cooled and pumped into the absorber.

8.3.3. Mass and energy balances.

The mass balance was derived from the mass balance developed by the IEA ALPS group [30] with similar adjustments as specified in Section 8.2.3. As for the present case model the energy balance is optimised by using hot product streams to preheat cold streams. An example of this is the product stream from the cracker, exiting the cyclone. The cracker product stream passes through the cracker feed preheater, the olefin recovery feed preheater and finally through a water cooled heat exchanger to reduce the stream temperature before it enters the primary separator. The off-gas produced from the process is sold as fuel gas to the utilities for use in the feed drying step.

8.3.4. Capital costs and utility requirements.

The capital cost for the present case step model was derived from the spreadsheet costings developed by the IEA [30] as described in Section 8.2.4. The pyrolysis and pretreatment sections of the spreadsheet were removed to leave the cost of the zeolite cracker and ancillary equipment. The derived cost expression resulting from this for the potential case model is given in equation {5} which gives a capital cost in 1993 Great Britain Pounds.



Stream Identification

1. Pyrolysis vapour feed
3. Reactor products
5. Cooled products
6. Separation gas stream
7. Separation waste water stream
8. Separation liquid stream
9. Olefin recovery feed
10. Olefin products
11. Liquid product to absorber
12. Absorber off-gas
13. Absorber product
14. Absorber product to distillation
15. Distillation gas stream
16. Heavy fuel oil
17. Distillation gas to separation
18. Separation off-gas
19. Waste water stream
20. Separation liquid stream
21. Fuel oil
22. Distillation gasoline stream
23. Distillation steam
24. Hot gasoline stream
25. Gasoline product stream
26. Fuel oil recycle
27. Hydrocarbon gas recycle
28. Gasol recycle

Figure 8.2. Flowsheet for the potential case zeolite cracking model.

Capital cost (1993 pounds) $= 15.85 \times 10^{6} \times \left(\frac{F_{in}}{1000}\right)^{0.7}$ {5}

where F_{in} is the d.a.f. feed input to the conversion plant i.e. the raw biomass feedstock (tonnes/day)

The utilities required by the potential case zeolite cracking model include oxygen, cooling water, power, steam and fuel gas. The oxygen requirement is based on the amount of oxygen required by the catalyst rejuvenation unit and is taken from the mass balance. The fuel gas requirement is based on the amount of fuel required by the cracker and the olefin feed preheaters. However, it is modified by the amount of off-gas which is exported to the utilities as fuel gas. The power requirement is based on the equipment specification from the ALPS work [30] and is given by equation {6}.

$$\text{Power required (MW)} \quad = 4.991 \times \left(\frac{F_{in}}{1000}\right) \qquad \{6\}$$

where F_{in} is the d.a.f. feed input to the conversion plant i.e. the raw biomass feedstock (tonnes/day)

In addition to the utilities mentioned above there are charges incurred for waste water treatment and catalyst rejuvenation these utilise the same algorithms described in Section 8.2.4.

8.4. The models for refining upgraded bio-oil.

The upgraded product from either the hydrotreating or present case zeolite cracking steps could be further upgraded by refining. It is common practice in the oil industry to pay a service charge to a refinery in order to have a crude feedstock refined to a more valuable product. As the products from both full hydrotreating and present case zeolite cracking are miscible with fossil fuels this would also be possible for bio-derived feedstocks.

The refining models in the program take fully hydrotreated bio-oil and crude aromatics (from present case zeolite cracking of crude bio-oil) to produce diesel and gasoline respectively. The product from the potential case zeolite cracking model is gasoline and requires no further refining, as described in Section 8.3. A 90% yield of refined product from the upgraded feedstock and a refining service charge of 35US$/tonne of refined product produced [30] are utilised to give the yields and costs for refining.

8.5. The power production model.

As well as liquid fuels, electricity is a potential product from biomass thermal conversion [156], and simple models have been incorporated in BLUNT. The power output from both engines and turbines are derived from correlations for the process efficiencies. The correlations were derived by another researcher who is modelling electricity production from biomass from information from private sources and published data [157]. Detailed designs or investigations have not been carried out on power generation as this was not part of the original scope of the project. The model has been incorporated into BLUNT to allow the program user to consider the possibility of using biomass derived products to produce power instead of transportation fuels.

The power production model work by firstly determining the potential energy output from the liquid product in MWth and from that determines the power production system utilised. If the potential energy output is below 5MWth diesel engines are selected and above 20MWth gas turbines are selected. These ranges have been based on accepted power generation techniques for example, above 20MW engines would not be utilised for power generation due to equipment size and efficiency considerations. Between 5 and 20MW the use of either engines or turbines is acceptable and so the program user is given an option of which to use. The first calculation in the power generation model determines the available energy entering the model and is determined by equation {7}.

$$E_{in} = \frac{F_{in} \times HHV}{t} \qquad \{7\}$$

where	E_{in}	energy input to the system (MWth)
	F_{in}	biomass derived feed input to the power generation plant (tonnes/day)
	HHV	higher heating value of the biomass derived feed (MW/tonne)
	t	conversion factor to convert from days in seconds (equivalent to 864000 seconds/day)

Simple relationships [157] linking the conversion efficiencies of engines and turbines with the energy input and are shown in equations {8} and {9}. The derived conversion efficiency of the system is then utilised to determine the actual power output from the liquid product using equation {10}. The capital cost for the system is derived from equation {11} [157].

$$S_e \text{ for engines} = 39.88 + 0.4343 \times \mathrm{Ln}(E_{in}) \qquad \{8\}$$

$$S_e \text{ for turbines} = 22.46 + 0.4343 \times \mathrm{Ln}(E_{in}) \qquad \{9\}$$

$$P_{out} = E_{in} \times \frac{S_e}{100} \qquad \{10\}$$

$$\text{Capital Cost (pounds)} = 466.43 \times 10^3 \times (P_{out})^n \qquad \{11\}$$

where	S_e	efficiency of power generation system (%)
	P_{out}	power output (MW) from power generation system
	n	scale factor, equal to 1 for engines and 0.7 for turbines [157].

The only utility used in the power generation model is power itself. The power consumption is equivalent to 10% of the electricity produced from the power generation system [157]. The uncertainty of the relationships detailed above must be regarded as high as the range of data utilised for their derivation is unknown. However, the power production model does provide some indication of the energy output possible from a biomass conversion system and the costs incurred and for this reason the model was been included in BLUNT

8.6. Summary.

This Chapter detailed the zeolite cracking, refining and power production models included in BLUNT. The zeolite cracking models are based on technoeconomic assessments carried out by the IEA on the NREL biomass to gasoline process. The models represent present and potential cases and produce crude aromatics suitable for blending and gasoline respectively. The refining model upgrades the products from the hydrotreating and zeolite cracking present case model to produce diesel and gasoline respectively and is based on paying a service charge for using an existing refinery. The power production model uses provided correlations to determine the cost and efficiency of producing electricity from biomass.

CHAPTER 9.

MACROECONOMIC MODELLING.

9.1. Introduction.

9.1.1. Background.

The previous Chapters in this thesis have dealt with the description and modelling of the technologies utilised for the thermochemical conversion of biomass to produce liquid fuels and electricity. The process modelling Chapters (Chapters 5 to 8) have dealt with the direct factors which affect a biomass conversion process for example, the feedstock (See Chapters 5 and 6), the pyrolysis temperature (See Chapter 6), the pyrolysis technology utilised (See Chapter 6) or the effect of upgrading (See Chapters 7 and 8). However, the process modelling factors are not the only elements which affect the economic viability of a biomass conversion process.

This Chapter considers some of the broader financial or "macroeconomic" issues which may affect a biomass conversion process and models their effects on the production cost of a biomass derived product. An example of a macroeconomic issue which might affect a biomass conversion process is the Non-Fossil Fuel Obligation (NFFO) [33] which is a tax to subsidise the use of non-fossil fuels e.g. nuclear or renewable.

9.1.2. The macroeconomic issues considered.

Specific macroeconomic issues which might influence the financial viability of a biomass conversion process were identified and are listed below. The issues are mainly policy related but some are determined by the market mechanisms and the physical nature of the biomass conversion processes. The issues can mainly be categorised into positive issues or "incentives" which reward the use of biomass derived or non-fossil fuels and energy and negative issues or "deterrents" which punish the use of fossil derived fuels and energy.

The issues which are "incentives" for using biomass derived or non-fossil fuels and energy are:

- Subsidies and/or credits to biomass converters for using low-sulphur feedstocks for example, wood. Although this is an incentive to biomass converters it is also a deterrent to fossil fuel users as it effectively increases the cost of fossil derived fuels and energy in relation to biomass derived fuels and energy.

- Subsidies and/or credits to biomass feedstock producers which are related to the area of land planted with an energy crop or as incentives to dispose of food crop by-products such as straw without burning or as incentives to "set-aside" land which was being utilised for food crops and either do not farm it or grow fuel crops.
- Allowances or credits to biomass feedstock producers or converters in respect of their input costs, for example employment credits which effectively reduce their labour costs and hence the production costs of the biomass derived fuels and energy.
- Pricing policy relative to competing fuels - direct price cross subsidies for fuels which the converter seeks to encourage. For example, a subsidy could be placed on the cost of biomass derived gasoline due to it's being produced by from an environmentally friendly feedstock.
- Allowances or credits to biomass converters for example, capital grants or allowances.
- Special electricity tariffs. For example, the Non-Fossil Fuel Obligation (NFFO) [33] incumbent upon the major power generators - minimum prices have been set by legislators to raise artificially the rates of return on biomass projects to encourage investment and/or offset the risk.

The issues which are "deterrents" for using fossil derived fuels and energy are:

- Measures to control the use of high sulphur fuels, for example a tax on high sulphur content fossil fuels. This is the same as the first issue under incentives as in addition to increasing the cost of fossil derived fuels and electricity it would indirectly decrease the cost of biomass derived fuels and electricity.
- Costs of compliance with current and future emissions standards. For example, penalties and taxes for high sulphur emissions from the use of fossil fuels.
- Taxes on competing fuels, for example if carbon taxes are applied on fossil fuels and biomass derived fuels and energy are exempt from the tax, there will be a significant influence in prices in favour of biomass derived products.

There are also some factors which are ambiguous in that they neither strictly encourage the use of a biomass derived fuel or discourage the use of fossil fuels. These factors include :

- Water and waste disposal charges. Increased water charges and waste disposal charges will affect both biomass converters and mainstream industry as both utilise cooling or process water and produce waste water streams.
- Water supply for feedstock production for example, the cost of securing a reliable supply and the risk attached to losing it. This could actually be a deterrent against using biomass derived fuels as an increase in the cost of water for feedstock production due to limited supply would directly increase the cost of the feedstock being used in a biomass conversion plant and hence the production cost of the biomass derived product.
- Non-specific subsidies and credits, relating to Development Zones. For example, incentives which encourage businesses to set up in areas that have high unemployment and could actually relate to both biomass converters and mainstream industry.
- Costs associated with planning, licensing and permitting applications for the biomass conversion plant. These costs could be a deterrent to biomass converters but since they are probably also applied to conventional industry could in fact be regarded as neutral.
- Set-up costs, for example connecting plant to power supply and connecting pipe to user (as for planning costs).
- Taxes on electricity generation or combustion plants. These taxes strictly speaking relate to the use of fossil fuels which are used to generate electricity. However, since the taxes will be passed on to the electricity consumers in the form of increased electricity charges they will relate to biomass converters i.e. electricity used for lighting. The costs will also be passed on to mainstream industry and so overall the taxes could be regarded as neutral.

The macroeconomic factors outlined above were evaluated in terms of their significance for determining the relative costs and benefits of biomass processes vis-à-vis other energy processes. Those factors which were considered to be neutral between different energy processes were considered to be less important than factors which were likely to significantly influence the viability of a process (measured by the delivered cost of a GJ) compared with other processes. From the range of factors considered four were selected for detailed analysis:-

1. measures to control the use of high sulphur fossil fuels;
2. carbon taxes on fossil fuels;
3. specific capital subsidies or credits; and
4. feedstock irrigation water resources.

ECOTEC Research and Consulting Ltd (United Kingdom) [33] were contracted to investigate the specific effects the four selected macroeconomic factors would have, if implemented, on the production costs of biomass derived products. The information generated by ECOTEC in respect of the four factors [33] was analysed and used to develop macroeconomic models for BLUNT. The background and models for each of the selected factors are considered in turn in Sections 9.2. - 9.5. Section 9.6. gives a sample calculation to demonstrate the effect of the macroeconomic factors on the production cost of a biomass derived product and Section 9.7. summarises the key issues raised in this Chapter.

9.2. The sulphur tax on fossil fuels model.

A sulphur tax model is included in the macroeconomics section of BLUNT in order to show the advantage of using biomass or low sulphur fuels with respect to fossil fuels. The model allows a true comparison between a sulphur taxed fossil fuel and a biomass derived and hence low sulphur renewable fuel.

9.2.1. Background.

The sulphur tax model is designed to measure the effect of controlling the use of high sulphur fuels. The impetus behind this macroeconomic factor was the European Commission's Large Combustion Plant Directive (LCPD) in 1988 [33] which set out national percentage reductions in sulphur emissions from power stations to be met by 2003. The ways in which the LCPD can be met are by using fuels with a lower sulphur content (e.g. biomass derived fuels) or by removing sulphur from the atmospheric emissions.

Sulphur taxes, a levy per unit on sulphur rich fuels used in power stations, are not so far used in the European Union. However, in 1991 the Swedish government instituted a 30 Swedish Kroner/kg sulphur emitted tax on sulphur emissions hoping to reduce emissions by 10-25 kilotonnes by the end of the decade.

9.2.2. Modelling.

The sulphur tax model works by imposes a tax on high sulphur content fossil fuels equivalent to an arbitrary default value of 50 Great Britain pounds/1% sulphur present in the fuel. The model then determines the effect the tax has on the price of electricity generated from the high sulphur fuels. This is utilised to determine the "price advantage" of using biomass derived (and hence low sulphur) fuel. The "price advantage" is the effective amount by which the sulphur tax decreases the biomass derived fuel production cost. The main information used by the sulphur tax model includes:

- the sulphur tax per tonne of fuel (default value of £50 per 1% sulphur in the fuel);
- the amount of each fossil fuel utilised by power stations in a year (default values of 76.64 million tonnes/year for Coal, 364.4 thousand barrels/year for Oil, 5.53 million m^3/year for Natural Gas based on UK usage in 1991);
- the pool electricity price (default value of 2.2 pence/kW based on UK costs in 1991);
- the total delivery of power from the power stations per year (default value of 248325 GWh/year based on UK usage in 1991); and
- the proportions of coal, oil and gas used to supply the power (default values of Coal: 61%, Oil: 19%, Gas: 5% based on UK usage in 1991).

The default values of the above parameters are given in BLUNT as a guide to the program user and may be altered by the program User if required. The remaining information required by the model includes the production cost of the desired liquid product per tonne, the total utilities power requirement for the conversion process and the product stream flowrate per day. This information is taken directly from the results of a simulation run by BLUNT and does not have to be entered by the program User. The calculations involved in determining the "price advantage" of a biomass fuel in the sulphur tax model are given in equations {1} to {4} below.

The first step in the calculation is to determine the Uplifted Pool Electricity price (U_{pp}) which is the new electricity price incorporating a sulphur tax on the coal, oil and gas utilised for power generation. This is shown in equation {1}. The second step in the calculation is to determine an Uplifted Fuel Cost (U_{fc}) for the biomass derived fuel based on the new pool electricity price. This is shown in equation {2}. The third part of the calculation is to recalculate the production cost of the biomass derived fuel to give the Reduced Fuel Cost (R_{fc}) as in equation {3}. Finally, the Price Advantage (Pa_{stax}) of using a biomass derived and hence low sulphur fuel is calculated in equation {4}.

$$U_{pp} \quad = \frac{S_{tax} \times (a+b+c)}{P_d \times 10^6} + E_{pp} \qquad \{1\}$$

when $a = (S_o \times R_o \times P_o \times 10)$, $b = (S_g \times R_g \times P_g \times 10^4)$ and $c = (S_c \times R_c \times P_c \times 10^6)$

$$U_{fc} \quad = F_{cost} + \left[\left(T_{power} \times F_{out}\right) \times \left(\left(U_{pp} - E_{pp}\right) \times 10\right)\right] \qquad \{2\}$$

$$R_{fc} \quad = U_{fc} - \frac{S_{tax} \times \left(\left(S_c \times P_c\right) + \left(S_o \times P_o / 100\right) + \left(S_g \times P_g / 100\right)\right)}{100} \qquad \{3\}$$

$$PA_{stax} \quad = F_{cost} - R_{fc} \qquad \{4\}$$

where	S_{tax}	is the sulphur tax for a one percent sulphur fuel (Great Britain pounds per tonne),
	subscript c, o or g	represent coal, oil and gas respectively
	S_x	is the sulphur content of a particular fossil fuel (%),
	R_x	is the quantity of a particular fossil fuel required by power stations per year (millions tonnes of coal, millions m^3 of gas or thousands barrels for oil),
	P_x	is the amount of a particular fossil fuel utilised by power stations as a percentage of the total fossil fuel used (%),
	P_d	is the power delivery by power stations using fossil fuels (GWh),
	E_{pp}	is the current electricity pool price (pence / kWh),

F_{cost}	is the biomass derived fuel production cost (Great Britain pounds / tonne),
T_{power}	is the total power usage in the biomass conversion process (MWh), and
F_{out}	is the quantity of biomass derived fuel produced in the conversion process (tonnes / hour)

9.3. The carbon tax on fossil fuels model.

A carbon tax model is included in the macroeconomics section of BLUNT due to the continuing concern over carbon dioxide emissions from the use of fossil fuels. Although the use of biomass derived fuels releases carbon dioxide this is offset by the amount of carbon dioxide fixed from the atmosphere when new biomass is growing [5]. Hence, biomass derived fuels are exempt from the carbon tax since they are in effect carbon dioxide neutral. The model allows a comparison between a carbon taxed fossil fuel and a biomass derived renewable fuel.

9.3.1. Background.

The carbon tax on fossil fuels model looks at measures to reduce carbon dioxide emissions. Carbon dioxide emissions in Europe are growing by 4% per annum with 31% of the emissions coming from power generation [33]. In 1990 at the European Joint Energy/Environment Council it was decided to stabilise carbon dioxide emissions at 1990 levels by 2000. To this end, the CEC proposed a package of fiscal and monetary measures aimed at improving energy efficiency and discouraging the use of fuels which generate large quantities of carbon dioxide when combusted. The proposed tax is composed of an Energy tax on all non-renewable fuels plus large scale hydro-schemes and a Carbon tax levied on all fossil fuels.

Non-energy use of fuels such as oil used as a feedstock for petrochemical manufacturers or in the production of man-made fibres are exempt from the tax. The tax has the heaviest impact on coal and oil and favours renewable fuels and nuclear power (non-fossil and so exempt from the tax). However, to date the carbon tax is still under discussion and has not been implemented.

9.3.2. Modelling.

The carbon tax model imposes a tax on fossil fuels and determines the effect this has on the price of electricity utilised by biomass converters. This is utilised to determine the "price advantage" of using biomass derived fuels. The "price advantage" is the effective amount by which the carbon tax decreases the biomass derived fuel production cost. The information required by the carbon tax model includes:

- the carbon tax imposed on each fossil fuel (default values of 1.71 Great Britain pounds/tonne for Coal, 20 Great Britain pounds/barrel for Oil, and 0.015 Great Britain pounds/m^3 for Natural Gas);
- the amount of each fossil fuels consumed by power stations per year (default values of 76.64 million tonnes/year for Coal, 364.4 thousand barrels/year for Oil and 5.53 million m^3/year for Natural Gas based on UK usage in 1991);
- the pool electricity price (default value of 2.2 pence/kWh based on UK costs in 1991);
- the total delivery of power from the power stations per year (default value of 248325 GWh/year based on UK usage in 1991); and
- the proportions of coal, oil and gas used to supply the power (default values of Coal: 61%, Oil: 19%, Gas: 5% based on UK usage in 1991).

The default values of the above parameters are given in BLUNT as a guide line to the program user and may be altered if required. The remaining information required by the model includes the production cost of the biomass derived product per tonne, the total utilities power requirement for the process and the product stream flowrate per day. This information is taken directly from the program by the carbon tax model and does not have to be entered manually by the program user. The calculations involved in determining the "price advantage" of a biomass fuel in the carbon tax model are given in equations {5} to {8} below and follow a similar process to those used in the sulphur tax model.

The first step in the calculation is to determine the Uplifted Pool Electricity price (U_{pp}) which is the new electricity price incorporating carbon taxes on the coal, oil and gas utilised for power generation. This is shown in equation {5}. The second step in the calculation is to determine an Uplifted Fuel Cost (U_{fc}) for the biomass derived fuel based on the new pool electricity price. This is shown in equation {6}. The third part of the calculation is to recalculate the production cost of the biomass derived fuel to give the Reduced Fuel Cost (R_{fc}) as in equation {7}. Finally, the Price Advantage (Pa_{ctax}) of using a biomass derived fuel is calculated in equation {8}.

$$U_{pp} = \frac{(a+b+c)}{P_d \times 10^6} + E_{pp} \qquad \{5\}$$

when $a = (C_o \times R_o \times P_o \times 10)$, $b = (C_g \times R_g \times P_g \times 10^4)$ and $c = (C_c \times R_c \times P_c \times 10^6)$

$$U_{fc} = F_{cost} + \left[\left(T_{power} \times F_{out}\right) \times \left(\left(U_{pp} - E_{pp}\right) \times 10\right)\right] \qquad \{6\}$$

$$R_{fc} = U_{fc} - \frac{\left(\left(C_c \times P_c\right) + \left(C_o \times P_o \times 7.3\right) + \left(C_g \times P_g \times 1379\right)\right)}{100} \qquad \{7\}$$

$$PA_{ctax} = F_{cost} - R_{fc} \qquad \{8\}$$

where	subscript c, o or g	represents coal, oil and gas respectively
	C_x	is the carbon tax on a particular fossil fuel (Great Britain pounds/tonne for Coal, Great Britain pounds/barrel for oil and great Britain pounds/m^3 for Natural Gas),
	R_x	is the quantity of a particular fossil fuel required by power stations per year (millions tonnes of coal, millions m^3 of gas or thousands barrels for oil),
	P_x	is the amount of a particular fossil fuel utilised by power stations as a percentage of the total fossil fuel used (%),
	P_d	is the power delivery by power stations using fossil fuels (GWh),
	E_{pp}	is the current electricity pool price (pence / kWh),
	F_{cost}	is the biomass derived fuel production cost (Great Britain pounds / tonne),
	T_{power}	is the total power usage in the biomass conversion process (MWh),

F_{out} is the quantity of biomass derived fuel produced in the conversion process (tonnes / hour)

9.4. The irrigation water pricing model.

The water pricing model has been included since it has a direct impact on the cost of a biomass feedstock. Water pricing is of particular interest in Southern Europe which has a shortage of water but grows large amounts of biomass for food or fuel. The water pricing model is double edged since the change in irrigation water cost may either be positive or negative i.e. the cost of water may increase or decrease allowing for location and water availability.

9.4.1. Background.

The irrigation water pricing model is designed to consider the effect of irrigation water pricing on a the production cost of a biomass derived fuel. Water availability and pricing policies vary across Europe and the water consumed by crop irrigation is a function of industrial development and climate. Southern States tend to have high levels of irrigated land as a proportion of total cultivated land. For example, 4.23% of cultivated land is irrigated in Germany while in Italy the figure is 23.6% [33]. In States with agriculture dominated economies a high proportion of total water withdrawals tend to be used for irrigation. Two types of water pricing policy exist:

- those based on the volume of water withdrawn for irrigation; and
- flat rate pricing structures.

The irrigation water pricing policy adopted for the model is based on the volume of water used for irrigation. Such systems are common in developed economies and typically include accurate measuring systems to ensure fair charging. Such systems also encourage efficient irrigation as users have a direct financial incentive to cut down their water usage.

9.4.2. Modelling.

The water pricing model assesses the effect of altering the cost of irrigation water on the cost of the biomass feedstock. This is utilised to determine the "price advantage" of using biomass derived fuels. The "price advantage" in this case is the effective amount by which the water pricing model decreases or increases the biomass derived fuel production cost. The information required by the irrigation water pricing model includes:

- the current (old) irrigation water cost including all fixed and variable charges (default value of 25 pence/m^3 based on UK prices in 1991);
- the area under cultivation of the feedstock (default value of 4000 hectares, arbitrary and alterable by the program user);
- the typical water requirement per hectare per year (default value of 1500 m^3/ha/year, arbitrary and alterable by the program user); and
- the new irrigation water cost (default value of 27 pence/m^3, arbitrary and alterable by the program user).

The default values of the above parameters are given in BLUNT as a guide to the program user and may be altered if required. The remaining information required by the model includes the feedstock cost, the production cost of the product per tonne, the feedstock input to the plant, the number of days per year that the plant operates and the product stream flowrate per day. This information is taken directly from the program by the irrigation water pricing model and does not have to be entered manually by the program user. The calculations involved in determining the "price advantage" of a biomass fuel in the irrigation water pricing model are given in equations {9} to {14}.

The first step in the calculation is to determine the Current Irrigation Water Cost (C_{wc}) per year in millions of Great Britain pounds, this is carried out in equation {9}. The second step is to calculate the Reduced Water Charge Burden (R_{wc}) which is the difference between the C_{wc} and the water cost per year using the new irrigation water charge. This is carried out in equation {10}. The third step is to calculate the Adjusted Feed Cost (A_{fc}) which is the cost of the feedstock to the biomass conversion process taking into account the change in irrigation water pricing. This is carried out in equation {11}. The fourth step in the process is to calculate the Adjusted Product Cost (A_{pc}) for the biomass derived product based on the A_{fc} as per equation {12}. The fifth step is to calculate the Uplifted Fuel Cost (U_{fc}) as shown in equation {13}. The sixth and final step is the calculate the Price Advantage (PA_{wat}) of biomass fuels based on the change in irrigation water pricing as per equation {14}.

$$C_{wc} = \left(\frac{C_c}{100}\right) \times \left(\frac{F_a \times 10^3 \times W_r}{10^6}\right) \quad \{9\}$$

$$R_{wc} = \left(\frac{N_{wc}}{100}\right) \times \left(\frac{F_a \times 10^3 \times W_r}{10^6}\right) - C_{wc} \quad \{10\}$$

$$A_{fc} = \left(\frac{R_{wc} \times 10^6}{F_{in} \times D_{op} \times 0.2}\right) + F_{cost} \quad \{11\}$$

$$A_{pc} = P_{cost} + \left(A_{fc} - F_{cost}\right) \times \left(\frac{F_{in}}{P_{out}}\right) \quad \{12\}$$

$$U_{fc} = A_{pc} - (A_{fc} - F_{cost}) \quad \{13\}$$

$$PA_{wat} = P_{cost} - U_{fc} \quad \{14\}$$

where	R_{wc}	is the reduced water charge burden (millions of Great Britain pounds),
	C_c	is the current cost of irrigation water per cubic meter (pence / m^3),
	F_a	is the area under cultivation of feedstock (thousands of hectares),
	W_r	is the water requirement per hectare each year (m^3 / ha / year),
	N_{wc}	is the new cost of irrigation water per cubic meter (pence / m^3),
	F_{in}	is the feedstock flowrate to the biomass conversion plant (tonnes/day, wet basis),
	F_{cost}	is the original cost of the feedstock (Great Britain pounds/tonne, dry basis),
	D_{op}	is the number of days per year for which the biomass conversion plant operates (days/year),
	P_{out}	is the amount of product produced by the biomass conversion process per day (tonnes/day)

P_{cost} is the production cost of the biomass derived product without any macroeconomic factors being applied (Great Britain pounds/tonne).

9.5. The regional assistance model.

The regional assistance model has been incorporated because it is one of the most significant programmes for bio-energy. It encourages biomass producers to grow non-food crops and encourages the building and running of biomass conversion facilities using grants and subsidies. Without such assistance biomass converters would not be able to compete with the low priced fossil derived fuels and energy available at present.

9.5.1. Background.

The regional assistance model considers the use of incentives which facilitate the establishment of biomass conversion plants. The incentives include grants which lower the amount of capital which the investor needs to raise and service. The total risk of the project is therefore lowered, allowing the construction of larger, more efficient units.

9.5.2. Modelling.

The regional assistance model assesses the effect of a range of local grants on the capital cost of a biomass project. This is utilised to determine the "price advantage" of using biomass derived fuels. The "price advantage is the effective amount by which the regional assistance model decreases the biomass derived fuel production costs. The information required by the regional assistance model includes:

- employment credit as a percentage of the annual labour costs (default value of 25%);
- the proportion of the workforce that is eligible for the employment credit (default value of 50%); and
- the amount of capital grant or allowance (default value of 2000 thousand Great Britain pounds).

The default values of the above parameters are given in BLUNT as a guide line to the program user and may be altered if required. The remaining information required by the model includes the labour cost, the plant life, the production cost of the product per tonne, the number of days per year that the plant operates and the product stream flowrate per day. This information is taken directly from the program by the regional assistance model and does not have to be entered manually by the program user. The calculations involved in the regional assistance model given in equation {15} to {18} below.

The first stage in the calculation is to determine the Capital Cost Effect (C_{ce}) i.e. the effect of a grant on the capital cost of the project. This is calculated in equation {15}. The second step is to determine the Process Cost Effect (P_{ce}) of the C_{ce} as per equation {16}. The third step is to calculate the Uplifted Fuel Cost (U_{fc}) as shown in equation {17}. The sixth and final step is the calculate the Price Advantage (PA_{reg}) of biomass fuels based on regional assistance as per equation {18}.

$$C_{ce} = P_{cost} - \left(\frac{\left(G_c \times 103\right) / P_t}{P_{out} \times D_{op}} \right) \qquad \{15\}$$

$$P_{ce} = C_{ce} - \left(\frac{\left(L_{cu} \times E_c / 100\right) \times \left(P_w / 100\right)}{P_{out} \times D_{op}} \right) \qquad \{16\}$$

$$U_{fc} = \left(1 - \left(1 - \left(\frac{P_{ce} / P_{hhv}}{P_{cost} / P_{hhv}}\right)\right)\right) \times \left(\frac{P_{ce}}{P_{hhv}}\right) \times P_{hhv} \qquad \{17\}$$

$$PA_{reg} = P_{cost} - U_{fc} \qquad \{18\}$$

where D_{op} is the number of days per year for which the biomass conversion plant operates (days/year),

P_{out} is the amount of product produced by the biomass conversion process per day (tonnes/day),

P_{cost} is the production cost of the biomass derived product without any macroeconomic factors being applied (Great Britain pounds/tonne),

G_c	is the amount of capital grant (thousands Great Britain pounds),
L_{cu}	is the cost of a labour unit (Great Britain pounds/year),
E_c	is the employment credit (%),
P_w	is the percentage of the workforce eligible for the employment credit (%)
P_{hhv}	is the higher heating value of the biomass derived product (GJ/tonne).

9.6. The effect of macroeconomic factors on product costs.

The macroeconomic factors modelled in BLUNT may be used singly (for example, a sulphur tax on fossil fuels) or in combination (for example, a carbon tax on fossil fuels coupled with regional assistance for the biomass converter).

To demonstrate the four macroeconomic models Table 9.1. gives the results for sample calculations carried out using the default information provided in the models. The other data required by the macroeconomic models is taken directly from the results from a sample simulation run by BLUNT.

Table 9.1.
The effect of the macroeconomic factors on the production cost of a biomass derived product.

	£/tonne	£/GJ
Production cost of biomass derived product (as produced and with no macroeconomic factors applied)	206.6	10.82
Price advantages of macroeconomic factors		
Sulphur tax on fossil fuels	-18.20	-0.95
Carbon tax on fossil fuels	-20.13	-1.05
Increase in irrigation water cost	+18.69	+0.98
Regional assistance	-18.06	-0.95
NEW PRODUCTION COST	168.60	8.85

As can be seen from Table 9.1. the sulphur tax, carbon tax and regional assistance models all reduce the production cost of a biomass conversion product because they favour biomass conversion processes. The irrigation water pricing model increases the production cost of a biomass conversion product due to the 2 pence increase in irrigation water costs used in the model. However, if there was a reduction in the irrigation water cost the Price Advantage of the irrigation water cost model would be negative and hence lower the production cost of the biomass derived product.

9.7. Summary

Environmental and macroeconomic effects are likely to become important in the near future. The actual type or size of taxes which will be implemented are as yet unknown and will almost certainly vary from region to region. The macroeconomics models produced and incorporated into BLUNT have been designed to be as flexible and wide ranging as possible to cover the likely aspects of any future legislation.

CHAPTER 10.

RESULTS.

10.1. Introduction.

10.1.1. The process variables available.

The previous Chapters have dealt with the technology and modelling of biomass flash pyrolysis, upgrading, refining and power production processes and the development of the dedicated computer program (BLUNT). The range of results obtainable from BLUNT includes the results of process optimisations, product yields, capital costs for conversion routes, production costs for the biomass derived products, sensitivity analyses on the financial variables and macroeconomic analyses. The main process variables which can be studied using BLUNT are the biomass feedstock, the conversion product obtained, the conversion process utilised, the biomass conversion plant size, the feedstock cost, the cost of utilities, financial variables and macroeconomic factors. The process variables are summarised in Table 10.1.

Table 10.1
Summary of BLUNT's main parameters and variables.

Variable	Possibilities	Notes / limits
Feedstock	3	Wood, sorghum bagasse and straw
Products	6	Pyrolysis oil, hydrotreated oil, diesel, crude aromatics, gasoline, electricity
Conversion processes	9	Ensyn, Fenosa, NREL & generic pyr., hydrotreatment, zeolite cracking pres. & pot., refining & power generation
Plant size	infinite	Between 25 and 1000 d.a.f. t/day
Feedstock cost	hundreds	Between 0 and 100 £/d.a.f. tonne (or equivalent in another currency)
Currencies	16	See Chapter 4 for details
Utilities	6	See Chapter 4 for details
Financial	8	See Chapter 4 for details
Macroeconomic	4	Sulphur and carbon taxes on fossil fuels, water pricing and regional assistance

The feedstock, products and conversion processes modelled are as shown in Table 10.1. above. There are a large number of combinations of feedstock and conversion processes to give the desired products. The conversion processes can be broken down into primary, secondary and tertiary processes. The primary process is the initial thermochemical conversion or flash pyrolysis process utilised, the secondary processes consist of any upgrading required and the tertiary processes consist of any refining or power production processes.

Table 10.2. gives the products and conversion routes available from sorghum bagasse and straw feedstocks. The only pyrolysis process modelled which will take straw as a feedstock is the Ensyn RTP system (See Section 2.4. and Chapter 6 for details). As can be seen from Table 10.2. the total number of conversion routes modelled in BLUNT for straw / sorghum bagasse is 9 considering all combinations of product and processing route.

Table 10.2.
Products and conversion routes modelled in BLUNT for straw or sorghum bagasse feedstocks.

Product	Primary conversion	Secondary conversion	Tertiary conversion
Crude pyrolysis oil	Ensyn RTP pyrolysis	none	none
Hydrotreated oil	Ensyn RTP pyrolysis	hydrotreating	none
Crude aromatics	Ensyn RTP pyrolysis	zeolite cracking (pres.)	none
Diesel	Ensyn RTP pyrolysis	hydrotreating	refining
Gasoline	Ensyn RTP pyrolysis	zeolite cracking (pres.)	refining
	Ensyn RTP pyrolysis	zeolite cracking (pot)	none
Electricity	Ensyn RTP pyrolysis	none	power gen.
	Ensyn RTP pyrolysis	hydrotreating	power gen.
	Ensyn RTP pyrolysis	zeolite cracking (pres.)	power gen.

The products and conversion routes available in BLUNT for wood and the combination of processes required are given in Table 10.3.

Table 10.3.
Products and conversion routes modelled in BLUNT for a wood feedstock.

Product	Primary conversion	Secondary conversion	Tertiary conversion
Crude pyrolysis oil	Generic flash pyrolysis	none	none
	Ensyn RTP pyrolysis	none	none
	Fenosa flash pyrolysis	none	none
	NREL vortex pyrolysis	none	none
Hydrotreated oil	Generic flash pyrolysis	hydrotreating	none
	Ensyn RTP pyrolysis	hydrotreating	none
	Fenosa flash pyrolysis	hydrotreating	none
	NREL vortex pyrolysis	hydrotreating	none
Crude aromatics	Generic flash pyrolysis	zeolite cracking (pres.)	none
	Ensyn RTP pyrolysis	zeolite cracking (pres.)	none
	Fenosa flash pyrolysis	zeolite cracking (pres.)	none
	NREL vortex pyrolysis	zeolite cracking (pres.)	none
Diesel	Generic flash pyrolysis	hydrotreating	none
	Ensyn RTP pyrolysis	hydrotreating	none
	Fenosa flash pyrolysis	hydrotreating	none
	NREL vortex pyrolysis	hydrotreating	none
Gasoline	Generic flash pyrolysis	zeolite cracking (pres.)	refining
	Ensyn RTP pyrolysis	zeolite cracking (pres.)	refining
	Fenosa flash pyrolysis	zeolite cracking (pres.)	refining
	NREL vortex pyrolysis	zeolite cracking (pres.)	refining
	Generic flash pyrolysis	zeolite cracking (pot)	none
	Ensyn RTP pyrolysis	zeolite cracking (pot)	none
	Fenosa flash pyrolysis	zeolite cracking (pot)	none
	NREL vortex pyrolysis	zeolite cracking (pot)	none
Electricity	Generic flash pyrolysis	none	power gen.
	Ensyn RTP pyrolysis	none	power gen.
	Fenosa flash pyrolysis	none	power gen.
	NREL vortex pyrolysis	none	power gen.
	Generic flash pyrolysis	hydrotreating	power gen.
	Ensyn RTP pyrolysis	hydrotreating	power gen.
	Fenosa flash pyrolysis	hydrotreating	power gen.
	NREL vortex pyrolysis	hydrotreating	power gen.
	Generic flash pyrolysis	zeolite cracking (pres.)	power gen.
	Ensyn RTP pyrolysis	zeolite cracking (pres.)	power gen.
	Fenosa flash pyrolysis	zeolite cracking (pres.)	power gen.
	NREL vortex pyrolysis	zeolite cracking (pres.)	power gen.

For wood there are four flash pyrolysis processes available - the Generic (See Chapter 6 for details), Ensyn RTP (See Section 2.4.), Fenosa (See Section 2.5.) and

NREL Vortex (See Section 2.6.) processes. As can be seen from Table 10.3. the total number of conversion routes possible for a wood feedstock as modelled in BLUNT is 36 considering all combinations of pyrolysis technology, product and processing route.

The utilities used in BLUNT include cooling water, steam, fuel gas, power, oxygen and hydrogen. The financial parameters which can be altered include interest and inflation rates, operating days per year, the maintenance and overhead costs and the project life. Details of the utility costs and financial parameters can be found in Chapter 4.

The macroeconomic factors considered in BLUNT include any combination of sulphur and carbon taxes on fossil fuels, irrigation water pricing and regional assistance. Details of the macroeconomic factors can be found in Chapter 9.

10.1.2. The process variables considered.

There are potentially millions of different analyses and combinations of results which could be considered however it is not viable to consider all of the possible combinations of the above process variables. This Chapter demonstrates the versatility and comprehensiveness of BLUNT without attempting to provide a thorough analysis of all the possibilities. In addition, disk copies of BLUNT are available to allow users to generate their own set of results for analysis. The results and analyses which are given in this Chapter demonstrate :

- the effect of feed cost variation on product costs;
- the effect of the choice of pyrolysis technology/process on product costs;
- the effect of macroeconomic factors on production costs;
- the effect of feedstock on product costs;
- the effect of feed moisture content on product costs;
- the effect of pyrolysis reactor temperature on product costs;
- the effect of pyrolysis energy sources on product costs;
- the effects of conversion technology on product costs;
- the effect of changes in utilities costs on product costs; and
- the effect of changes in the base financial variables on product costs.

Section 10.2. gives the base case data which was used to generate the results given in this Chapter. Section 10.3. gives the results from the program expressed in graphical form for easy interpretation and show the effect of feedstock cost, in pounds per d.a.f. tonne, on the product cost in pounds per unit energy (product cost per GJ of energy in the product). The sensitivity analyses on utility costs and financial parameters are expressed as a percentage change in the value of the cost or parameter with respect to the cost of the biomass derived product in pounds per unit energy. A summary of the results is given in Section 10.4.

10.2. Base case data.

The base case financial and economic data used in the derivation of the results is for the values shown in Table 10.4. unless otherwise stated. All of the results and the variables and data utilised in the derivation of the results are in 1993 Great Britain pounds.

Table 10.4.
Base case data used in BLUNT.

Variable	**Default value**
Project life (years)	10
Days of operation per year (days/year)	330
Real interest rate (%)	5.0
Inflation rate (%)	5.0
Feedstock cost (pounds/d.a.f. tonne)	30.00
Maintenance (% Capital cost/year)	5.0
Overheads (% Capital cost / year)	7.0
Utilities costs	
Low pressure steam (pounds/tonne)	6.00
Hydrogen (pounds/tonne)	360.00
Cooling water (pounds/tonne)	0.05
Oxygen (pounds/tonne)	50.00
Fuel gas (pounds/GJ)	4.00
Power (pounds/MWh)	50.00

To allow the results obtained from BLUNT to be viewed in perspective comparisons are made with conventional fossil fuels. In Section 10.3. biomass pyrolysis oil is compared with medium fuel oil, bio-naphtha (hydrotreated pyrolysis oil) with fossil fuel naphtha, bio-diesel with fossil derived diesel and bio-gasoline with fossil fuel derived gasoline. Table 10.5. gives the comparison costs of the fossil derived liquids, ex-Rotterdam January 1995.

Table 10.5.
Typical fossil fuel costs for 1994/1995.
(Based on typical spot prices CIF Northern Europe, January 1995)

Commodity	**Cost**
Medium fuel oil (£/tonne)	73.2
Naphtha (£/tonne)	103.9
Diesel (£/tonne)	97.4
Gasoline (£/tonne)	106.5

10.3. Results from the program.

10.3.1. The effect of choice of feedstock.

Figure 10.1. shows the effect the choice of feedstock on the production costs of crude pyrolysis oil with respect to feedstock costs between 0 and 100 Great Britain pounds per dry ash free tonne. As explained earlier in Chapter 6, there are three feedstocks available in BLUNT, wood, straw and sorghum bagasse. Straw and sorghum bagasse may only be pyrolysed in an Ensyn RTP pyrolysis system and so this is utilised for comparison of the feedstocks. The straw, bagasse and wood (with a moisture content of 10% dry basis) feedstocks were pyrolysed under the process conditions given in Chapter 6 in a 100 tonnes/day Ensyn RTP plant.

As can be seen in Figure 10.1. the costs for all three feedstocks fall within a close range however it is apparent that the choice of feedstock does affect the product cost. For a typical United Kingdom feedstock cost of 30 Great Britain pounds per tonne the production costs of a crude pyrolysis oil are 6 pounds per GJ for straw, 7 pounds per GJ for sorghum bagasse and 7.6 pounds per GJ for wood.

For feedstock costs below 50 Great Britain pounds per tonne wood gives the highest product cost although this changes to sorghum bagasse when the feedstock cost rises above 50 Great Britain pounds per tonne. A straw feedstock gives the lowest production costs for a crude pyrolysis oil product. As described in Chapter 5, the pretreatment processes for the straw and bagasse feedstocks are the same.

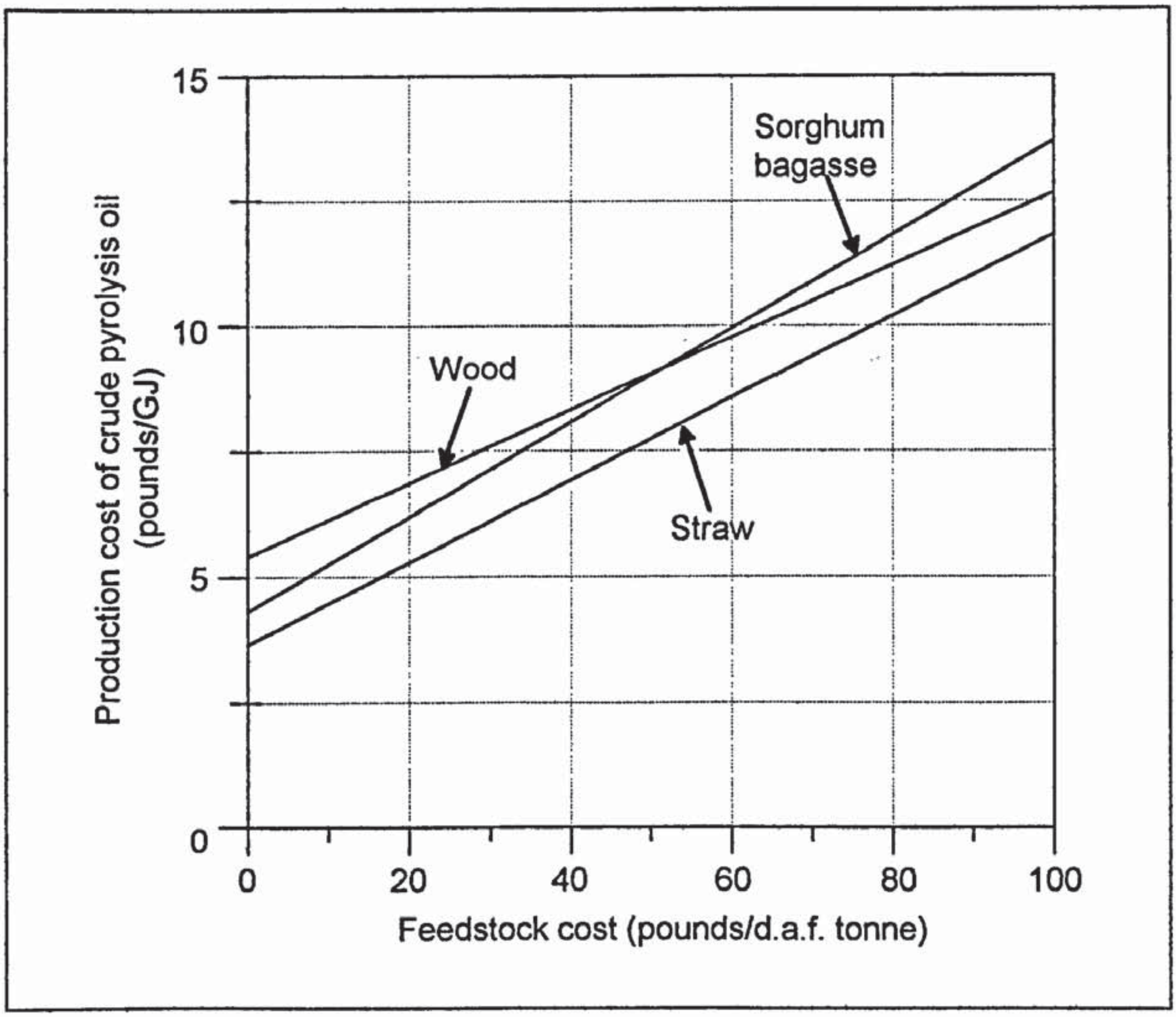


Figure 10.1. The effect of choice of feedstock on the production costs of crude pyrolysis oil.

The main difference is the feedstock characteristics i.e. ash and moisture contents and the yields obtained from the pyrolysis process itself - bagasse gives lower bio-oil yields than straw (See Chapter 6). These variations would explain the difference in the bagasse and straw production costs.

10.3.2. The effect of pyrolysis process.

Figure 10.2. shows the effect of the choice of flash pyrolysis process on the production cost of crude pyrolysis oil with respect to feedstock costs between 0 and 100 Great Britain pounds per dry ash free tonne. A wood feedstock with a moisture content of 10% (dry basis) was pyrolysed in the Generic, Ensyn RTP, Union Fenosa and NREL processes, for the process conditions given in Chapter 6. The plant size of the flash pyrolysis processes utilised to generate the results was 100 d.a.f. tonnes/day.

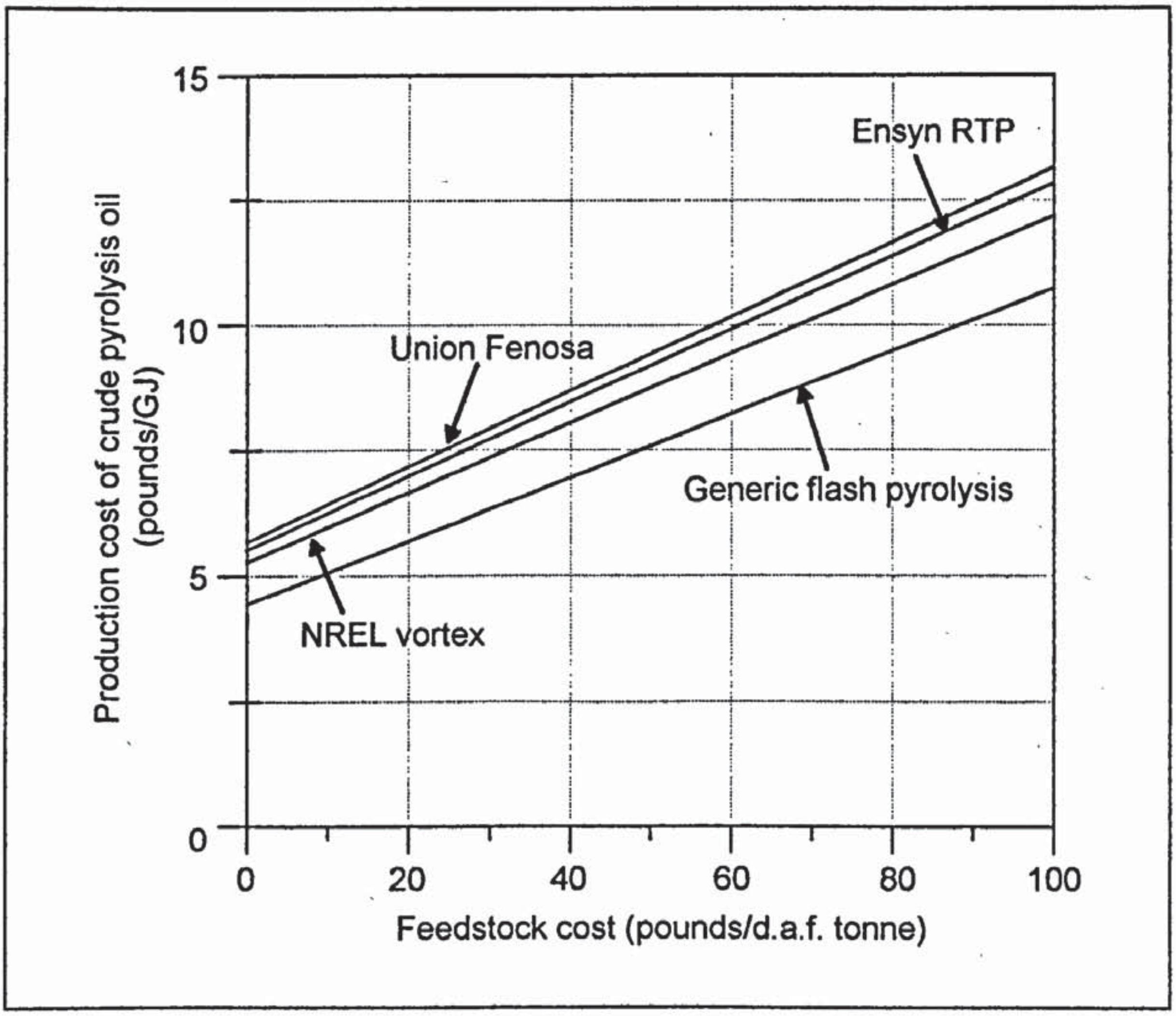


Figure 10.2. The effect of choice of pyrolysis conversion process on the production costs of crude pyrolysis oil.

As can be seen from Figure 10.2. the NREL, Ensyn and Fenosa production costs all fall within the same narrow range. This is to be expected as the pyrolysis models all use the same plant capital cost and have similar product yields. However, the Fenosa process gives the highest product cost probably due to it's slightly lower product yields (See Chapter 2 for further details). The product yields from the Generic flash pyrolysis process are lower than those of the other processes despite the same capital cost being utilised and the small variation in product yields (See Chapter 6). The lower production costs for the generic process are probably due to it's optimised energy balance - allowing the use of the char as fuel with the remainder being exported and credited as a by-product. In addition the off-gas is exported and credited as fuel gas to the utilities (See Chapter 6 for further details on the Generic flash pyrolysis model's energy optimisation). However, since comprehensive data on the Ensyn, Union Fenosa and NREL processes was not available when the models were written the higher production costs could be explained by the general assumptions made when modelling. Further investigation and more complete data could result in a reduction in these costs. For a typical United Kingdom feedstock cost of 30 Great Britain pounds per tonne the production costs of a crude pyrolysis oil are 6 pounds per GJ for the Generic flash pyrolysis process, 7.6 pounds per GJ for the Ensyn RTP system, 7.3 pounds per GJ for the NREL vortex system and 7.8 pounds per GJ for the Fenosa system.

10.3.3. The effect of conversion technology.

Figure 10.3. shows the effect of feedstock cost changes between 0 and 100 Great Britain pounds per dry ash free tonne on the crude and upgraded products from biomass conversion. The products considered are crude pyrolysis oil produced from a 10% moisture content wood feedstock (produced from the generic flash pyrolysis process for 1000 d.a.f. tonnes/day of wood at a pyrolysis temperature of 500°C), bio-naphtha produced from hydrotreating the pyrolysis oil to a deoxygenation extent of 98% and crude aromatics produced from present case zeolite cracking of the vapours from the generic pyrolysis process.

Two set of data are presented for hydrotreating representing: Option 1 which has not hydrogen recovery and Option 2, a process which utilised recovery and recycling of hydrogen (See Chapter 7 for details on the options). As can be seen from Figure 10.3. the production costs of the upgraded products (bio-naphtha and crude aromatics) are in the same region at high feedstock costs. The costs for the bio-naphtha from the hydrotreating option 2 system and the crude aromatics are the same at a feedstock cost around 75 pounds/tonne. Below a feedstock cost of 75 pounds per tonne the production cost of bio-naphtha for both options is higher than that of crude aromatics. The cost of bio-naphtha from Option 1 (no hydrogen recovery) is more expensive than from Option 2 (membrane separation and hydrogen recovery) and the lines are parallel since the products are of the same value.

The bio-fuel cost lines on Figure 10.3. are not parallel for crude aromatics, crude pyrolysis oil and bio-naphtha because the products are of different qualities and the incremental cost of upgrading cannot be simply added on to the cost of producing crude pyrolysis oil. For a typical United Kingdom wood feedstock cost of 30 pounds per d.a.f. tonne crude pyrolysis oil costs 4 pounds per GJ to produce, bio-naphtha from Option 1 costs 8 pounds per GJ and from Option 2 costs 9 pounds per GJ to produce and crude aromatics 7 pounds per GJ to produce.

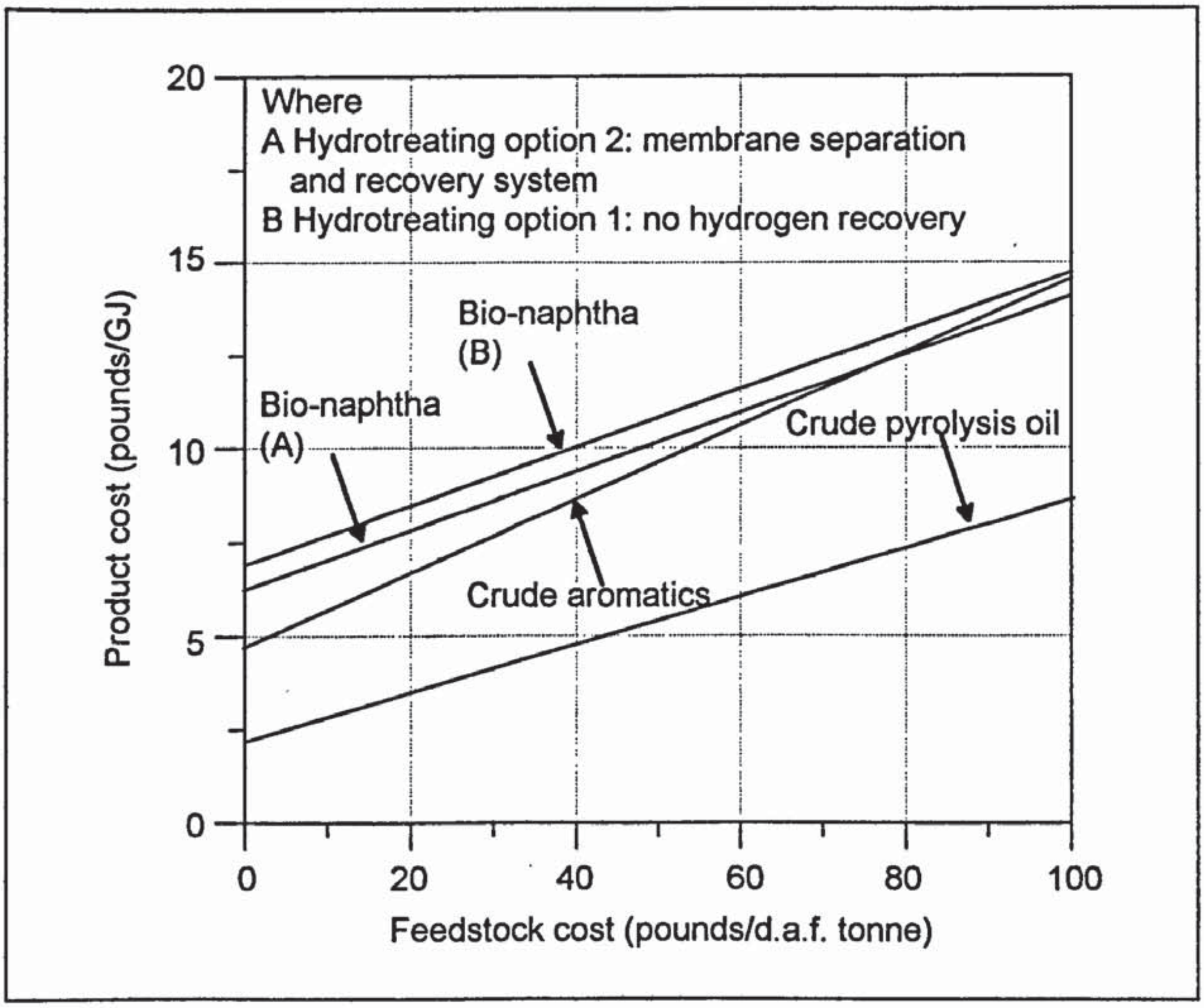


Figure 10.3. The effect of feed cost on the production costs of the crude and upgraded products from biomass conversion.

Fossil fuel costs have not been included in Figure 3 due to the large amount of information already on the figure. However, for fossil fuel derived naphtha of around 2.6 pounds/GJ, the production of bio-naphtha from the hydrotreating of biomass pyrolysis oils cannot compete, even for low cost feedstocks.

Figure 10.4. compares the production costs of diesel and gasoline derived from the thermochemical conversion of wood. On the graph the cost of gasoline produced from the potential case zeolite cracking process is compared with that of gasoline produced by present case zeolite cracking followed by refining. The diesel is produced by refining fully hydrotreated pyrolysis oil using a hydrotreating process incorporating hydrogen recovery and recycling (Option 2 as in Chapter 7). All of the processes utilised a 10% moisture content wood feedstock pyrolysed in a 1000 tonnes/day generic flash pyrolysis process at a pyrolysis temperature of 500°C..

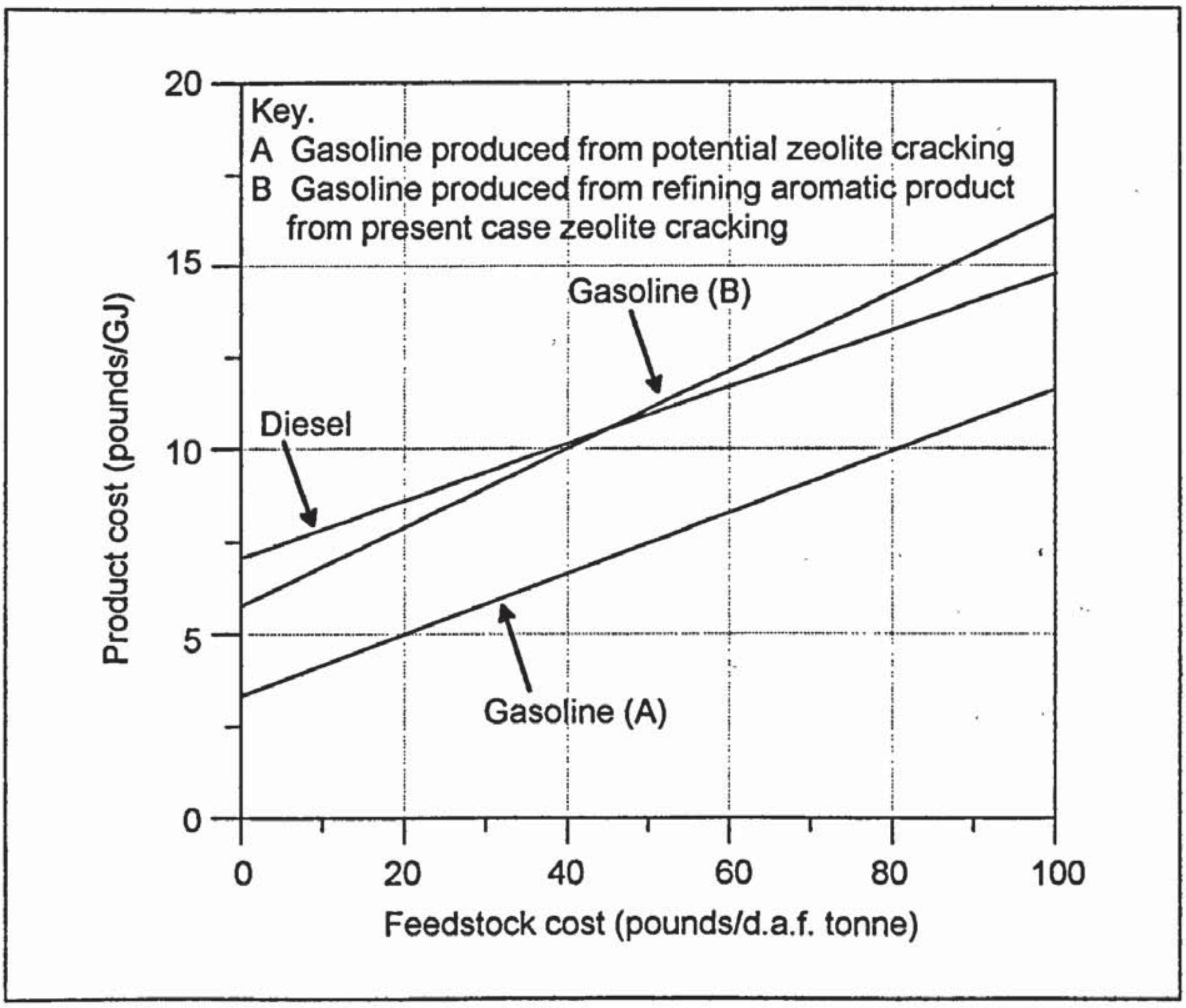


Figure 10.4. The effect of conversion process on the production costs of crude, upgraded and refined products of biomass conversion.

As on Figure 10.3. the cost lines for the production costs of the different products shown in Figure 10.4. are not parallel due to the differences in product quality. The production costs of diesel and gasoline (B) (produced from the zeolite cracking present process plus refining) are in a similar range. This is to be expected since the bio-naphtha and crude aromatics costs shown in Figure 10.3. were similar and the diesel and gasoline costs just assume a small service charge for refining and a 90% yield (See Chapter 8 for details). There is a substantial difference between the production costs of gasoline produced from the potential case zeolite cracking process and that from refining the aromatics produced from the present case zeolite cracking process. The capital costs for the two zeolite cracking processes are similar (See Chapter 8) so the difference is probably due to the difference in product quality from the processes combined with the additional charge for refining added to the present case costs. From the graph it would appear that the potential case zeolite cracking process is an extremely viable option. However, it must be remembered that the potential case assumes substantial improvements in technology and extensive research to achieve the process yields and products. A comparison with fossil fuels has not been included on the graph. For fossil derived gasoline costs around 2.5 pounds/GJ the bio-derived gasoline cannot compete, similarly for fossil diesel costs around 2.3 pounds/GJ bio-derived diesel also cannot compete.

Figure 10.5. shows the effect of feedstock cost changes between 0 and 100 Great Britain pounds per dry ash free tonne on the cost of producing electricity from crude pyrolysis oil, hydrotreated oil and crude aromatics. The crude pyrolysis oil was produced from a 10% moisture content wood feedstock in the generic flash pyrolysis process at a pyrolysis temperature of 500°C and for 100 tonnes/day of biomass feedstock. The bio-naphtha was produced from hydrotreating the pyrolysis oil to a deoxygenation extent of 98% using hydrogen recovery and recycling and the crude aromatics were produced from present case zeolite cracking of the vapours from the generic pyrolysis process.

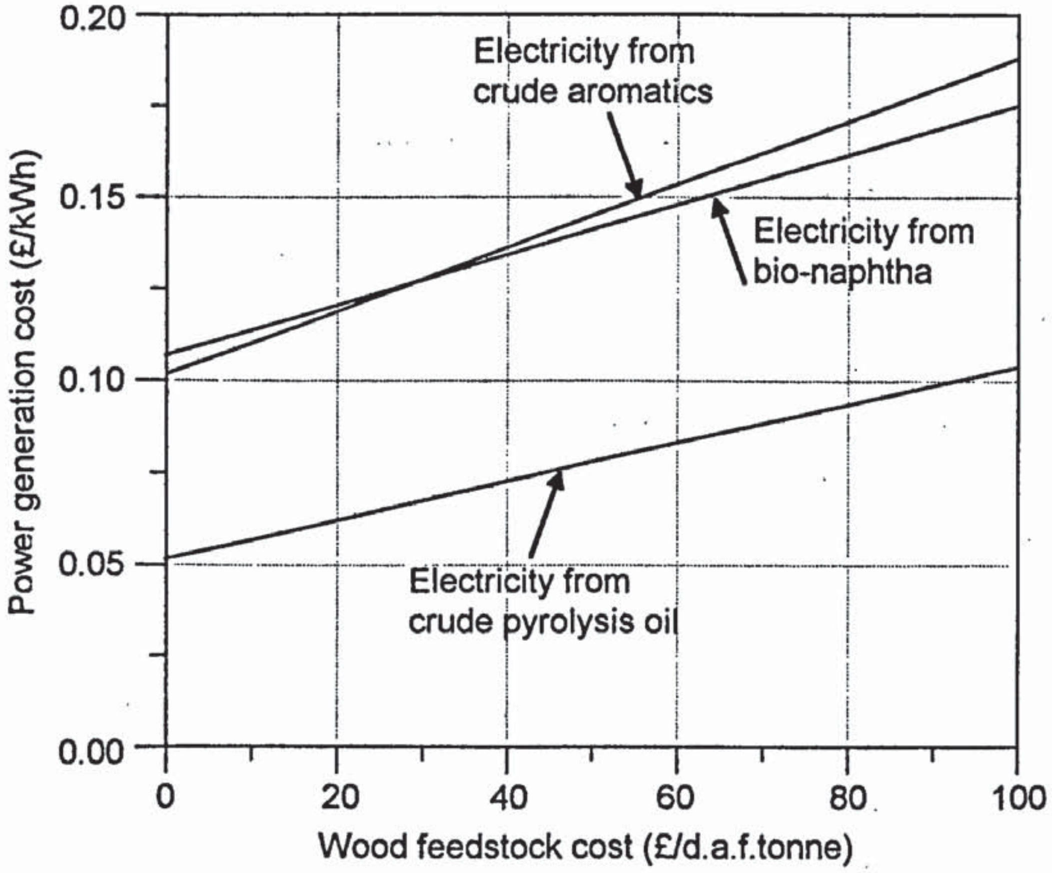


Figure 10.5. The effect of conversion process on the production costs of electricity.

As would be expected from the results shown in Figure 10.3., the lines on Figure 10.5. are not parallel because the liquid products used to generate the electricity are of different qualities. The cost of producing electricity from crude aromatics and that using bio-naphtha are in a similar range. The cost of producing electricity from crude pyrolysis oil is significantly lower than those of bio-naphtha and crude aromatics as would be expected from the production costs of the liquid products shown in Figure 10.3.

10.3.4. The effect of flash pyrolysis temperature.

Figure 10.6. demonstrates the effect flash pyrolysis temperature has on the production cost of crude pyrolysis oil with respect to feedstock costs between 0 and 100 Great Britain pounds per dry ash free tonne. A wood feedstock with a moisture content of 10% (dry basis) was pyrolysed in the Generic process for the conditions given in Chapter 6 and a feed throughput of 100 d.a.f. tonnes/day. The effects of flash pyrolysis temperatures between 400 and 650°C are shown.

The results shown in Figure 6 have an unusual grouping. The lines for 600 and 400°C are grouped together as are those of 550, 450 and 500°C. These results can be explained by the product yields shown in Chapter 6 where the optimum liquid yield is produced at 500°C, hence the lowest costs. As the liquid yield curve is dome shaped (See Figure 6.2., Chapter 6), the liquid yields at 550 and 450°C will be similar as are the costs although the costs at 550°C are slightly higher. This could be perhaps explained by the increased temperature and hence energy demands at 550°C which will increase the operating cost of the system. The costs for 400 and 600°C are higher than those for the previously explained group, due to the lower liquid yields. The product costs at 650°C are considerably higher than the others. The explanation is probably the low liquid yields, low char yields for use as fuel (See Chapter 6) and high temperature increasing the operating cost of the system.

10.3.5. The effect of feed moisture content.

The effect of feed moisture content on the production cost of crude pyrolysis oil is described here. A wood feedstock was pyrolysed in the Generic process for the conditions given in Chapter 6 at 500°C and a feed throughput of 100 d.a.f. tonnes/day.

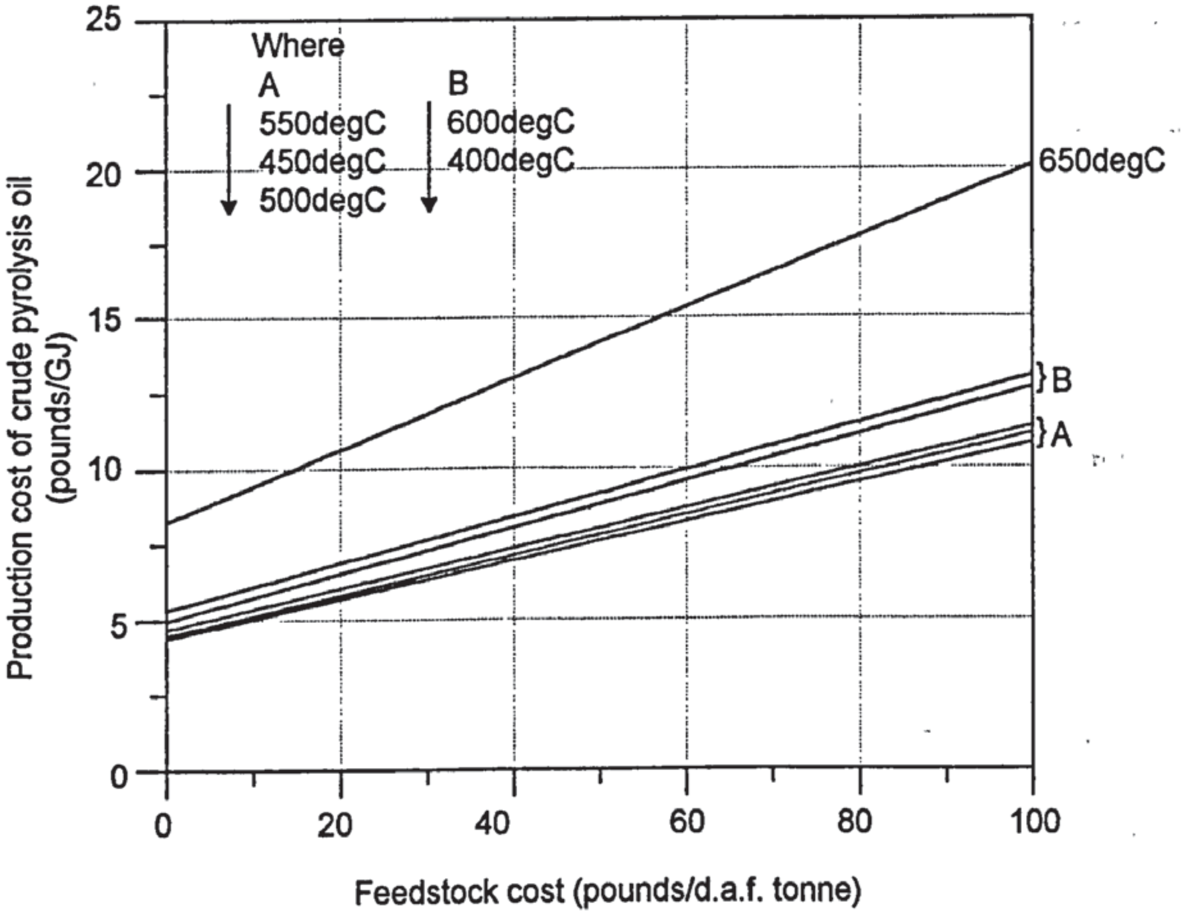


Figure 10.6. The effect of flash pyrolysis temperature on the production costs of crude pyrolysis oil.

Table 10.6. shows the effect of feed moisture contents between 5 and 25% (dry basis) with respect to feedstock costs between 0 and 100 Great Britain pounds per dry ash free tonne. The data is expressed in tabular form rather than graphical because the results are too close together to be seen clearly on a graph.

Table 10.6.
The effect of feed moisture content on the production costs of crude pyrolysis oil in £/GJ.

Feedstock cost (£/t)	**0**	**20**	**40**	**60**	**80**	**100**
Feed moisture content						
5wt% dry basis	4.42	5.70	6.98	8.26	9.54	10.83
10wt% dry basis	4.40	5.69	6.97	8.26	9.54	10.83
15wt% dry basis	4.37	5.65	6.94	8.23	9.52	10.80
20wt% dry basis	4.31	5.60	6.88	8.17	9.45	10.73
25wt% dry basis	4.27	5.55	6.83	8.12	9.40	10.68

As can be seen from Table 10.6. the effect of feed moisture content on pyrolysis oil production costs is slight. There is a lowering of production costs with increasing moisture content but the effect is only 0.15 £/GJ between a 5% moisture content feedstock and a 25% moisture content feedstock. The slight lowering of production costs with respect to feed moisture content will be due to the increased water content reducing the energy content of the crude pyrolysis oil product (See Chapter 6).

10.3.6. The effect of pyrolysis energy options.

The effect of pyrolysis fuel options on the production cost of crude pyrolysis oil is described here. A wood feedstock was pyrolysed in the Generic process for the conditions given in Chapter 6 at 500°C and a feed throughput of 100 d.a.f. tonnes/day. The five fuel options for flash pyrolysis were described in Chapter 6 and are:

1. char followed by imported fuel gas to make up the deficit, all the off-gas and any surplus char are exported as by-products;
2. char followed by the off-gas, any surplus char or off-gas is exported as by-product;
3. off-gas followed by imported fuel gas to make up the deficit, all char is exported as a by-product;
4. off-gas followed by char, any surplus char is exported as by-product; and
5. imported fuel gas only, all char and off-gas are exported as by-products.

Table 10.7. shows the effect of the fuel option (for char and gas combustion efficiencies of 75%) with respect to feedstock costs between 0 and 100 Great Britain pounds per dry ash free tonne. The data is expressed in tabular form rather than graphical because the results are too close together to be seen clearly on a graph.

Table 10.7.
The effect of pyrolysis fuel option on the production costs of crude pyrolysis oil in £/GJ.

Feedstock cost (£/t)	**0**	**20**	**40**	**60**	**80**	**100**
Pyrolysis fuel						
Option 1	4.40	5.69	6.97	8.26	9.54	10.83
Option 2	4.40	5.69	6.97	8.26	9.54	10.83
Option 3	4.76	6.04	7.33	8.61	9.90	11.19
Option 4	4.42	5.71	6.99	8.28	9.57	10.85
Option 5	4.76	6.04	7.33	8.61	9.90	11.19

As can be seen from Table 6.7. the production costs for Options 1 and 2 are the same. This is due to the way in which the pyrolysis model is set up i.e. the off-gas produced is sold as fuel gas to the utilities. Options 3 and 5 are also equivalent for the same reason. The lowest costs are for Options 1 and 2 where the char is utilised as the primary fuel source, this is due to the energy values of the char with respect to off-gas as given in Chapter 6 and the amount of char produced. Option 4 has slightly higher costs as the off-gas is utilised first and the energy deficit is made up by combusting the char. Exporting all of the char and using either fuel gas or the off-gas and fuel gas gives the highest product costs due to the relatively low energy value of the gas with respect to char. However, all of the costs shown in Table 10.7. are in a similar range with the difference in product costs between the highest and lowest fuel options being only 0.36£/GJ.

10.3.7. The effect of macroeconomic factors.

Figures 10.7. and 10.8. show the effect of the macroeconomic factors detailed in Chapter 9 on the production costs of biomass derived diesel and gasoline. The results which are expressed in Figure 10.7. are for biomass derived diesel obtained from fully hydrotreating (using the hydrogen recovery and recycling process, see Chapter 7 for further details) and refining pyrolysis oil obtained from the generic flash pyrolysis process for a wood input of 100 tonnes/day. The results which are expressed in Figure 10.8. are for biomass derived gasoline obtained from the potential case zeolite cracking process using the vapours from the generic flash pyrolysis process for a wood input of 100 tonnes/day. In both cases the generic flash pyrolysis process utilised a wood feedstock of 10% moisture content (dry basis) at a pyrolysis temperature of 500°C. The macroeconomic factors used to generate the results were carbon and sulphur taxes on fossil fuels, an increase in irrigation water pricing and capital grants or regional assistance using the default data in Chapter 9. Both Figures 10.7. and 10.8. demonstrate that the use of macroeconomic factors such as carbon or sulphur taxes on fossil fuels or regional assistance lowers the effective cost of a biomass derived product and could potentially make it more comparable with a fossil derived fuel. The irrigation water pricing factor increases the effective cost of a biomass derived product and has the most dramatic effect on the overall cost as it directly alters the cost of the feedstock entering the biomass conversion process. This effect could also potentially reduce the effective cost of a biomass derived fuel if

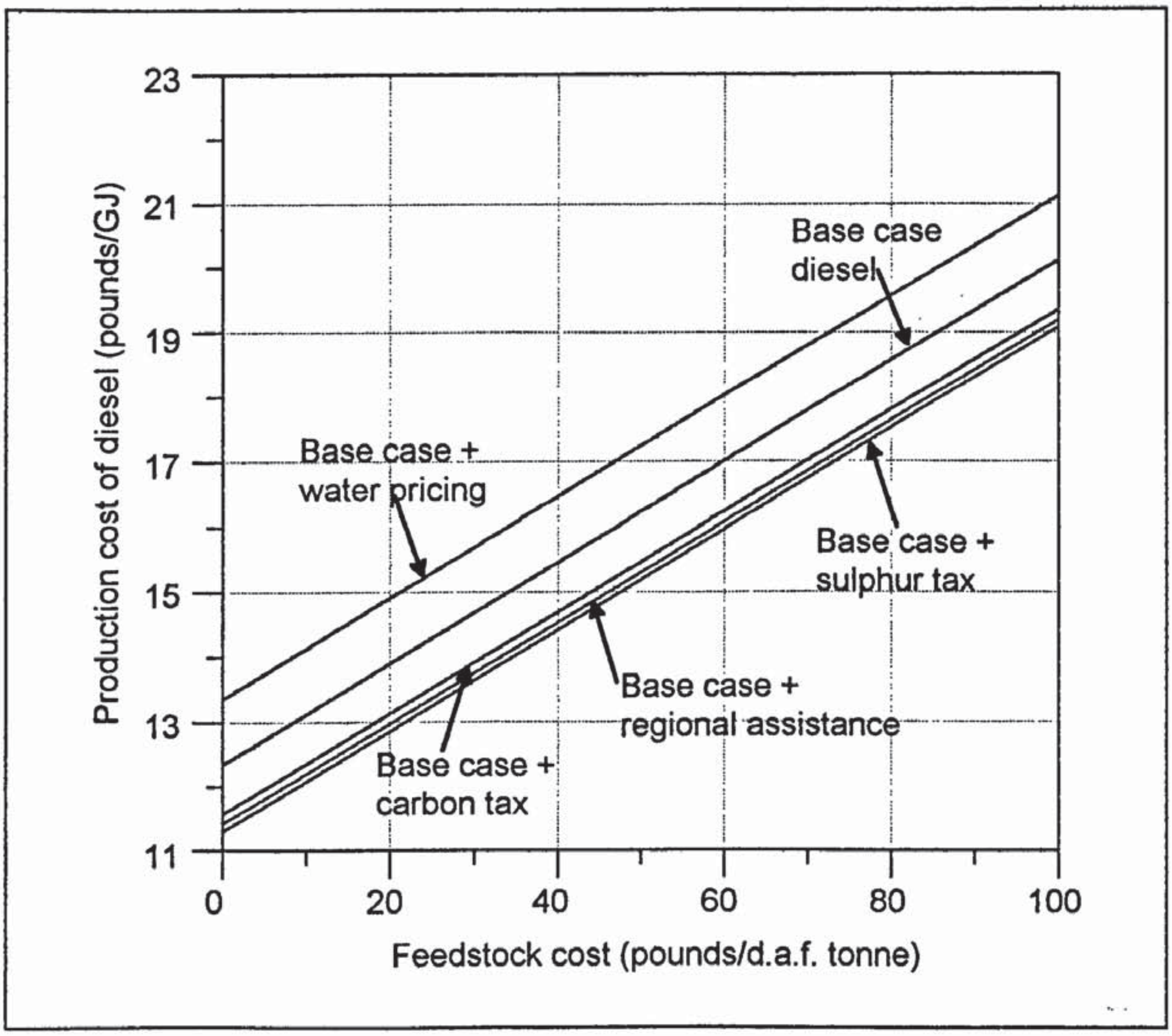


Figure 10.7. The effect of macroeconomic factors on the production cost of biomass derived diesel.

the irrigation water cost is lowered rather than increased (See Chapter 9 for further details).

10.3.8. The effect of changes in utility costs.

Figures 10.9. and 10.10. illustrate the effects of altering the costs of the utilities used in the production of crude pyrolysis oil and bio-naphtha respectively. Figure 10.9. shows the results from a 100 d.a.f. tonnes/day generic flash pyrolysis process utilising wood with a 10% moisture content (dry basis) as the feedstock converted at a pyrolysis temperature of 500°C. The utilities utilised by the flash pyrolysis process are cooling water, fuel gas and power (See Chapter 6 for further details) and their default costs shown in Table 10.4. were increased and decreased by 10 and 20% to generate the sensitivity analyses.

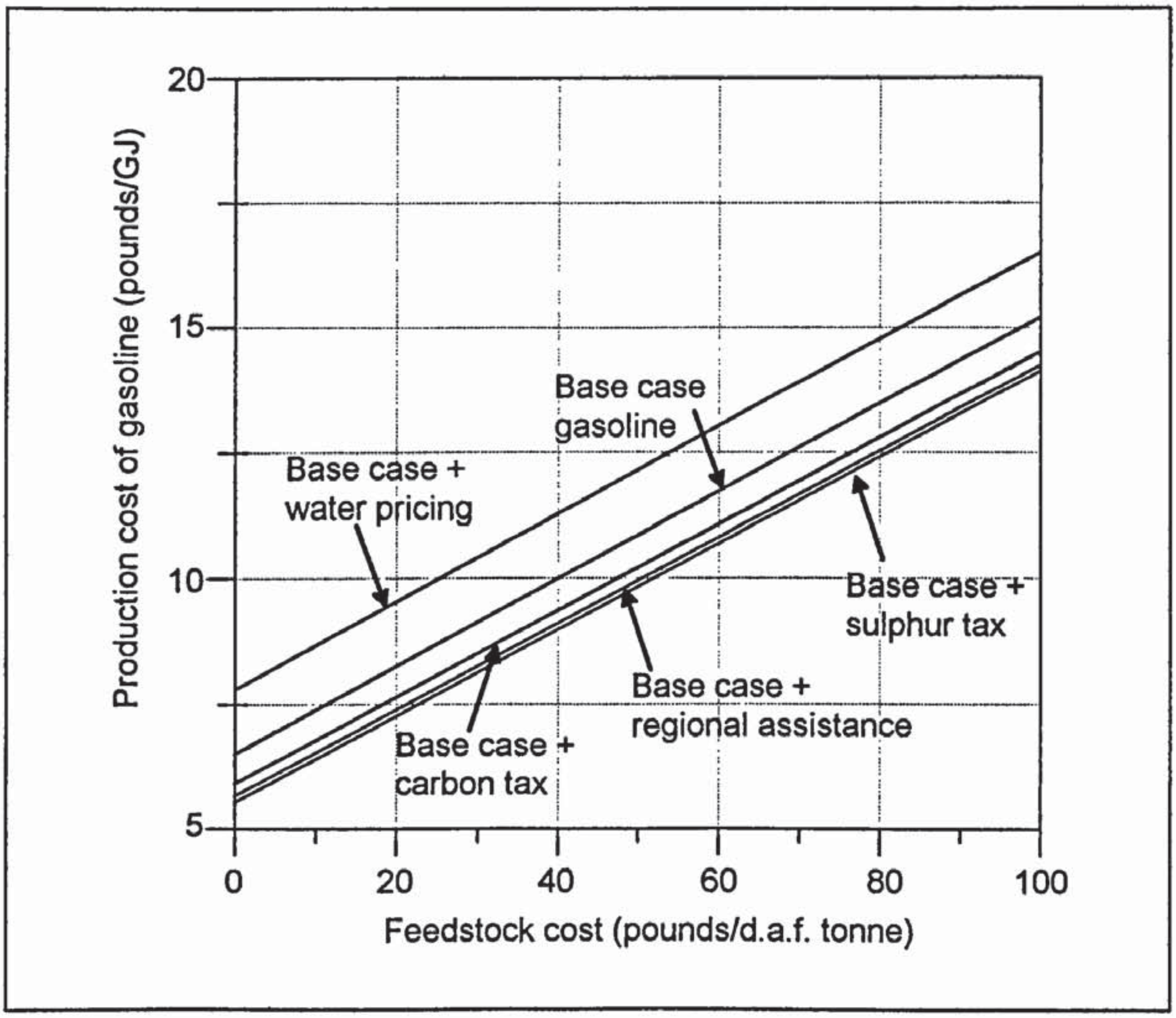


Figure 10.8. The effect of macroeconomic factors on the production cost of biomass derived gasoline.

From Figure 10.9. it can be seen that changes in fuel gas, power and cooling water costs do alter the cost of the crude pyrolysis oil with all the trends being in the same direction i.e. a lowering in fuel gas, power and cooling water costs lower the effective cost of the pyrolysis oil. However, the effects of power and cooling water costs on the overall production cost are slight in comparison with the effect of fuel gas which is required for drying the feedstock and supplying any energy required by the pyrolysis process itself.

Figure 10.10. illustrates the results from the production of bio-naphtha obtained from fully hydrotreating crude pyrolysis oil from the generic process using the hydrogen recovery and recycling hydrotreating process. The utilities utilised by the hydrotreating process are cooling water, fuel gas, hydrogen and power (See Chapter 7 for further details) and their default costs shown in Table 10.4. were increased and decreased by 10 and 20% to generate the sensitivity analyses.

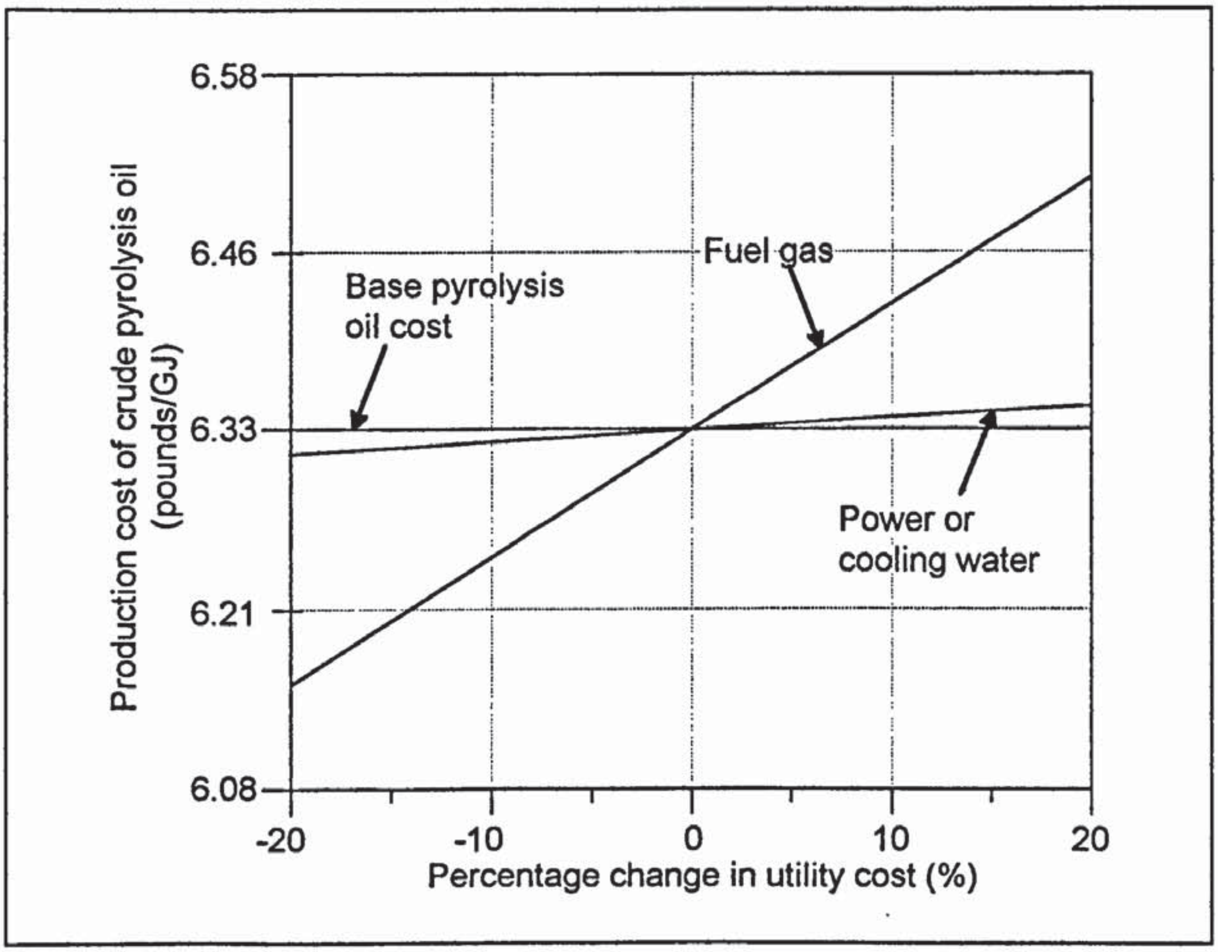


Figure 10.9. The effect of changes in utility costs on the production costs of crude pyrolysis oil.

Figure 10.10. demonstrates that the cost of bio-naphtha is sensitive to changes in the cost of cooling water, power, fuel gas and hydrogen with the trends for cooling water and power being similar to those in Figure 10.9. i.e. an increase in cooling water and power costs increase the effective cost of the bio-naphtha. However, the trend for fuel gas is reversed i.e. a lowering in fuel gas cost increases the cost of the bio-naphtha. This apparent anomaly can be explained by the high fuel gas exports and hence credits from the hydrotreating process to the utilities pool (See Chapter 7 for details) which obviously have a significant impact on the production cost of bio-naphtha. Hydrogen is utilised by the hydrotreating process and although a hydrogen recovery and recycling system is used to minimise the hydrogen requirements the cost of hydrogen has a significant impact on the production cost of bio-naphtha. Hydrogen follows the expected trend i.e. a lowering in hydrogen costs lowers the effective cost of the bio-naphtha.

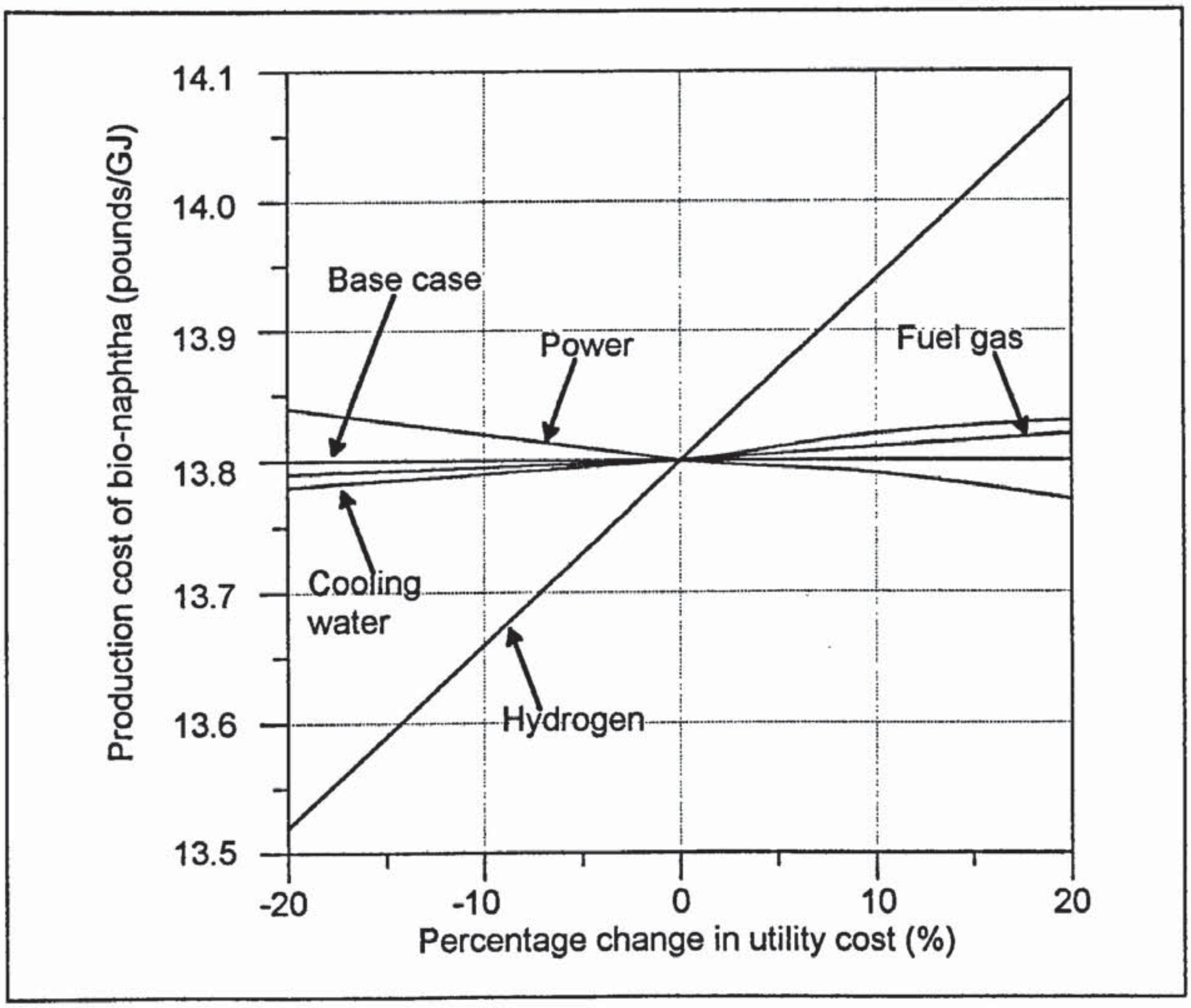


Figure 10.10. The effect of changes in utility costs on the production costs of bio-naphtha.

10.3.9. The effect of changes in the base financial parameters.

Figures 10.11. and 10.12. illustrate the effects of altering the base financial parameters used in the cost estimations i.e. project life, operational days, interest and inflation rates and maintenance and overhead costs. The default values of the financial parameters as shown in Table 10.4. were increased and decreased by 10 and 20% to generate the sensitivity analyses.

The results illustrated in Figure 10.11. utilise a 100 d.a.f. tonnes/day generic flash pyrolysis process utilising wood with a 10% moisture content (dry basis) as the feedstock converted at a pyrolysis temperature of 500°C. The results in Figure 10.12. illustrate the production of bio-naphtha obtained from fully hydrotreating crude pyrolysis oil from the generic process using the hydrogen recovery and recycling hydrotreating process.

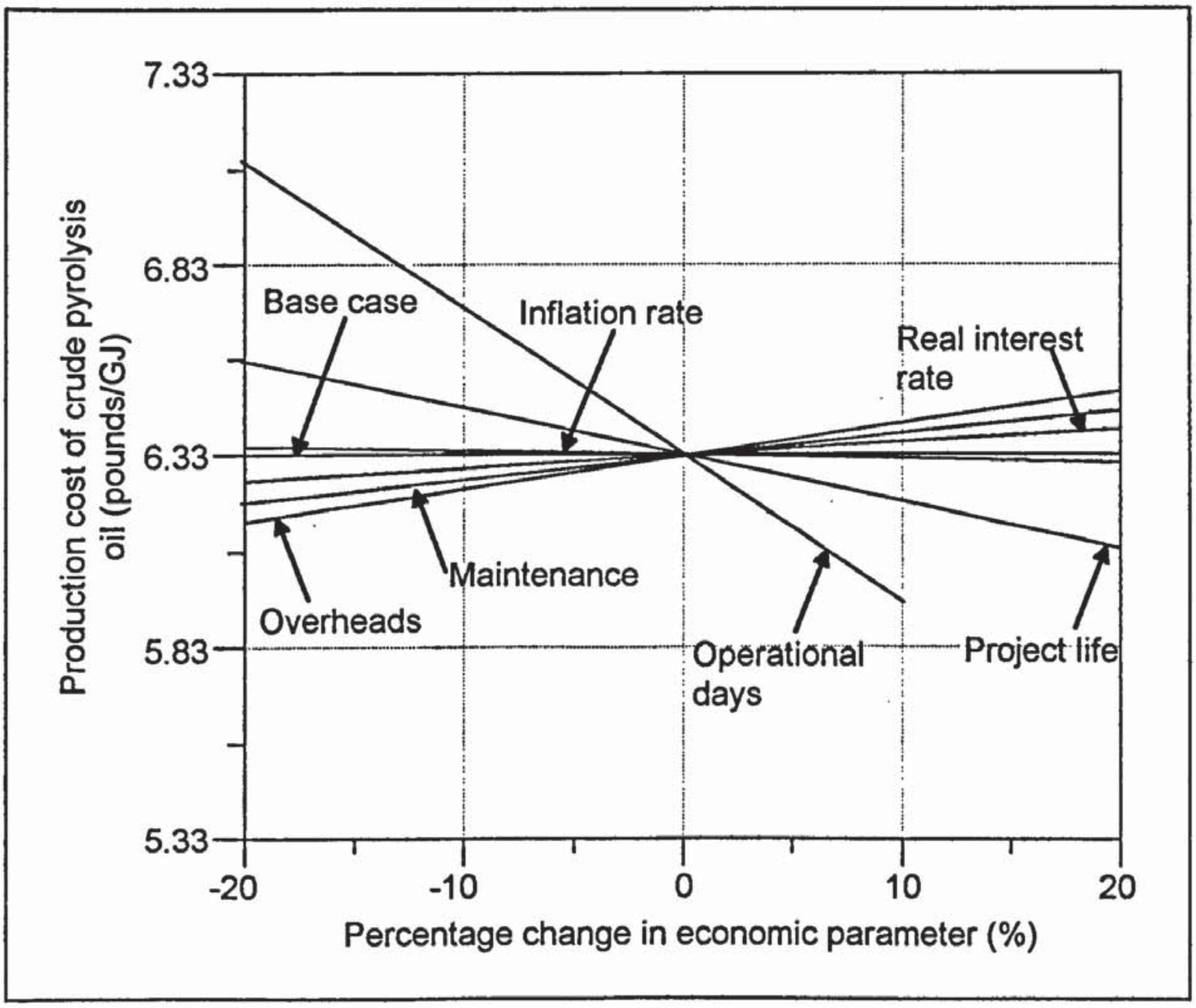


Figure 10.11. The effect of changes in financial parameters on the production costs of crude pyrolysis oil.

The results shown in Figure 10.11. and Figure 10.12. demonstrate that the production cost of the biomass derived products are sensitive to changes in the financial parameters although to varying extents. Overheads, maintenance and real interest rate all follow an expected trend i.e. increases in the value of any of these parameters increase the cost of the crude pyrolysis oil or bio-naphtha products. However, changes in the operational days per year and the project life of the plant have the most dramatic effects and their effects is a reversal of the trend demonstrated by the other financial parameters. An increase in project life increases the number of years over which the original capital investment has to be paid back, decreasing the amount of amortised capital which has to be paid back each year and hence lowering the production cost of the biomass derived product.

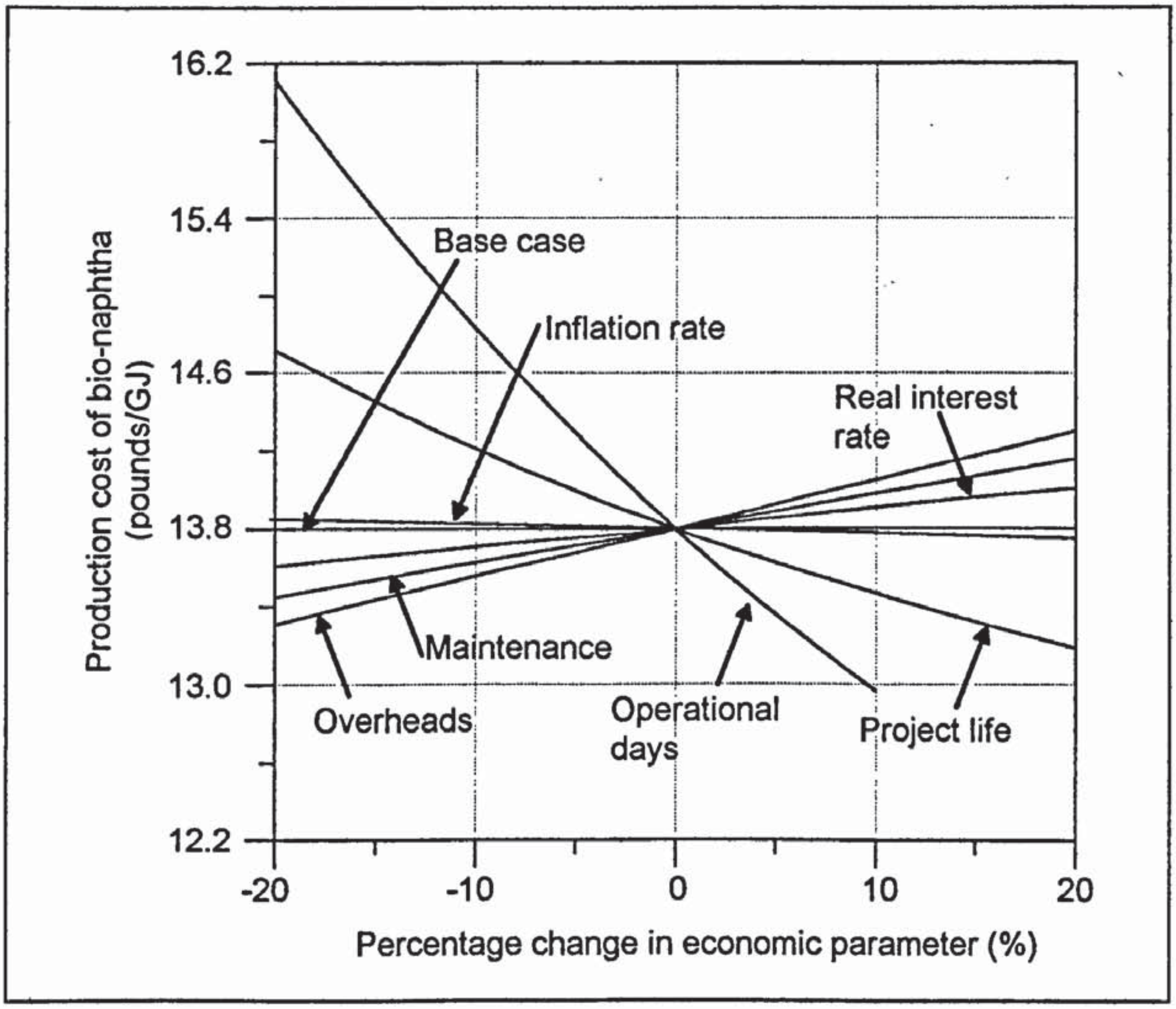


Figure 10.12. The effect of changes in financial parameters on the production costs of bio-naphtha.

Increasing the operational days per year increases the availability of the plant for production of the product and hence reduces the production cost of the biomass derived products. The other products available in BLUNT i.e. crude aromatics, diesel, gasoline and electricity would be expected to follow similar trends to those illustrated in Figures 10.11. and 10.12.

10.4. Summary.

Section 10.3. has given a sample of the types of results available from BLUNT and clearly shows that biomass derived processes are sensitive to changes in feedstock cost and to the nature of the feedstock itself. The generic flash pyrolysis process produces pyrolysis oil at a lower cost than the other flash pyrolysis processes modelled. It is cheaper to produce gasoline by the zeolite cracking potential case process than by coupling the present case process with a refining step and it is also cheaper to produce electricity from crude pyrolysis oil than from upgraded biomass products. The production costs of the biomass derived products are sensitive to changes in the financial parameters specifically to the availability and project life of conversion plants. The temperature of a flash pyrolysis process has a dramatic impact on the production costs of crude pyrolysis oil due to the effect of temperature on product yields. Feed moisture content and pyrolysis fuel options have small but quantifiable effects on the production cost of a pyrolysis oil product.

CHAPTER 11.

CONCLUSION.

11.1. Introduction.

The aims of the project were to compare selected biomass flash pyrolysis conversion, upgrading and power production technologies in terms of cost and performance, to develop new relationships and models for these technologies and to identify promising technologies and areas of further research. This has been achieved by analysing the selected technologies, determining the critical process parameters and utilising them to develop models based within a dedicated computer program. The models are not intended to be detailed chemical engineering models of a particular process but to represent "black box" approximations of the processes. Detailed engineering models are not possible due to a sparcity of data available on biomass thermochemical conversion processes and products and "black box" models have been utilised previously [28, 135] with some success.

The computer program written, BLUNT, estimates the performance and production costs of flash pyrolysis, upgrading, refining and power production processes on a consistent basis. The program allows the user to look at the effects of different feedstocks, flash pyrolysis and upgrading processes, macroeconomic factors and a range of specific process variables on the wood drying, generic flash pyrolysis, hydrotreating and power production models. The results from the model have been verified by comparison with published information where available and consultation with relevant sources.

The points raised in this Chapter summarise the issues which arose during modelling of the biomass conversion processes (Section 11.2.) and the economic results and analyses obtained from BLUNT (Section 11.3.).

11.2. Modelling.

11.2.1. The computer program.

The computer program written for the project incorporates the latest computer graphics to allow users not familiar with computers to make use of the program. It has multiple screens accessed using push down buttons and data entry options including text and pull down combo or list boxes. A range of help screens are included which have summarised information on the program operation and the technologies modelled including flowsheets for the conversion processes. The program operates in a range of currencies to enable it to be used not only in the United Kingdom but also in Europe and North America. BLUNT provides a visual display of the simulation results and also allows the user to printout or save the results to file. The outputs from BLUNT are:

- a visual summary of results;
- a visual display of the capital cost breakdown;
- a visual display of the summary of utilities usage;
- a visual display of the breakdown of annual costs;
- a visual display of the summary of additional costs;
- a printout of the results summary;
- a printout of the capital cost breakdown, utilities usage, annual costs, additional costs and the results summary;
- a results file containing the capital cost breakdown, utilities usage, annual costs, additional costs and the results summary; and
- a results file containing the full mass balance for the processing route.

11.2.2. Flash pyrolysis.

The flash pyrolysis models were developed for the processes of Ensyn, Union Fenosa and NREL using published data. These models are representative of the current status of flash pyrolysis development and the technologies utilised. The other flash pyrolysis model is that of the Generic flash pyrolysis process which was designed as a flexible tool to assist in the understanding of the process variables which affect a flash pyrolysis process. With this aim in mind the model was designed to allow both the mass and energy balances of the process to be controlled by the program User (See Chapter 6 for details).

There has been both experimental and theoretical work carried out on flash pyrolysis however, the modelling work carried out in this project has highlighted several areas in which there is little data available. Vapour residence time was identified in Chapter 2 as having a significant affect on the flash pyrolysis product yields but there was insufficient information available to allow this to be modelled and so the product yield algorithms in generic flash pyrolysis model were based around the pyrolysis temperature. The flash pyrolysis temperature was demonstrated in Chapters 2 and 6 to affect the quantity and relative proportions of the products from flash pyrolysis with the optimum temperature for liquid yields being around 500°C.

Feed moisture content was identified in Chapter 2 as being one factor which affected a biomass flash pyrolysis process (Section 2.3.1.). The modelling of flash pyrolysis in Chapter 6 re-emphasised this issue and demonstrated how the feed moisture content affected both the energy balance of the pyrolysis process and on the water content of the bio-oil product.

The use of the char and off-gas by-products from flash pyrolysis was shown in Chapter 6 to have a significant impact on the energy balance of the process. At temperatures around 500°C sufficient energy is available from combusting the char by-product to enable the flash pyrolysis process itself to be energy self-sufficient and to allow the off-gas to be exported as fuel gas for use in the feed drying process . The utilisation of the by-products increases the self-sufficiency of a biomass conversion process and reduces the production costs of the final liquid product.

11.2.3. Upgrading by hydrotreatment.

A hydrotreating model has been included in the program based on a design by Battelle PNL [130] and is described in Chapter 7. The deoxygenation extent of a crude oil in the hydrotreating model has been shown to have a significant impact on the quality and quantity of upgraded products, the hydrogen requirements and the size of the hydrotreating plant (See Chapter 7 for details). The oxygen content of the crude pyrolysis oil feedstock going into a hydrotreating process must be known in order to determine the oxygen content of the hydrotreated product. To enable the oxygen content of a crude pyrolysis oil to be determined, new algorithms were developed relating the carbon to oxygen, carbon to nitrogen and carbon to hydrogen ratios of the biomass feedstock to that of the pyrolysis product (See Chapter 6 for details).

The hydrotreating model incorporates new algorithms relating the deoxygenation of a crude pyrolysis feedstock to the quantity of the upgraded products produced, for deoxygenation extents between 0 and 100%. Chapter 7 demonstrated that the deoxygenation extent affects the quality of the upgraded product where product quality is measured by the amount of water remaining in the upgraded product. The higher the deoxygenation extent, the lower the amount of water remaining in the upgraded product and hence the higher it's quality. The amount of waste water and the contamination of that waste water by organic compounds was also demonstrated to be a function of deoxygenation extent. It has been demonstrated that increasing the deoxygenation extent increases the amount of waste water produced and lowers the contamination of the waste water by organic compounds. Additionally, the deoxygenation extent affects the amount of hydrogen required for hydrotreating, i.e. the higher the deoxygenation extent the more hydrogen is required.

11.2.4. Upgrading by zeolite cracking, power production and the refining of upgraded products..

Models were developed for the zeolite cracking of crude pyrolysis oil based on processes investigated by the IEA [30] (See Chapter 8) representing present and potential cases. Algorithms were developed relating the products obtained from the zeolite cracking processes to the pyrolysis vapour feed and the models were optimised for energy self-sufficiency. The processes modelled potentially allow the production of crude aromatics and gasoline from crude bio-oil without the hydrogen required for hydrotreatment.

Simple models for refining hydrotreated oil and crude aromatics from zeolite cracking (present case) were developed to give diesel and gasoline products respectively. A model representing the production of electricity from crude pyrolysis oil, hydrotreated oil and crude aromatics using engines or turbines was incorporated into the program.

11.2.5. The modelling of macroeconomic factors.

A section on macroeconomics was included in BLUNT because it allows the influence of outside factors on the viability of a biomass conversion process to be determined. The macroeconomic factors modelled included sulphur and carbon taxes on fossil fuels, irrigation water pricing and regional assistance. The implementation of carbon and sulphur taxes on fossil fuels has been proposed but the actual type or size of the taxes are as yet unknown. The macroeconomics models produced and incorporated into BLUNT have been designed to be as flexible and wide ranging as possible to cover the likely aspects of any future legislation.

11.3. Economics.

The choice of feedstock has a noticeable affect on product costs, as shown in Chapter 10. Feedstock cost itself is generally the highest single cost item in the annual cost breakdown of a biomass conversion plant and has a significant effect on the production cost of a biomass derived product. The production costs of biomass derived products, in particular the upgraded and refined products, are considerably higher than those of fossil derived fuels. The use of low cost feedstocks and macroeconomic factors could lower the cost of biomass derived products in the future and allow them to compete with fossil derived fuels. However, the future of biomass thermochemical conversion could lie with the direct use of the crude pyrolysis oil as a chemical feedstock or for power production although considerable research and development is still required to overcome some of the characteristics associated with pyrolysis oils. For biomass to compete with fossil fuels in electricity production the crude pyrolysis liquid would have to be utilised to minimise the costs incurred for biomass conversion.

11.4. Summary.

The computer program developed, BLUNT, utilises the latest programming languages to give a user friendly environment for the modelling and cost estimations. It was designed and written specifically for this purpose, taking into account the limitations of earlier computer packages. This Chapter has summarised the modelling work carried out in the development of the program and highlighted the major conclusions drawn from it. The models derived were validated by the appropriate researchers in the field where possible and papers on the modelling have been both delivered at conferences and published in journals. One of the published papers is given in Appendix C and Appendix D gives a list of published papers and conference presentations. The algorithms derived for the generic flash pyrolysis and hydrotreating models are original and have contributed to a wider understanding of the processes. The costs of the upgrading models in particular have already altered the perceptions of biomass derived products and directed research away from high cost transportation fuels towards lower cost, higher value chemicals recovery.

CHAPTER 12.
RECOMMENDATIONS.

12.1. Introduction

The recommendations resulting from the work reported in this thesis fall into two categories: those dealing with the modelling and process design of the biomass technologies (detailed in Section 12.2.) and those dealing with the development of the simulation program, BLUNT which models the technologies (detailed in Section 12.3.).

12.2. Recommendations for further work on the modelling / process design aspects.

There are several areas which have been highlighted by the modelling work in which information is scarce. The recommendations for work on the process design and research side of biomass thermochemical conversion are given below.

- Vapour residence time has been highlighted as a significant variable in the flash pyrolysis of biomass. However, there has been little research carried out on it's affect on product yields at pyrolysis temperatures between 400 and 650°C. It is recommended that trials be carried out at laboratory scale to generate data and allow correlations for the affect of both residence time and temperature to be generated. This will lead to a more detailed understanding of the links between product yields, pyrolysis temperature and vapour residence time.

- The preliminary results from the hydrotreating model indicate that if hydrotreating is to be used in the future to deoxygenate crude pyrolysis oil the costs must be reduced. One option to reduce the costs is to consider the partial hydrotreatment or stabilisation of the pyrolysis oil. It is recommended that work should be carried out at low deoxygenation extents to determine the process requirements, products and physical characteristics of the products from partial hydrotreatment to determine whether this is a viable option.

- There are two hydrogen source / off-gas disposal options which have been incorporated into the hydrotreating model. Since the cost of the required hydrogen is a major contributor to the production cost of hydrotreated oil, options which recover and recycle hydrogen will reduce the costs. Other options which recommended for investigation and incorporation into BLUNT are steam reforming and hydrogen recovery of the off-gas and the generation of hydrogen by gasification of biomass.

- Plant sizes between 25 and 100 tonnes/day have been incorporated in BLUNT with a scaling of the equipment and costs. However, the costs do not allow for the minimal equipment and increased labour requirements utilised by small scale operations. It is recommended that the equipment and requirements for small scale operation be investigated and the models presently incorporated in BLUNT be modified accordingly.

- The upgrading of pyrolysis vapours by zeolite cracking is regarded as a potential method of providing gasoline from renewable sources, although like hydrotreating the costs at present are quite high and the process is not economic. However, in the future the costs of biomass derived products are expected to fall in relation to fossil products as fossil resources are depleted or taxation on sulphur and carbon emissions are implemented. With this in mind it is recommended that further work is carried out on zeolite cracking to determine the process variables and the viability of this process from an engineering point of view.

12.3. Recommendations for further work on the computational aspects of the project or the development of BLUNT.

The computer program written during this project utilised the most advanced programming software available at the time and the criticisms and comments on earlier modelling and simulation programs [28,135] were considered and acted upon when BLUNT was designed. However, due to time or resource constraints at the time of programming some ideas were not implemented and the following recommendations for updating BLUNT are made.

- A sub program could be written to allow the program user to modify and permanently update the data files used by BLUNT for the input of chemical and financial data. This will allow for example, the User to update the values of the currency exchange rates or incorporate currencies not presently available in BLUNT.

- A sub program could be added to allow the program user to perform sensitivity analyses on the production costs of the biomass derived products. This could be incorporated into BLUNT in a similar fashion to the macroeconomic factors (described in Chapter 9) as an additional feature after the main computational section of the program. The sub program would take a selected parameter such as the cost of a specific utility or the value of a financial parameter and generate new values for the production cost based on user determined increases or decreases in the parameter. The resulting data could be displayed and printed in graphical form. The ability to plot graphs is available in Visual Basic and BLUNT would have to be modified to incorporate this facility.

- In the current version of BLUNT the cost of utilities are included as a unit cost for example, the cost of hydrogen in Great Britain pounds/tonne. The default values chosen for these costs are intended to include all production costs for the particular utility. However, to enable the model to be more internally consistent and the biomass conversion process to be self-sufficient in terms of utilities, the use of feedstock to generate power, hydrogen or fuel gas should be incorporated allowing the cost of the specific utility to vary with the cost of feedstock. For example, a feed gasification system could be incorporated to provide the hydrogen required for hydrotreating.

- In the current version of BLUNT although the results from a simulation run can be saved to a text file, this file is overwritten when the next simulation run is performed. This applies to both the results summary and the mass balance results files. This is due to the way in which the graphical interface is set up i.e. the name of the results file has to be specified in the source code of the program to allow the results to be sent and the file printed. It is hoped that if further investigation is carried out on how Visual Basic works, this problem can be overcome, possibly by writing a routine which can change the name of the results file one produced to a file name specified by the program User.

- Due to the way in which BLUNT is configured, the source files for the program must be in a specific place on the hard disk of a computer i.e. in the BLUNT directory on the C-drive of a computer. If the specific files required by BLUNT are not present in the correct directory the program crashes. This configuration means that BLUNT can only be run from the hard disk and does not work on floppy disk or networked systems. Further investigation into the Visual Basic programming system is recommended to modify this problem and allow BLUNT to operate from any location.

- The addition of extra step models for conversion processes into the model is difficult due to the way in which the program is configured. However, additional processes for example gasification, could be added by incorporating them in a self-contained Pascal or similar program linked and initiated by BLUNT's graphical interface. The graphical interface would require some modification but this would be simpler than rewriting the present source code for the program to incorporate new processes. It is recommended that the gasification and biochemical conversion models in AMBLE [28] be added to BLUNT in this way to give a package which considers the complete range of biomass conversion options.

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BLUNT

Biomass Liquefaction and Upgrading using Novel Technologies.

Program operating manual.

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Sponsored by the Commission of the European Communities under the JOULE programme.

ABOUT BLUNT.

What is BLUNT?

BLUNT was developed at the Department of Chemical Engineering and Applied Chemistry, Aston University. The development of BLUNT was sponsored by the Commission of the European Communities under the JOULE (Joint Opportunities for Unconventional or Long-term Energy Supply) programme.

BLUNT is a computer simulation package which assesses and compares the economic and technical opportunities for producing and upgrading crude pyrolysis liquids into higher quality fuels or electricity.

BLUNT's system requirements.

BLUNT can be run on any IBM compatible computer with Microsoft Windows Version 3.0 or later. The program must be installed correctly on a hard disk using the provided installation programme - it will not run from a floppy disk.

When installed, BLUNT will run like any other Windows application, allowing the user to use either keyboard or mouse for entering data and switching between screens.

What to do if you have any problems running or installing BLUNT.

It is recommended that you thoroughly read this manual before attempting to either install or operate BLUNT.

If you have any problems or queries not dealt with by this manual contact:-

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United Kingdom.

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M-L Cottam
July 1994

CONTENTS.

7. APPENDICES.

SECTION 1.
THE BLUNT OPERATING MANUAL.

1.1. How to use this manual.

The BLUNT operating manual has been designed to give non-computer literate users an easy to follow guide to the program. Users familiar with computers may be tempted to skip Sections or attempt to run the program without first reading this manual. This is not recommended. Before attempting to install and run BLUNT, you are advised to read Sections 2, 3, 4 and 5 of this manual.

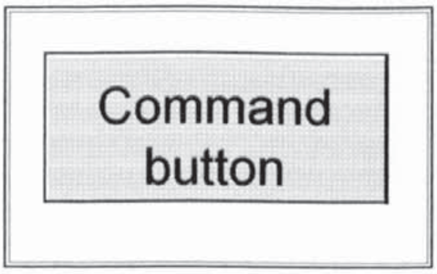
Section 2	describes the purpose behind the program (what it aims to achieve and how it is carried out);
Section 3	describes how BLUNT is installed on your computer;
Section 4	describes the graphical screens of which the program is composed; and
Section 5	details a sample simulation run.

Once you are more familiar with the operation and technologies modelled in BLUNT, Section 6 will enable you to use the more advanced features built into the program.

1.2. Terminology.

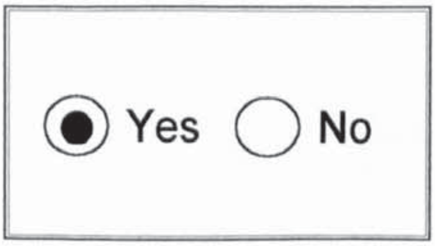
The BLUNT Operating Manual uses terminology with which an inexperienced computer user may not be familiar. The following section describes some of the more common phrases which are used.

Command buttons



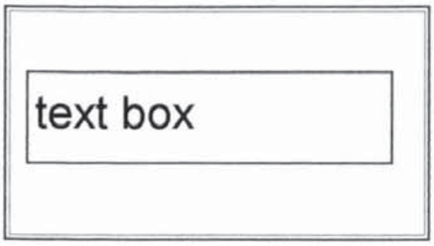
Command buttons are raised, rectangular buttons with text on the face. They are activated by clicking with the mouse or pressing the <RETURN> key when the focus is on the button. Command buttons are referred to in the manual by the text on the button e.g. *RUN*.

Radio buttons



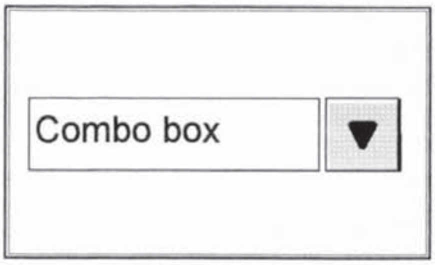
Radio buttons are small round buttons with text on the right hand side. They are activated by clicking with the mouse or pressing the <RETURN> key when the focus is on the button. Radio buttons are referred to in the manual by the text next to the button e.g. *YES*.

Text boxes



Text boxes are rectangular boxes containing numbers or text. The contents of the text box may altered by the program user by clicking on, highlighting and overwriting. Text boxes are referred to in the manual by the text on the left of the box i.e. their titles e.g. Plant feed input.

Combo boxes.



Combo boxes are text boxes with an arrow on the right hand side. Clicking on the arrow or box activates a list within the combo box and allows the user to scroll down the list to the item to be selected. Combo boxes are referred to in the manual by the text on the left of the combo box i.e. their titles e.g. Product.

SECTION 2.
INTRODUCTION TO THE BLUNT COMPUTER PROGRAM.

2.1. Introduction.

The thermochemical conversion of biomass by flash pyrolysis produces a crude liquid product that can be used directly to substitute for conventional fossil fuels or upgraded to a higher quality fuel. Both the crude and upgraded products may be utilised for power generation.

BLUNT models the flash pyrolysis of biomass with subsequent upgrading and refining of the crude liquid products to produce higher value and more marketable liquid fuel products. The program assesses and compares the economic and technical opportunities for upgrading crude pyrolysis liquids into the higher quality fuels demanded by more advanced technologies. Complete integrated processes from wood delivered to the factory gate through to an end product are modelled.

The computer program works by building up a selected processing route from a number of process steps through which the material passes sequentially. The process steps modelled in a process route include the feedstock reception and storage followed by an appropriate combination of thermochemical conversion, upgrading of the crude liquid product, refining of the upgraded product, liquid product storage and electricity production from the crude or upgraded product. Each process step has a step model that calculates the mass and energy balances, the utilities usage and the capital cost for that step of the process. The results of the step models are combined to determine the performance of the whole conversion route.

2.2. The scope of BLUNT.

The feedstocks modelled in BLUNT are:

- wood
- straw
- sorghum bagasse

To give products of:

- crude pyrolysis oil
- hydrotreated oil (hydrotreated pyrolysis oil)
- crude aromatics (from zeolite cracking (present case model) of pyrolysis vapours)
- diesel (hydrotreated pyrolysis oil)
- gasoline (from refined crude aromatics or zeolite cracking (potential model) of pyrolysis vapours)
- electricity (from crude pyrolysis oil, hydrotreated oil or crude aromatics)

The program works for biomass feed inputs to the pretreatment section between 25 and 1000 tonnes/day on a dry ash free (d.a.f.) basis.

The products and conversion routes available in the program for wood and the combination of primary, secondary and tertiary conversion processes required are given in Table 2.1. For wood there are four flash pyrolysis processes available - the Generic, Ensyn RTP, Fenosa and NREL Vortex processes. As can be seen from Table 2.1. the total number of conversion routes possible for a wood feedstock as modelled in BLUNT is 36 considering all combinations of pyrolysis technology, product and processing route.

Table 2.1.
Products and conversion routes modelled in BLUNT for a wood feedstock.

Product	Primary conversion	Secondary conversion	Tertiary conversion
Crude pyrolysis oil	Generic flash pyrolysis	none	none
	Ensyn RTP pyrolysis	none	none
	Fenosa flash pyrolysis	none	none
	NREL vortex pyrolysis	none	none
Hydrotreated oil	Generic flash pyrolysis	hydrotreating	none
	Ensyn RTP pyrolysis	hydrotreating	none
	Fenosa flash pyrolysis	hydrotreating	none
	NREL vortex pyrolysis	hydrotreating	none
Crude aromatics	Generic flash pyrolysis	zeolite cracking (pres)	none
	Ensyn RTP pyrolysis	zeolite cracking (pres)	none
	Fenosa flash pyrolysis	zeolite cracking (pres)	none
	NREL vortex pyrolysis	zeolite cracking (pres)	none
Diesel	Generic flash pyrolysis	hydrotreating	none
	Ensyn RTP pyrolysis	hydrotreating	none
	Fenosa flash pyrolysis	hydrotreating	none
	NREL vortex pyrolysis	hydrotreating	none
Gasoline	Generic flash pyrolysis	zeolite cracking (pres)	refining
	Ensyn RTP pyrolysis	zeolite cracking (pres)	refining
	Fenosa flash pyrolysis	zeolite cracking (pres)	refining
	NREL vortex pyrolysis	zeolite cracking (pres)	refining
	Generic flash pyrolysis	zeolite cracking (pot)	none
	Ensyn RTP pyrolysis	zeolite cracking (pot)	none
	Fenosa flash pyrolysis	zeolite cracking (pot)	none
	NREL vortex pyrolysis	zeolite cracking (pot)	none
Electricity	Generic flash pyrolysis	none	power gen.
	Ensyn RTP pyrolysis	none	power gen.
	Fenosa flash pyrolysis	none	power gen.
	NREL vortex pyrolysis	none	power gen.
	Generic flash pyrolysis	hydrotreating	power gen.
	Ensyn RTP pyrolysis	hydrotreating	power gen.
	Fenosa flash pyrolysis	hydrotreating	power gen.
	NREL vortex pyrolysis	hydrotreating	power gen.
	Generic flash pyrolysis	zeolite cracking (pres)	power gen.
	Ensyn RTP pyrolysis	zeolite cracking (pres)	power gen.
	Fenosa flash pyrolysis	zeolite cracking (pres)	power gen.
	NREL vortex pyrolysis	zeolite cracking (pres)	power gen.

Table 2.2. gives the products and conversion routes available from sorghum bagasse and straw. The only pyrolysis processs modelled which will take straw as a feedstock is the Ensyn RTP system (See Appendix for further details). As can be seen from Table 2.2. the total number of conversion routes modelled in BLUNT for straw / sorghum bagasse is 9 considering all combinations of product and processing route.

Table 2.2.
Products and conversion routes modelled in BLUNT for straw or sorghum bagasse feedstocks.

Product	Primary conversion	Secondary conversion	Tertiary conversion
Crude pyrolysis oil	Ensyn RTP pyrolysis	none	none
Hydrotreated oil	Ensyn RTP pyrolysis	hydrotreating	none
Crude aromatics	Ensyn RTP pyrolysis	zeolite cracking (pres)	none
Diesel	Ensyn RTP pyrolysis	hydrotreating	refining
Gasoline	Ensyn RTP pyrolysis	zeolite cracking (pres)	refining
	Ensyn RTP pyrolysis	zeolite cracking (pot)	none
Electricity	Ensyn RTP pyrolysis	none	power gen.
	Ensyn RTP pyrolysis	hydrotreating	power gen.
	Ensyn RTP pyrolysis	zeolite cracking (pres)	power gen.

Figure 2.1 shows how the step models in the program fit together to form a conversion route. For example, the processing steps that make up the conversion route for crude pyrolysis oil include pretreatment (reception, storage, drying and size reduction) of the feedstock, flash pyrolysis (by either generic, Ensyn, Fenosa or NREL processes) and liquid product storage.

Several of the step models - wood drying, generic flash pyrolysis, hydrotreating and power production have variables which may be altered by the user to customise the models. Details of these models can be found in the Appendix and the graphics screens used to customise the models are described in Section 4.

The financial parameters used in the program (currency unit and exchange rate, utilities costs and financial variables) can be modified by the user to customise the program. Details of how to alter the financial parameters can be found in Section 4.

2.3. Program output.

Blunt provides a visual display of the simulation results and also allows the user to printout or save the results to file. The outputs from BLUNT are:

1. a visual summary of results;
2. a visual display of the capital cost breakdown;

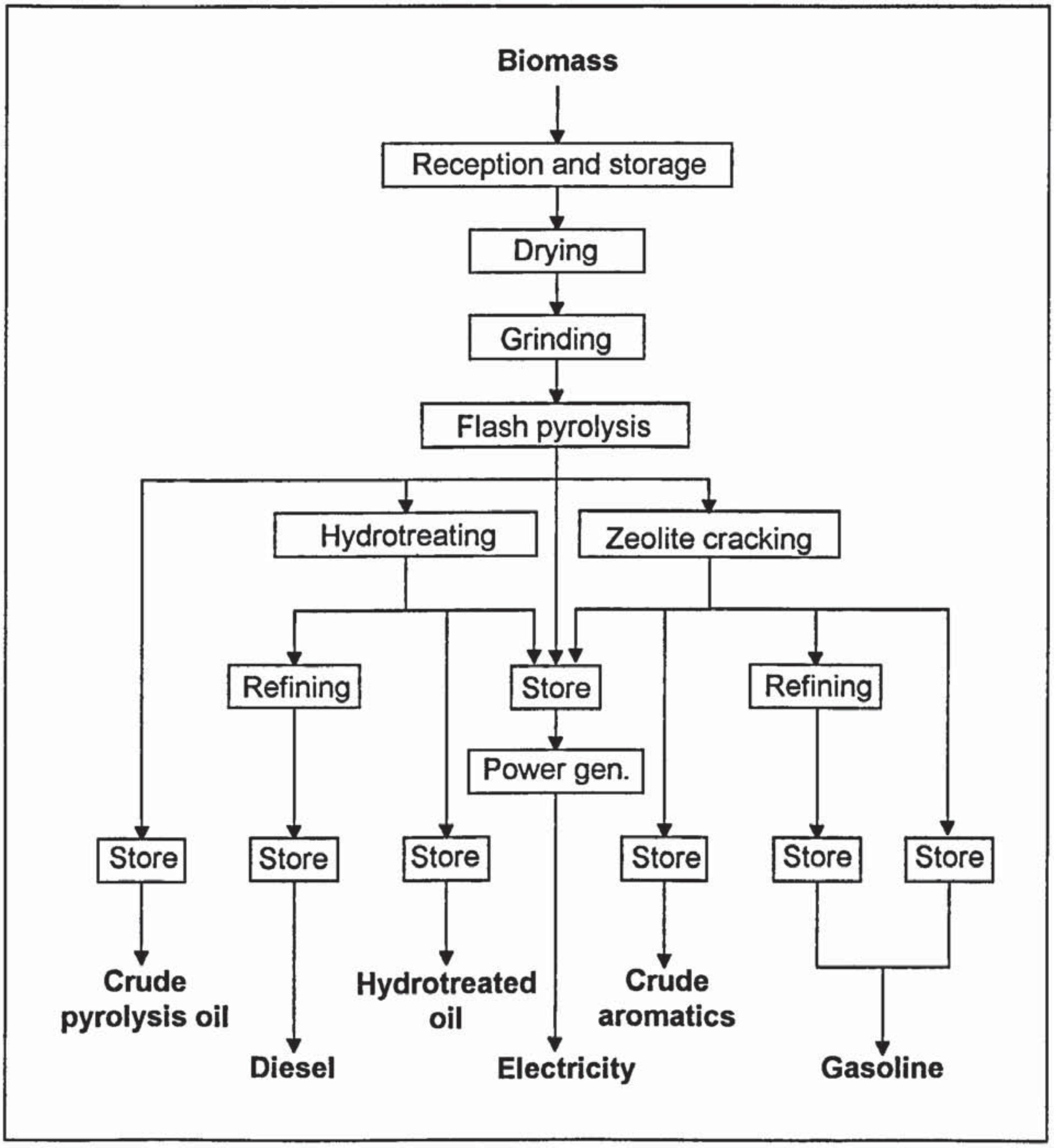


Figure 2.1. Conversion routes modelled in the computer program.

3. a visual display of the summary of utilities usage;
4. a visual display of the breakdown of annual costs;
5. a visual display of the summary of additional costs;
6. a printout of the results summary;
7. a printout of the capital cost breakdown, utilities usage, annual costs, additional costs and the results summary;
8. a results file containing the same information as (7); and
9. a results file containing the full mass balance for the processing route.

SECTION 3.
PROGRAM INSTALLATION.

3.1. Introduction.

The files which make up the BLUNT program are supplied on the floppy disk which accompanies this manual. The files are:-

• Setup.exe	BLUNT's installation file.
• Blunt.exe	BLUNT's executable file.
• Blunt.dll	BLUNT's procedure library.
• Fin2.dat	a datafile used by BLUNT.
• Stepord.dat	a datafile used by BLUNT.
• Compinfo.dat	a datafile used by BLUNT.
• Stepinfo.dat	a datafile used by BLUNT.
• Basicinf.dat	a datafile used by BLUNT.
• Vbrun300.dll	a graphical file required by BLUNT.
• Threed.dll	a graphical file required by BLUNT.

Before installing BLUNT on your computer please check that all these files are present on the installation disk. If any of the files are missing contact **Dr A.V. Bridgwater** at the address given at the beginning of this manual.

WARNING!

DO NOT COPY THE FILES CONTAINED ON THE INSTALLATION DISK DIRECTLY TO YOUR HARD DISK. THE FILES REQUIRED BY BLUNT MUST BE IN THE CORRECT DIRECTORIES ON YOUR HARD DISK FOR THE PROGRAM TO WORK. THE INSTALLATION PROGRAM SUPPLIED MUST BE USED TO INSTALL BLUNT AS IT WILL COPY THE FILES TO THE CORRECT PLACES ON YOUR HARD DISK. THE PROGRAMMER AND DISTRIBUTOR OF BLUNT TAKE NO LIABILITY FOR INCORRECT INSTALLATION OF BLUNT IF THE INSTALLATION PROGRAM SUPPLIED IS NOT USED.

3.2. Installing BLUNT.

3.2.1. Important information.

There are three pieces of information required by the BLUNT installation program and it is recommended that you check them before you begin to install BLUNT. The information required is:-

- the path for the Windows System directory;
- the source drive for the BLUNT files; and
- the destination drive for the BLUNT files.

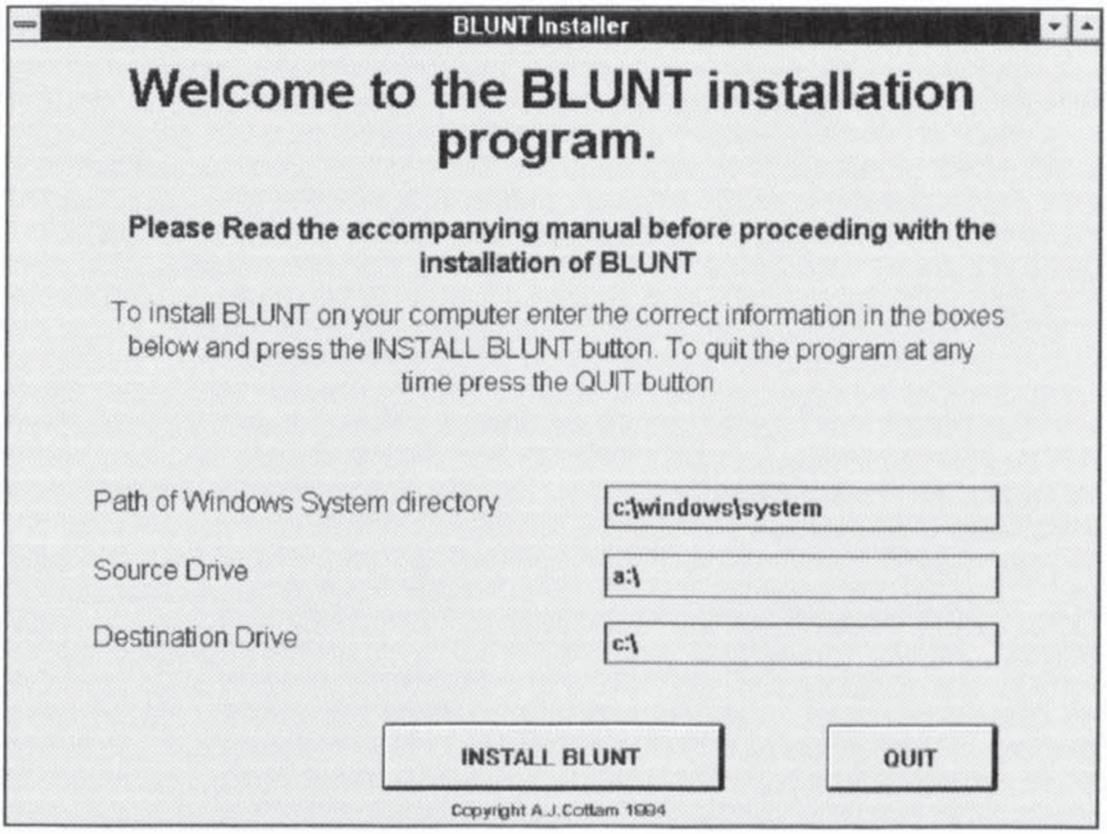
3.2.2. Step by step installation of BLUNT.

To install BLUNT on your computer follow the steps outlined below.

1. Start Windows;
2. Insert the BLUNT installation disk in the appropriate disk drive;
3. Click on the **File** item on the **Windows Program Manager** main menu to open the **File Menu.**
4. Click in the **Run** item on the **File** Menu to open the **Run Dialog Box.**

5. Click on the **Browse** button on the **Run Dialog Box** to open the **Browse Dialog Box**.
6. Select the floppy disk drive for the disk containing the BLUNT files.
7. Click on the filename **SETUP.EXE** displayed in the program list box to select the installation program.
8. Click on the **OK** button to accept the **SETUP.EXE** file and return to the **Browse** dialog box.
9. Click on the **Run** button to close the Run Dialog Box. and run **SETUP.EXE**. The BLUNT installation screen will appear as shown in Figure 3.1.
10. installation program's screen to ensure that the information as detailed in Section 3.2.1. is correct.
11. When you are satisfied with the information in the text boxes press the **INSTALL BLUNT** button to start the installation procedure. If you wish to check any information or decide not to install BLUNT press the **QUIT** button.
12. When the BLUNT files have been installed a dialog box appears with the message "BLUNT installation complete". You may now check the entries in the text boxes on the leave the installation program.

Steps 1-12 above deal with the installation of the BLUNT files on your computer. To allow easy access to BLUNT the BLUNT icon can be placed in one of the Windows Groups in the Windows Program Manager. The procedure to do this is outlined in steps 13-23.



Welcome to the BLUNT installation program.

Please Read the accompanying manual before proceeding with the installation of BLUNT

To install BLUNT on your computer enter the correct information in the boxes below and press the INSTALL BLUNT button. To quit the program at any time press the QUIT button

Path of Windows System directory: c:\windows\system

Source Drive: a:\

Destination Drive: c:\

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Figure 3.1. The screen for the BLUNT installation program.

13. Open a Windows Group such as **Applications** by double clicking on the appropriate icon.
14. Click on the **File** item on the **Windows Program Manager** main menu to open the **File Menu**.
15. Click in the **New** item on the **File Menu** to open the **New Program Object** dialog box.
16. Click on the **Program Item** radio button to install a new item to the **Applications** Group.
17. Press **OK** to accept a new Program Item and to open the **Program Item Properties** dialog box.
18. Click on the **Browse** button on the **Program Item Properties** dialog box to open the **Browse Dialog Box**
19. Select the disk drive on which the BLUNT files were installed.
20. Double click on the **C:** prompt in the directories list box to show all the directories on the selected disk drive.
21. Double click on the BLUNT directory listed in the directory list to open the directory.
22. Click on the filename **BLUNT.EXE** displayed in the program list box to select the program.
23. Click on the **OK** button to install the BLUNT.EXE file in the **Applications** Group.

If you have followed steps 1-22 correctly a sun-shaped icon should appear in the selected Windows Group. BLUNT is now ready for use. You are advised to read Sections 4 and 5 of this manual before running BLUNT.

SECTION 4.
THE GRAPHICAL SCREENS.

4.1. Initialising BLUNT.

BLUNT may be initialised by double-clicking on the icon that is generated when the program is installed under Windows (See Section 3 for further details). When BLUNT is initialised, the program activates the title screen as shown in Figure 4.1. The BLUNT title screen has three command buttons. The functions of the buttons are:-

START THE PROGRAM	closes the title screen and opens the primary process inputs screen described in Section 3.2.
HELP	opens the help screen detailing more information about the program.
QUIT	closes BLUNT and returns the program user to the Windows Program Manager

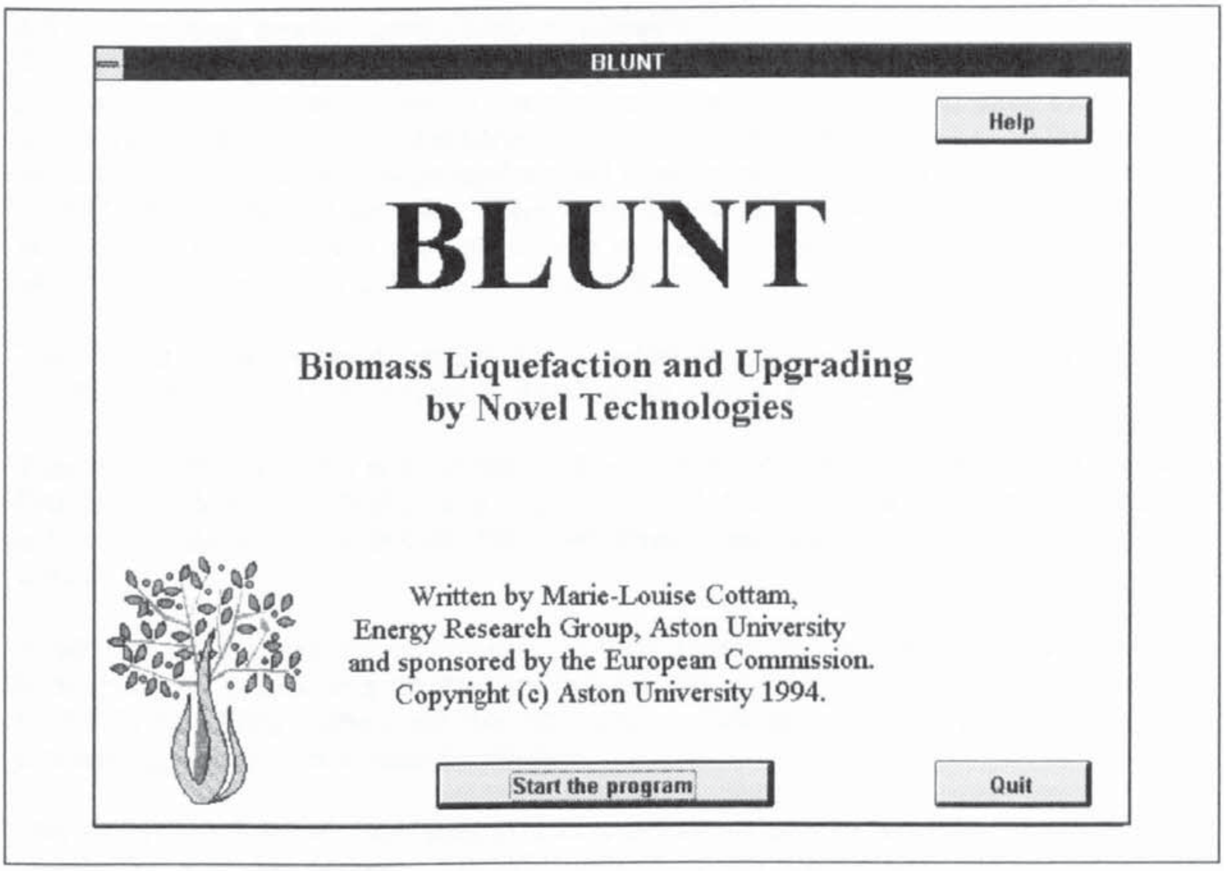
4.2. The primary process inputs screen.

The Primary Process Inputs (PPI) screen allows the user to select the essential information required for running BLUNT. The primary data consists of the feedstock, product, conversion route and the feed input to the plant. The PPI screen is shown in Figure 4.2. The command buttons available on the PPI Screen are *HELP*, *QUIT*, *RUN* and *CUSTOMISE*. The function of these buttons are as follows:-

QUIT	closes BLUNT and returns the user to the Windows Program Manager.
RUN	allows the user to run the program with the process inputs and access the results display screens.
CUSTOMISE	allows the user to access the screens which permit the customisation of the step models and financial parameters.
HELP	allows the user to access the program assistance screens which give details on the scope of BLUNT and the processes modelled.

The feed, product and conversion route are selected from a list of items which appear when the mouse is clicked on the relevant combo box. The user is not permitted to type in an extra item to the lists because only the items present in the combo boxes are permitted by the program. Selection of the feedstock and product determines the items which appear in the conversion route combo box. For example, selecting a wood feedstock and crude pyrolysis oil as the product would give conversion routes of generic flash pyrolysis, Ensyn RTP pyrolysis, NREL vortex pyrolysis and Fenosa flash pyrolysis in the conversion route list box. If the feedstock is changed to straw the only item appearing in the conversion route list box would be Ensyn RTP pyrolysis as this is the only pyrolysis process in BLUNT which can take a straw feedstock. Defaults for the feed, product and conversion route of wood, crude pyrolysis oil and generic flash pyrolysis respectively have been preset in the program.

The feed input or throughput has a default setting of 100 d.a.f. tonnes/day. This can be changed by the program user to accept any values between 25 and 1000 d.a.f. tonnes/day by altering the number which appears in the relevant text box.



BLUNT

Biomass Liquefaction and Upgrading
by Novel Technologies

Written by Marie-Louise Cottam,
Energy Research Group, Aston University
and sponsored by the European Commission.
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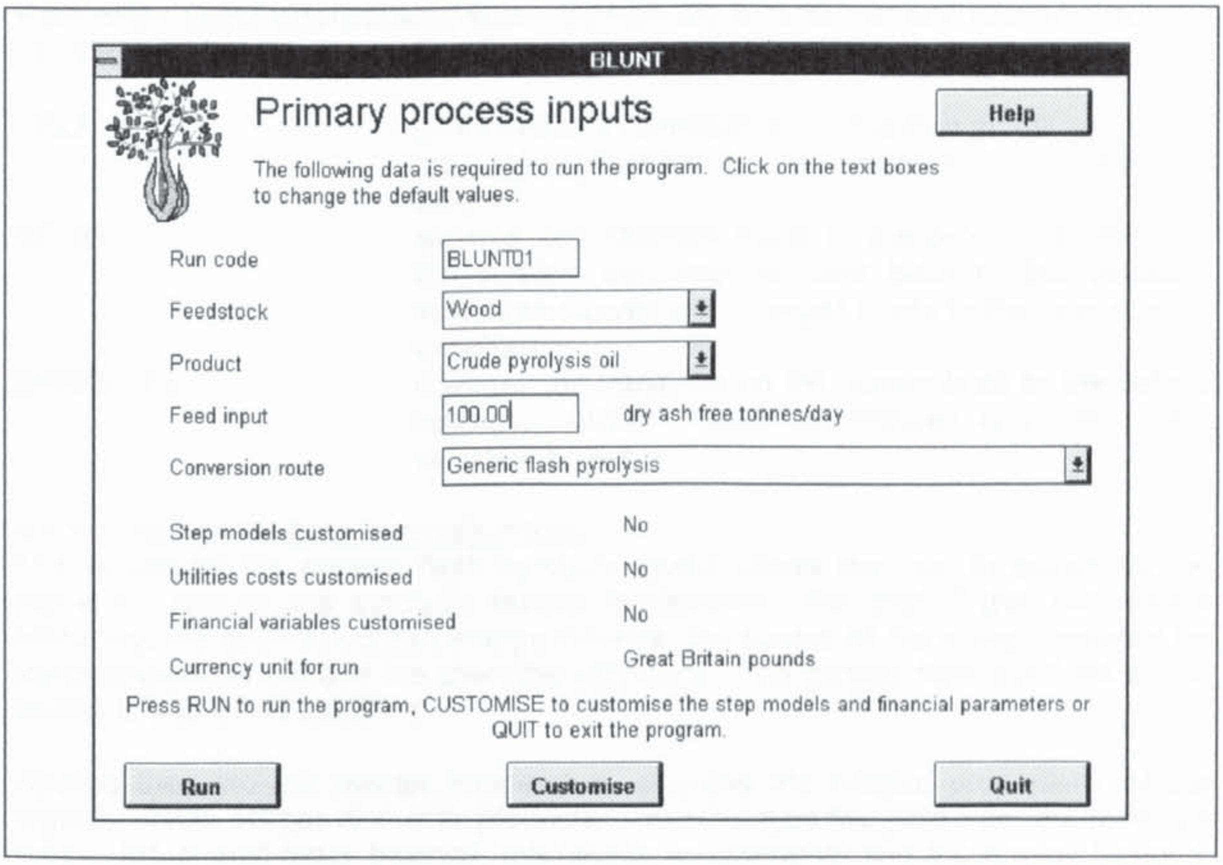
Figure 4.1. The BLUNT title screen.



Primary process inputs

The following data is required to run the program. Click on the text boxes to change the default values.

Run code	BLUNT01
Feedstock	Wood
Product	Crude pyrolysis oil
Feed input	100.00 dry ash free tonnes/day
Conversion route	Generic flash pyrolysis
Step models customised	No
Utilities costs customised	No
Financial variables customised	No
Currency unit for run	Great Britain pounds

Press RUN to run the program, CUSTOMISE to customise the step models and financial parameters or QUIT to exit the program.

Figure 4.2. The Primary Process Inputs Screen.

4.3. The step model customisation screens.

The step models for generic flash pyrolysis, drying, hydrotreating and power production have variables which the program user can change to alter the ways in which the models behave. The screen for selecting the step models (and for selecting the financial parameters) to be customised is accessed from the PPI screen when the *CUSTOMISE* button is pressed. The selection screen is shown in Figure 4.3. The selection of the feed and product on the primary process inputs screen determines which of the step model buttons are available.

The drying model only operates for a wood feedstock so selection of straw or sorghum bagasse as the feedstock deactivates the *DRYING* button.

If generic flash pyrolysis is selected as the pyrolysis process from the list of Generic, Ensyn, Fenosa and NREL the *GENERIC FLASH PYROLYSIS* button will be activated. Selection of any of the other three pyrolysis processes deactivates the button.

If either hydrotreated oil or diesel or power (when the processing route involves hydrotreating) is selected as the product on the primary process inputs screen the *HYDROTREATING* button will be activated. Selection of any other products or processing routes deactivates this button.

The *POWER PRODUCTION* button is only activated when electricity is selected as a product on the PPI screen. Clicking on any of the step model buttons which are activated sends the user to the appropriate screen. Clicking on the *RETURN* button at the bottom of the screen send the user back to the PPI screen.

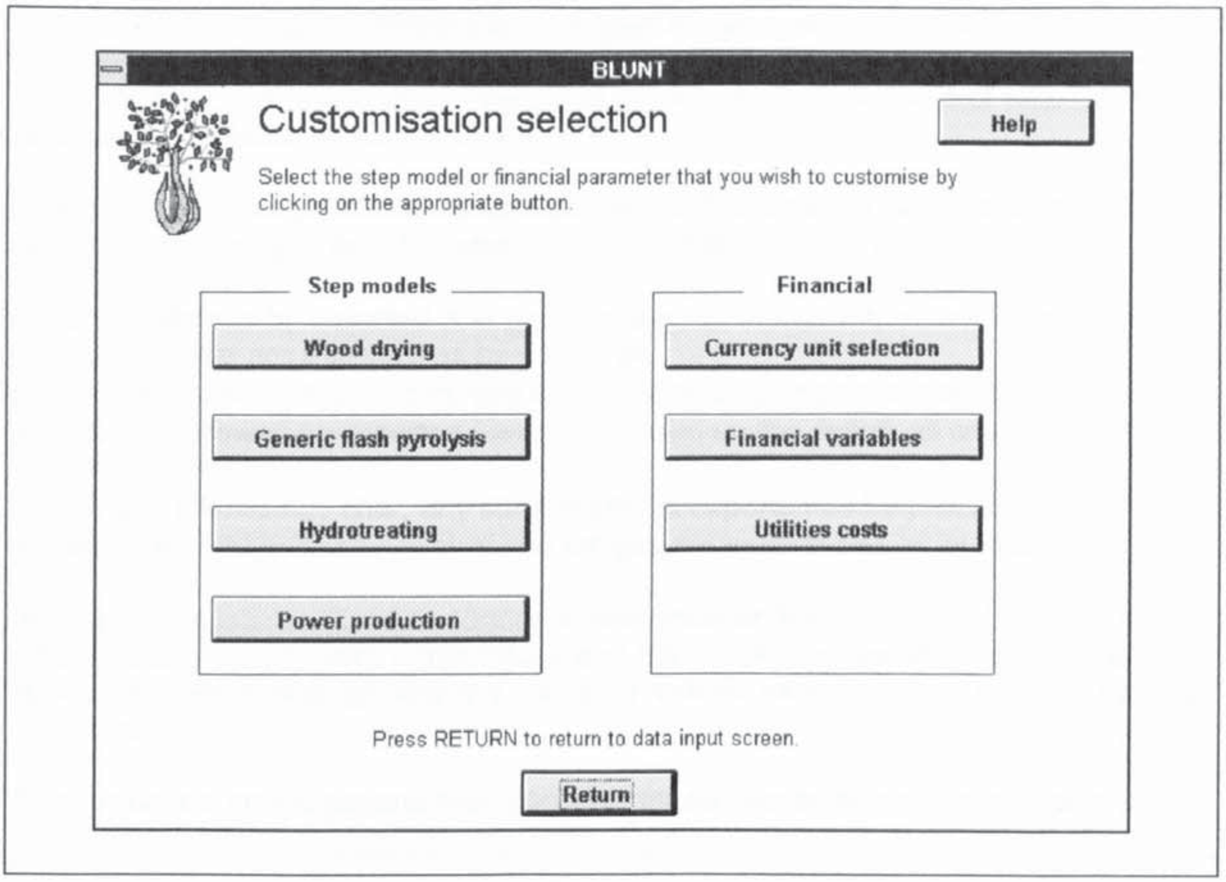
Each of the step model screens accessed from the selection screen has three buttons - *HELP*, *DEFAULTS* and *ACCEPT*. Details of each button is given below.

HELP	gives further information about the step model and allows the user to view the process flowsheet for the selected model.
ACCEPT	accepts any changes made to the default variables on the screen, switches the user back to the selection screen and saves any changes made to the step model's variables.
DEFAULTS	changes the variables on the screen back to the default settings. *ACCEPT* must be pressed to return to the selection screen.

4.3.1. The generic flash pyrolysis model.

The screen for the generic flash pyrolysis model allows the user to customise the model by altering the pyrolysis reactor temperature, the char/off-gas combustion efficiency, the quench column energy balance, the source of the energy required for the pyrolysis process and the char filter efficiency. The generic flash pyrolysis model screen is shown in Figure 4.4.

Altering the pyrolysis reactor temperature changes the relative proportions of the organic, water, off-gas and char products. This changes the yield from the pyrolysis model, the overall mass balance, the utilities requirements and the energy balance. The char and off-gas produced from the pyrolysis process are used to provide the pyrolysis energy requirements.

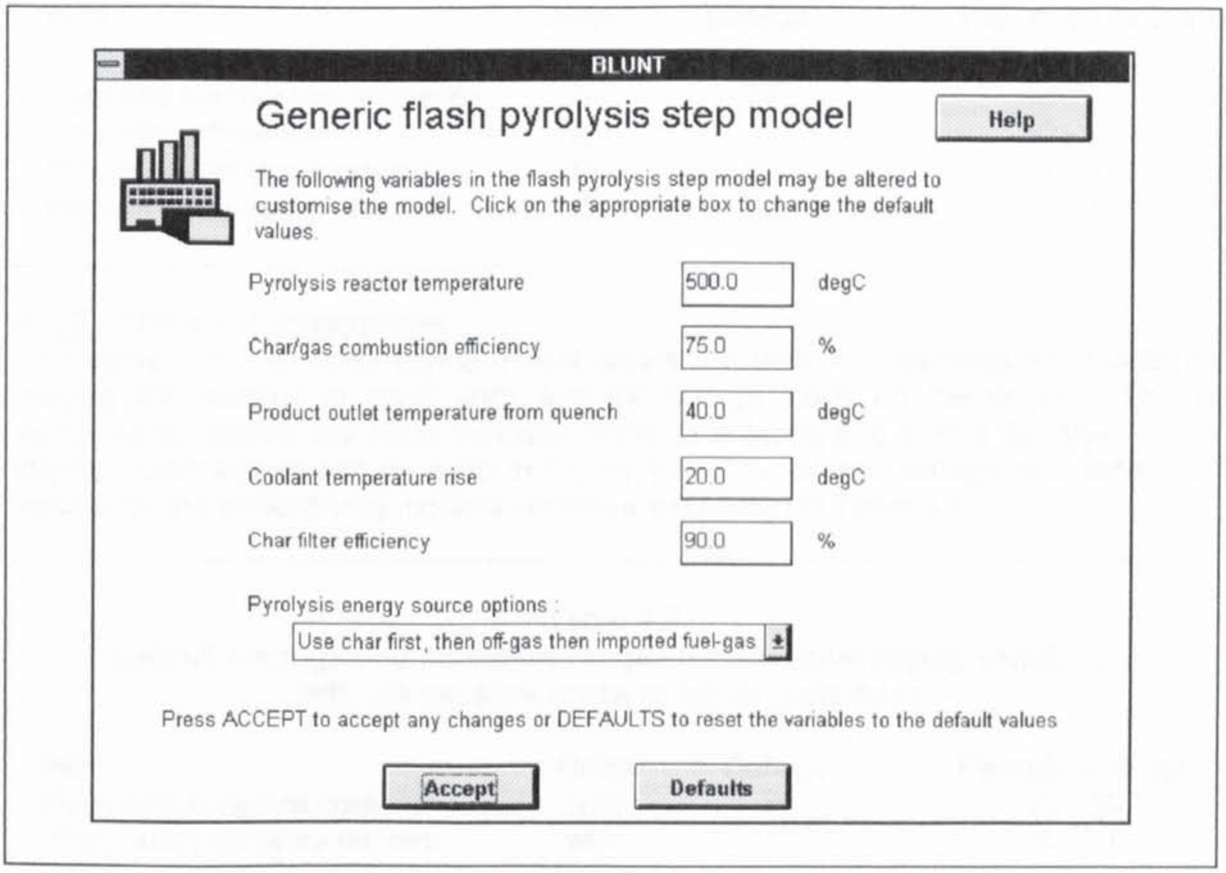


Customisation selection

Select the step model or financial parameter that you wish to customise by clicking on the appropriate button.

Press RETURN to return to data input screen.

Figure 4.3. The step models and financial parameters selection screen.



Generic flash pyrolysis step model

The following variables in the flash pyrolysis step model may be altered to customise the model. Click on the appropriate box to change the default values.

Pyrolysis reactor temperature 500.0 degC

Char/gas combustion efficiency 75.0 %

Product outlet temperature from quench 40.0 degC

Coolant temperature rise 20.0 degC

Char filter efficiency 90.0 %

Pyrolysis energy source options : Use char first, then off-gas then imported fuel-gas

Press ACCEPT to accept any changes or DEFAULTS to reset the variables to the default values

Figure 4.4. The screen for the generic flash pyrolysis model.

The available energy is affected by changes in pyrolysis temperature, the char filter efficiency and the char/off-gas combustion efficiency. The char filter efficiency affects the amount of char remaining in the liquid product and the amount of char recovered and hence available for combustion.

The pyrolysis energy source option determines the order in which the char, off-gas and imported fuel gas are combusted. The available options are:-

- char followed by imported fuel gas to make up the deficit, all the off-gas and any surplus char are exported as by-products;
- char followed by the off-gas, any surplus char or off-gas is exported as by-product;
- off-gas followed by imported fuel gas to make up the deficit, all char is exported as a by-product;
- off-gas followed by char, any surplus char is exported as by-product; and
- imported fuel gas only, all char and off-gas are exported as by-products.

The quench system utilised to condense and recover the product oil is customised by altering the product outlet temperature and the coolant temperature rise. Table 4.1. gives the default settings and the range of values permitted for the flash pyrolysis variables.

Further details on the generic flash pyrolysis model can be found in the Appendix.

Table 4.1.
Default settings and permitted ranges for the flash pyrolysis variables.

Item	Units	Default	Permitted range
Pyrolysis temperature	°C	500	400 - 650
Char/gas combustion efficiency	%	75	40 - 100
Char filter efficiency	%	90	50 - 100
Product outlet temperature	°C	40	20 - 70
Coolant temperature rise	°C	20	10 - 50

4.3.2. The wood drying model.

The screen for the wood drying model allows the user to customise the model by varying the number of dryer units and the energy loads on the dryers. This is achieved by altering the wood moisture contents entering and exiting the dryer . The drying model screen can be seen in Figure 4.5. The default settings and ranges of values for the wood drying model's variables are shown in Table 4.2.

Table 4.2.
Default settings and permitted ranges for the wood drying variables.
(NB. All moisture contents are on a dry basis)

Item	Units	Default	Permitted range
Feed inlet moisture content	wt%	50	25 - 50
Feed outlet moisture content	wt%	7	5 - 30

4.3.3. The hydrotreating model.

The screen for the hydrotreating model allows the user to change the way in which the model reacts by altering the deoxygenation extent, gas combustion efficiency, cooling medium temperature rise, heat exchanger efficiencies, product outlet temperature and the hydrogen source/off-gas disposal option. The hydrotreating model screen is shown in Figure 4.6.

The hydrotreating deoxygenation extent is the most important variable as is affects the hydrogen requirement for the process and hence the operating cost, the liquid hourly space velocity and hence the reactor size and capital cost and the oxygen content of the product and hence quantities of all products produced and quality of the final product. If the deoxygenation extent is below 98% the product can no longer be refined to produce diesel. The energy balance and utilities requirements for the hydrotreating model are affected by the gas combustion efficiency, the cooling medium temperature rise in the heat exchangers, the percentage heat loss in the heat exchangers and the product outlet temperature. Gas is used to preheat the crude pyrolysis oil feed for the first stage and to preheat the stabilised oil feeding into the second stage of the hydrotreating process. The default settings and permitted ranges for the hydrotreating model's variables are given in Table 4.3.

Table 4.3.
Default settings and permitted ranges for the hydrotreating variables.

Item	Units	Default	Permitted range
Deoxygenation extent	%	98	0 - 100
Gas combustion efficiency	%	75	40 - 100
Heat loss in the heat exchangers	%	90	0 - 100
Coolant temperature rise	°C	20	10 - 50
Product outlet temperature	°C	73	20 - 80

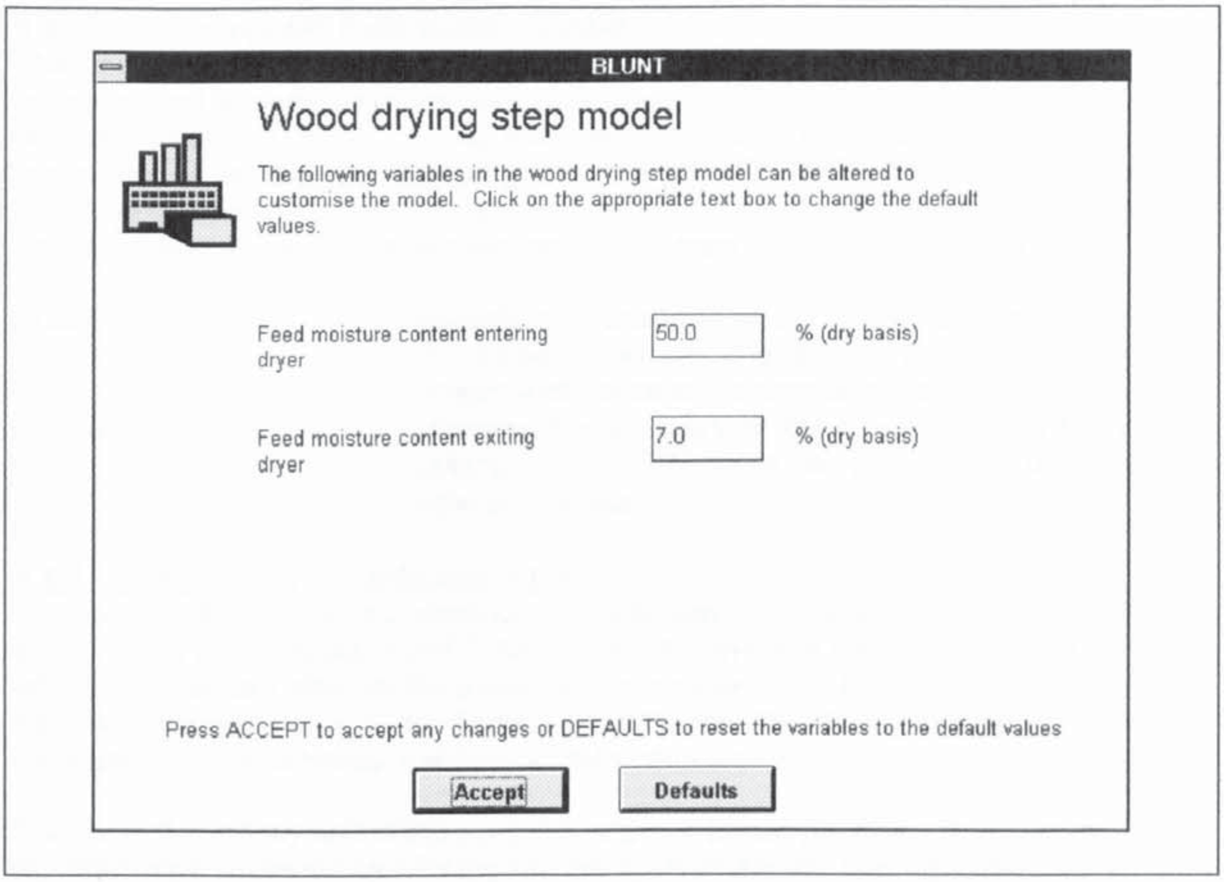
The off-gas disposal/hydrogen source options available are given in a combo box opened by clicking on the arrow on the right hand side of the box. Only the items given in the list can be used by the program. The hydrogen source/off-gas disposal options are:-

- sell all the off-gas as fuel gas to the utilities, importing all of the required hydrogen; and
- separate and recycle the surplus hydrogen, import the excess from the utilities and sell the remaining off-gas as fuel gas.

Further details on the hydrotreating model can be found in the Appendix.

4.3.4. The power production model.

The screen for the power production model allows the user to select the technology used to generate electricity from the crude or upgraded liquid product. The user is given the choice of using an engine or turbine. However, if the power output from the liquid product is less than 5 MWe the default technology is an engine and above 20MWe the default technology is a turbine. Figure 4.7. shows the screen for the power production model.



Wood drying step model

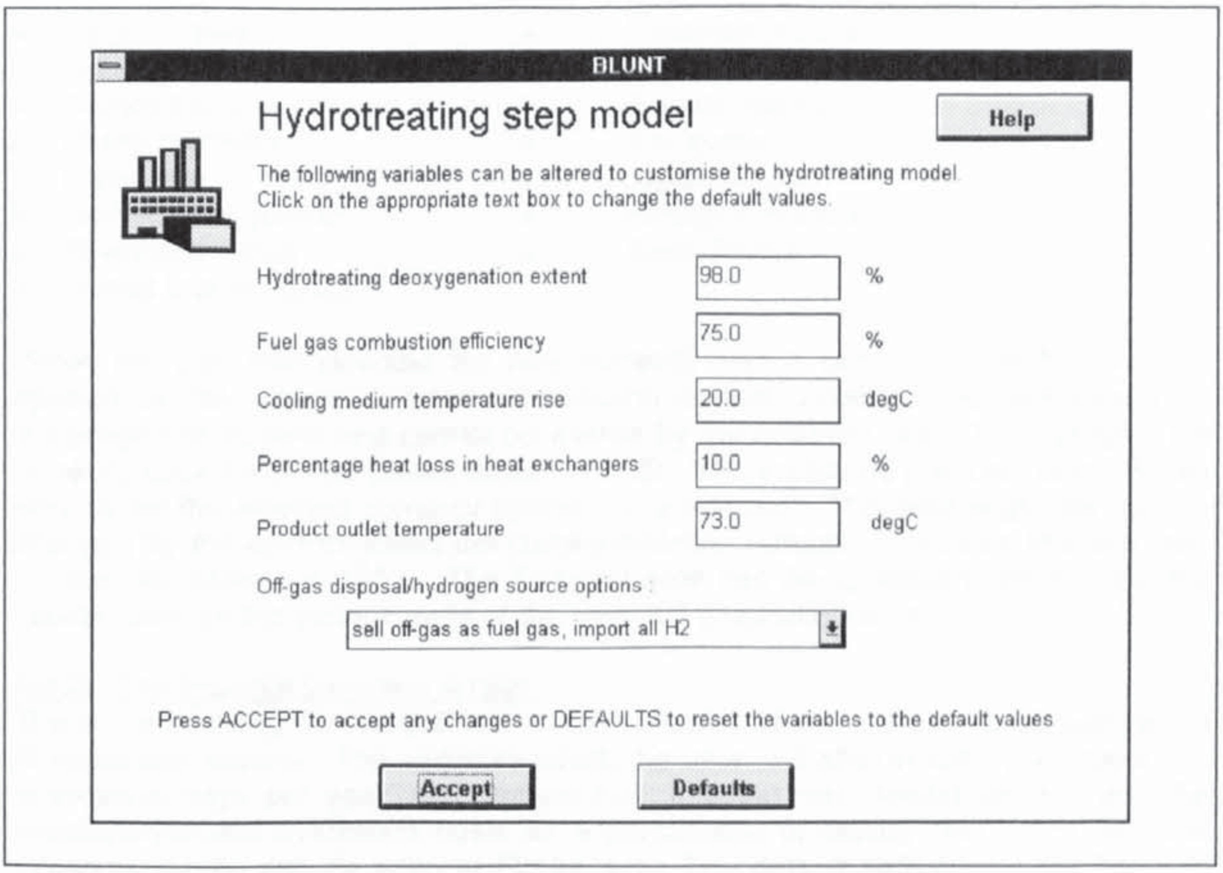
The following variables in the wood drying step model can be altered to customise the model. Click on the appropriate text box to change the default values.

Feed moisture content entering dryer 50.0 % (dry basis)

Feed moisture content exiting dryer 7.0 % (dry basis)

Press ACCEPT to accept any changes or DEFAULTS to reset the variables to the default values

Figure 4.5. The screen for the wood drying model.



Hydrotreating step model

The following variables can be altered to customise the hydrotreating model. Click on the appropriate text box to change the default values.

Hydrotreating deoxygenation extent 98.0 %

Fuel gas combustion efficiency 75.0 %

Cooling medium temperature rise 20.0 degC

Percentage heat loss in heat exchangers 10.0 %

Product outlet temperature 73.0 degC

Off-gas disposal/hydrogen source options :

sell off-gas as fuel gas, import all H2

Press ACCEPT to accept any changes or DEFAULTS to reset the variables to the default values

Figure 4.6. The screen for the hydrotreating model.

4.4. Financial base case customisation.

The financial base case for the program i.e. the currency unit in use, the financial variables and the utilities costs can be altered by the program user to customise the financial aspect of BLUNT. The financial base case screens are accessed in the same way as for the step models by pressing *CUSTOMISE* on the PPI screen and selecting the parameter to be customised. Each of the financial base case screens accessed from the selection screen has two buttons - *DEFAULTS* and *ACCEPT*.

ACCEPT	accepts any changes made to the default variables on the screen, switches the display back to the selection screen and saves any changes made.
DEFAULTS	changes the variables on the screen back to the default settings. *ACCEPT* must be pressed to return to the selection screen.

4.4.1. The currency unit selection screen.

The screen dealing with the currency unit selection allows the program user to change the currency unit in which BLUNT carries out the financial calculations. The variables which the user can alter on the screen are the currency unit, the exchange rate of the selected currency with Great Britain pounds and the financial year for the cost calculations. The currency selection screen can be seen in Figure 4.8.

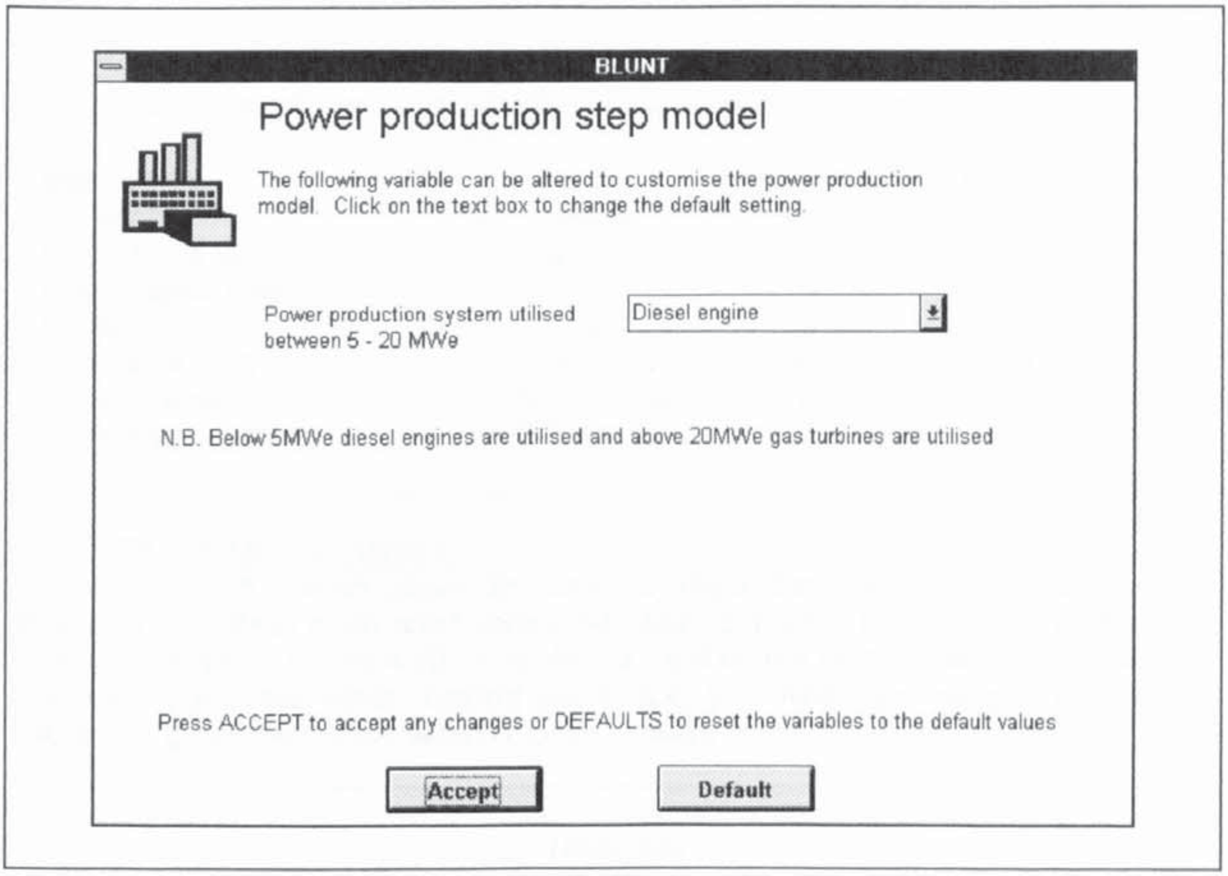
The default currency unit in the program is Great Britain pounds. The user can select an alternative currency by clicking on the appropriate list box and selecting an item from the list that is revealed. The user cannot add any other items to the list as only the currencies included in the list are recognised by the program. The currencies that the program recognises are:

- Belgium francs
- Danish kroner
- French francs
- Greek drachma
- Italian lira
- Netherlands guilder
- Spanish pesetas
- United States dollars.
- Canadian dollars
- European currency units
- German marks
- Irish punts
- Japanese yen
- Portuguese escudos
- Swiss francs

When the user has selected the new currency unit a currency code for that unit appears on the screen. It is the abbreviation for that currency which will appear on the program's screens and cannot be altered by the program user. For example, the currency code for United States dollars is USD. The exchange rate with Great Britain pounds for the selected currency appears in a text box. This exchange rate can be changed by the user to reflect the current financial climate. The base financial year for the calculations is 1993. The financial year can be updated to recalculate the capital costs for the various parts of the selected processing route.

4.4.2. The financial variables screen.

The screen dealing with BLUNT's financial variables allows the user to customise the financial calculations. The variables which the user can alter include the project life, operational days per year, real interest rate, inflation rate, feedstock cost and the maintenance and overheads costs as a percentage of capital cost. The financial variables screen can be seen in Figure 4.9. The default settings for the financial variables are shown in Table 4.4.



Power production step model

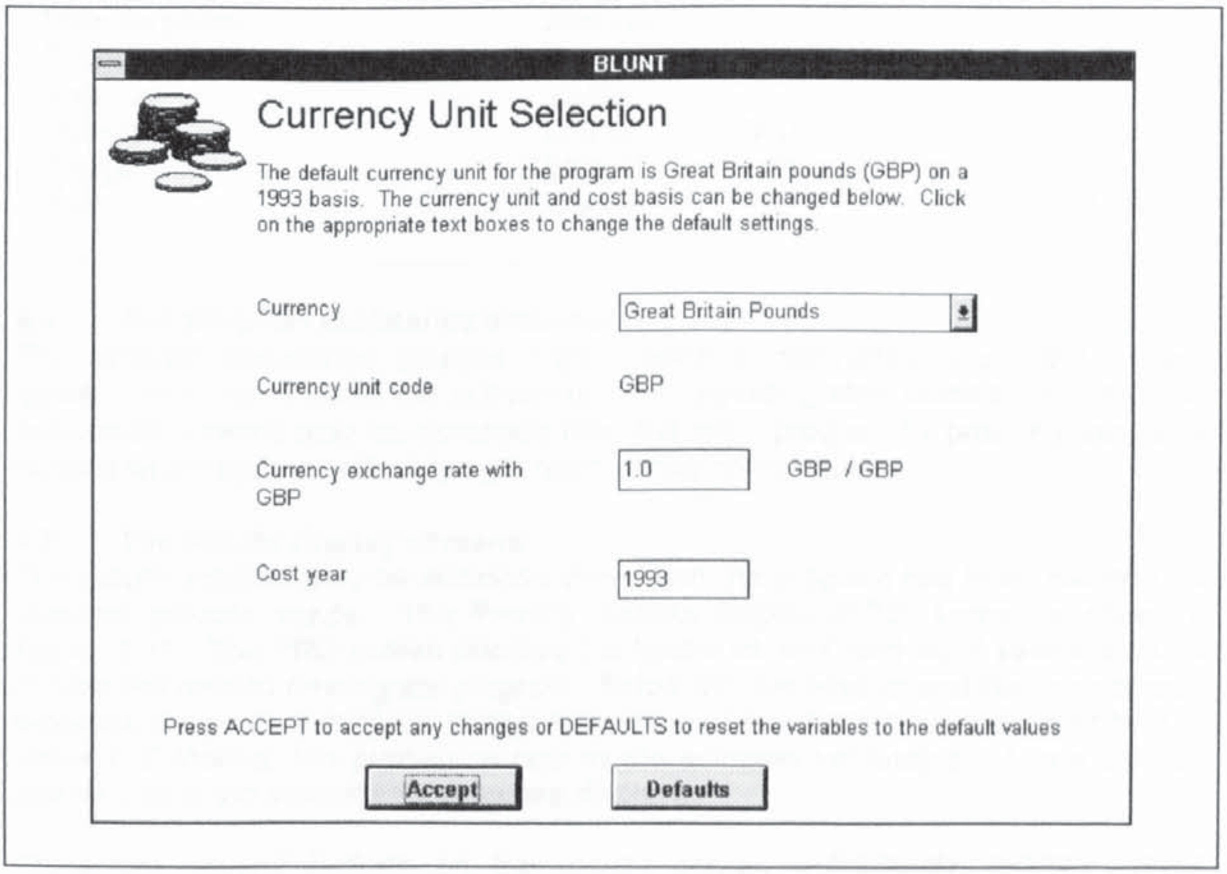
The following variable can be altered to customise the power production model. Click on the text box to change the default setting.

Power production system utilised between 5 - 20 MWe: Diesel engine

N.B. Below 5MWe diesel engines are utilised and above 20MWe gas turbines are utilised

Press ACCEPT to accept any changes or DEFAULTS to reset the variables to the default values

Figure 4.7. The screen for the power production model.



Currency Unit Selection

The default currency unit for the program is Great Britain pounds (GBP) on a 1993 basis. The currency unit and cost basis can be changed below. Click on the appropriate text boxes to change the default settings.

Currency: Great Britain Pounds

Currency unit code: GBP

Currency exchange rate with GBP: 1.0 GBP / GBP

Cost year: 1993

Press ACCEPT to accept any changes or DEFAULTS to reset the variables to the default values

Figure 4.8. The screen for currency unit selection.

Table 4.4.
Default settings and permitted ranges for the financial variables

Item	Units	Default	Permitted range
Project life	years	10	1 - 50
Operational days	days/year	330	1 - 365
Real interest rate	%	5	1 - 100
Inflation rate	%	5	1 - 100
Feedstock cost	£/d.a.f. t	30	0 - 100
Maintenance cost	% Cap Cost	5	1 - 100
Overheads cost	% Cap Cost	7	1 - 100

4.4.3. The utilities cost screen.

The utilities cost screen allows the user to affect the viability of the process by changing the utilities costs and hence production cost of the product. The utilities cost screen is shown in Figure 4.10. The utilities used by the program are low pressure (5 bar) steam, process water, cooling water, fuel gas, hydrogen, oxygen and power. Table 4.5. gives the default settings for the utilities costs.

Table 4.5.
Default settings and permitted ranges for the utilities costs.

Item	Units	Default	Permitted range
Low pressure steam	£/tonne	5	> 0
Process water	£/tonne	1	> 0
Cooling water	£/tonne	0.04	> 0
Fuel gas	£/GJ	3	> 0
Hydrogen	£/tonne	360	> 0
Oxygen	£/tonne	40	> 0
Power	£/MWh	30	> 0

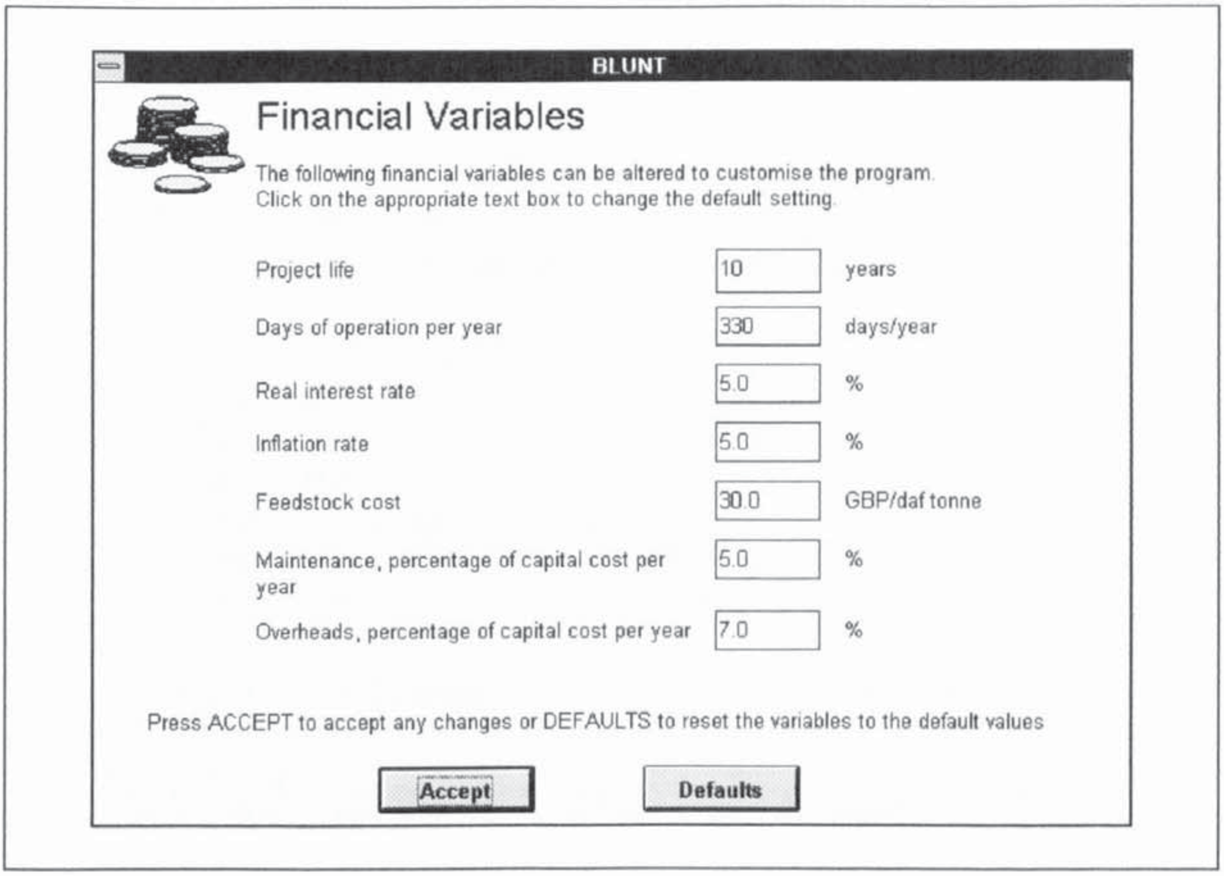
4.4. The program assistance screens.

The program assistance screens display general information about BLUNT and detailed information about the conversion and upgrading step models. The program assistance screens may be accessed from the main program by pressing the *HELP* buttons which appear at the top right hand corner of the screens.

4.5. The results display screens.

The results screens may be accessed only when the program has been run with the selected process inputs. The Primary Results Display (PRD) screen is shown in Figure 4.11. The PRD screen displays the feedstock and feed input selected on the PPI screen prior to running the program. Below this the product and conversion route selected, the product output in tonnes/day, product moisture content, product heating value in GJ/tonne, the production cost in the selected currency per tonne and the capital cost in the selected currency are displayed.

There are several buttons on the results screen - *RETURN*, *MORE*, *PRINT*, *MACROECON*, *SAVE* and *QUIT*. The function of these buttons is given :



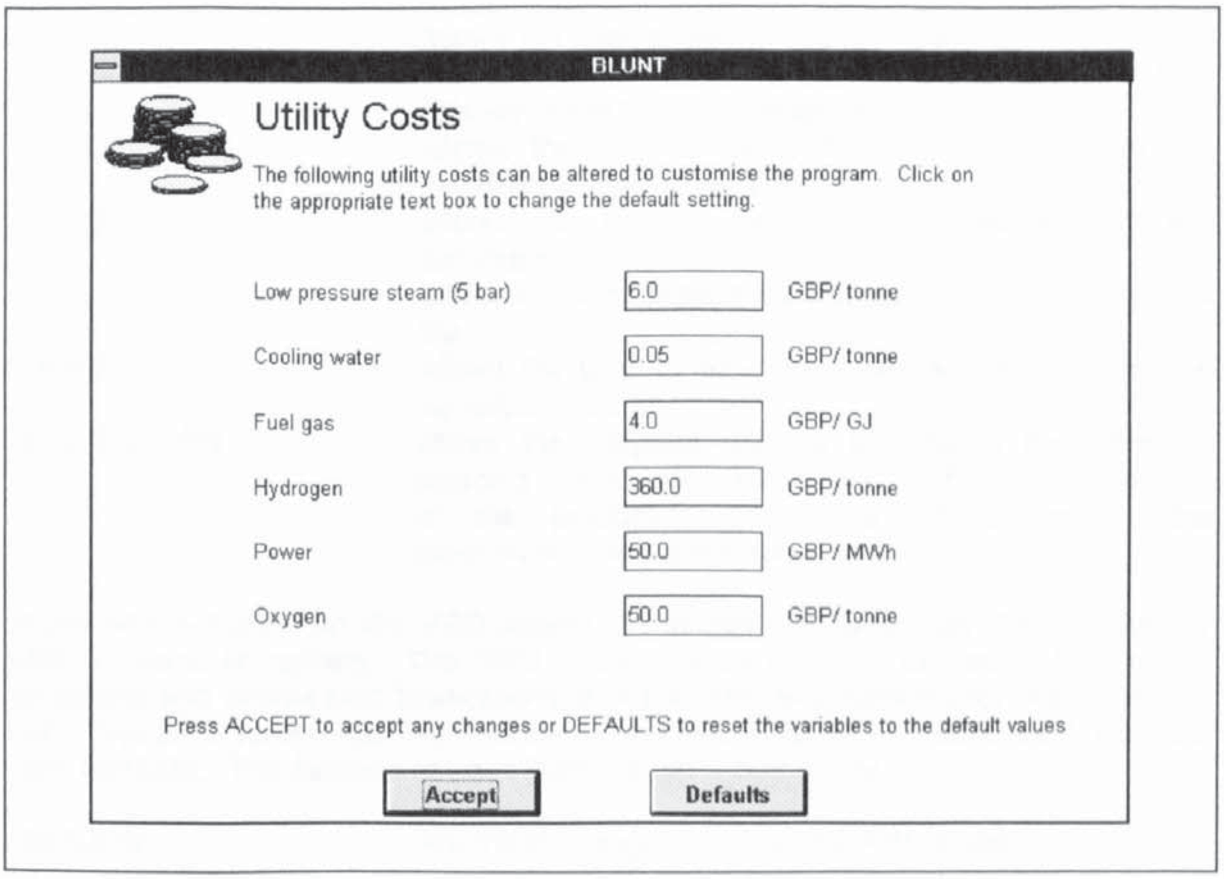
Financial Variables

The following financial variables can be altered to customise the program. Click on the appropriate text box to change the default setting.

Project life: 10 years
Days of operation per year: 330 days/year
Real interest rate: 5.0 %
Inflation rate: 5.0 %
Feedstock cost: 30.0 GBP/daf tonne
Maintenance, percentage of capital cost per year: 5.0 %
Overheads, percentage of capital cost per year: 7.0 %

Press ACCEPT to accept any changes or DEFAULTS to reset the variables to the default values

Figure 4.9. The screen for customising the financial variables.



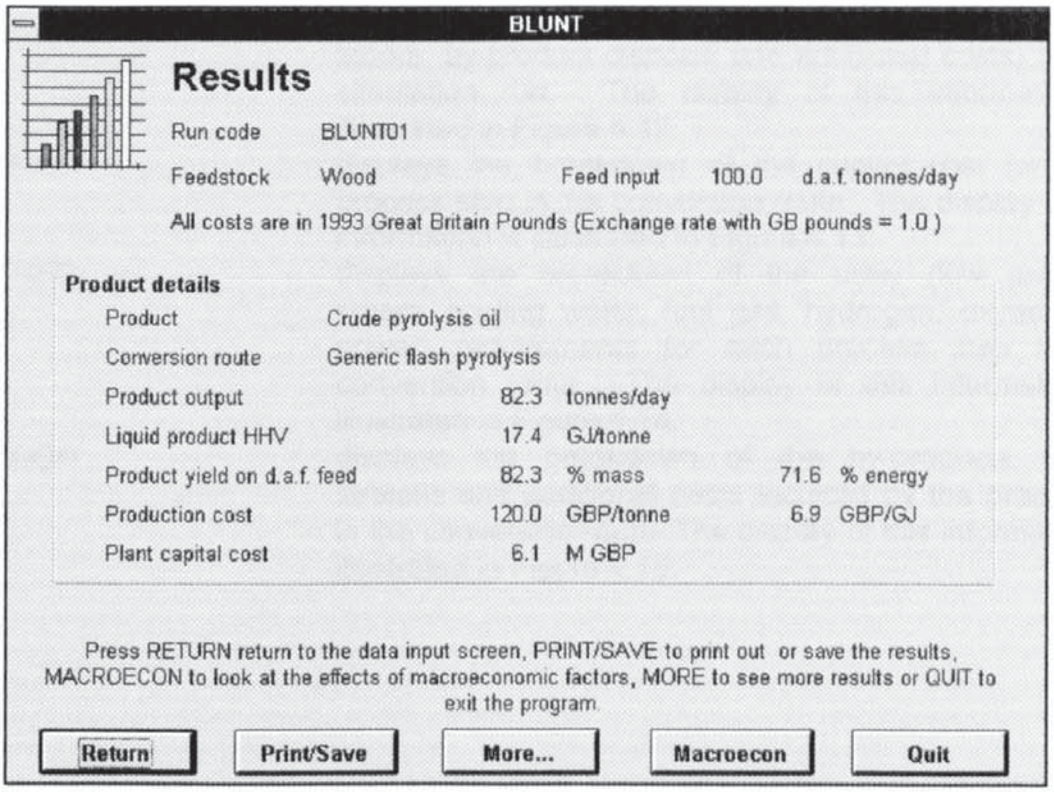
Utility Costs

The following utility costs can be altered to customise the program. Click on the appropriate text box to change the default setting.

Low pressure steam (5 bar): 6.0 GBP/ tonne
Cooling water: 0.05 GBP/ tonne
Fuel gas: 4.0 GBP/ GJ
Hydrogen: 360.0 GBP/ tonne
Power: 50.0 GBP/ MWh
Oxygen: 50.0 GBP/ tonne

Press ACCEPT to accept any changes or DEFAULTS to reset the variables to the default values

Figure 4.10. The screen for customising the utilities costs.



Results

Run code BLUNT01

Feedstock Wood Feed input 100.0 d.a.f. tonnes/day

All costs are in 1993 Great Britain Pounds (Exchange rate with GB pounds = 1.0)

Product details

Product	Crude pyrolysis oil			
Conversion route	Generic flash pyrolysis			
Product output	82.3	tonnes/day		
Liquid product HHV	17.4	GJ/tonne		
Product yield on d.a.f. feed	82.3	% mass	71.6	% energy
Production cost	120.0	GBP/tonne	6.9	GBP/GJ
Plant capital cost	6.1	M GBP		

Press RETURN return to the data input screen, PRINT/SAVE to print out or save the results, MACROECON to look at the effects of macroeconomic factors, MORE to see more results or QUIT to exit the program.

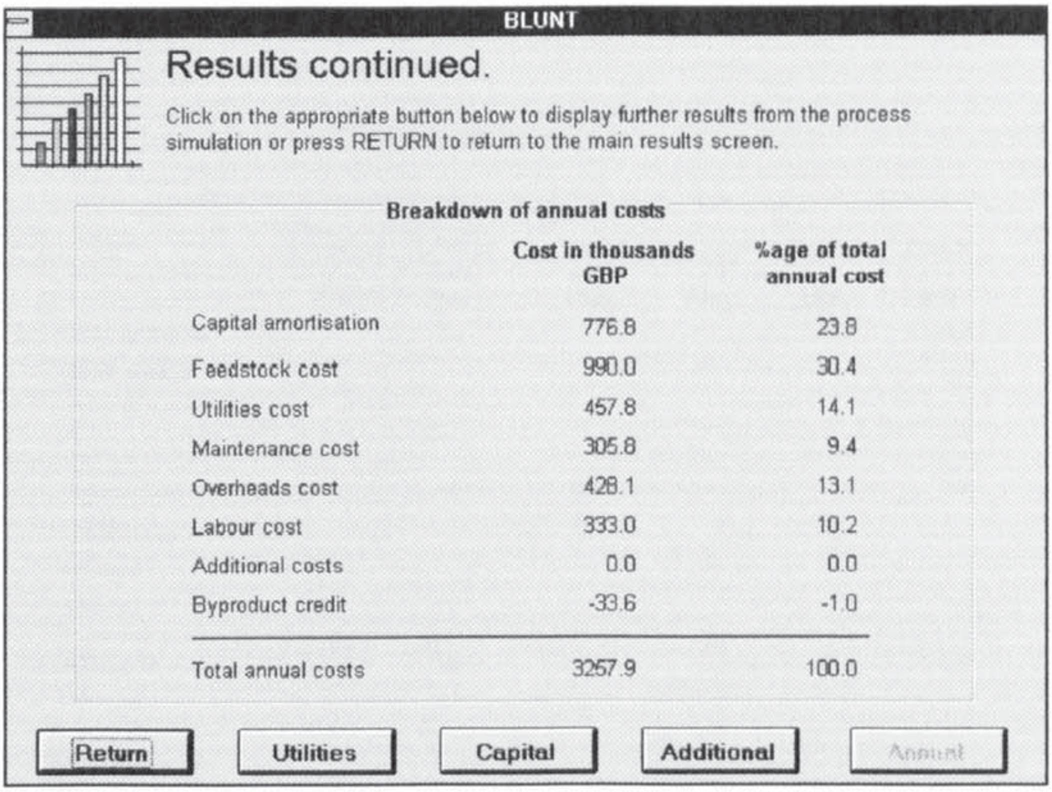
Figure 4.11. The Primary Results Display Screen.

RETURN	allows the user to return to the PPI screen to check the process inputs or to change the variables in the step models or the financial parameters.
QUIT	allows the user to close BLUNT and return to the Windows Program Manager.
PRINT	allows the user to print out the results from the simulation run.
SAVE	allows the user to save the simulation results to a results file.
MORE	allows the user to view more detailed results from the simulation run.
MACROECON	allows the program user to investigate the effect of selected macroeconomic factors on the production cost of the product. For more information on the macroeconomic factors see Section 6.

If the *MORE* button on the PRD screen is pressed the Secondary Results Display (SRD) screen is opened. The SRD screen allows the user to display the capital, additional and annual cost breakdowns and the utility requirements for the simulation run. The SRD screen has five buttons - *CAPITAL*, *ADDITION*, *RETURN*, *UTILITIES* and *ANNUAL*. The function of each button is described below.

RETURN	returns the program user to the PRD screen.

ANNUAL	displays the breakdown of the annual costs (feedstock, capital amortisation, overheads, utilities, maintenance, labour, by-product disposal and additional costs) for the simulation run. The display of this information is illustrated in Figure 4.12.
CAPITAL	displays the breakdown of the capital cost for each process step in the conversion route. The display of this information is illustrated in Figure 4.13.
UTILITIES	displays the breakdown of the utility (low pressure steam, cooling water, fuel gas, hydrogen, oxygen and power) requirements for each process step in the conversion route. The display of this information is illustrated in Figure 4.14.
ADDITION	displays the breakdown of the by-products, waste streams and additional costs incurred by the processes in the conversion route. The display of this information is illustrated in Figure 4.15.



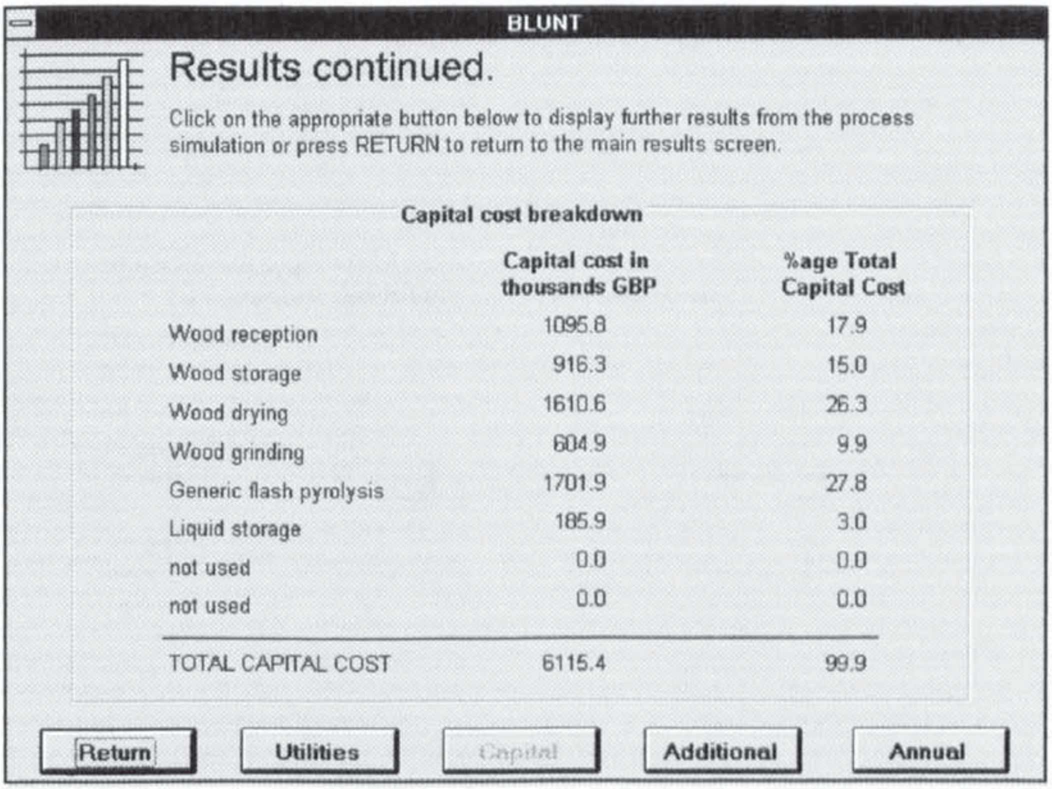
Results continued.

Click on the appropriate button below to display further results from the process simulation or press RETURN to return to the main results screen.

Breakdown of annual costs

	Cost in thousands GBP	%age of total annual cost
Capital amortisation	776.8	23.8
Feedstock cost	990.0	30.4
Utilities cost	457.8	14.1
Maintenance cost	305.8	9.4
Overheads cost	428.1	13.1
Labour cost	333.0	10.2
Additional costs	0.0	0.0
Byproduct credit	-33.6	-1.0
Total annual costs	3257.9	100.0

Figure 4.12. The SRD screen displaying the breakdown of annual costs.



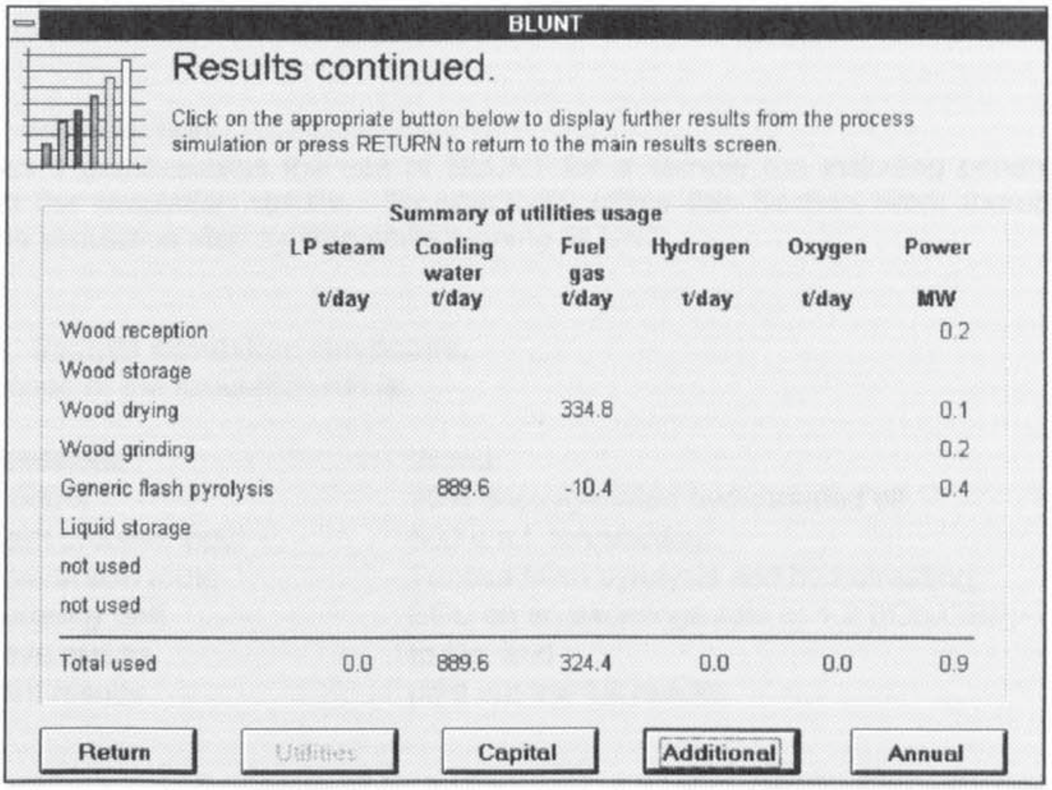
Results continued.

Click on the appropriate button below to display further results from the process simulation or press RETURN to return to the main results screen.

Capital cost breakdown

	Capital cost in thousands GBP	%age Total Capital Cost
Wood reception	1095.8	17.9
Wood storage	916.3	15.0
Wood drying	1610.6	26.3
Wood grinding	604.9	9.9
Generic flash pyrolysis	1701.9	27.8
Liquid storage	185.9	3.0
not used	0.0	0.0
not used	0.0	0.0
TOTAL CAPITAL COST	6115.4	99.9

Figure 4.13. The SRD screen displaying the breakdown of capital costs.



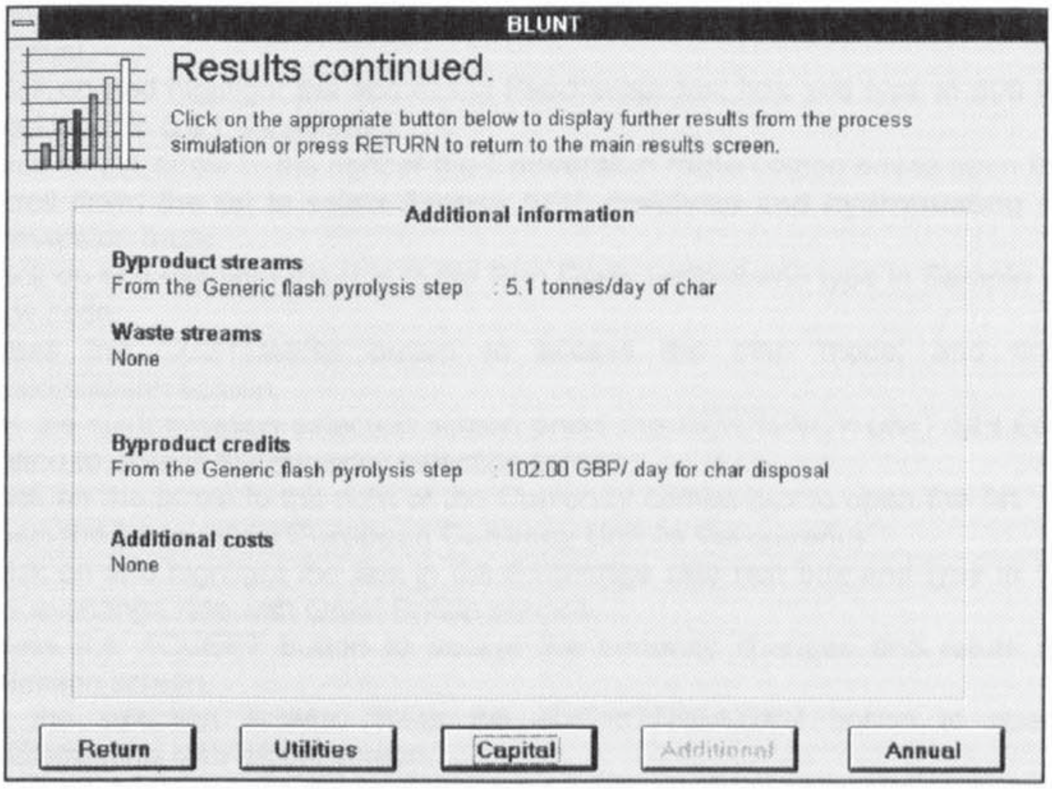
Results continued.

Click on the appropriate button below to display further results from the process simulation or press RETURN to return to the main results screen.

Summary of utilities usage

	LP steam t/day	Cooling water t/day	Fuel gas t/day	Hydrogen t/day	Oxygen t/day	Power MW
Wood reception						0.2
Wood storage						
Wood drying			334.8			0.1
Wood grinding						0.2
Generic flash pyrolysis		889.6	-10.4			0.4
Liquid storage						
not used						
not used						
Total used	0.0	889.6	324.4	0.0	0.0	0.9

Figure 4.14. The SRD screen displaying the breakdown of utilities requirements.



Results continued.

Click on the appropriate button below to display further results from the process simulation or press RETURN to return to the main results screen.

Additional information

Byproduct streams
From the Generic flash pyrolysis step : 5.1 tonnes/day of char

Waste streams
None

Byproduct credits
From the Generic flash pyrolysis step : 102.00 GBP/ day for char disposal

Additional costs
None

Figure 4.15. The SRD screen displaying the breakdown of additional costs.

SECTION 5.
A SAMPLE SIMULATION RUN.

5.1. Introduction.

Section 5 demonstrates the use of BLUNT for a sample run including printing and saving the simulation results. To effectively utilise this Section, work through the sample simulation step by step while running BLUNT.

5.2. Sample simulation run scope.

The scope of the simulation run is:

• Feedstock	Wood;
• Product	90% deoxygenated hydrotreated oil;
• Plant process input	500 d.a.f. tonnes/day;
• Conversion route	Fenosa flash pyrolysis and hydrotreating;
• Currency unit	ECU on an exchange rate of 1.2 ECU/GBP;
• Save results	to file; and
• Print results	print out the full results.

5.3. Running BLUNT for the sample simulation.

Follow the steps below to work through the sample simulation run.

1. Initialise BLUNT by double clicking on the icon and start BLUNT by pressing the *START THE PROGRAM* button.
2. On the PPI screen click on the arrow to the right of the **Product** combo box to open the product list. Scroll down the list to select **Hydrotreated oil** as the product.
3. Click on and highlight the text in the **Feed Input** text box and type in **500** for the feed input in d.a.f. tonnes/day.
4. Click on the arrow to the right of the **Conversion route** combo box to open the list. Scroll down the list to select **Fenosa flash pyrolysis and hydrotreating** as the conversion route.
5. Click on and highlight the text in the **Run Code** textbox and type in **Sample** as the new code.
6. Press the *CUSTOMISE* button to access the step model and financial customisation screen.
7. On the customisation selection screen press the *CURRENCY UNIT SELECTION* button to access the currency selection screen.
8. Click on the arrow to the right of the **Currency** combo box to open the list. Scroll down the list to select **European Currency Unit** as the currency.
9. Click on and highlight the text in the **Exchange rate** text box and type in **1.2** for the exchange rate with Great Britain pounds.
10. Press the *ACCEPT* button to accept the currency changes and return to the selection screen.
11. At the selection screen, press the *HYDROTREATING* button to open the hydrotreating step model screen.
12. Click on and highlight the text in the **Hydrotreating deoxygenation extent** text box and type in **90** for the deoxygenation percentage.

13. Click on the arrow to the right of the **Off-gas disposal/hydrogen source option** combo box to open the list. Scroll down the list to select **Separate and recycle the surplus hydrogen.**
14. Press the *ACCEPT* button to accept the changes to the hydrotreating variables and return to the selection screen.
15. At the selection screen press the *RETURN* button to return to the PPI screen.
16. Having followed steps 1-15 press the *RUN* button to run BLUNT for the selected inputs and open the PRD screen.
17. The results summary from the simulation are displayed as shown in Figure 5.1. Click on the MORE button to look at the utilities, capital costs, annual costs and additional costs information.

5.4. Printing the simulation results.

BLUNT is set up to utilise the default printer as detailed in the Windows Print Manager. Set the printer to print out in Portrait mode on A4 paper - printing in Landscape or in any other paper size will disrupt the way in which the results print out. The Windows Program Manager can be accessed from BLUNT without quitting the program. Clicking on the icon on the top left hand corner of the screen and using SWITCH TO allows access to the Windows Program Manager.

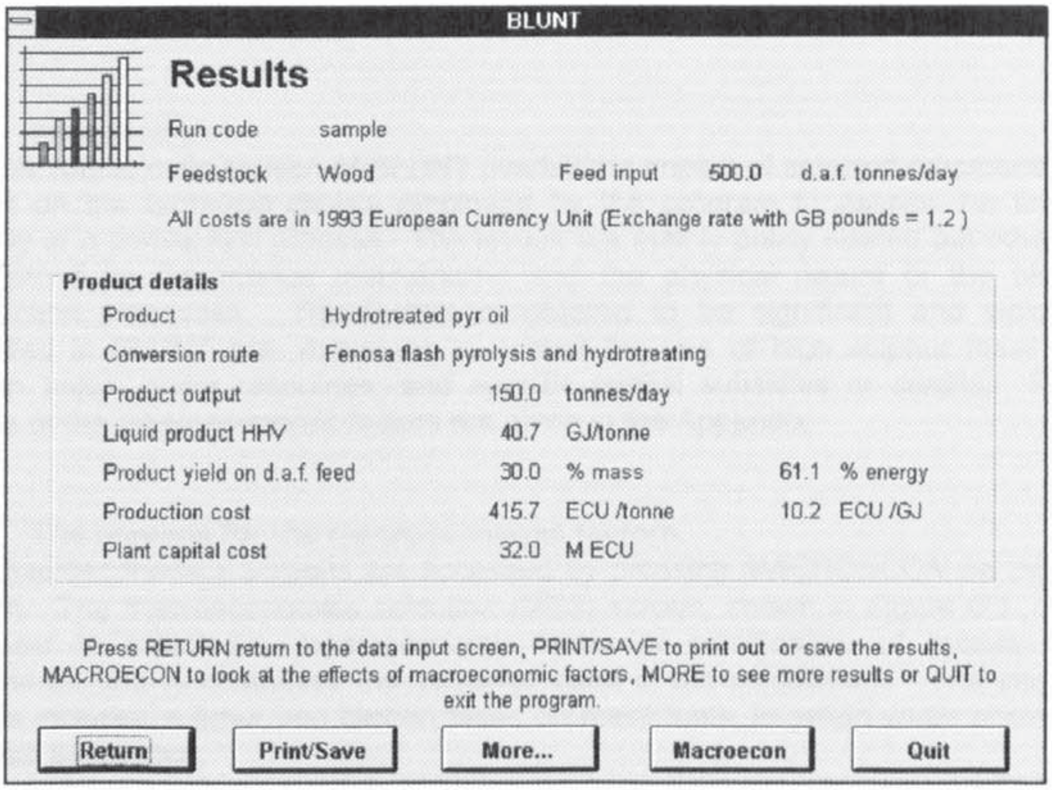
The results from the process simulation run may be printed out from the PRD screen by pressing the *PRINT/SAVE* button. On pressing the button, a dialog box appears asking whether a summary is required or a full results printout. The **Print/Save** dialog box is shown in Figure 5.2. The print dialog box has four button: *RETURN*, *SUMMARY*, *SAVE* and *ALL*.

SUMMARY	prints out a summary sheet containing the results shown on the results screen, the utilities costs and financial variables used for the simulation run and any alterations made to the default settings of the step models.
ALL	prints out the results summary sheet, the utilities usage breakdown, an annual cost breakdown, a breakdown of the capital cost and a summary of the additional costs.
SAVE	saves the full set of simulation results to file.
RETURN	returns the user to the PRD screen.

Pressing the *ALL* button will send the results from the simulation run to the default printer. A message will appear on the screen when the results have been sent to the printer. An example of the printout is given in the Appendix. The **Print/Save** dialog box is closed by pressing the RETURN button which returns the program user to the PRD screen. The mass balance results file **Massbal.res** cannot be printed from BLUNT due to the large amount of data which it contains. To access and printout the mass balance for the simulation run quit BLUNT and use a file management program such as XTreeGold or a word-processing package such as Word for Windows.

5.5. Saving the simulation results.

The results from the simulation run can be saved to a text file by pressing the *SAVE* button on the **Print/Save** dialog box. The results file generated will be saved in the BLUNT directory as **BLUNT.res**. The results file generated will contain the same information as the full results printout generated in Section 5.4.



Results

Run code sample

Feedstock Wood Feed input 500.0 d.a.f. tonnes/day

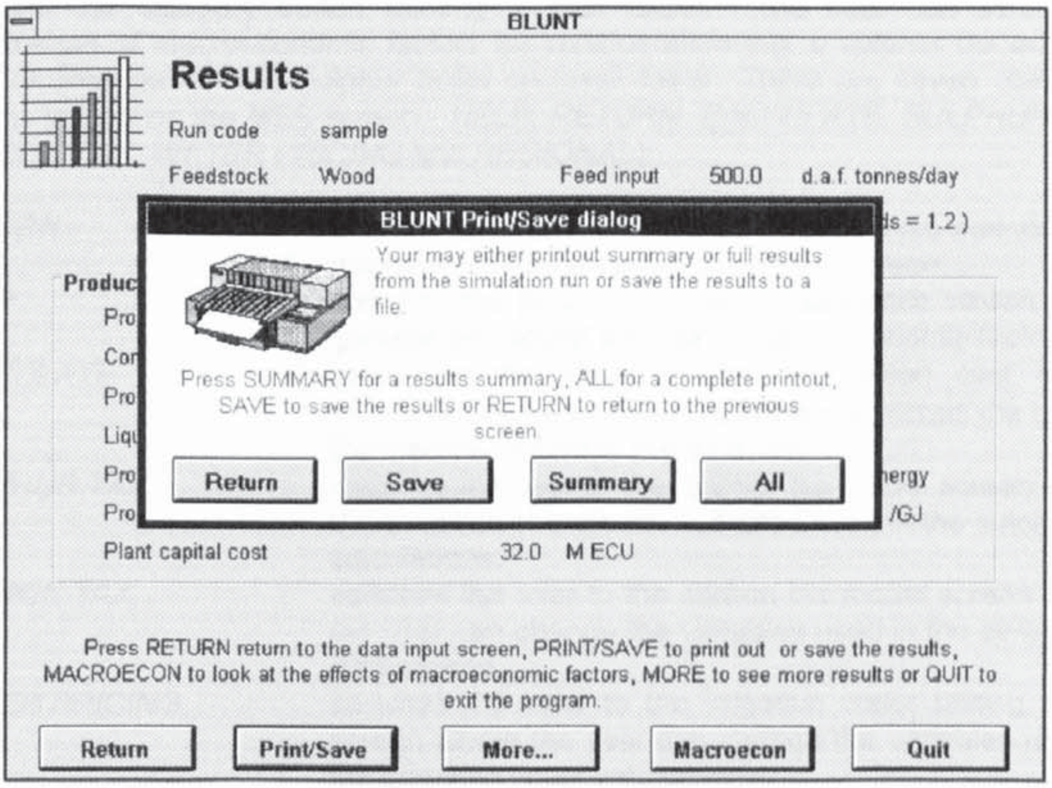
All costs are in 1993 European Currency Unit (Exchange rate with GB pounds = 1.2)

Product details

Product	Hydrotreated pyr oil			
Conversion route	Fenosa flash pyrolysis and hydrotreating			
Product output	150.0	tonnes/day		
Liquid product HHV	40.7	GJ/tonne		
Product yield on d.a.f. feed	30.0	% mass	61.1	% energy
Production cost	415.7	ECU /tonne	10.2	ECU /GJ
Plant capital cost	32.0	M ECU		

Press RETURN return to the data input screen, PRINT/SAVE to print out or save the results, MACROECON to look at the effects of macroeconomic factors, MORE to see more results or QUIT to exit the program.

Figure 5.1. Results summary screen for the sample simulation run.



Results

Run code sample

Feedstock Wood Feed input 500.0 d.a.f. tonnes/day

Your may either printout summary or full results from the simulation run or save the results to a file.

Press SUMMARY for a results summary, ALL for a complete printout, SAVE to save the results or RETURN to return to the previous screen.

Plant capital cost 32.0 M ECU

Press RETURN return to the data input screen, PRINT/SAVE to print out or save the results, MACROECON to look at the effects of macroeconomic factors, MORE to see more results or QUIT to exit the program.

Figure 5.2. The print/save dialog box.

SECTION 6.
ADVANCED PROGRAM OPERATION.

6.1. Introduction.

The macroeconomic section of BLUNT overlay the impact of selected macroeconomic issues on the technical criteria employed by the program to assess the financial viability of a conversion process. The issues are mainly policy related but others are determined by the market mechanisms and the physical nature of the biomass conversion processes. The factors considered to be significant and which are modelled in BLUNT are: measures to control the use of high sulphur fossil fuels; carbon taxes; water resources; and specific capital subsidies or credits. Further details of the macroeconomic factors are given in the Appendix.

6.2. The screens for the macroeconomic factors.

The macroeconomics screens are accessed by pressing *MACROECON* on the PRD screen. The macroeconomics selection (MES) screen, shown in Figure 6.1., allows the user to select the macroeconomic factor or combination of factors to be considered and to customise the variables used in the calculations. The choice of factors includes sulphur and carbon taxes on fossil fuels, irrigation water pricing and regional assistance.

The MES screen consists of four frames each headed by the name of one of the macroeconomic factors. Inside each frame are two radio buttons marked **YES** and **NO** respectively and a category button relating to the macroeconomic factor. Initially the radio buttons are set to **NO** i.e. the relevant macroeconomic factors are not selected. Clicking on a **YES** radio button activates that macroeconomic factor and activates the category button relating to that factor. The user can select any combination of macroeconomic factors for consideration e.g. a sulphur tax on fossil fuels or both carbon and sulphur taxes on fossil fuels. There are seven command buttons in total on the MES screen - *HELP*, *RETURN*, *CALCULATE*, *SULPHUR TAX*, *CARBON TAX*, *WATER PRICING* and *REGIONAL*.

RETURN	unloads the macroeconomics screens from memory and returns the program user to the PRD screen.
HELP	switches the user to a program assistance screen giving general help about the use of macroeconomic factors.
CALCULATE	calculates the effect on the production cost of the selected macroeconomic factors and switches the user to the macroeconomics results screen.
SULPHUR TAX	switches the user to the sulphur tax model screen where the user can change the variables used in the sulphur tax calculations.
CARBON TAX	switches the user to the carbon tax model screen where the user can change the variables used in the carbon tax calculations.
WATER PRICING	switches the user to the irrigation water pricing model screen where the user can change the variables used in the irrigation water calculations.
REGIONAL	switches the user to the regional assistance model screen where the user can change the variables used in the regional assistance calculations.

Each of the four screens on which the variables for the macroeconomic factors can be customised have three buttons - *HELP*, *DEFAULTS* and *ACCEPT*.

HELP	gives further information about the macroeconomic model.
ACCEPT	accepts any changes made to the default variables on the screen, switches the display back to the MES screen and saves any changes made to the macroeconomic model's variables.
DEFAULTS	changes the variables on the screen back to the default settings. *ACCEPT* must be pressed to return to the MES screen.

6.2.1. The sulphur tax model screen.

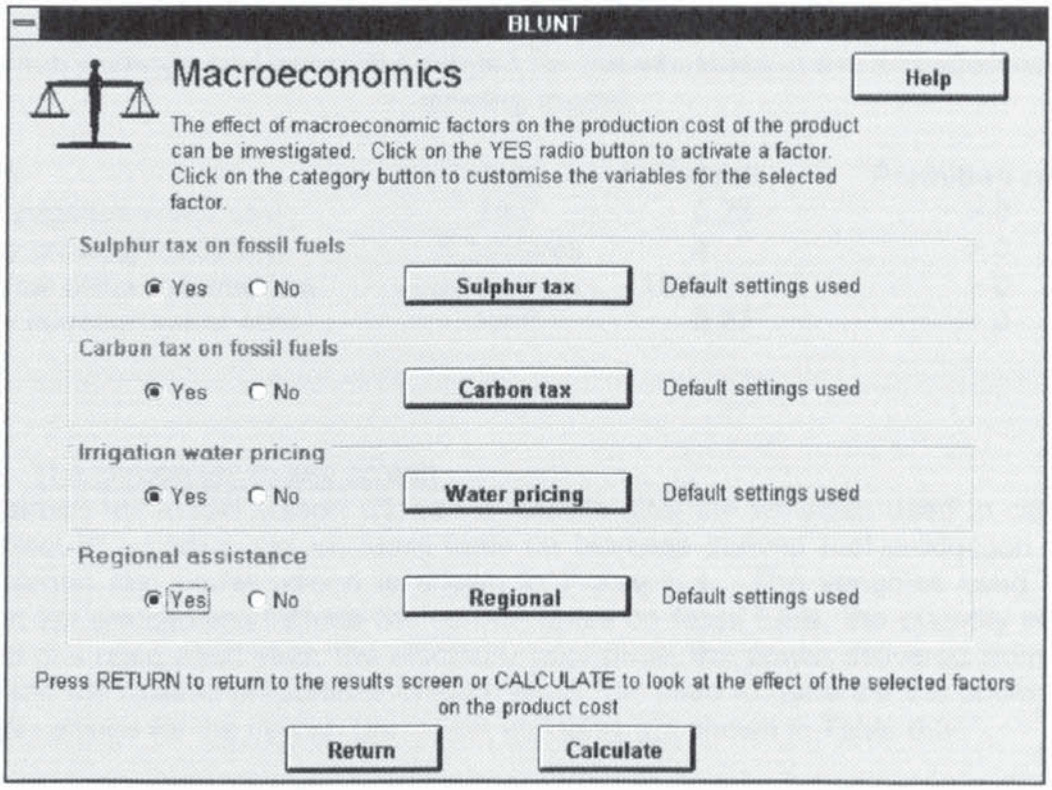
The sulphur tax model screen allows the user to alter the variables used to calculate the effect of a sulphur tax on fossil fuels on the biomass derived fuel production costs. The variables used in the sulphur tax calculations include the sulphur tax on fossil fuels, the quantity of coal, oil and gas used each year, the electricity pool price, the power delivered from fossil fuels and the relative proportions of each fossil fuel used to generate the power. The sulphur tax model screen is shown in Figure 6.2. The default settings for the sulphur tax model variables are shown in Table 6.1.

Table 6.1.
Default settings and permitted ranges for the variables in the sulphur tax model

Item	Units	Default	Permitted range
Sulphur tax on fossil fuels	£/1% sulphur	50	> 0
Coal requirement	M t/year	76.64	> 0
Oil requirement	K barrels/year	364.4	> 0
Gas requirement	M m^3/year	5.53	> 0
Electricity pool price	£/kWh	0.022	> 0
Power delivery	GWh/year	248325	> 0
Proportion coal used	%	61	1 - 100
Proportion oil used	%	19	1 - 100
Proportion gas used	%	5	1 - 100

6.2.3. The water pricing model screen.

The water pricing model looks at the effect of changing the irrigation water cost on the feedstock cost and hence the biomass derived fuel production cost. The irrigation water pricing screen is shown in Figure 6.3. The variables used are the old and new irrigation water costs, the area growing the feedstock and the typical water requirement. The default settings for the variables used in the irrigation water pricing model are shown in Table 6.2.



Macroeconomics

The effect of macroeconomic factors on the production cost of the product can be investigated. Click on the YES radio button to activate a factor. Click on the category button to customise the variables for the selected factor.

Sulphur tax on fossil fuels
● Yes ○ No — Default settings used

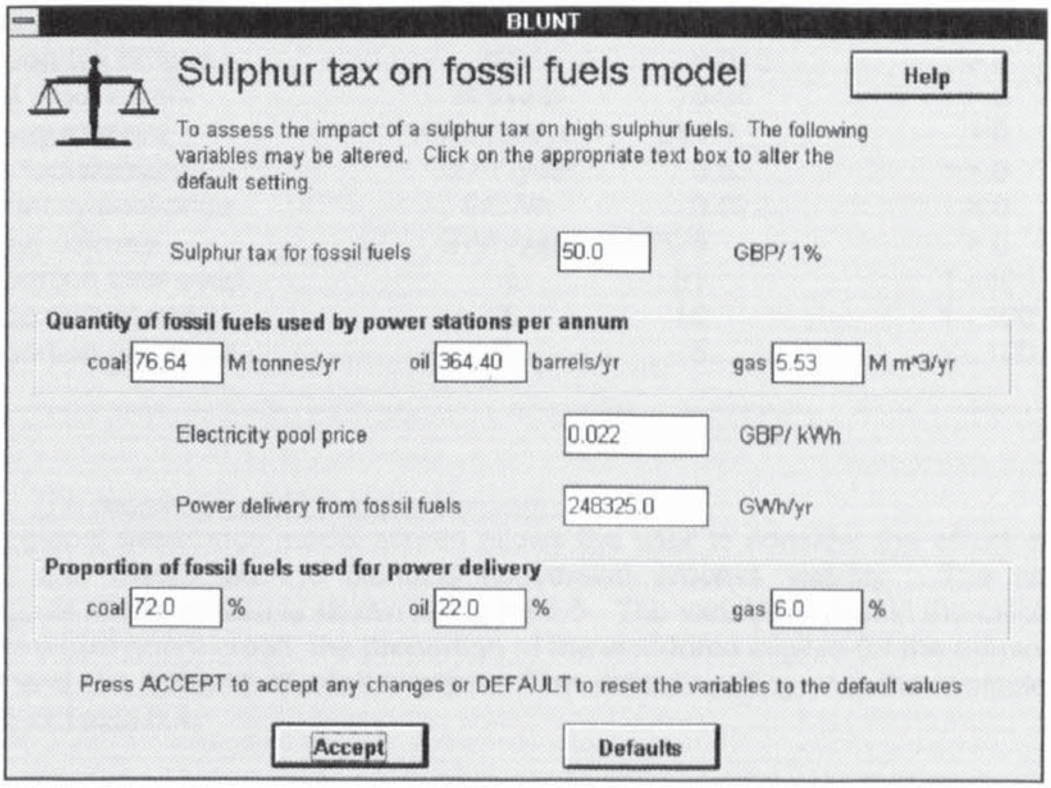
Carbon tax on fossil fuels
● Yes ○ No — Default settings used

Irrigation water pricing
● Yes ○ No — Default settings used

Regional assistance
● Yes ○ No — Default settings used

Press RETURN to return to the results screen or CALCULATE to look at the effect of the selected factors on the product cost

Figure 6.1. The macroeconomic selection screen.



Sulphur tax on fossil fuels model

To assess the impact of a sulphur tax on high sulphur fuels. The following variables may be altered. Click on the appropriate text box to alter the default setting.

Sulphur tax for fossil fuels 50.0 GBP/ 1%

Quantity of fossil fuels used by power stations per annum

coal 76.64 M tonnes/yr
oil 364.40 barrels/yr
gas 5.53 M m^3/yr

Electricity pool price 0.022 GBP/ kWh

Power delivery from fossil fuels 248325.0 GWh/yr

Proportion of fossil fuels used for power delivery

coal 72.0 %
oil 22.0 %
gas 6.0 %

Press ACCEPT to accept any changes or DEFAULT to reset the variables to the default values

Figure 6.2. The sulphur tax model screen.

Table 6.2.
Default settings and permitted ranges for the variables in the irrigation water pricing model.

Item	Units	Default	Permitted range
Old irrigation water cost	£/m^3	0.25	> 0
Area growing feedstock	K hectares	4	> 0
Typical water requirement	m^3/ha/year	1500	> 0
New irrigation water cost	£/m^3	0.27	> 0

6.2.3. The carbon tax model screen

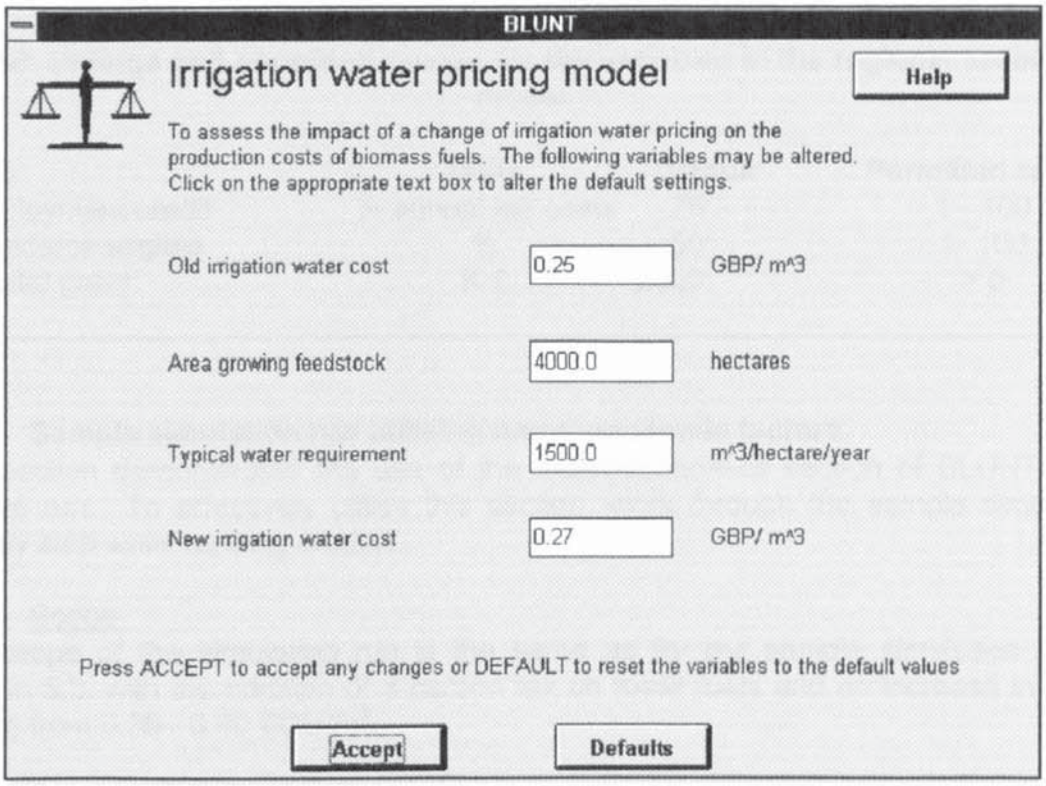
The carbon tax model screen allows the user to alter the variables used to calculate the effect of a carbon tax on fossil fuels on biomass derived fuel production costs. The carbon tax model screen is shown in Figure 6.4. The variables used in the carbon tax calculations include the carbon taxes on fossil fuels, the quantity of coal, oil and gas used each year, the electricity pool price, the power delivered from fossil fuels and the relative proportions of each fossil fuel used to generate the power. The default settings for the carbon tax model variables are shown in Table 6.3.

Table 6.3.
Default settings and permitted ranges for the variables in the carbon tax model.

Item	Units	Default	Permitted range
Carbon tax on coal	£/tonne	1.71	> 0
Carbon tax on oil	£/barrel	20	> 0
Carbon tax on gas	£/m^3	0.015	> 0
Coal requirement	M t/year	76.64	> 0
Oil requirement	K barrels/year	364.4	> 0
Gas requirement	M m^3/year	5.53	> 0
Electricity pool price	£/kWh	0.022	> 0
Power delivery	GWh/year	248325	> 0
Proportion coal used	%	61	1 - 100
Proportion oil used	%	19	1 - 100
Proportion gas used	%	5	1 - 100

6.2.4. The regional assistance model screen.

The regional assistance model screen allows the user to consider the effect of local grants and credits on the biomass conversion process viability. The regional assistance model screen is shown in Figure 6.5. The variables used in the calculation are the employment credit, the proportion of the workforce eligible for the employment credit and the amount of capital grant. The default settings for the variables are shown in Table 6.4.



Irrigation water pricing model

To assess the impact of a change of irrigation water pricing on the production costs of biomass fuels. The following variables may be altered. Click on the appropriate text box to alter the default settings.

Old irrigation water cost: 0.25 GBP/ m^3

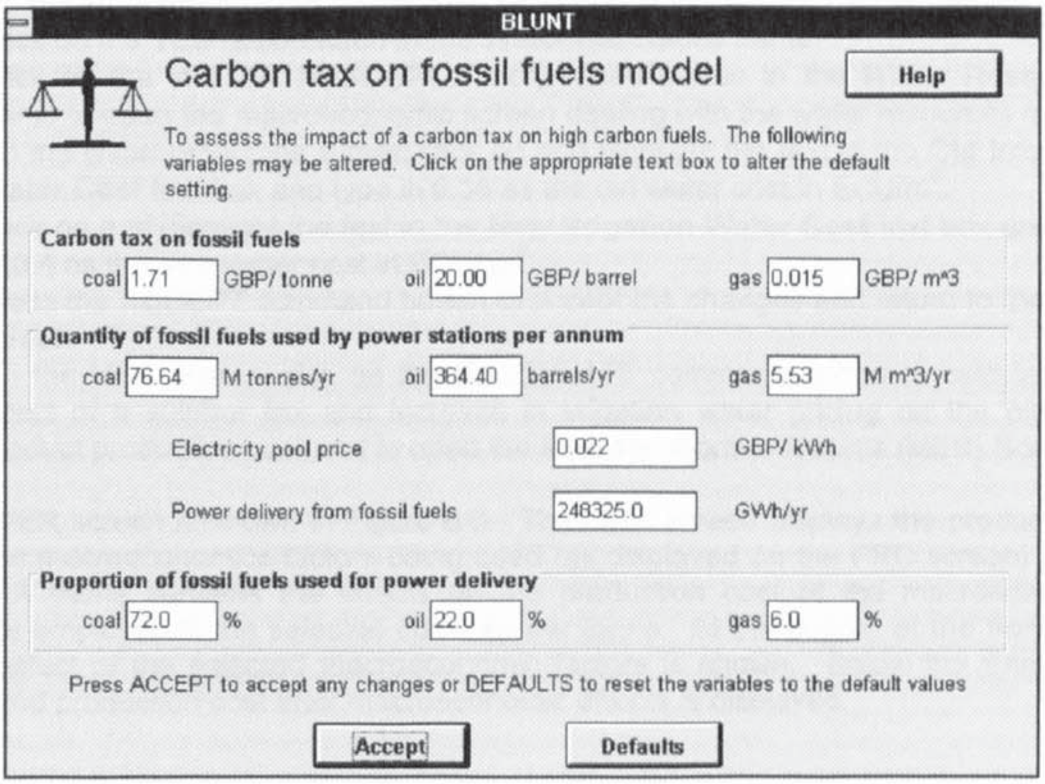
Area growing feedstock: 4000.0 hectares

Typical water requirement: 1500.0 m^3/hectare/year

New irrigation water cost: 0.27 GBP/ m^3

Press ACCEPT to accept any changes or DEFAULT to reset the variables to the default values

Figure 6.3. The water pricing model screen.



Carbon tax on fossil fuels model

To assess the impact of a carbon tax on high carbon fuels. The following variables may be altered. Click on the appropriate text box to alter the default setting.

Carbon tax on fossil fuels

coal 1.71 GBP/ tonne; oil 20.00 GBP/ barrel; gas 0.015 GBP/ m^3

Quantity of fossil fuels used by power stations per annum

coal 76.64 M tonnes/yr; oil 364.40 barrels/yr; gas 5.53 M m^3/yr

Electricity pool price: 0.022 GBP/ kWh

Power delivery from fossil fuels: 248325.0 GWh/yr

Proportion of fossil fuels used for power delivery

coal 72.0 %; oil 22.0 %; gas 6.0 %

Press ACCEPT to accept any changes or DEFAULTS to reset the variables to the default values

Figure 6.4. The carbon tax model screen.

Table 6.4.
Default settings and permitted ranges for the variables in the regional assistance model.

Item	Units	Default	Permitted range
Employment credit	% annual lab costs	25	1 - 100
Workforce eligible	%	50	1- 100
Capital grant	K £	2000	> 0

6.3. Sample simulation run utilising macroeconomic factors.

This section demonstrates the use of the macroeconomics section of BLUNT for a sample run. To effectively utilise this section, work through the sample simulation step by step while running BLUNT.

6.3.1. Scope.

The scope of the simulation run is the same as for the sample simulation run in Section 5.3. with the addition of a carbon tax on fossil fuels and an increase in water pricing from 0.36 - 0.40 ECU/m^3.

6.3.2. Sample macroeconomic simulation.

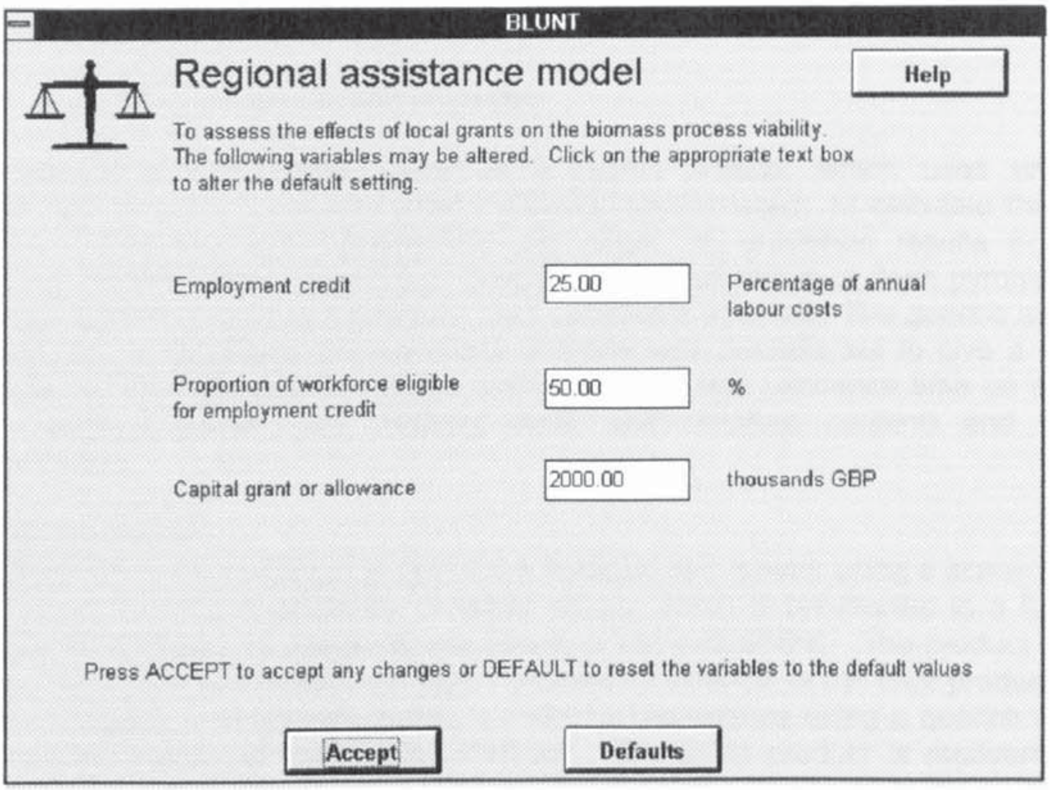
Follow the steps below to work through the sample simulation run.

1. Run BLUNT as demonstrated in the sample simulation in Section 5.3.
2. Having completed the simulation run on the PRD screen press the *MACROECON* button to open the macroeconomics section of BLUNT.
3. On the MES screen click on the **YES** radio button in the **Carbon tax** frame.
4. Click on the **YES** radio button in the Water Resources frame.
5. Click on the *WATER RESOURCES* command button in the **Water Resources** frame to open the macroeconomic screen dealing with the water resources model.
6. On the water resources screen click on and highlight the text in the **Old Irrigation Water Cost** text box and type in **0.36** as the old water cost in ECU/m^3.
7. Click on and highlight the text in the **New Irrigation Water Cost** text box and type in **0.4** as the new water cost in ECU/ m^3.
8. Press the *ACCEPT* command button to accept the changes and return to the MES screen.
9. On the MES screen click on the *CALCULATE* command button to calculate the effect of a sulphur tax and increase in irrigation water pricing on the biomass product production cost and to open the Macroeconomics Results (MER) Screen.

The MER screen is shown in Figure 6.6. The MER screen displays the product cost without macroeconomics factors being used (as displayed on the PRD screen) at the top. A frame contains the effects on the production cost of the macroeconomic factors employed in the selected currency per tonne. At the bottom of the frame the total effect of the selected macroeconomic factors is shown. Below the frame the adjusted production cost after macroeconomic effects is displayed.

There are two buttons on the MER screen - *MACROECON* and *RETURN*.

MACROECON	returns the user to the MES screen where another combination of macroeconomic factors may be selected or the screens for customising the variables can be accessed.
RETURN	returns the user to the PRD screen.



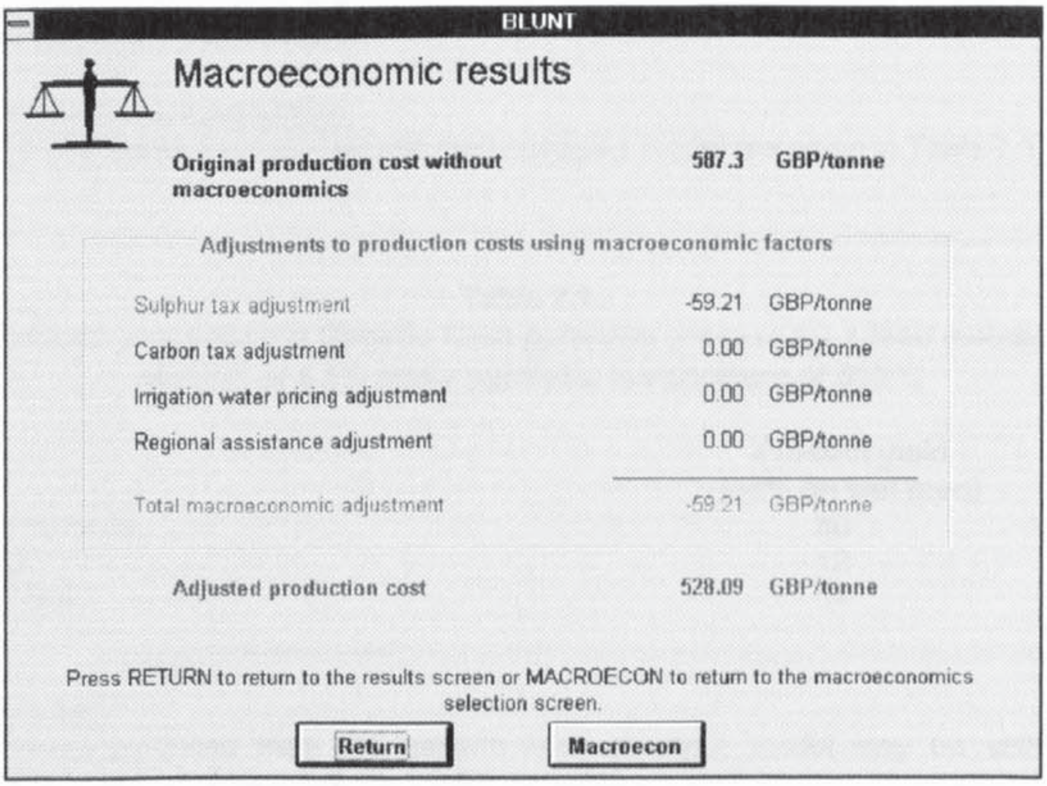
Regional assistance model

To assess the effects of local grants on the biomass process viability.
The following variables may be altered. Click on the appropriate text box to alter the default setting.

Employment credit	25.00	Percentage of annual labour costs
Proportion of workforce eligible for employment credit	50.00	%
Capital grant or allowance	2000.00	thousands GBP

Press ACCEPT to accept any changes or DEFAULT to reset the variables to the default values

Figure 6.5. The regional assistance model screen.



Macroeconomic results

Original production cost without macroeconomics	**587.3**	**GBP/tonne**

Adjustments to production costs using macroeconomic factors

Sulphur tax adjustment	-59.21	GBP/tonne
Carbon tax adjustment	0.00	GBP/tonne
Irrigation water pricing adjustment	0.00	GBP/tonne
Regional assistance adjustment	0.00	GBP/tonne
Total macroeconomic adjustment	-59.21	GBP/tonne
Adjusted production cost	**528.09**	**GBP/tonne**

Press RETURN to return to the results screen or MACROECON to return to the macroeconomics selection screen.

Figure 6.6. The macroeconomics results screen.

APPENDIX SECTION 7.
7.1. THE CONVERSION AND UPGRADING MODELS.

7.1.1. The generic flash pyrolysis model.

The generic flash pyrolysis model is a hybrid process which uses empirical relationships for each product's yield, including reaction water, to calculate the mass balance. The empirical relationships are based on published results from the University of Waterloo because their process is representative of flash pyrolysis and they possess the most comprehensive and consistent data set. The generic model is not intended to represent the University of Waterloo's process but to give a flexible tool that can demonstrate the effects of temperature and residence time on product yields, product quality (e.g. heating value and moisture content) and energy requirements.

Process description.

The prepared wood feedstock is fed to the fluidised bed reactor using a screw feeder. The reactor's bed is fluidised by recycled off-gas which is pre-heated in a furnace. The reactor operates for temperatures between 450 and 650°C. The product stream leaves the reactor and passes through a cyclone for removal of the char product. The pyrolysis vapours and gas are cooled on leaving the cyclone using a quench column utilising old product as the cooling medium. The liquid product is recovered and stored while some of the off-gas is returned to the reactor for use as fluidising gas and the remainder is removed from the system. The energy required for the pyrolysis process is met by utilising either the char, off-gas or imported fuel gas as fuel in the furnace. A simplified flowsheet for the process is given in Figure 7.1.

Feedstocks processed.

The feedstock allowed in the generic flash pyrolysis model is wood of any moisture content between 5 and 50wt%

Product yields and properties.

The product yields from the generic flash pyrolysis model are given in Table 7.1.

Table 7.1.
Product yields for the Generic flash pyrolysis process for a feed moisture content of 6.5% and a pyrolysis temperature of 500°C

	Product yield (wt% on wet feed)
Pyrolysis oil	80
Char	12
Off-gas	14

Product use.

The bio-oil produced from the generic flash pyrolysis model may be utilised to generate power or upgraded via zeolite cracking or hydrotreating processes to give crude aromatics, gasoline or diesel.

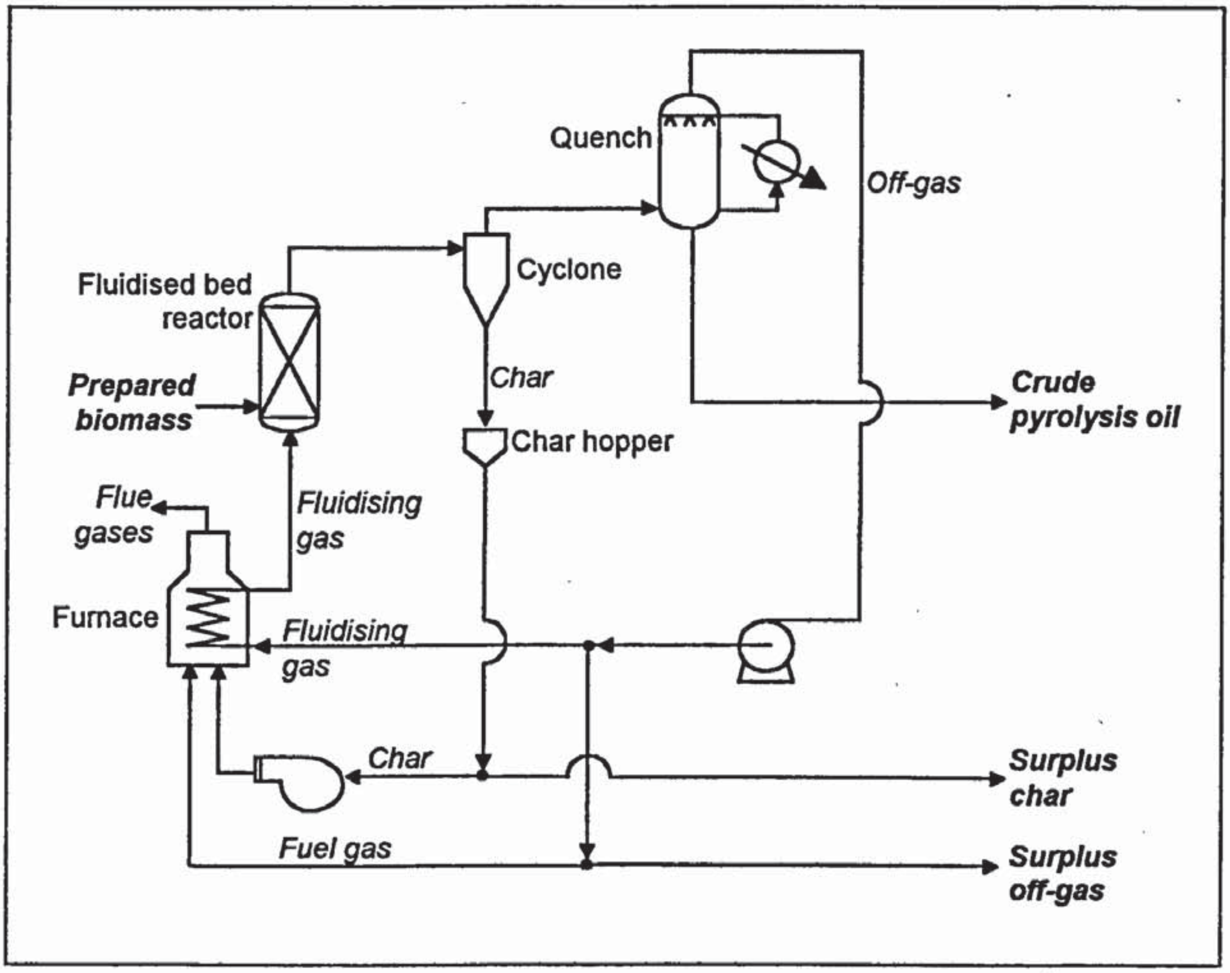


Figure 7.1. Flowsheet for the generic flash pyrolysis step model.

7.1.2. The Ensyn RTP pyrolysis model.

Ensyn Technologies Inc. (Ottawa, Canada) was incorporated in 1984 to develop and market flash pyrolysis technology for the production of chemicals. The Ensyn process utilises technology developed from the Ultrapyrolysis work carried out at the University of Western Ontario in the late 1970's. The Ensyn recirculating, transported-bed system is designed to allow rapid heat transfer to the feedstock with control of the short residence time. Biomass and fossil fuel feedstocks have been processed in Rapid Thermal Processing (RTP) plants with capacities from 0.3 - 1000 kg/hour, reactor residence times of 0.03 - 1.5 seconds and temperatures between 400 - 950°C.

Process description.

In the Ensyn system, feed passes from a hopper, through a metering and feed screws into the base of the reactor where it is mixed with the solid heat carrier. The product leaves the reactor and passes through a primary cyclone where the solid heat carrier is removed. The heat carrier is reheated before being recirculated to carry heat to fresh feedstock. A second cyclone removes the char and inorganic fines consisting of ash and entrained sand from the primary product stream. The secondary cyclone may be coupled to an in-plant recovery system which utilises the char for process heat generation. The product stream passes through condensers, filters and demisters where the liquid product is condensed and recovered.

Further details on the Ensyn system are not available due to commercial confidentiality. A flowsheet for the Ensyn process is given in Figure 7.2.

Feedstocks processed.

The RTP reactors require feedstocks with maximum particle sizes of 6mm and with moisture contents below 10% (wet basis). Biomass feed materials which have been processed in the Ensyn plants include wood (for example, Red Maple, IEA Poplar), cellulose, lignins, pulping residues and agricultural residues.

Product yields and properties.

The maximum total liquid yields obtained from dry poplar with a 10wt% moisture content is 83% at 500°C and 250ms. Table 7.2 gives typical product yields for the Ensyn RTP process.

Table 7.2.
Product yields for the Ensyn RTP pyrolysis process.

	Product yield (wt% on dry feed)
Pyrolysis oil	67
Char	8
Off-gas	26

The bio-oil produced from the Ensyn process is a complex liquid consisting of water, depolymerised lignin, carboxylic acids, aldehydes, carbohydrates, alcohols, ketones and other chemical groups. The products of interest from the Ensyn process include liquid fuel oils for use in boiler and turbines, speciality chemicals, commodity chemicals, polymers including copolymers and resins, petrochemicals including ethylene and other olefins and liquid transportation fuels. Table 7.3. gives some properties of wood derived pyrolysis oils from the Ensyn process.

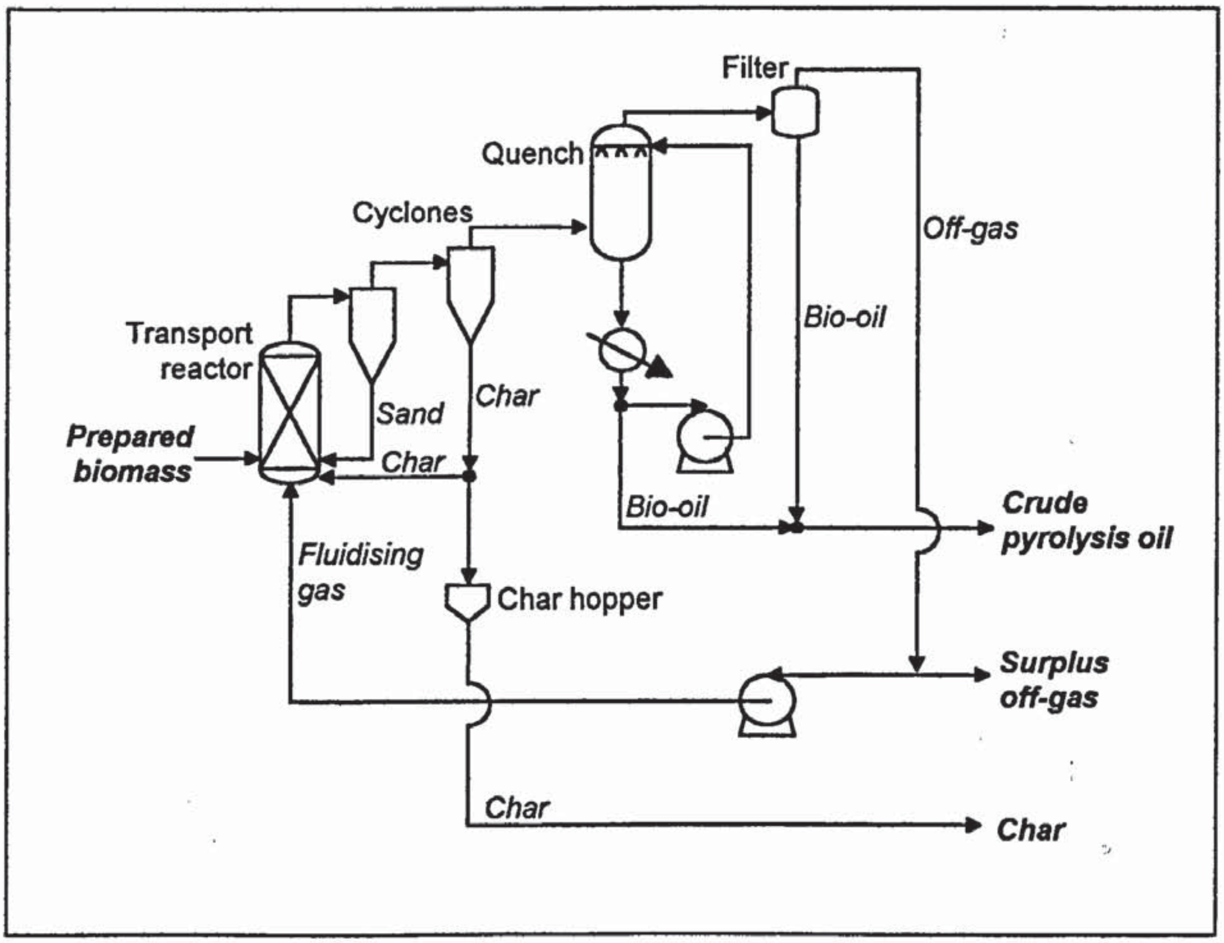


Figure 7.2. Flowsheet for the Ensyn RTP pyrolysis step model.

Product use.
A 25 tonnes/day Ensyn RTP system has been in commercial operation since 1992. The crude liquid product produced is delivered to existing extraction processes where speciality chemicals are isolated and recovered for commercial sale. The residue after chemical extraction has been test fired in a 2 GJ/hr fire-tube, single-pass Kewanee boiler normally fuelled by natural gas. A two-phase air-atomisation nozzle assembly with large fuel passages was designed and fitted to the boiler to allow the use of filtered pyrolysis oil. The boiler was operated for 2.5 hours during which the required steam output was maintained.

A series of flame tunnel tests were conducted on Ensyn RTP oil to compare their combustion and heat transfer performance to commercial grade heating oils. The filtered bio-oil was preheated in-line to 90°C and completed a three-hour test burn with flue gas oxygen levels maintained at 5.5%. The thermal heat transfer and gas temperature profiles of the bio-oil were comparable with those of fossil fuels. However, the particulate and carbon monoxide levels in the bio-oil combustion flue gas were higher than for fossil fuel flue gases and were probably due to incomplete burnout of the char.

Table 7.3.
Ensyn RTP bio-oil characteristics.

Property	Typical values
Moisture content (%)	16
Acidity (pH)	2.5
Specific gravity/density (kg/l)	1.21
Higher heating value, dry basis (MJ/kg)	23
Higher heating value as produced (MJ/kg)	19.3
Viscosity (cSt @ 40°C)	51
Pour point (°C)	-23
C/H ratio	9.1
Elemental analysis (%)	
Carbon	56.4
Hydrogen	6.2
Nitrogen	0.2
Sulphur	< 0.01
Ash	0.1
Oxygen (by difference)	37.1

RTP III oil has been tested in a single cylinder Petter engine with 5% cetane enhancer. Although there was a greater ignition delay than for diesel, the bio-oil burned rapidly. The combustion duration and exhaust emissions are reported to be similar to conventional diesel. The oil has also successfully fired in a dual fuel mode in a modified four cylinder industrial diesel engine.

7.1.3. The NREL vortex pyrolysis model.

The National Renewable Energy Laboratory, NREL have been developing an ablative flash pyrolysis system since 1980. The NREL system consists of a vortex reactor, utilising solid conductive heat transfer, which can process 20-25 kg/hr of feedstock. The technology is being scaled by Interchem Industries Inc. to convert 32.7 dry tonnes per day of sawdust into fuel oil replacements and charcoal. Construction of the plant started in the Spring of 1990 and it has undergone testing and redesign since then.

Process description.

The feedstock enters the system where it is entrained and mixed with a high temperature, high velocity carrier gas stream, typically nitrogen or steam. The feed particles enter the vortex reactor tangentially at speeds of over 100m/s and are forced to the reactor wall by high centrifugal forces. Raised helical ribs along the wall of the reactor force the particles into tighter helical paths than would normally occur and increases the contact of the particles with the reactor wall. The reactor wall temperature is limited to a maximum of 625°C to ensure production of a liquid film between the wall of the reactor and the particle, which then vaporises and leaves the reactor. Higher wall temperatures result in solid coke formation on the wall which prevents rapid particle movement and quickly fouls the heat-transfer surface of the reactor. The 625°C wall temperature results in vapour exit temperature and residence time of 525°C and 0.3 seconds respectively and gives the maximum yield of vapours for most biomass feedstocks. A schematic diagram of the process is shown in Figure 7.3.

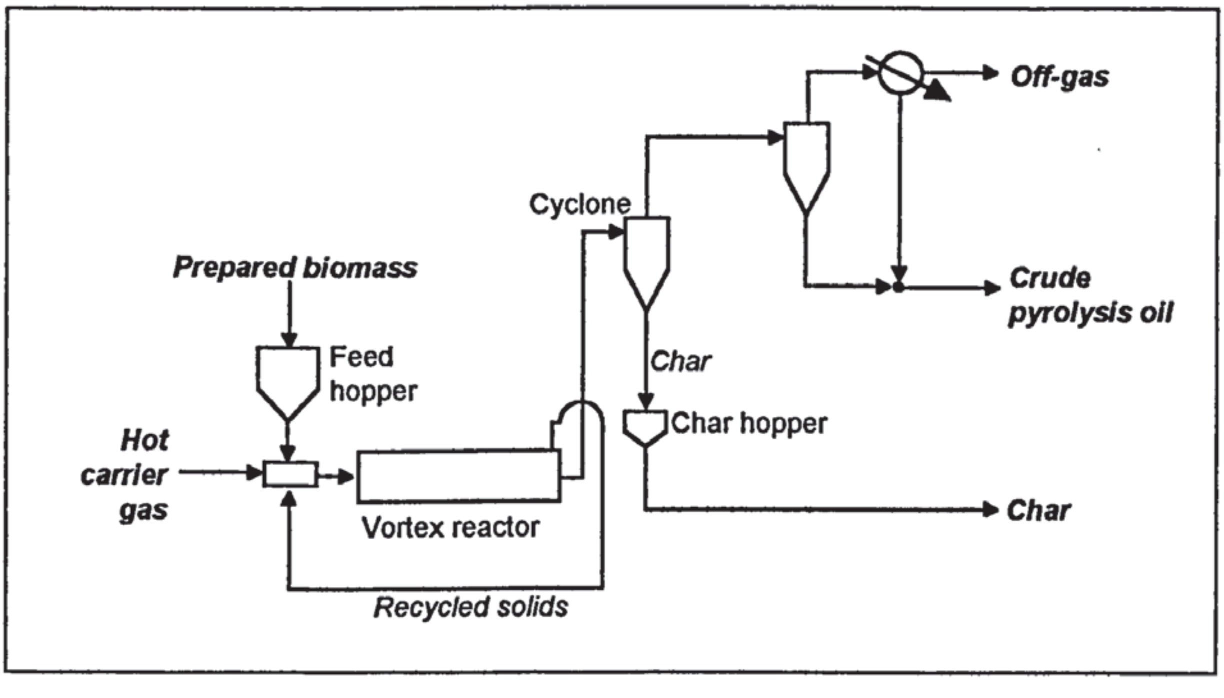


Figure 7.3. Flowsheet for the NREL vortex pyrolysis step model.

An insulated recycle loop at the exit of the reactor returns partially pyrolysed feedstock and any large char particles to the reactor entrance. Fine char particles (< 50 micrometers in size), gases and vapours leave the reactor through an axial exit. The product stream passes through a char cyclone, maintained above 125°C, where the char is removed as a dry powder. The vapours pass to the first heat exchanger where the condensed liquids and water are retained in the receiver. The cooled gas stream then passes through a series of heat exchangers before being flared.

Feedstocks processed.

Oak, southern pine and switchgrass have been processed to generate oils for combustion testing. The feedstocks were milled and screened through a 2.175mm ($^1/_8$ inch) sieve and dried at 105°C to moisture free conditions.

Product yields and properties.

Typical product yields from the NREL vortex system are given in Table 7.4.

Table 7.4.

Product yields for the NREL vortex pyrolysis process.

	Product yield (wt% on dry feed)
Pyrolysis oil (wet)	74.0
Char	14.0
Off-gas (by difference)	12.0

Table 7.5 gives some properties of wood derived pyrolysis oils from the NREL system.

Table 7.5.

NREL fast pyrolysis wood-oil characteristics.

	Dry basis	**As produced**
Moisture content (%)		15
H/C ratio (atomic)	1.78	1.26
O/C ratio (atomic)	0.68	0.55
Density (g/cm^3 @ 55°C)	1.23	
Heating value, dry basis (MJ/kg)	22.3	20.3
Viscosity (cp. @ 30°C)	90	90
Ash (%)		0.05
Elemental analysis (wt% on feed)		
Carbon	48.6	54.4
Hydrogen	7.2	5.7
Nitrogen	< 0.1	< 0.1
Oxygen	44.2	39.8

Product use.

The vapour stream from the reactor (prior to being condensed) has been cracked using zeolite catalysts to produce aromatics and gasoline.

7.1.4. The Fenosa flash pyrolysis model.

Union Fenosa (Spain) have constructed a 200 kg/hour biomass flash pyrolysis pilot plant which has been in operation since 1992. The process is based on the fluidised bed flash pyrolysis process developed by the University of Waterloo (Canada). The Fenosa plant encountered several difficulties on start up and modifications to equipment were required. The plant has been operating for 300 hours since July 1993 and is currently processing 160 kg/hour of dry biomass.

Process description.
The wood feedstock is dried utilising hot air produced by a propane burner before being ground in a shredder. The feed is classified with oversized particles being returned for regrinding. The prepared feed passes through a cyclone to remove it from the gas stream before being conveyed to the reactor feed hopper. The feed is fed to the reactor through a screw conveyor which is used to control the feed rate and hence the temperature of the fluidised bed. The fluidised bed reactor is cylindrical with a fine sand bed located above the nozzle plate. The bed is fluidised by recycled process gas which is heated by a propane burner. The reactor operates at 500°C with a fluidising gas temperature of 750°C and vapour residence time of 0.5-1.0 seconds.

The product stream leaves the reactor at a temperature of 366°C and passes through a cyclone for removal of the char product. The pyrolysis vapours and gas are cooled on leaving the cyclone to just above the liquid dew point using three water cooled heat exchangers in series. The liquid product is recovered and stored while some of the off-gas is returned to the reactor for use as fluidising gas and the remainder is removed from the system. A simplified flowsheet for the Fenosa process is given in Figure 7.4.

Feedstocks processed.
To date wood waste and forestry biomass have been processed with raw moisture contents between 30-45%.

Product yields and properties.
The product yields reported for the process are given in Table 7.6.

Table 7.6.
Product yields for the Fenosa flash pyrolysis process.

	Product yield (wt% on dry feed)
Pyrolysis oil (crude, wet)	60.0
Char	15.0
Off-gas (by difference)	25.0

Product use.
There were early operating problems at low throughputs which delayed the analysis and use of the Fenosa oils. Consequently, apart from the early preliminary product characteristics, there have been no published results or applications of the Fenosa pyrolysis oil.

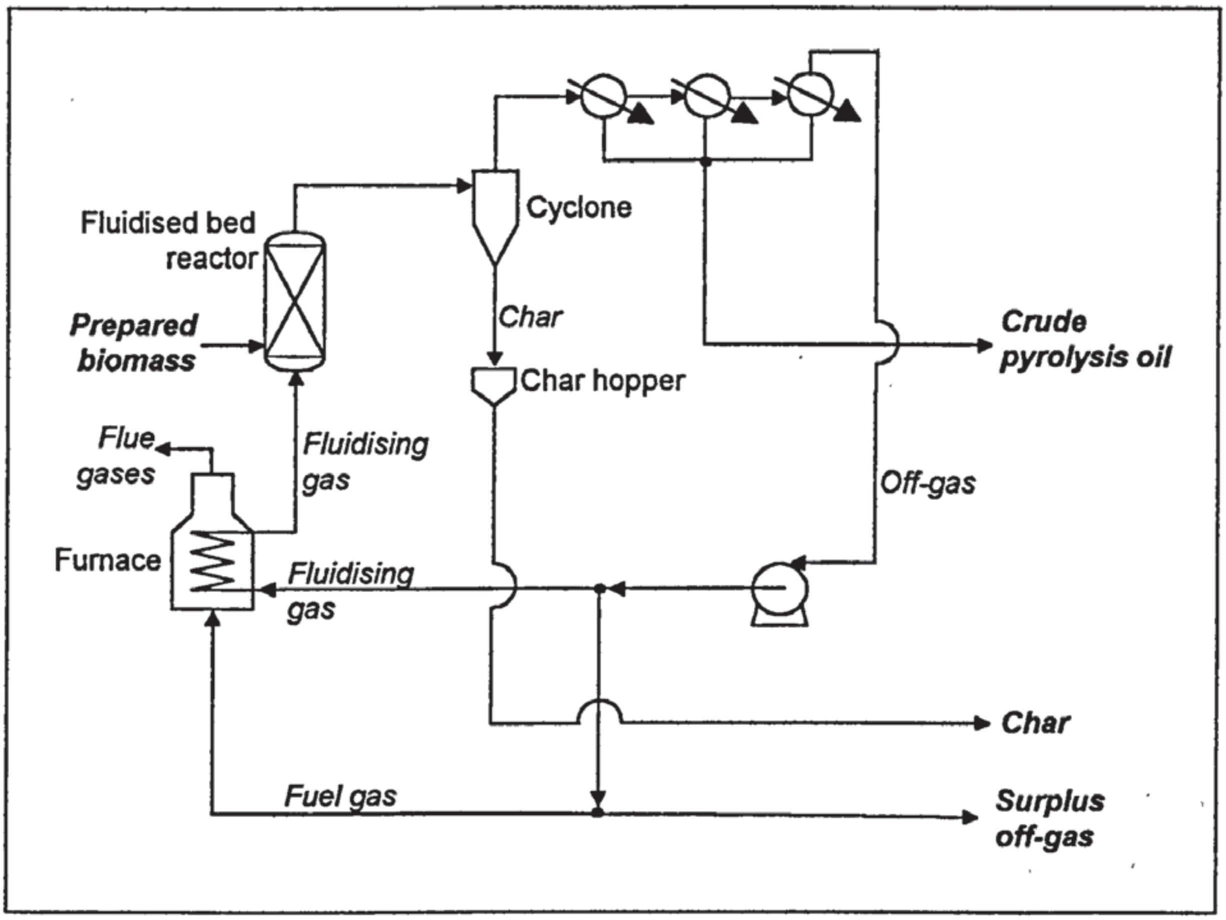


Figure 7.4. Flowsheet for the Fenosa flash pyrolysis step model.

7.1.5. Hydrotreating step model.

Upgrading of crude pyrolysis oil by hydrotreatment produces a more stable product which is a naphtha like hydrocarbon with properties similar to light distillate or kerosene. The hydrotreating step model included in BLUNT was based on a pilot plant design commissioned from Battelle PNL. The hydrotreating pilot plant design is based on their early work on the catalytic hydrotreatment of biomass pyrolysis oil.

Process description.
The Battelle plant incorporates two hydrotreatment reactors. The first reactor is used to provide a low temperature, 250-300°C, catalytic treatment of the pyrolysis oil and is significantly smaller than the second reactor. The oil leaving the first stage reactor is more stable and can be suitably upgraded at a higher temperature, 350-400°C, in the second stage reactor. Operating pressures are in the 170-190 bar range. It is assumed that the pyrolysis oil entering the plant is deoxygenated by 98% in the process.

The upgraded product from the second stage reactor is cooled to 70-80°C to condense most of the water.

There are two hydrogen source/off-gas disposal option available.

1. A hydrogen recycle is utilised to minimise the amount of hydrogen required for operation. The hydrogen is separated from the product oil and water in a high-pressure separation vessel at about 134 bar. Carbon dioxide, water and light organics are removed from the hydrogen rich vapour in a membrane separation unit. The purified hydrogen is combined with make-up hydrogen at 20°C and 1 bar and compressed to the reactor pressure. The remaining off-gas can be used as fuel gas or flared.

2. There is no recycling of hydrogen and all the hydrogen required for the hydrotreating process is imported from the utilities. The off-gas is separated from the product oil and water and sold to the utilities as high value fuel gas.

A schematic flow sheet of the base hydrotreating process modelled can be seen in Figure 7.5.

Feedstocks processed.
The feedstock for the hydrotreating process is crude pyrolysis oil.

Product yields and properties.
The yield for a 98% deoxygenated hydrotreated product is 34wt% on dry wood. The properties of the product depends on the deoxygenation extent of the feedstock. For deoxygenation extents below 98% the hydrotreated product is still miscible with water and above 98% the product is immiscible with water and miscible with conventional fossil fuels.

Product use.
The hydrotreated product may be utilised to produce electricity. If a deoxygenation extent of greater than 98% was utilised the product may be blended in a refinery to produce diesel.

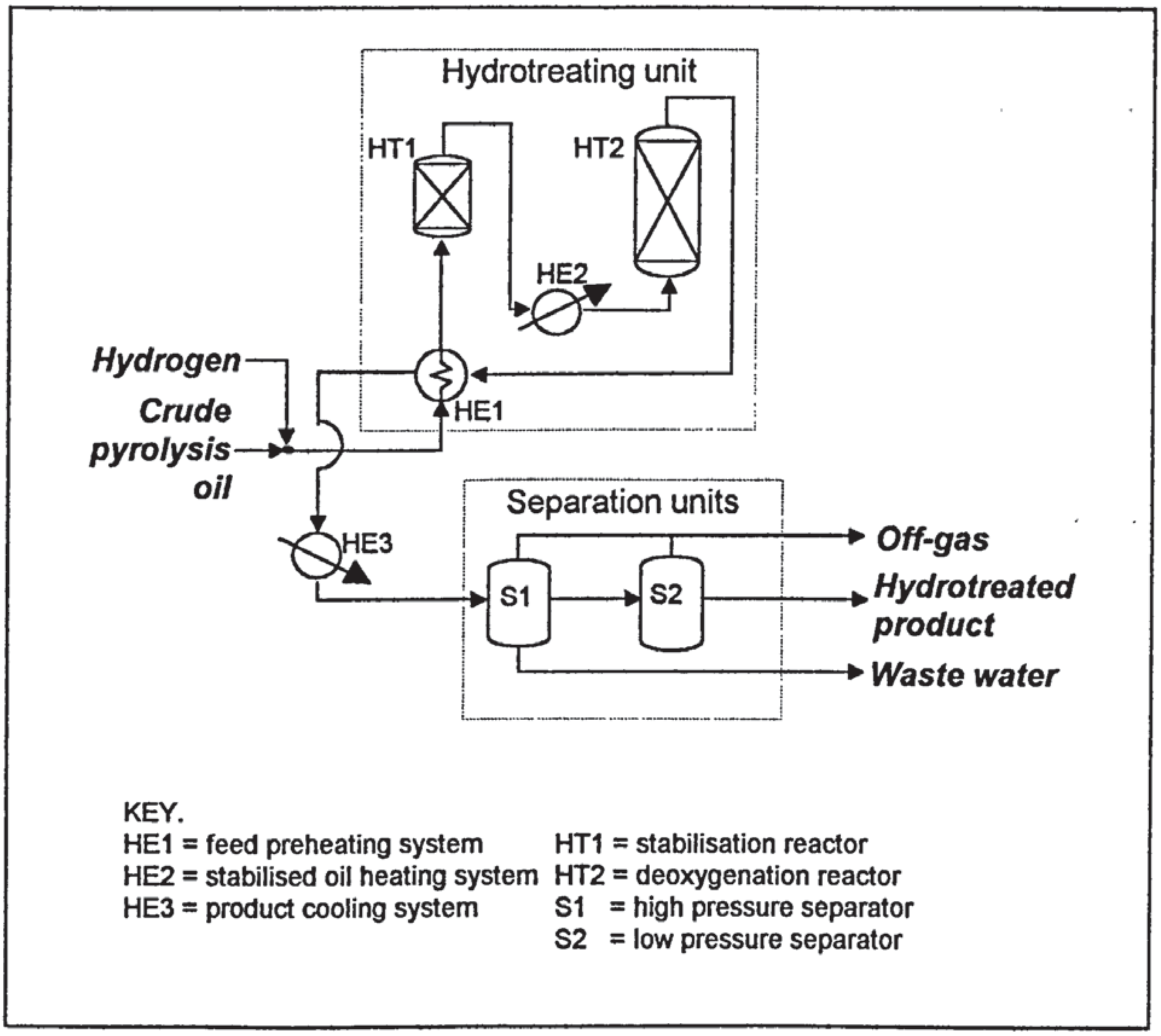


Figure 7.5. Flowsheet for the hydrotreating step model.

7.1.6. Zeolite cracking present case model.

The Liquefaction Group of the IEA Biomass Agreement carried out technoeconomic assessments of the NREL biomass to gasoline processes. The process was analysed as both present and potential cases. The present case represents product yields determined experimentally and commercially demonstrated processing techniques. The mass balances and cost estimations derived by the IEA ALPS Group were used to develop the models detailed below. The model is fixed, i.e. have no internal variables which can be changed by the program user, due to the sparcity of information on how changes in processing conditions would affect product yields. The only variability in the models comes from the crude vapours and gas feeding from the pyrolysis models.

Process description.

The hot vapours and gases from pyrolysis module (after char removal) enter the catalytic cracking reactor after being preheated with the product stream from the cracker. The cracker consists of a recirculating entrained bed with a 1 - 2 second contact between the pyrolysis vapours and the zeolite catalyst e.g. Mobil's HZSM-5. The temperature of the cracking reactor is between 500-525°C with a temperature rise of 20 to 50°C due to the exothermic reactions. Attached to the cracking reactor is a fluidised bed catalyst regenerator in which the coke formed on the catalyst is burned off using air in an oxygen-starved environment. The flowsheet for the present case model is shown in Figure 7.6.

After the process stream leaves the cracking reactor, it is cooled to condense the hydrocarbons heavier than benzene and any steam. The product stream is split into waste water, gases and condensed organics in a separator. The waste water is removed for treatment and disposal. The gas stream, consisting of permanent gases, gaseous olefins and a small amount of benzene vapour, is fed to the fixed-bed secondary catalytic reactor (olefin recovery) where the olefins alkylate to form cumene. The organic liquids stream is cooled further before being recombined with the olefin stream in the absorber.

The absorber product is a crude hydrocarbon mixture, rich in alkylated monocyclic aromatics but low in benzene. It is separated from the waste water which is removed for treatment and disposal. The crude aromatics product is suitable for use in a refinery to produce gasoline.

Feedstocks processed.

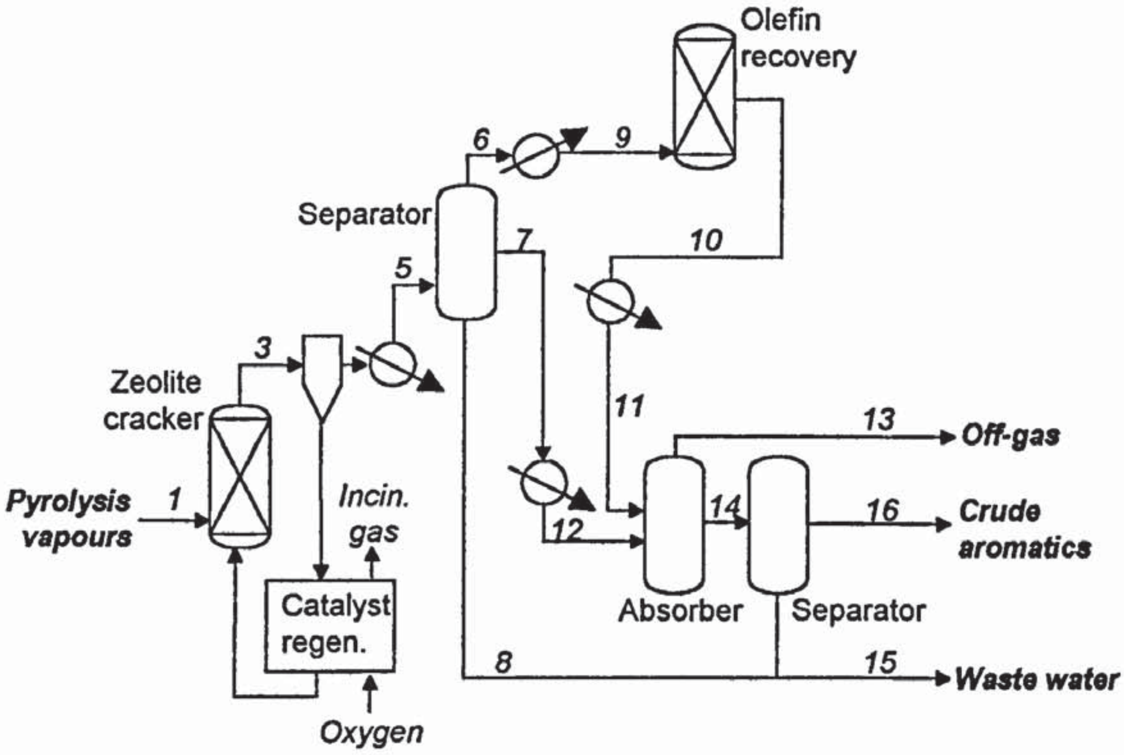
The feedstock for the zeolite cracking process are the vapours from the flash pyrolysis process.

Product yields and properties.

The product yield from the present case zeolite model is 21wt% on dry wood.

Product use.

The products from the present case zeolite cracking model are crude aromatics which may be used directly as chemical feedstocks, for power production or as a refinery feedstock to produce gasoline.



Stream Identification

1. Pyrolysis vapour feed
3. Reactor products
5. Cooled products
6. Separation gas stream
7. Separation waste water stream
8. Separation liquid stream
9. Olefin recovery feed
10. Olefin products
11. Olefin products to absorber
12. Liquid product to absorber
13. Absorber off-gas
14. Absorber liquid product
15. Waste water stream
16. Crude aromatic product

Figure 7.6. Flowsheet for the present case zeolite cracking model.

7.1.7. ZEOLITE CRACKING POTENTIAL CASE MODEL.

The Liquefaction Group of the IEA Biomass Agreement carried out technoeconomic assessments of the NREL biomass to gasoline processes. The process was analysed as both present and potential cases. The potential case assumes progress in research and development programs to attain the improved yields and modified process. The mass balances and cost estimations derived by the IEA ALPS Group were used to develop the models detailed below. The model is fixed, i.e. have no internal variables which can be changed by the program user, due to the sparcity of information on how changes in processing conditions would affect product yields. The only variability in the models comes from the crude vapours and gas feeding from the pyrolysis models.

Process description.
The pyrolysis vapours and gases enter the catalytic cracking reactor after passing through the char cyclones. The recirculating entrained bed reactor has a 1 - 2 second contact between the pyrolysis vapours and the zeolite catalyst. The temperature of the reactor is between 500 and 525°C with a temperature rise of 25 to 50°C from the exothermic reactions. The catalyst regenerator is as for the present case model. The flowsheet for the future case model is shown in Figure 7.7.

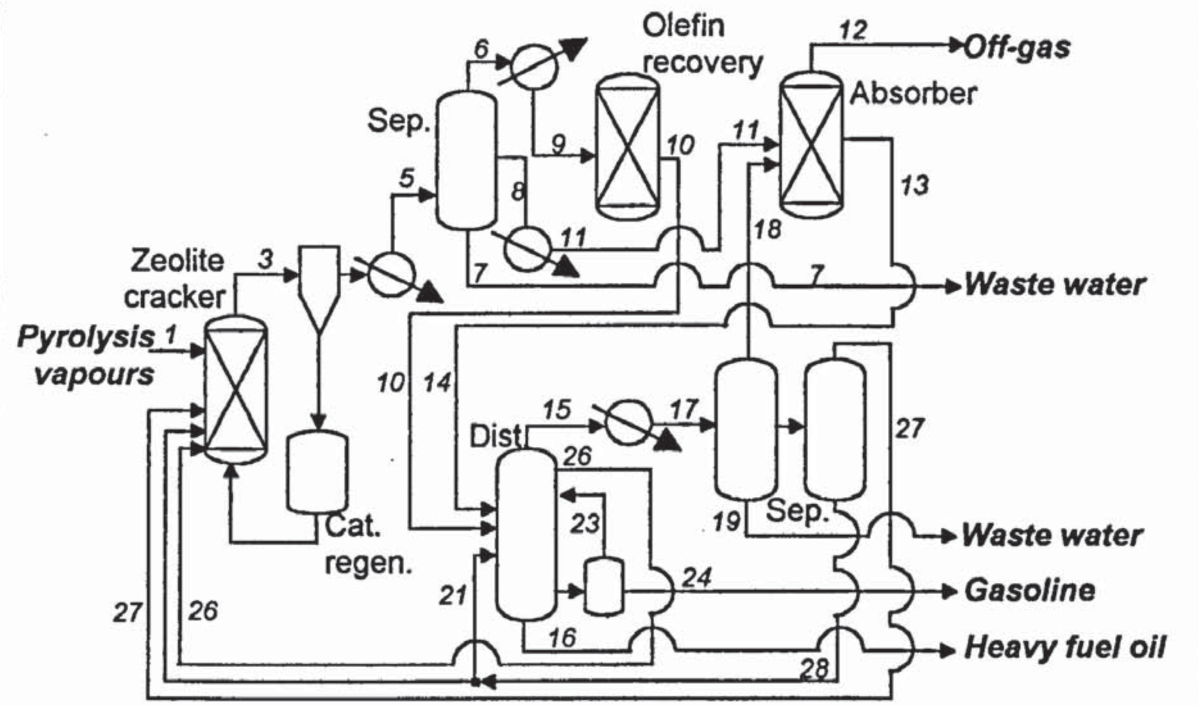
After the process stream leaves the reactor it is cooled to condense most of the steam and the hydrocarbons heavier than benzene. The permanent gases, gaseous olefins and benzene vapour are compresses to 100 psig before entering the secondary catalytic reactor. In the fixed bed secondary reactor the olefins alkylate to form cumene. The hot vapours from the secondary reactor enter the distillation column where they supply the heat for distillation. Unreactive butanes and pentanes along with gas oil are recycled to the cracking unit to be transformed into reactive intermediates. A small amount of polycyclic aromatic oil may be drawn off the bottom of the distillation tower and can be burned for fuel, hydrocracked to gasoline in a refinery or sold as fuel oil.

The volatile hydrocarbons are recovered by absorption or condensation from the noncondensible gases in the process stream. The hydrocarbon product condensed at atmospheric pressure is cooled and pumped into the absorber.

Feedstocks processed.
The feedstock for the zeolite cracking process are the vapours from the flash pyrolysis process.

Product yields and properties.
The product yield from the potential case zeolite model is 25wt% on dry wood.

Product use.
The product from the potential case zeolite process is gasoline which can be used as a substitute for fossil transportation fuels.



Stream Identification

1. Pyrolysis vapour feed
3. Reactor products
5. Cooled products
6. Separation gas stream
7. Separation waste water stream
8. Separation liquid stream
9. Olefin recovery feed
10. Olefin products
11. Liquid product to absorber
12. Absorber off-gas
13. Absorber product
14. Absorber product to distillation
15. Distillation gas stream
16. Heavy fuel oil
17. Distillation gas to separation
18. Separation off-gas
19. Waste water stream
20. Separation liquid stream
21. Fuel oil
22. Distillation gasoline stream
23. Distillation steam
24. Hot gasoline stream
25. Gasoline product stream
26. Fuel oil recycle
27. Hydrocarbon gas recycle
28. Gasol recycle

Figure 7.7. Flowsheet for the potential case zeolite cracking model.

7.2. THE MACROECONOMIC MODELS.

7.2.1. The sulphur tax on fossil fuels model.

The sulphur tax model is designed to measure the effect of controlling the use of high sulphur fuels. The impetus behind this macroeconomic factor was the European Commission's Large Combustion Plant Directive in 1988 which set out national percentage reductions in sulphur emissions from power stations to be met by 2003. The ways in which the LCPD can be met are:

- to use fuels with a lower sulphur content (e.g. biomass derived fuels); or
- to remove the sulphur from the atmospheric emissions.

Sulphur taxes, a levy per unit on sulphur rich fuels used in power stations, are not so far used in the European Union. However, in 1991 the Swedish government instituted a 30 SKR/kg tax on sulphur emissions hoping to reduce sulphur emissions by 10-25 kilotonnes by the end of the decade.

The sulphur tax model imposes a tax on high sulphur content fossil fuels equivalent to 50 GBP/1% sulphur present. The model then determines the effect the tax has on the price of electricity generated from the high sulphur fuels and utilised by biomass converters. This is utilised to determine the "price advantage" of using biomass derived (and hence low sulphur) fuels. The "price advantage" is the effective amount by which the sulphur tax decreases the biomass derived fuel production cost.

The information required by the sulphur tax model includes:

- a sulphur tax on fossil fuels in GBP/1% sulphur;
- the amount of coal required for power generation in M t/year;
- the amount of oil required for power generation in K barrels/year;
- the amount of gas required for power generation in M m^3/year;
- the electricity pool price in £/kWh;
- the power station power delivery in GWh/year;
- the proportion coal used to generate the power , %;
- the proportion oil used to generate the power , %; and
- the proportion gas used to generate the power , %.

7.2.2. The carbon tax on fossil fuels model.

The carbon tax model looks at measures to reduce carbon dioxide emissions. Carbon dioxide emissions in Europe are growing by 4% per annum with 31% of the emissions coming from power generation. In 1990 at the European Joint Energy/Environment Council it was decided to stabilise carbon dioxide emissions at 1990 levels by 2000. To this end, the CEC proposed a package of fiscal and monetary measures aimed at improving energy efficiency and discouraging the use of fuels which generate large quantities of carbon dioxide when combusted. The proposed tax has:

- an Energy Tax on all non-renewable fuels plus large scale hydro-schemes; and
- a Carbon Tax levied on all fossil fuels.

Non-energy use of fuels such as oil used as a feedstock for petrochemical manufacturers or in the production of man-made fibres are exempted from the tax. The tax has the heaviest impact on coal and oil and favours renewable fuels and nuclear power.

The carbon tax model imposes a tax on fossil fuels and determines the effect this has on the price of electricity utilised by biomass converters. This is utilised to determine the "price advantage" of using biomass derived fuels. The "price advantage" is the effective amount by which the carbon tax decreases the biomass derived fuel production cost.

The information required by the carbon tax model includes:

- the carbon tax on coal in £/tonne;
- the carbon tax on oil in £/barrel;
- the carbon tax on gas in £/m^3;
- the amount of coal required for power generation in M t/year;
- the amount of oil required for power generation in K barrels/year;
- the amount of gas required for power generation in M m^3/year;
- the electricity pool price in £/kWh;
- the power station power delivery in GWh/year;
- the proportion coal used to generate the power , %;
- the proportion oil used to generate the power , %; and
- the proportion gas used to generate the power , %.

7.2.3. The water resources model.

Water availability and pricing policies vary across Europe and the water consumed by crop irrigation is a function of industrial development and climate. Southern states tend to have high levels of irrigated land as a proportion of total cultivated land. For example, 4.23% of cultivated land is irrigated in Germany while in Italy the figure is 23.6%.

In states with agriculture dominated economies a high proportion of total water withdrawals tend to be used for irrigation. Two types of water pricing policy exist:

- those based on the volume of water withdrawn for irrigation; and
- flat rate pricing structures.

The irrigation water pricing policy adopted for the model is based on the volume of water used for irrigation. Such systems are common in developed economies and typically include accurate measuring systems to ensure fair charging. Such systems also encourage efficient irrigation as users have a direct financial incentive to cut down their water usage.

The water pricing model assesses the effect of altering the cost of irrigation water on the cost of the biomass feedstock. This is utilised to determine the "price advantage" of using biomass derived fuels. The "price advantage" is the effective amount by which the water pricing model decreases or increases the biomass derived fuel production cost.

The information required by the water resources model includes:

- the old irrigation water cost in £/m3;
- the area growing feedstock in K hectares;
- the typical water requirement in m3/ha/year; and
- the new irrigation water cost in £/m3.

7.2.4. The regional assistance model.

The regional assistance model considers the use of incentives which facilitate the establishment of biomass conversion plants. The incentives include grants which lower the amount of capital which the investor needs to raise and service. The total risk of the project is therefore lowered, allowing the construction of larger, more efficient units. The most important programme of grants in this category is the five year THERMIE programme established by the Directorate-General for Energy to promote new projects.

In relation to biomass, THERMIE funding is available for projects which involve :

- the direct or indirect use of biomass and all plant, animal, urban, and industrial waste; or
- the use of technology to produce and use biomass for energy purposes.

The regional assistance model assesses the effect of a range of local grants on the capital cost of a biomass project. This is utilised to determine the "price advantage" of using biomass derived fuels. The "price advantage is the effective amount by which the regional assistance model decreases the biomass derived fuel production costs.

The information required by the regional assistance model includes:

the employment credit as a percentage of the annual lab costs;
the proportion of the workforce eligible; and
the amount of the capital grant in K GBP.

7.3. Sample results summary printout from BLUNT

BLUNT RESULTS SUMMARY.

Run date : 15.Dec.1994
Run identifier : BLUNT01

All costs are in 1993 Great Britain Pounds (Exchange rate with GB pounds = 1.0)

Product : Crude pyrolysis oil
Conversion route : Generic flash pyrolysis
Product output (t/day) : 82.3
Product yield (mass %) : 82.3

Production cost (tonnes/day) : 119.9
Capital cost (M GBP) : 6.1

BASED ON THE FOLLOWING INFORMATION

Feed data

Feedstock : Wood
Feed input (d.a.f. t/day) : 100.0
Feed moisture content (wt% dry) : 7.0
Feedstock cost (GBP/d.a.f. tonne) : 30.0

Financial base case

Financial variables

Project life (years) : 10.0
Operational days per year (days) : 330.0
Real interest rate (%) : 5.0
Inflation rate (%) : 5.0
Maintenance (% capital cost / year) : 5.0
Overheads (% capital cost / year) : 7.0

Utilities costs

LP steam (GBP/ tonne) : 6.0
Oxygen (GBP/ tonne) : 50.0
Cooling water (GBP/ tonne) : 0.1
Fuel gas (GBP/ GJ) : 4.0
Hydrogen (GBP/ tonne) : 360.0
Power (GBP/ Mwh) : 50.0

Customised step models

Wood drying model

Moisture content entering (wt%) : 50.0
Moisture content exiting (wt%) : 7.0

Generic flash pyrolysis model

Reactor temperature (degC) : 500.0
Char/off-gas combustion efficiency (%) : 75.0
Char filter efficiency (%) : 90.0
Coolant temperature rise (degC) : 20.0
Product outlet temperature (degC) : 40.0
Pyrolysis energy source option : Use char first, then off-gas then imported fuel-gas

ANNUAL COST BREAKDOWN

(All costs are in thousands of GBP)

Capital amortisation	: 776.80	(23.84 %)
Feedstock	: 990.00	(30.39 %)
Utilities	: 457.80	(14.05 %)
Maintenance	: 305.80	(9.39 %)
Overheads	: 428.10	(13.14 %)
Labour	: 333.00	(10.22 %)
Additional	: 0.00	(0.00 %)
Byproduct disposal	: -33.60	(-1.03 %)
TOTAL ANNUAL COSTS	: 3257.90 K GBP	

CAPITAL COST BREAKDOWN

(All costs are in thousands of GBP on a 1993 basis.)

Step		
Wood reception	: 1095.80	(17.92 %)
Wood storage	: 916.30	(14.98 %)
Wood drying	: 1610.60	(26.34 %)
Wood grinding	: 604.90	(9.89 %)
Generic flash pyrolysis	: 1701.90	(27.83 %)
Liquid storage	: 185.90	(3.04 %)
not used	: 0.00	(0.00 %)
not used	: 0.00	(0.00 %)
TOTAL CAPITAL COST	: 6115.4 K GBP	

UTILITIES USAGE SUMMARY

Step	LP steam t/day	Cooling water t/day	Fuel gas GJ/day	Hydrogen t/day	Oxygen t/day	Power MWh
Wood reception						0.20
Wood storage						0.00
Wood drying			334.80			0.10
Wood grinding						0.20
Generic flash pyrolysis		889.60	-10.40			0.40
Liquid storage						
not used						
not used						
TOTAL USAGE	0.00	889.60	324.40	0.00	0.00	0.90

ADDITIONAL INFORMATION

Byproduct streams

From the Generic flash pyrolysis step : 5.1 tonnes/day of char

Waste streams

none

Byproduct credits

From the Generic flash pyrolysis step : 102.00 GBP/ day for char disposal

Additional costs

none

BLUNT Simulation Mass Balance

<u>OUTLET STREAM from step 101 Wood reception</u>
Flow = 201.6129 tonnes/day
Composition :

Wood	Water	Ash
0.4960	0.5000	0.0040

Flows of each component

Wood	Water	Ash
100.0000	100.8065	0.8065

<u>OUTLET STREAM from step 111 Wood storage</u>
Flow = 201.6129 tonnes/day

Mass fraction

Wood	Water	Ash
0.4960	0.5000	0.0040

Flows of each component

Wood	Water	Ash
100.0000	100.8065	0.8065

<u>OUTLET STREAM from step 211 Drying</u>
Flow = 108.3940 tonnes/day

Mass fraction

Wood	Water	Ash
0.9226	0.0700	0.0074

Flows of each component

Wood	Water	Ash
100.0000	7.5876	0.8065

<u>OUTLET STREAM from step 221 Wood grinding</u>
Flow = 108.3940 tonnes/day

Mass fraction

Wood	Water	Ash
0.9226	0.0700	0.0074

Flows of each component

Wood	Water	Ash
100.0000	7.5876	0.8065

<u>OUTLET STREAM from step 301 Generic Flash pyrolysis</u>
Flow = 83.8392 tonnes/day

Mass fraction

Pyr oil	Char
0.9818	0.0182

Flows of each component

Pyr oil	Char
82.3100	1.5292

OUTLET STREAM from step 701 Liquid product storage
Flow = 83.8392 tonnes/day

Mass fraction

Pyr oil	Char
0.9818	0.0182

Flows of each component

Pyr oil	Char
82.3100	1.5292

Appendix B. Sample results printouts from BLUNT

BLUNT RESULTS SUMMARY.

Run date : 15.Dec.1994
Run identifier : BLUNT01

All costs are in 1993 Great Britain Pounds (Exchange rate with GB pounds = 1.0)

Product : Hydrotreated pyr oil
Conversion route : Generic flash pyrolysis and hydrotreating
Product output (t/day) : 32.0
Product yield (mass %) : 32.0

Production cost (tonnes/day) : 517.5
Capital cost (M GBP) : 12.1

BASED ON THE FOLLOWING INFORMATION

Feed data

Feedstock : Wood
Feed input (d.a.f. t/day) : 100.0
Feed moisture content (wt% dry) : 7.0
Feedstock cost (GBP/d.a.f. tonne) : 30.0

FINANCIAL BASE CASE

Financial variables

Project life (years) : 10.0
Operational days per year (days) : 330.0
Real interest rate (%) : 5.0
Inflation rate (%) : 5.0
Maintenance (% capital cost / year) : 5.0
Overheads (% capital cost / year) : 7.0

Utilities costs

LP steam (GBP/ tonne) : 6.0
Oxygen (GBP/ tonne) : 50.0
Cooling water (GBP/ tonne) : 0.05
Fuel gas (GBP/ GJ) : 4.0
Hydrogen (GBP/ tonne) : 360.0
Power (GBP/ MWh) : 50.0

CUSTOMISED STEP MODELS

Wood drying model

Moisture content entering (wt%) : 50.0
Moisture content exiting (wt%) : 7.0

Generic flash pyrolysis model

Reactor temperature (degC) : 500.0
Char/off-gas combustion efficiency (%) : 75.0
Char filter efficiency (%) : 90.0
Coolant temperature rise (degC) : 20.0
Product outlet temperature (degC) : 40.0
Pyrolysis energy source option : Use char first, then off-gas then imported fuel-gas

Hydrotreating model

Deoxygenation extent (%) : 98.0
Fuel gas combustion efficiency (%) : 75.0
Coolant temperature rise (degC) : 20.0
Heat exchanger heat loss (%) : 10.0
Product outlet temperature (degC) : 73.0
Hydrogen source/off-gas disposal option : sell off-gas as fuel gas, import all H2

ANNUAL COST BREAKDOWN

(All costs are in thousands of GBP)

Capital amortisation	: 1537.00	(28.12 %)
Feedstock	: 990.00	(18.11 %)
Utilities	: 880.40	(16.11 %)
Maintenance	: 605.00	(11.07 %)
Overheads	: 847.00	(15.50 %)
Labour	: 416.30	(7.62 %)
Additional	: 223.40	(4.09 %)
Byproduct disposal	: -33.60	(-0.61 %)

TOTAL ANNUAL COSTS : 5465.50 K GBP

CAPITAL COST BREAKDOWN

(All costs are in thousands of GBP on a 1993 basis.)

Step		
Wood reception	: 1095.80	(9.06 %)
Wood storage	: 916.30	(7.57 %)
Wood drying	: 1610.60	(13.31 %)
Wood grinding	: 604.90	(5.00 %)
Generic flash pyrolysis	: 1701.90	(14.07 %)
Hydrotreating	: 6098.50	(50.40 %)
Liquid product storage	: 72.00	(0.60 %)
not used	: 0.00	(0.00 %)

TOTAL CAPITAL COST : 12100 K GBP

UTILITIES USAGE SUMMARY

Step	LP steam t/day	Cooling water t/day	Fuel gas GJ/day	Hydrogen t/day	Oxygen t/day	Power MWh
Wood reception						0.20
Wood storage						0.00
Wood drying			334.80			0.10
Wood grinding						0.20
Generic flash pyrolysis		889.60	-10.40			0.40
Hydrotreating		2327.90	-868.30	12.80		0.50
Liquid product storage						
not used						
TOTAL USAGE	0.00	3217.50	-543.90	12.80	0.00	1.40

ADDITIONAL INFORMATION

Byproduct streams

From the Generic flash pyrolysis step : 5.1 tonnes/day of char

Waste streams

From the hydrotreating step : 37.5 tonnes/day of waste water contaminated with 0.0 tonnes/day of organics

Byproduct credits

From the Generic flash pyrolysis step : 102.00 GBP/ day for char disposal

Additional costs

From the hydrotreating step : 677.0 GBP/day for catalyst rejuvenation

From the hydrotreating step : 0.0 GBP/day for waste water disposal

BLUNT Simulation Mass Balance

OUTLET STREAM from step 101 Wood reception

Flow = 201.6129 tonnes/day

Mass fraction

Wood	Water	Ash
0.4960	0.5000	0.0040

Flows of each component

Wood	Water	Ash
100.0000	100.8065	0.8065

OUTLET STREAM from step 111 Wood storage

Flow = 201.6129 tonnes/day

Mass fraction

Wood	Water	Ash
0.4960	0.5000	0.0040

Flows of each component

Wood	Water	Ash
100.0000	100.8065	0.8065

OUTLET STREAM from step 211 Drying

Flow = 108.3940 tonnes/day

Mass fraction

Wood	Water	Ash
0.9226	0.0700	0.0074

Flows of each component

Wood	Water	Ash
100.0000	7.5876	0.8065

OUTLET STREAM from step 221 Wood grinding

Flow = 108.3940 tonnes/day

Mass fraction

Wood	Water	Ash
0.9226	0.0700	0.0074

Flows of each component

Wood	Water	Ash
100.0000	7.5876	0.8065

OUTLET STREAM from step 301 Generic Flash pyrolysis

Flow = 83.8392 tonnes/day

Mass fraction

Pyr oil	Char
0.9818	0.0182

Flows of each component

Pyr oil	Char
82.3100	1.5292

OUTLET STREAM from step 401 Pyr oil hydrotreating
Flow = 32.4698 tonnes/day

Mass fraction

H/T oil	Water	Methane	CO2	C2+ H/c
0.9856	0.0000	0.0003	0.0017	0.0124

Flows of each component

H/T oil	Water	Methane	CO2	C2+ H/c
32.0020	0.0000	0.0096	0.0557	0.4026

OUTLET STREAM from step 701 Liquid product storage
Flow = 32.4698 tonnes/day

Mass fraction

H/T oil	Water	Methane	CO2	C2+ H/c
0.9856	0.0000	0.0003	0.0017	0.0124

Flows of each component

H/T oil	Water	Methane	CO2	C2+ H/c
32.0020	0.0000	0.0096	0.0557	0.4026

BLUNT RESULTS SUMMARY.

Run date : 15.Jan.1995
Run identifier : BLUNT01

All costs are in 1993 Great Britain Pounds (Exchange rate with GB pounds = 1.0)

Product : Gasoline
Conversion route : Generic flash pyrolysis and zeolite cracking (pot)
Product output (t/day) : 26.3
Product yield (mass %) : 26.3

Production cost (tonnes/day) : 413.3
Capital cost (M GBP) : 9.2

BASED ON THE FOLLOWING INFORMATION

Feed data

Feedstock : Wood
Feed input (d.a.f. t/day) : 100.0
Feed moisture content (wt% dry) : 7.0
Feedstock cost (GBP/d.a.f. tonne) : 30.0

FINANCIAL BASE CASE

Financial variables

Project life (years) : 10.0
Operational days per year (days) : 330.0
Real interest rate (%) : 5.0
Inflation rate (%) : 5.0
Maintenance (% capital cost / year) : 5.0
Overheads (% capital cost / year) : 7.0

Utilities costs

LP steam (GBP/ tonne) : 6.0
Oxygen (GBP/ tonne) : 50.0
Cooling water (GBP/ tonne) : 0.05
Fuel gas (GBP/ GJ) : 4.0
Hydrogen (GBP/ tonne) : 360.0
Power (GBP/ MWh) : 50.0

CUSTOMISED STEP MODELS

Wood drying model

Moisture content entering (wt%) : 50.0
Moisture content exiting (wt%) : 7.0

Generic flash pyrolysis model

Reactor temperature (degC) : 500.0
Char/off-gas combustion efficiency (%) : 75.0
Char filter efficiency (%) : 90.0
Coolant temperature rise (degC) : 20.0
Product outlet temperature (degC) : 40.0
Pyrolysis energy source option : Use char first, then off-gas then imported fuel-gas

ANNUAL COST BREAKDOWN

(All costs are in thousands of GBP)

Capital amortisation	: 1162.30	(32.36 %)
Feedstock	: 990.00	(27.56 %)
Utilities	: -92.60	(-2.58 %)
Maintenance	: 457.50	(12.74 %)
Overheads	: 640.50	(17.83 %)
Labour	: 416.30	(11.59 %)
Additional	: 51.60	(1.44 %)
Byproduct disposal	: -33.60	(-0.94 %)
TOTAL ANNUAL COSTS	: 3592.00 K GBP	

CAPITAL COST BREAKDOWN

(All costs are in thousands of GBP on a 1993 basis.)

Step		
Wood reception	: 1095.80	(11.98 %)
Wood storage	: 916.30	(10.01 %)
Wood drying	: 1610.60	(17.60 %)
Wood grinding	: 604.90	(6.61 %)
Generic flash pyrolysis	: 1701.90	(18.60 %)
Zeol cracking (pot)	: 3162.50	(34.56 %)
Liquid product storage	: 58.40	(0.64 %)
not used	: 0.00	(0.00 %)
TOTAL CAPITAL COST	: 9150.4 K GBP	

UTILITIES USAGE SUMMARY

Step	LP steam t/day	Cooling water t/day	Fuel gas GJ/day	Hydrogen t/day	Oxygen t/day	Power MWh
Wood reception						0.20
Wood storage						0.00
Wood drying			334.80			0.10
Wood grinding						0.20
Generic flash pyrolysis						0.40
Zeol cracking (pot)	5.30	2364.20	-544.60		6.80	0.50
Liquid product storage not used						
TOTAL USAGE	5.30	2364.20	-209.80	0.00	6.80	1.40

ADDITIONAL INFORMATION

Byproduct streams

From the Generic flash pyrolysis step : 5.1 tonnes/day of char

Waste streams

From the zeolite cracking step : 0.0 tonnes/day of waste water contaminated with 6.1 tonnes/day of organics

Byproduct credits

From the Generic flash pyrolysis step : 102.00 GBP/ day for char disposal

Additional costs

From the zeolite cracking step : 95.5 GBP/day for catalyst rejuvenation
From the zeolite cracking step : 60.9 GBP/day for waste water disposal

BLUNT Simulation Mass Balance

OUTLET STREAM from step 101 Wood reception

Flow = 201.6129 tonnes/day

Mass fraction

Wood	Water	Ash
0.4960	0.5000	0.0040

Flows of each component

Wood	Water	Ash
100.0000	100.8065	0.8065

OUTLET STREAM from step 111 Wood storage

Flow = 201.6129 tonnes/day

Mass fraction

Wood	Water	Ash
0.4960	0.5000	0.0040

Flows of each component

Wood	Water	Ash
100.0000	100.8065	0.8065

OUTLET STREAM from step 211 Drying

Flow = 108.3940 tonnes/day

Mass fraction

Wood	Water	Ash
0.9226	0.0700	0.0074

Flows of each component

Wood	Water	Ash
100.0000	7.5876	0.8065

OUTLET STREAM from step 221 Wood grinding

Flow = 108.3940 tonnes/day

Mass fraction

Wood	Water	Ash
0.9226	0.0700	0.0074

Flows of each component

Wood	Water	Ash
100.0000	7.5876	0.8065

OUTLET STREAM from step 301 Generic Flash pyrolysis

Flow = 94.6314 tonnes/day

Mass fraction

Pyr oil	Char	Off gas
0.8698	0.0162	0.1140

Flows of each component

Pyr oil	Char	Off gas
82.3100	1.5292	10.7921

OUTLET STREAM from step 403 Zeolite cracking pot

Flow = 26.3370 tonnes/day

Mass fraction

Gasoline
1.0000

Flows of each component

Gasoline
26.3370

OUTLET STREAM from step 701 Liquid product storage

Flow = 26.3370 tonnes/day

Mass fraction

Gasoline
1.0000

Flows of each component

Gasoline
26.3370

BLUNT RESULTS SUMMARY.

Run date : 17.Dec.1994
Run identifier : BLUNT01

All costs are in 1993 Great Britain Pounds (Exchange rate with GB pounds = 1.0)

Product : Diesel
Conversion route : Ensyn RTP pyrolysis, hydrotreating and refining
Product output (t/day) : 25.9
Product yield (mass %) : 25.9

Production cost (tonnes/day) : 679.9
Capital cost (M GBP) : 12.1

BASED ON THE FOLLOWING INFORMATION

Feed data

Feedstock : Wood
Feed input (d.a.f. t/day) : 100.0
Feed moisture content (wt% dry) : 7.0
Feedstock cost (GBP/d.a.f. tonne) : 30.0

FINANCIAL BASE CASE

Financial variables

Project life (years) : 10.0
Operational days per year (days) : 330.0
Real interest rate (%) : 5.0
Inflation rate (%) : 5.0
Maintenance (% capital cost / year) : 5.0
Overheads (% capital cost / year) : 7.0

Utilities costs

LP steam (GBP/ tonne) : 6.0
Oxygen (GBP/ tonne) : 50.0
Cooling water (GBP/ tonne) : 0.05
Fuel gas (GBP/ GJ) : 4.0
Hydrogen (GBP/ tonne) : 360.0
Power (GBP/ MWh) : 50.0

CUSTOMISED STEP MODELS

Wood drying model

Moisture content entering (wt%) : 50.0
Moisture content exiting (wt%) : 7.0

Hydrotreating model

Deoxygenation extent (%) : 98.0
Fuel gas combustion efficiency (%) : 75.0
Coolant temperature rise (degC) : 20.0
Heat exchanger heat loss (%) : 10.0
Product outlet temperature (degC) : 73.0
Hydrogen source/off-gas disposal option : sell off-gas as fuel gas, import all H2

ANNUAL COST BREAKDOWN

(All costs are in thousands of GBP)

Capital amortisation	: 1540.40	(26.51 %)
Feedstock	: 990.00	(17.04 %)
Utilities	: 1087.40	(18.71 %)
Maintenance	: 606.40	(10.44 %)
Overheads	: 848.90	(14.61 %)
Labour	: 416.30	(7.16 %)
Additional	: 374.00	(6.44 %)
Byproduct disposal	: -52.60	(-0.91 %)

TOTAL ANNUAL COSTS : 5810.80 K GBP

CAPITAL COST BREAKDOWN

(All costs are in thousands of GBP on a 1993 basis.)

Step		
Wood reception	: 1095.80	(9.04 %)
Wood storage	: 916.30	(7.56 %)
Wood drying	: 1610.60	(13.28 %)
Wood grinding	: 604.90	(4.99 %)
Ensyn RTP pyrolysis	: 1701.90	(14.03 %)
Hydrotreating	: 6140.00	(50.63 %)
Refining	: 0.00	(0.00 %)
Liquid product storage	: 57.40	(0.47 %)

TOTAL CAPITAL COST : 12126.9 K GBP

UTILITIES USAGE SUMMARY

Step	LP steam t/day	Cooling water t/day	Fuel gas GJ/day	Hydrogen t/day	Oxygen t/day	Power MWh
Wood reception						0.20
Wood storage						0.00
Wood drying			334.80			0.10
Wood grinding						0.20
Ensyn RTP pyrolysis		575.00	159.80			0.40
Hydrotreating		2486.30	-748.80	11.40		0.50
Refining						
Liquid product storage						
TOTAL USAGE	0.00	3061.30	-254.20	11.40	0.00	1.40

ADDITIONAL INFORMATION

Byproduct streams

From the Ensyn RTP pyrolysis step : 8.0 tonnes/day of char

Waste streams

From the hydrotreating step : 40.3 tonnes/day of waste water contaminated with 0.0 tonnes/day of organics

Byproduct credits

From the Ensyn RTP pyrolysis step : 160.00 GBP/ day for char disposal

Additional costs

From the hydrotreating step : 600.0 GBP/day for catalyst rejuvenation

From the hydrotreating step : 0.0 GBP/day for waste water disposal

From the refining step : 533.2 GBP/day for refining upgraded product

BLUNT Simulation Mass Balance

OUTLET STREAM from step 101 Wood reception

Flow = 201.6129 tonnes/day

Mass fraction

Wood	Water	Ash
0.4960	0.5000	0.0040

Flows of each component

Wood	Water	Ash
100.00	100.8065	0.8065

OUTLET STREAM from step 111 Wood storage

Flow = 201.6129 tonnes/day

Mass fraction

Wood	Water	Ash
0.4960	0.5000	0.0040

Flows of each component

Wood	Water	Ash
100.0000	100.8065	0.8065

OUTLET STREAM from step 211 Drying

Flow = 108.3940 tonnes/day

Mass fraction

Wood	Water	Ash
0.9226	0.0700	0.0074

Flows of each component

Wood	Water	Ash
100.0000	7.5876	0.8065

OUTLET STREAM from step 221 Wood grinding

Flow = 108.3940 tonnes/day

Mass fraction

Wood	Water	Ash
0.9226	0.0700	0.0074

Flows of each component

Wood	Water	Ash
100.0000	7.5876	0.8065

OUTLET STREAM from step 303 Ensyn RTP pyrolysis

Flow = 81.2682 tonnes/day

Mass fraction

Pyr oil	Char
0.9855	0.0145

Flows of each component

Pyr oil	Char
80.0876	1.1806

OUTLET STREAM from step 401 Pyr oil hydrotreating

Flow = 28.7771 tonnes/day

Mass fraction

H/T oil	Water	Methane	CO2	C2+ H/c
0.9856	0.0000	0.0003	0.0017	0.0124

Flows of each component

H/T oil	Water	Methane	CO2	C2+ H/c
28.3625	0.0000	0.0085	0.0494	0.3568

OUTLET STREAM from step 601 Refining H/treated oil

Flow = 25.8994 tonnes/day

Mass fraction

Diesel
1.0000

Flows of each component

Diesel
25.8994

OUTLET STREAM from step 701 Liquid product storage

Flow = 25.8994 tonnes/day

Mass fraction

Diesel
1.0000

Flows of each component

Diesel
25.8994

Appendix C. Paper presented at the American Chemical Society Meeting, Denver USA March 1993

TECHNO-ECONOMIC MODELLING OF BIOMASS FLASH PYROLYSIS AND UPGRADING SYSTEMS.

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The thermochemical conversion of biomass by flash pyrolysis or liquefaction produces a crude liquid product that can be used directly to substitute for conventional fossil fuels or upgraded to a higher quality fuel. Both the crude and upgraded products may be utilised for power generation. A computer program is used to model the flash pyrolysis of biomass with subsequent upgrading and refining of the crude liquid products to produce higher value and more marketable liquid fuel products. The structure and scope of the program, the hydrotreating model and the development of an energy self-sufficient flash pyrolysis model are detailed. The program is used to assess and compare the economic and technical opportunities for upgrading crude pyrolysis liquids into the higher quality fuels demanded by more advanced technologies.

INTRODUCTION.

The techno-economic simulation program is being developed to carry out objective and consistent comparisons and evaluations of the alternative flash pyrolysis and upgrading technologies[1]. A complete integrated process from wood delivered to the factory gate through to an end product is modelled for capacities between 25 and 1000 tonnes/day dry ash free (d.a.f.) biomass.

The computer program works by building up a selected processing route from a number of process steps through which the material passes sequentially. The process steps modelled in a process route include the feedstock reception and storage followed by an appropriate combination of thermochemical conversion, upgrading of the crude liquid product, refining of the upgraded product, liquid product storage and electricity production from the crude or upgraded product. The processing routes modelled in the computer program are shown in Figure 1.

Each process step has a step model that calculates the mass and energy balances, the utilities usage and the capital cost for that step of the process. The results of the step models are combined to determine the performance of the whole conversion route. An executive program controls the selection of the feedstock, product and conversion route, the input and output of data from each step model, the passing of information between step models, the economic evaluation and summation of results. The step and component information required for the running of the program is stored in data files which are accessed by the executive program as required. The utilities requirements for each step model are handled by "importing" requirements or "exporting" surpluses into a utilities "pool", controlled by the executive program, where it is debited or credited respectively at the specified market price. The passage of data from one part of the program to another is illustrated in Figure 2.

THE GENERIC FLASH PYROLYSIS MODEL.

The generic flash pyrolysis model uses empirical relationships[1] for each product's yield, including reaction water, to calculate the mass balance. The empirical relationships are based on published results from the University of Waterloo[2,3,4,5] because their process is representative of flash pyrolysis and they possess the most comprehensive and consistent data set. The generic model is not intended to represent the University of Waterloo's process but to give a flexible tool that can demonstrate the effects of temperature and residence time on product yields, product quality (e.g. heating value and moisture content) and energy requirements. Other models have been included in the program to represent the Ensyn RTP[6], NREL vortex[7] and Fenosa (based on Waterloo) pyrolysis processes. The generic flash pyrolysis model shown in Figure 3 works within the limits given:

- pyrolysis reactor temperatures between 400 and 650°C;
- the yields of organics, reaction water, char and off-gas are determined by empirical relationships incorporating the reactor temperature and vapour residence time;
- the quantity of water in the final product is determined by the yield of reaction water and the moisture content of the feedstock (limited to any values between 0 and 30% as selected in the drying model);
- the quantity of ash/char in the liquid product is assumed to be 5% of the total produced by the process, the remainder being removed for fuel (which is comparable to the latest reported char contents of Ensyn RTP oils[6]);
- the energy requirement of the pyrolyser is met by burning the char, off-gas and imported fuel gas as explained later; and
- product oil and off-gas are cooled by a direct quench system, as favoured by Ensyn[6], Egemin[8] and Fenosa, with pyrolysis oil from previous runs.

Product yields.

To illustrate the results obtained from the program, Table 1 shows predicted product yields from the generic flash pyrolysis model for the case of 100 d.a.f. tonnes/day wood feed with a moisture content of 6.5% at a range of reaction temperatures.

From the predicted results it can be seen that the yields of organic liquids increase to a maximum around 500°C and then decrease as the temperature rises significantly above that. The yields of char decrease and the yields of gas increase with increasing temperature. At a pyrolysis temperature of 500°C the yield of crude, wet pyrolysis oil obtained is 80.4% (dry feed basis) which is close to the results reported by Ensyn for industrial RTP operation[6].

Table 1.
Predicted product yields from the flash pyrolysis of 100 d.a.f. tonnes/day of wood feed.
(Basis: 6.5% moisture content feedstock (wet basis))

Reactor temperature, °C	400	450	500	550	600	650
Derived product yields, (t/day)						
Organic liquid, dry	55.1	60.9	61.7	58.2	48.9	29.7
Total water	12.4	16.9	18.7	18.4	15.8	9.9
Char	33.1	20.5	13.8	10.0	7.6	6.0
Gas	5.9	8.2	12.4	20.0	34.2	60.9
Moisture content pyrolysis oil product (%)	18.4	21.8	23.3	24.0	24.4	25.0

Energy self-sufficiency.

One of the objectives of the generic flash pyrolysis model was that it should be as energy self-sufficient as possible. To this end the char and gas by-products are utilised to satisfy the pyrolyser energy requirements. The process modelled in the program utilises the char first because of it's high energy value, followed by off-gas and finally imported fuel gas. Surplus fuel is exported from the pyrolyser model for use in feed drying. The recoverable energy values for the off-gas assume that it is at 60°C. Typical energy values of 0.0015 GJ m^{-3} and 30 GJ/tonne have been assumed for the off-gas and char respectively.

Figure 4 shows energy balances over the pyrolyser for feed moisture contents between 5 and 25 % (wet basis) for temperatures between 450 and 650 °C. The balances were derived using the energy requirements of the pyrolyser and the energy recoverable from the char and off-gas at 75% combustion efficiency. All of the values shown are in GJ/tonne d.a.f. feed and are based on the yields shown in Table 1.

The figure demonstrates quite clearly the temperatures below which the pyrolysis model is energy self-sufficient. For example, for a 10% moisture content feedstock (wet basis) the utilisation of the char and off-gas will be sufficient to satisfy the energy demand for temperature below 600°C. For an extreme case of a 25% moisture content feedstock (wet basis), the energy demand can be satisfied at temperatures below 580°C. In all cases where the energy surplus is negative fuel gas must be imported from the utilities to satisfy the energy demand. The energy balances follow an expected trend - there is an energy surplus in the lower temperature region where the char yields are relatively high (as demonstrated in Table 1) and a deficit in the high temperature region where the char yields are low. Although the off-gas yields increase with an increase in temperature (as demonstrated in Table 1), the off-gas has a lower energy value than the char and contributes less to the overall energy balance.

Figure 5 shows the maximum pyrolyser temperatures possible for energy self-sufficiency for char and off-gas combustion efficiencies between 20 and 100% and feed moisture contents between 5 and 25% (wet basis). The graph was derived using the energy requirements of the pyrolyser and the energy recoverable from the char and off-gas at a range of pyrolyser temperatures, char/gas combustion efficiencies and feedstock moisture contents. Char/off-gas combustion efficiencies of below 50% could represent a system with inefficient combustion or indirect heat transfer e.g. a system heated indirectly with external fuel combustion. Although a 100% combustion efficiency is perhaps unrealistic, a figure close to it would be possible with in-situ "combustion" where the heat losses are minimal.

For example, using a 5% moisture content (wet basis) feedstock at a char/off-gas combustion efficiencies of 70% the maximum temperature possible for energy self-sufficiency is 580°C. By increasing char/gas combustion efficiencies to 80% the maximum temperature shifts to 620°C. For a 25% moisture content (wet basis) feedstock at char/gas combustion efficiencies of 80% the maximum possible temperature for energy self-sufficiency is 550°C. This can be increased to 565°C if a 20% moisture content (wet basis) feedstock is utilised. Figure 5 demonstrates that the maximum pyrolysis temperature possible for energy self-sufficiency can be increased by improving the char/off-gas combustion efficiency or decreasing the feedstock moisture content.

Residence time.

The yield and quality of the products from flash pyrolysis are affected by the vapour residence time and the reaction temperature. The data utilised to determine the product yields includes a residence time effect but temperature is the only variable. Some data is available on the interaction between residence time, temperature and product yields[9] but is not directly applicable to the model as it is for high temperature gas production. At the present time there is insufficient data for the temperature range utilised in the model for residence time to be included as a separate variable.

CHO analysis.

It is necessary to be able to relate the products from flash pyrolysis to the feed specifications for the hydrotreating model if the models are to be used in a processing route. The methodology used is to determine the CHO analysis of the flash pyrolysis product and to use this in conjunction with the severity of the hydrotreating deoxygenation process to determine the oxygen content of the hydrotreated product. The extent of deoxygenation also determines the amount of separation of the hydrotreated product from the reaction water and the quantity of reaction water produced. The CHO of the flash pyrolysis product can be related to that of the respective feedstock using algorithms derived from data collated from published sources. The following provisional correlations have been derived for C/H, C/O and C/N.

$$\text{Pyrolysis oil C/H} = -9.7950 + 2.0099 * (\text{Feed C/H}) \qquad r^2 = 0.912 \qquad \{1\}$$

$$\text{Pyrolysis oil C/O} = -0.65441 + 1.6146 * (\text{Feed C/O}) \qquad r^2 = 0.945 \qquad \{2\}$$

$$\text{Pyrolysis oil C/N} = -283.47 + 9.0906 * (\text{Feed C/N}) \qquad r^2 = 0.9 \qquad \{3\}$$

THE HYDROTREATING MODEL.

Upgrading of crude pyrolysis oil by hydrotreatment produces a more stable product which is a naphtha like hydrocarbon with properties similar to light distillate or kerosene. Hydrotreating has been successfully demonstrated at laboratory scale for crude pyrolysis oils[10,11]. Battelle PNL was commissioned to design and cost a hydrotreating pilot plant[12] for Aston University based on their early work on the catalytic hydrotreatment of biomass pyrolysis oil. This design is being utilised to develop the hydrotreating model in the program.

Figure 6 shows a schematic flow sheet of the hydrotreating process for the following model design:

- a two-stage up-flow reactor, the first stage at 250-300°C for stabilisation and the second at 350-400°C for hydrotreating;

- the main variable in the model is the deoxygenation extent of the crude pyrolysis oil feedstock, controlling:
  - the hydrogen requirement for the process and hence the operating cost,
  - the liquid hourly space velocity and hence the reactor size and capital cost.
  - oxygen content of the hydrotreated product and hence quantities of all products produced and quality of the final product;

- user determined choice of hydrogen sources and off-gas disposal options including:
  a. the hydrogen rich off-gas is sold as fuel gas to the utilities and all of the required hydrogen is imported from the utilities. This process has a high utilities cost because all the hydrogen has to be bought from the utilities but no additional capital cost.
  b. hydrogen is produced by separation of the hydrogen rich off-gas using for example a membrane and the remaining off-gas is sold as fuel gas to the utilities. The amount of additional hydrogen required from the utilities to make up the deficit depends on the efficiency of the membrane separation. This process involves a potential reduction in the utilities costs because less hydrogen is imported from the utilities than in option (a) but a higher capital expenditure.
  c. the hydrogen rich off-gas is reformed and separated to recover as much as possible of the hydrogen. Any deficit is made up from hydrogen imported from the utilities or fuel gas imported from the utilities which is reformed and separated. This option although similar to (b) will incur a high capital cost from the steam reforming equipment but will however provide a higher hydrogen recovery.

Data on the effect of the severity of the deoxygenation process on liquid hourly space velocity, water separation and product quality has been obtained from Veba Oel who conducted hydrotreating trials on Ensyn RTP oil[13]. Battelle PNL have also supplied data on the hydrotreated product oil oxygen content and liquid hourly space velocity.

The extent of the crude pyrolysis oil hydrotreatment within the hydrotreating model is user determined and could be either full or partial. Partial hydrotreatment would involve partial removal of the oxygen present in the crude pyrolysis oil to give a more stable, higher quality product which could be more readily substituted for conventional fuels. However, the process is unlikely to be cost effective as the hydrotreating equipment is still required. The product would also be of lower quality than fully hydrotreated oil due to the oxygen content reducing heating values, apart from any unknown adverse physical and chemical properties. The choice of whether to partially or fully hydrotreat the crude bio-oil depends on the application for which the upgraded product is destined with a trade off between high oxygen contents but lower costs and low or zero oxygen contents and high costs respectively.

SYSTEM PERFORMANCE AND COST.

Mass balances and production cost estimates for upgrading by hydrotreating are given in Table 2. The costs are based on chipped wood delivered to the factory gate and include all costs for a complete process including reception, storage, handling, pre-treatment, conversion, upgrading and product storage.

The hydrotreating process option used to generate the results in Table 2 is for a hydrogen self-sufficient system using separation and steam reforming of the off-gases. From using such a process there is a potential reduction in the utilities costs but an increase in the capital expenditure. The trade off between reforming (high capital and low operational costs) will be compared in the model with importing all required hydrogen (high hydrogen and minimal capital costs) and recovering all surplus hydrogen and importing the hydrogen deficit (medium capital and operating costs).

The incremental production cost of diesel is provisionally calculated by taking a refining service charge of 35 US $/tonne of refined product produced[14], assuming a 10% loss of product during refining.

Table 2.
Predicted production costs and yields for the products from upgrading biomass pyrolysis oil by hydrotreating.
(Basis: 1000 tonnes/day 7% moisture content wood feed (wet basis) at 43.2 $/tonne.)

	Yields t/day	$/t product	$/GJ product
Wood feed d.a.f.	1000	43.2	2.2
Crude wet pyrolysis oil (produced at 500°C)	804	78.4	4.1
Hydrotreating products			
Bio-naphtha (hydrotreated bio-oil, 98% deoxygenated)	316	213.5	5.1
Diesel (refined bio-naphtha)	284	272.2	6.2

Figure 7 shows the effect of feedstock cost variation between 0 and 100 $/d.a.f. tonne on the production costs of the crude, upgraded and refined products from hydrotreating crude pyrolysis oil. The bio-fuel cost lines on the graph are not parallel because the products are of different qualities. The incremental cost of upgrading based on a hydrotreated oil product cannot be simply added to the pyrolysis oil cost as the bases are different. The same applies to refining although the effect is less noticeable as the products are of similar quality.

Figure 7 also compares the production costs of the biomass derived fuels with their fossil fuel equivalents. Crude pyrolysis oil is compared with medium fuel oil, bio-naphtha (98% hydrotreated pyrolysis oil) with fossil fuel derived naphtha and diesel (refined bio-naphtha) with fossil fuel derived diesel. From the figure it can be seen that crude pyrolysis oil can support a feedstock cost of 24 \$/d.a.f. tonne when compared to medium fuel oil, bio-naphtha can support 31 \$/d.a.f. tonne when compared to fossil fuel naphtha and diesel (refined bio-naphtha) can support 46 \$/d.a.f. tonne when compared to fossil fuel derived diesel. The figure shows that the cost of upgrading is more than justified by the additional value of the product.

CONCLUSIONS.

The results obtained from the flash pyrolysis model indicate that an energy self-sufficient system is feasible utilising the char and off-gas as fuels for combustion efficiencies of 75% for feed moisture contents between 5 and 25% (wet basis) for pyrolyser temperatures below 540°C. In addition, the maximum pyrolysis temperature possible for energy self-sufficiency can be increased by improving the char/off-gas combustion efficiency or decreasing the feedstock moisture content or both.

The performance and cost estimates of the crude, upgraded and refined biomass derived products compared with their equivalent fossil fuel derived products show that they are extremely sensitive to feedstock cost fluctuations and are only economic at feedstock costs below 24 \$/d.a.f. tonne for crude pyrolysis oil, 31 \$/d.a.f. tonne for bio-naphtha and 46 \$/d.a.f. tonne for refined bio-naphtha. Biomass derived fuels could be made more economic by utilising waste materials with their low inherent cost as feedstocks.

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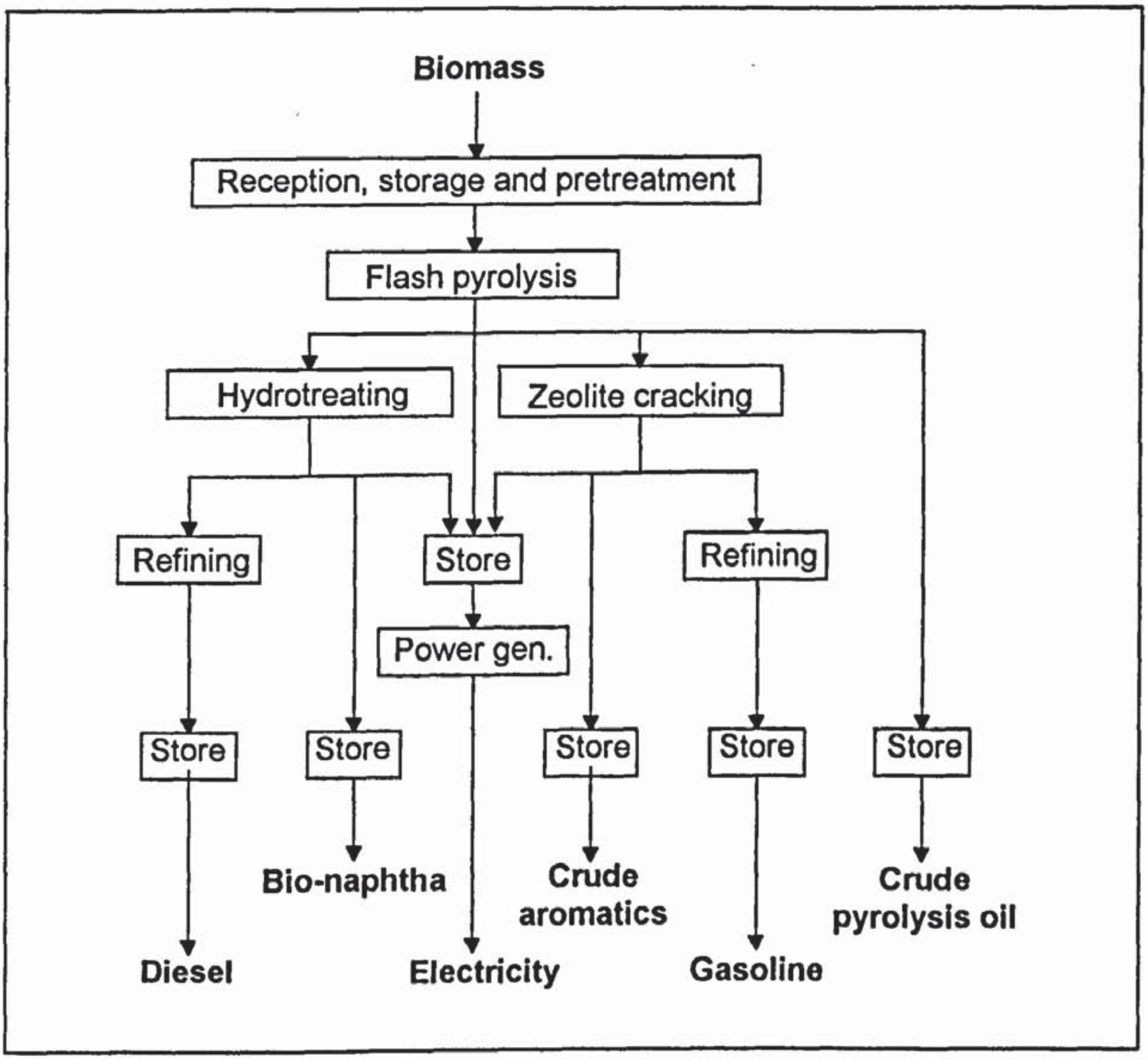
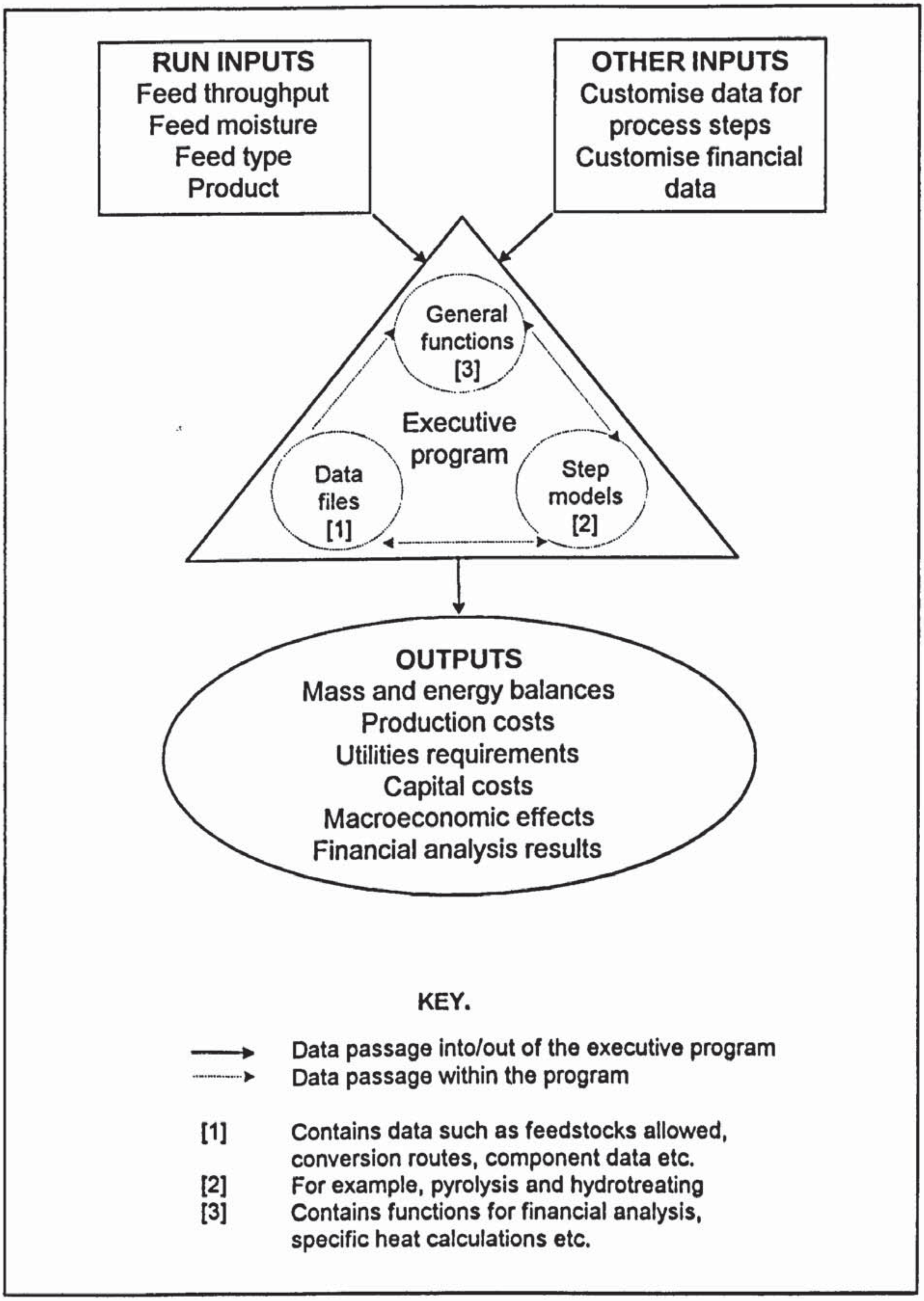


Figure 1.
Schematic diagram to show the conversion routes modelled in the program.



KEY.

→ Data passage into/out of the executive program
⇢ Data passage within the program

[1] Contains data such as feedstocks allowed, conversion routes, component data etc.
[2] For example, pyrolysis and hydrotreating
[3] Contains functions for financial analysis, specific heat calculations etc.

Figure 2.
Schematic diagram to illustrate the passage of data within the computer program.

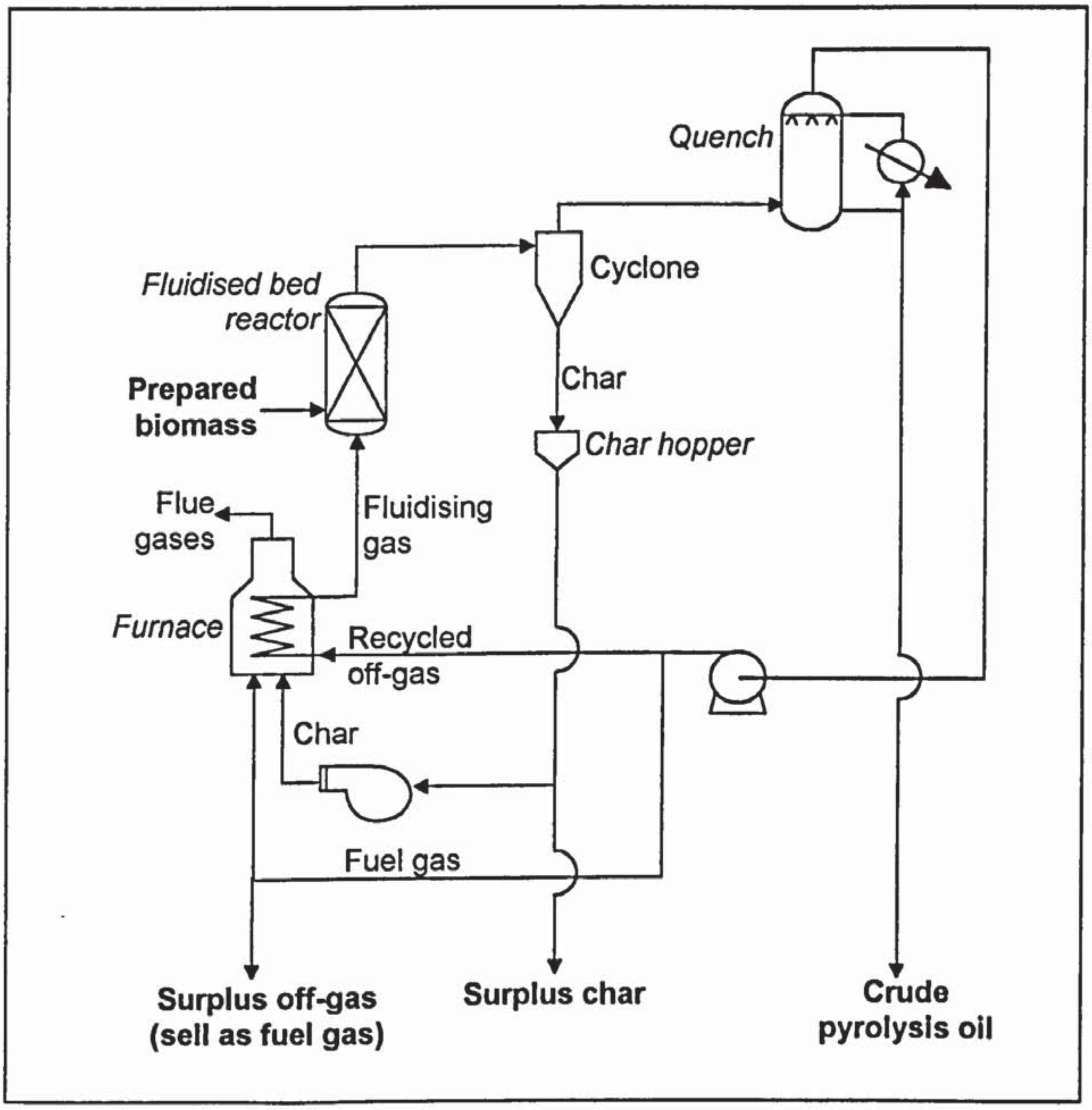


Figure 3.
Simple flowsheet for the generic flash pyrolysis model.

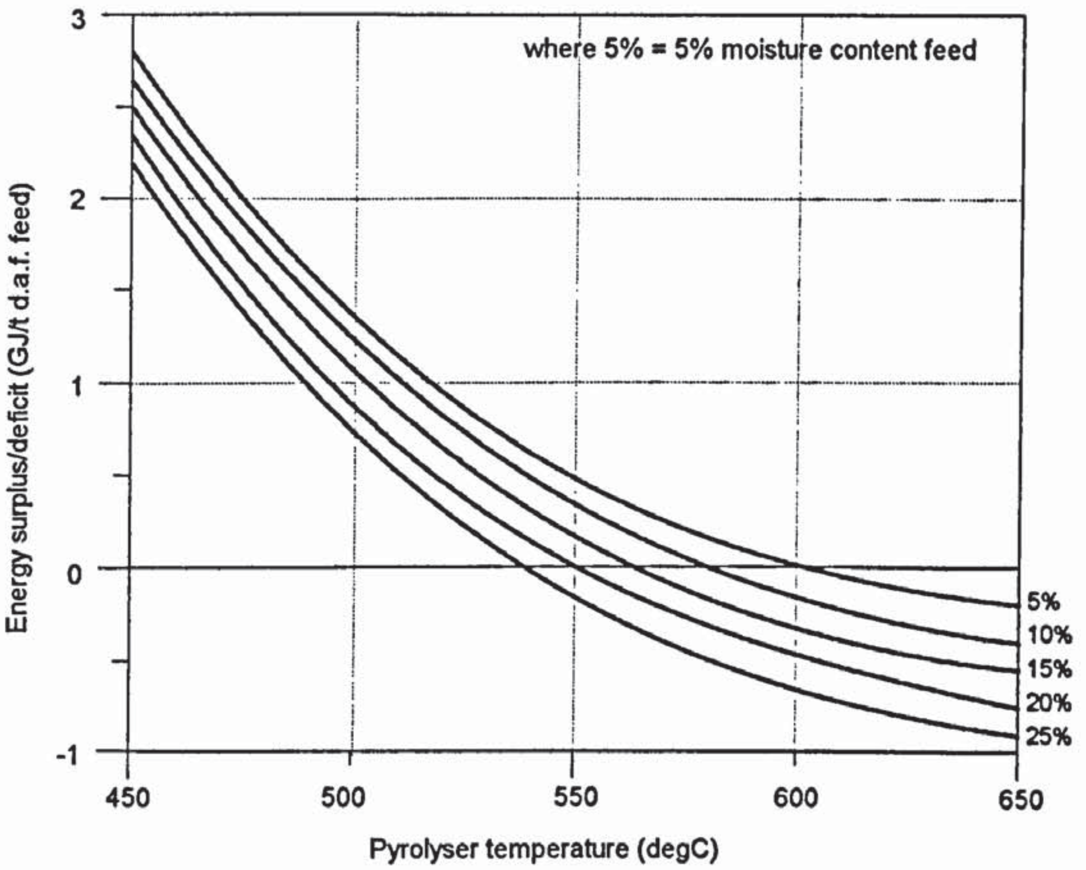


Figure 4. Energy balance for the flash pyrolysis model for 75% char/off-gas combustion efficiencies and feed moisture contents between 5 and 25% (wet basis).

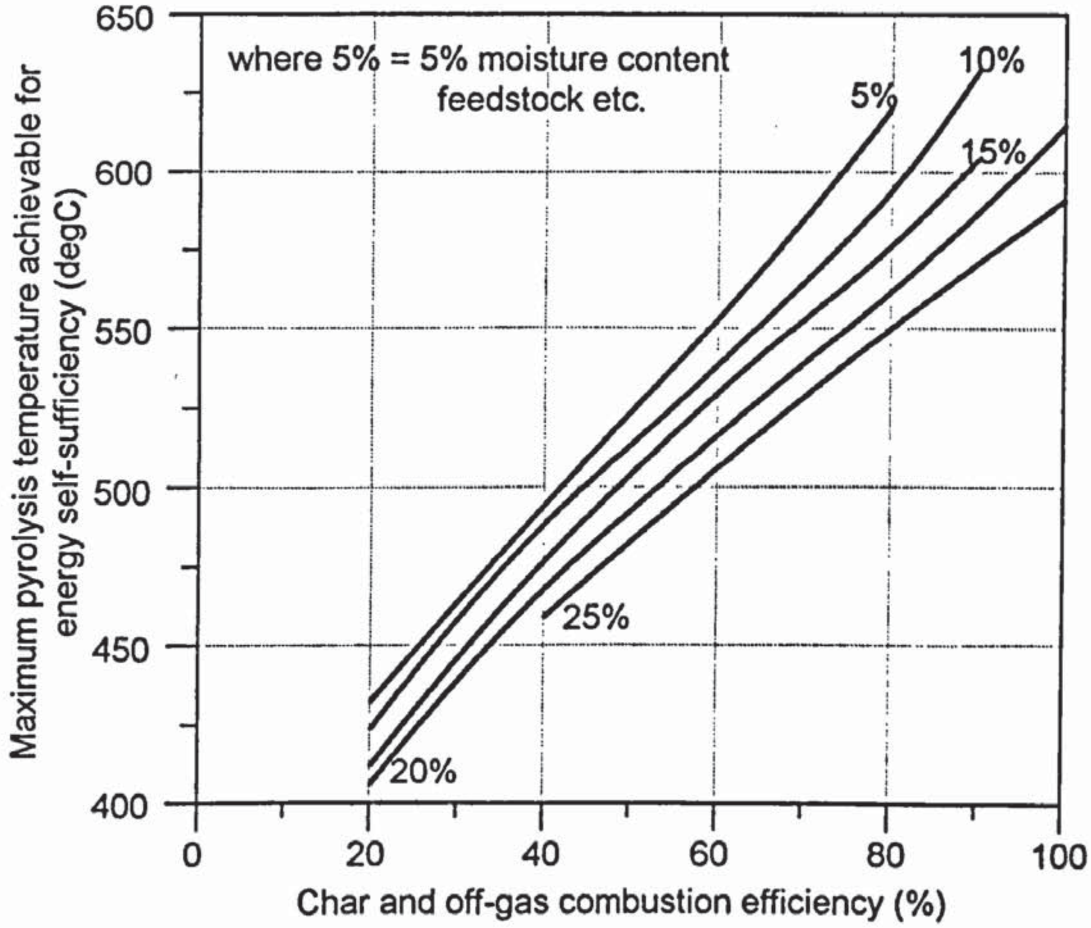


Figure 5.
The effect of char and off-gas combustion efficiencies on the maximum pyrolysis temperature achievable for energy self-sufficiency for feed moisture contents between 5 and 25% (wet basis).

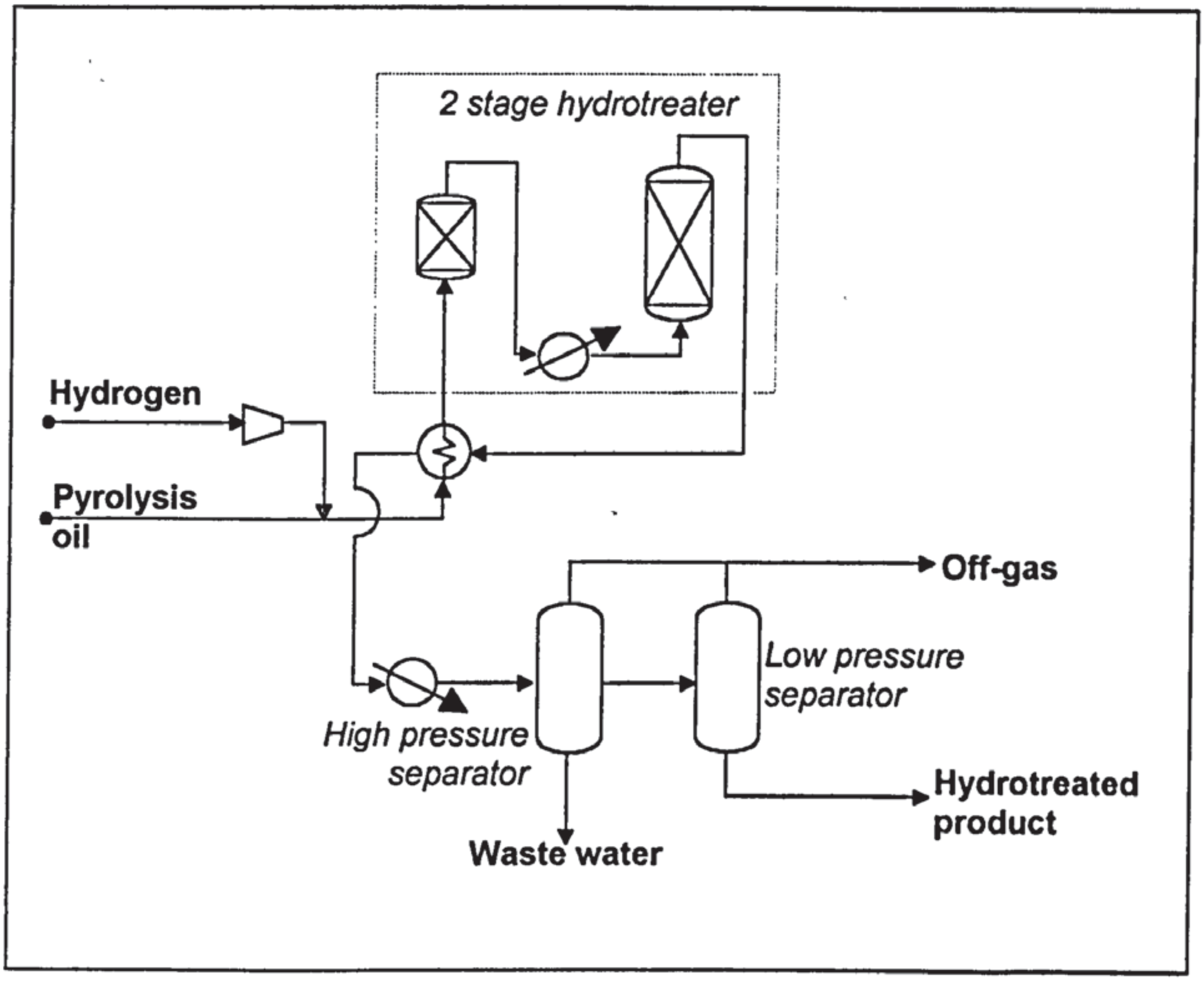


Figure 6.
Simple flowsheet for the hydrotreating model.

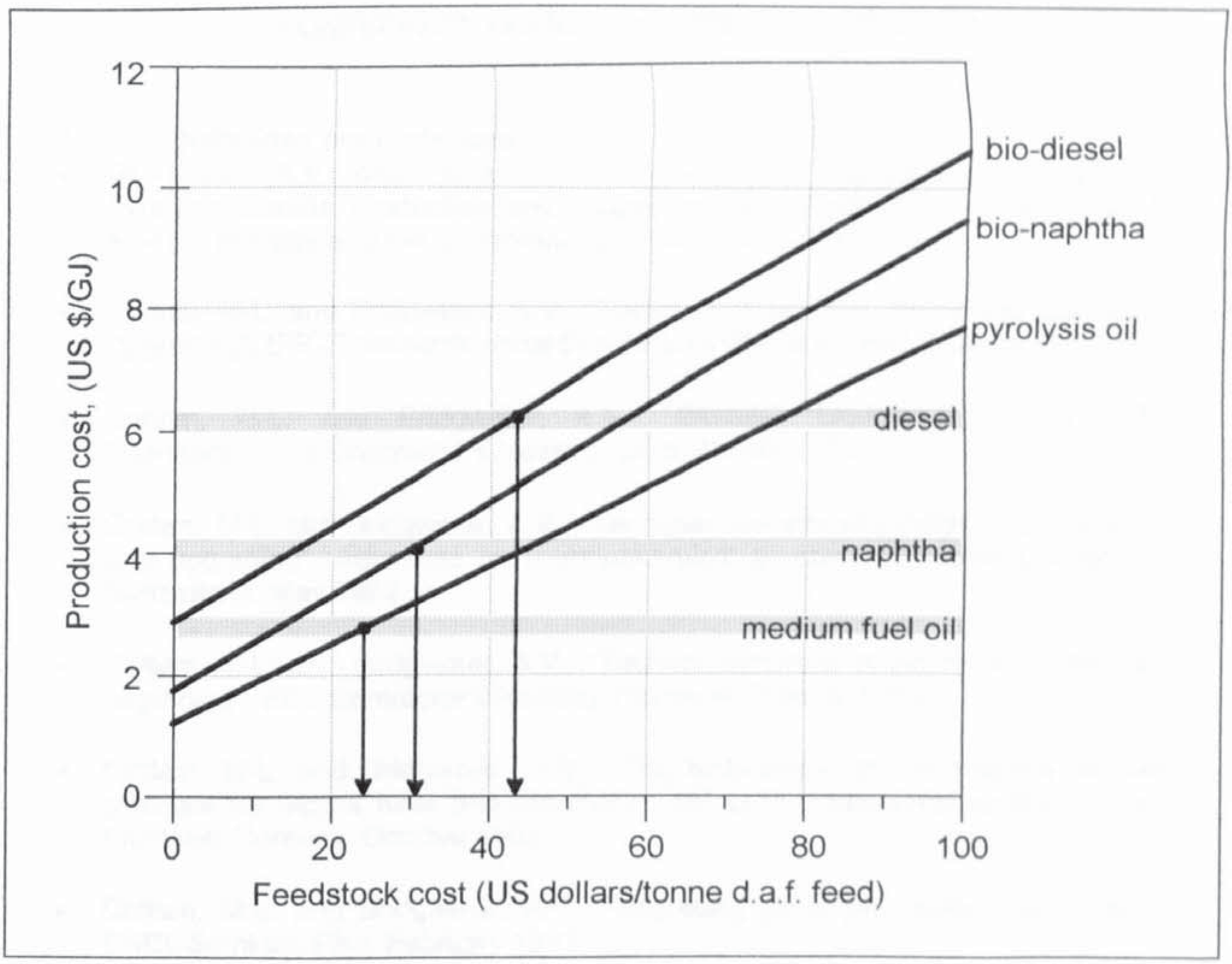


Figure 7.
Predicted production costs for the products from upgrading biomass pyrolysis oil by hydrotreatment.

Appendix D. List of Authors published papers and presentations.

AUTHORS CONFERENCE PRESENTATIONS AND PUBLISHED PAPERS.

1. Conference presentations.

- Bridgwater, A.V. and Cottam, M-L., "Costs and Opportunities for Biomass Pyrolysis Liquids Production and Upgrading", 6th Conference on Biomass for Energy, Industry and the Environment, Athens, April 1991.

- Cottam, M-L. and Bridgwater, A.V., "Technoeconomics of Bio-oil Production and Upgrading", EEC Thermochemical Contractor's Meeting, Gent, October 1991.

- Cottam, M-L. and Bridgwater, A.V., "Biomass Liquids Upgrading", EEC Thermochemical Contractor's Meeting, Gent, October 1991.

- Cottam, M-L. and Bridgwater, A.V., "Technoeconomics of pyrolysis oil production and upgrading", Advances in Thermochemical Biomass Conversion, Interlaken Switzerland, May 1992.

- Cottam, M-L. and Bridgwater, A.V., "Technoeconomics of bio-oil production and upgrading", EEC Contractor's Meeting, Florence, October 1992.

- Cottam, M-L. and Bridgwater, A.V., "The technology and economics of flash pyrolysis for liquids fuels and electricity", 7th EEC Conference on Energy from Biomass, Florence, October 1992.

- Cottam, M-L. and Bridgwater, A.V., "Upgrading pyrolysis liquids - economics", ENEL Seminar, Pisa, February 1993.

- Cottam, M-L. and Bridgwater, A.V., "Techno-economic modelling of biomass flash pyrolysis and upgrading systems", American Chemical Society Annual Meeting - Fast Pyrolysis Symposium, Denver USA, March 1993.

- Cottam, M-L. and Bridgwater, A.V., "Techno-economic assessment of thermochemical liquid fuels production systems and modelling of pyrolysis processes", CEC Joule Contractor's Meeting, Athens Greece, June 1993.

2. Published papers.

- Bridgwater, A.V. and Cottam, M-L, "Opportunities for Biomass Pyrolysis Liquids Production and Upgrading", pp113-120 in Energy and Fuels, Vol. 6 No. 2 (1992).

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- Cottam, M-L. and Bridgwater, A.V., "Technoeconomics of pyrolysis oil production and upgrading", pp1343-1357 in Advances in Thermochemical Biomass Conversion. Eds., Bridgwater, A.V. (Blackie, 1993).

- Cottam, M-L. and Bridgwater, A.V., "Process simulation and techno-economic modelling of biomass thermal liquefaction and upgrading processes", pp927-929 in Proceedings of 1994 IChemE Research Event (IChemE, 1994).

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