SUSTAINABLE ENERGY FROM PAPER INDUSTRY WASTES

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THESIS SUMMARY

Secondary fibre paper mills are significant users of both heat and electricity which is mainly derived from the combustion of fossil fuels. The cost of producing this energy is increasing year upon year. These mills are also significant producers of fibrous sludge and reject waste material which can contain high amounts of useful energy. Currently the majority of these waste fractions are disposed of by landfill, land-spread or incineration using natural gas. These disposal methods not only present environmental problems but are also very costly.

The focus of this work was to utilise the waste fractions produced at secondary fibre paper mills for the on-site production of combined heat and power (CHP) using advanced thermal conversion methods (gasification and pyrolysis), well suited to relatively small scales of throughput. The heat and power can either be used on-site or exported.

The first stage of the work was the development of methods to condition selected paper industry wastes to enable thermal conversion. This stage required detailed characterisation of the waste streams in terms of proximate and ultimate analysis and heat content. Suitable methods to dry and condition the wastes in preparation for thermal conversion were also explored.

Through trials at pilot scale with both fixed bed downdraft gasification and intermediate pyrolysis systems, the energy recovered from selected wastes and waste blends in the form of product gas and pyrolysis products was quantified.

The optimal process routes were selected based on the experimental results, and implementation studies were carried out at the selected candidate mills. The studies consider the pre-processing of the wastes, thermal conversion, and full integration of the energy products.

The final stage of work was an economic analysis to quantify economic gain, return on investment and environmental benefits from the proposed processes.

Keywords; Gasification, Pyrolysis, De-inking sludge, Rejects

Bismillah (In the name of Allah)

The most beneficent, the most merciful.

All praise and thanks be to Allah, the Lord of all creation.

The only owner and only ruling judge of the day of recompense.

You alone we worship, and you alone we ask for help.

Guide us to the straight way, the way of those on whom you have bestowed your grace,

not the way of those who have earned your anger, nor of those who have gone astray.

[Holy Quran 1:1]

May Allah's peace and blessings be on the prophet Mohammed

To my family

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LIST OF ABBREVIATIONS

AN	Aylesford Newsprint Ltd (Kent UK)
ASTM	American Standard Test Method
BCFC	Break Specific Fuel Consumption
BD	Biodiesel
CASE	Co-operative Award in Science and Engineering
CEN	European Committee for Standardisation
CHP	Combined Heat and Power
CI	Compression Ignition
DAF	Dissolved Air Flotation
DFM	Digital Flow Meter
DRX	X-Ray Diffraction
DSPO	De-inking Sludge Pyrolysis Oil
DSPO-BD	De-inking Sludge Pyrolysis Oil - Biodiesel
DSPO-FD	De-inking Sludge Pyrolysis Oil - Fossil Diesel
DTG	Derivative Weight Loss
EBRI	European Bioenergy Research Institute
ECN	Energy Research Centre Netherlands
ECU	European Currency Unit
EPSRC	Engineering and Physical Science Research Council
FC	Fixed Capital
FD	Fossil Diesel
FTIR	Fourier Transform Infrared Spectroscopy
FW	Food Waste
GC	Gas Chromatograph
GC-MS	Gas Chromatography Mass Spectrometry
GC-TCD	Gas Chromatography Thermal Conductivity Detector
GCU	Gasifier Control Unit
GEK	Gasifier Experimenters Kit
HDPE	High Density Polyethylene
HHV	Gross Heating Value
IC	Internal Combustion
ICP	Inductively Coupled Plasma
IR	Infrared
KC	Kimberly Clark (Flint UK)
LEFR	Laminar Entrained Flow Reactor
MSW	Municipal Solid Waste
OCC	Old Corrugated Containers
ONP	Old Newsprint Paper
OC	Operating Cost
PCE	Equipment Total Purchase Cost
PET	Polyethylene Terephthalate
PPC	Total Physical Plant Cost
PPM	Parts Per Million
PVC	Polyvinyl Chloride
Py-GC/MS	Pyrolysis - Gas Chromatography Mass Spectrometry
RDF	Refuse Derived Fuel
RHI's	Renewable Heat Incentives
RPF	Residue Plastic Fuel
ROC's	Renewable Obligation Certificates
SEM	Scanning Electron Microscopy
SOFC	Solid Oxide Fuel Cell

SK	Smurfit Kappa
SSK	Smurfit Kappa (Birmingham UK)
TCC	Total Capital Cost
TGA	Thermal Gravimetric Analysis
WC	Working Capital

1.0 INTRODUCTION

Since the industrial revolution, the world has been consuming ever-increasing amounts of energy from fossil fuels. This is now placing immense pressure on our planet's natural resources and the CO_2 produced from the combustion of fossil fuels is widely believed to be the principal cause of climate change. Furthermore, there are increasing concerns over the price and security of the supply of petroleum and natural gas. Therefore any attempt to decrease the demand on fossil fuels is considered to be environmentally beneficial.

The paper industry is a significant user of energy in the form of electricity and heat to power machinery and to dry paper sheets. According to the UK department of Energy and Climate Change [1], in 2008 the pulp and paper industry within the UK was ranked 19th out of the 93 highest fuel consuming industrial sectors, and collectively consumed a crude oil equivalent of 1.8 million tonnes annually. Furthermore the total energy input at UK based paper mills accounts for approximately 13 % of all operational costs [2]. As the cost for producing this energy increases year upon year, many UK based paper mills are finding it increasingly difficult to remain profitable, and this has led to the closure of lower tonnage operations that manufacture commodity grade paper and board products. It is extremely difficult to reduce the high energy demand from this industry without compromising the rate of production or quality of the paper making process itself. Therefore there is increasing interest in alternative lower cost methods of providing the energy.

In recent years there have been significant efforts by paper manufacturers to make their process more sustainable by increasing the use of recycled paper and board (secondary fibre) as a feedstock. The amount of recovered paper used by UK paper mills as a proportion of their total output stands at approximately 72 %, which is a figure believed to be close to the maximum achievable [2]. However, the use of recovered feedstocks results in large volumes of wastes such as rejected plastics, fibres and other coarse materials, "stickies" (adhesive residues), and in the case of some mills de-inking sludge (fibres, minerals and ink). These have hitherto been disposed of by landfill, land-spreading or incineration (combustion).

Landfilling waste in the UK is fast becoming an unacceptable method of disposal and in some countries such as Germany and the Netherlands the disposing of paper mill wastes in this way has already been prohibited [3]. Furthermore landfilling costs are increasing year on year, making this disposal method less feasible economically [3]. Some paper mills have tried to reduce this cost by either paying farmers to land-spread a proportion of their sludge or by supplying it to smaller firms for further processing to become cattle bedding. However it is possible that pressure may be brought to bear from government and industry to reduce these practices in future.

If UK paper mills wish to further improve the sustainability of their process, then a cost effective waste management strategy that reduces dependency on fossil fuels and environmental impact must be implemented. One possibility is to utilise those waste streams that have a calorific value as a source of energy for generating combined heat and power (CHP), to contribute to the demands of the mill. Of particular interest would be the possibility of using advanced thermal conversion technologies such as gasification and pyrolysis, to take advantage of their enhanced efficiency and environmental performance, and also their appropriate scale in the context of UK mill operations.

1.1 Project scope

This project was funded by the Engineering and Physical Science Research Council (EPSRC) and was a Co-operative Award in Science and Engineering (CASE) Studentship in collaboration with Smithers Pira Ltd., and three UK secondary fibre paper mills, (i) Smurfit Kappa Birmingham (SSK), (ii) Aylesford Newsprint Ltd (AN), and (iii) Kimberly Clark Flint (KC).

The scope of this work, determined jointly by Aston University, the funding bodies and the industrial partners, was the investigation of the possibility of utilising the wastes produced by each of the secondary fibre mills as an energy source in a site-based combined heat and power (CHP) system based on the advanced thermal conversion technologies pyrolysis and gasification. The work includes both experimental investigation (such as feedstock characterisation and conversion system performance) and design analysis (such as system techno-economic performance and environmental impact). Further details on the specific project aims and objectives is presented in Chapter 4.

2.0 BACKGROUND

2.1 Chapter summary

This chapter gives a general overview of the three paper mills that have been investigated during the course of this project: (i) the Smurfit Kappa mill at Birmingham, Nechells, West Midlands, UK (referred to henceforth as SSK) which is a secondary fibre brown paper mill which manufactures brown paper sheets for cardboard production; (ii) the Aylesford Newsprint Ltd mill at Aylesford, Kent, UK (referred to henceforth as AN) a secondary fibre newsprint mill that uses recycled newsprint to manufacture new low grade newsprint paper; and (iii) the Kimberly Clark mill at Flint, Clwyd, UK (referred to henceforth as KC), a tissue mill which recycles waste office grade paper for the manufacture of soft and hard tissue paper and wet cleansing wipes. A general description of each manufacturing process is given along with the general types and quantities of wastes each mill produces and the locations within their processes from where these waste fractions are created. The chapter also identifies and justifies both fixed bed downdraft gasification and intermediate pyrolysis as the two advanced thermal conversion techniques which have been explored to process these wastes for energy recovery.

2.2 Background to the paper industry

The word “paper“ comes from the latin “papyrus” and the early Egyptians were its inventors. Paper has played an important role throughout history, cataloguing and recording important findings, research, events, historic and religious texts [4].

Modern paper, was developed by the Chinese around 105 AD. From China the knowledge of paper production travelled westwards and spread to the Middle East, North Africa and then onwards to Europe. By the end of the 18th century there was more correspondence through letters and newspapers and at the same time education began to grow increasing the demand for more books and writing materials. This led to further demand for paper and thus more paper mills began to spring up throughout Europe. At the end of the 18th century there were 416 registered paper mills in England and Wales, 49 in Scotland and 60 in Ireland [4]. All the paper was made by hand so, although the quality was usually high, output was low. Therefore, attempts were made to replace the hand craft by machinery. The first paper machine was invented in Holland in 1680 and towards the end of the 19th century paper production methods using wood pulp were developed [4]. Improvements to wires and felts led to higher machine speeds and improved efficiencies.

Peak employment within the paper industry was reached in 1959, when 100,000 people were employed [4]. Since then, as the industry has moved more towards internet and email based technologies the demand for paper has decreased therefore employment within this industry has declined along with the number of mills.

In 1960 UK paper consumption exceeded 4 million tonnes per year. By 2010 in the UK, there were 49 mills producing an estimated 4.3 million tonnes of paper and board. There are currently approximately 9,400 people employed in the UK paper and board manufacturing industry [4].

The global pulp and paper industry in 2012 was estimated to be worth 575 billion US dollars, with the UK share estimated to be worth around 21 billion US dollars [4].

2.3 Secondary fibre paper mills

Secondary fibre paper mills utilise recycled paper as a feedstock for the production of new paper sheets. The three paper mills which have been explored in this study include a secondary fibre board mill that produces brown paper sheets for the production of cardboard (SSK), a secondary fibre newsprint mill that produces low grade newsprint paper (AN) and a secondary fibre tissue mill which manufacturers tissue products from recycled office grade paper (KC). The paper manufacturing process at each mill employs essentially similar processing steps, although there exist two main aspects that differentiate each process.

The first aspect is the initial quality and type of the recovered paper. A board mill will only receive recovered fibre comprised of old corrugated containers (OCC) with lower amounts of contaminants, mainly stickies and plastic film (soft plastics), and unusable short length paper fibres [84]. A tissue mill will receive mainly office grade paper and a newsprint mill will receive a much broader range of waste paper grades such as old newsprint paper (ONP), magazines, office grade paper and packaging paper [85, 86]. Accepting a wider range of feedstock paper in this way means that the manufacturing process at a newsprint mill usually requires more rigorous paper conditioning to that at a board mill or tissue mill, as the removal of higher amounts of adhesives and stickie residues is required [6]. Similarly as the recovered fibre is mainly recycled by households, there is a higher potential for it to contain a broader range of other more generic contaminants. For example a newsprint mill will often see increased levels of glass, metals, textiles, mixed plastics (soft and hard), and other general household and business wastes (rubber, staples, paper clips, cardboard etc.) and the composition of these wastes will vary from day to day [86].

The second aspect is that newsprint and tissue paper production requires extra de-inking and bleaching processing steps to remove ink, increase the brightness and improve the overall appearance of the final paper or tissue product [85, 86]. These steps are not necessary for packaging grade paper and so are not implemented by board mills [84].

2.3.1 An overview of a board mill's paper manufacturing process

The board mill explored in this study (SSK) produces approximately 192,000 tonnes per year of packaging grade paper which is sent to corrugators to produce cardboard boxes [84]. Once produced the paper is of dark brown appearance, and is multi-layered for increased strength and flexibility. The multi-layering of this paper is applied by producing a bottom ply and top ply, and the manufacturing process to produce each ply is almost identical and the process is carried out using duplicate machinery [84].

The full stock preparation flow system of a board mill is shown in Figures 1- 4. The first stage of the brown paper production process begins with the initial recovered fibres being brought into the mill by the lorry load. The feedstock is mainly comprised of OCC grade cardboard and is usually brought into the mill by local authorities and other contracted companies. The feedstock usually comprises large tonnages of mixed waste cardboard along with lower amounts of other contaminant materials, mainly stickies, soft plastics and unusable fibres. The feedstock is stored at the mill's feedstock holding bay until further processing.

Figure 1 A board mill’s stock preparation process flow sheet for the production of bottom ply thick stock [courtesy of SSK] [84]

Figure 2 A board mill's stock preparation process flow sheet for the production of bottom ply thin stock [courtesy of SSK] [84]

Figure 3 A board mill's stock preparation process flow sheet for the production of top ply thick stock [courtesy of SSK] [84]

Illustration removed for copyright restrictions

Figure 4 A board mill's stock preparation process flow sheet for the production of bottom ply thin stock [courtesy of SSK] [84]

Key to Figures 1-4

Name	**Tag**
Base ply pulper	PU41
Top ply pulper	PU42
Base ply fiberiser	FIB 1&2
Top ply fiberiser	F1
Base ply drum screens	4DRB1&2
Top ply drum screens	DRT
Base ply fibre sorter	F4B1&2
Top ply fibre sorter	4F3
Base ply reject sorter	RS2B1&2
Top ply ominifractor	OS8
Omega screen base ply	BOS
Omega screen top ply	TOS
Base ply screen	SCB3
Base ply tampella screen	SCB
Top ply tampella screen	SCT
Dilution water screen base ply	SCB4

The next series of processing steps is given the term “stock preparation” or “approach flow system” and this is where paper pulp is formed and conditioned before it is transferred to the paper making machine [5]. The stock preparation begins with the transport of waste cardboard on large conveyors into pulpers that are essentially high speed mechanical churning devices filled with chemicals, usually sodium hydroxide, hydrogen peroxide, sodium silicate, soap and water [5]. The pulper serves the purpose of de-fibreising the recycled paper back into the state of pulp. One batch of paper pulp usually requires a residence time of 30-60 minutes in the pulper [6]. During this stage unwanted materials such as course rejects which consist of mainly plastic bails, plastic wraps and fibres are removed from the pulp by the use of a ragger rope. This is a long wire rope suspended within the pulper that collects the rejected fibres and plastics as the pulp is agitated. The rejected material is then transferred to large skips for landfill disposal [84].

The next stage of the process consists of further removing course rejects in a unit known as a fibreiser, which also serves the purpose of mixing dilution water back into the pulper. After the pulping stage the paper pulp is given the term “thick stock” and is transferred to a high density cleaner that has the capability of removing rejected staples, glass paper clips, small stones, bolts and knots. Positioning the high density cleaner close to the beginning of the process helps to protect downstream equipment from damage and rapid wear caused by reject wastes [5].

The thick stock is transferred into large storage tanks called dump towers, and sufficient residence time allows for further chemical reactions to occur. After such time, the thick stock is transferred to a fibre sorter and reject sorter where grit, sand and other light weight contaminants are removed [5]. The bottom ply thick stock then passes into a series of tanks consisting of a mixing chest, buffer chest, constant header and wire silos, which ensures complete mixing of all stock components as well as remixing in stock recovered from elsewhere in the mill. The top ply thick stock passes through a wire press where the viscosity increases and dissolved contaminants are removed. A disperger heats the material to remove stickies and inks by plate rotation. At this stage the pulp is given the term “thin stock” and is passed through a series of hydro-cyclones where light- and heavy-weight contaminants are cleaned in multi-stages [5,6]. The contaminants, mainly grit and sand, are de-watered by the use of a screw press before being transferred into large skips for land-spread disposal [84].

The last stage in stock preparation consists of the thin stock passing into a series of ominfractor and tampella screens; these are essentially slot screens that efficiently remove residual impurities within the stock. At this point the thin stock is sufficiently pre-conditioned and is ready to be turned back into paper, and so is transferred to the headbox of the paper making machine.

After the stock preparation system the stock is transferred into the paper making machine, and this is where the paper sheets are formed. There are several stages of the paper making machine namely the headbox, wire section, press section, dryer section, calendar section and pope reel section [5]. The paper making machine is extremely large and occupies a significant amount of space within a mill. This machine also requires a significant amount of energy in the form of electricity and heat in order to transport and dry the paper sheets. Figure 5 shows a typical paper making machine.

Figure 5 Paper making machine [7]

When the stock enters the paper making machine the first stage it encounters is the headbox. This is a piece of equipment that discharges a uniform amount of paper making stock onto a moving fabric [5]. The importance of this device is primarily to spread the stock evenly across the width of the machine, and to level out cross currents [5]. After the paper has been spread evenly onto the running fabric it enters the wire section, and here paper is formed on an endless fabric wire running in a loop across foils and various other de-watering elements. This process is called a Fourdrinier system [8]. A duo former is used for de-watering the paper and the web strength of the paper increases with increasing dryness [8]. After the wire section the paper web is pressed by multiple rollers in the press section, a bi-nip press followed by a straight press, and as the paper is pressed further drying occurs. Hydrogen bonding of the paper fibres also occurs at this stage, at around 40 % dryness [8]. Fibre surfaces are pressed close enough together that oxygen atoms of hydroxyl groups from adjoining surfaces come within reaction distance of each other. In this situation two oxygen atoms can share a proton of a hydrogen atom thus producing a hydrogen bond [5]. Figure 6 shows the fibre structure of paper with hydrogen bonding.

Figure 6 Fibre structure of paper with hydrogen bonding [5]

After the press section the paper passes through a number of drying cylinders. The dryers are driven by fabric wires and the heat is provided by the use of steam at 250-750 kPa [8]. Steam is supplied to each cylinder and condensate is removed. Some hydrogen bonding also takes place at the dryer section as surface tension during drying collapses the lumen and changes the tubular fibres into bonds [5]. After the drying section the paper is pressed in the calendar section by calendar rolls. The pressing effect of the calendar section provides an ironing effect of the paper and the rolls themselves may be heated or unheated depending on the design. After the calendar section paper winding takes place in a pope reel section. The finished paper rolls are then transferred to a large warehouse where they are stored before being shipped off-site to corrugators [5].

2.3.2 An overview of a newsprint mill's paper manufacturing process

The newsprint mill explored in this study (AN) produces on average 400,000 tonnes per year of low grade newsprint paper sheets which are sent to newspaper printers [86]. Once produced the paper is of white appearance. All secondary fibre newsprint mills follow similar manufacturing processing steps and may only differ slightly in configuration of equipment or type of equipment used.

Figures 7-10 shows stock preparation flow sheets and block diagrams for a typical secondary fibre newsprint mill. The first stage of this process consists of paper waste feedstock being brought into the mill by local authorities and other contracted companies. The paper waste grade brought in will contain ONP, magazines, office grade paper, packaging paper and contaminants such as mixed plastics (soft and hard), metals, glass, textiles and other general household waste in smaller quantities [86]. The paper is transported on large conveyors into the first unit of the process which is the pulper. Here the paper is churned with chemicals and water as previously described in Section 2.3.1. After the pulping process the stock is transferred into a fibreiser which removes course reject materials such as heavy plastics, metals, and glass. From here the stock is transferred into dump chests to settle before being routed through a series of high density cleaners. Smaller rejected fibres and plastic contaminants are removed at this stage and transferred to skips for landfill disposal.

The stock is then routed into a series of hole screens where alkaline reagents are added to remove de-inking sludge particles; these are then transferred to a sludge tower for disposal. After hole screening the stock then moves into flotation de-inking cells where alkaline chemical additives are added to enhance the brightness and to efficiently remove stickie residues [6]. Flotation machines comprised of cells in series and an aeration element generates a vacuum which sucks in air and forms bubbles in the stock. The bubbles remove ink particles suspended in the stock and transport them to the top of the cell, where a foam layer is formed which is then skimmed off the surface and de-watered before being transferred to a sludge tower [6]. The cleaner stock suspension is then transferred to the next cell where the process is repeated. After this series of de-inking stages, the stock is transferred to a disc filter where the pulp is de-watered [8]. The recovered water is then reused at various locations in the process. The filter consists of low volume discs mounted on a horizontal shaft with conical flow channel. The de-watered stock is removed from the discs by jets of water and the filter cloth is continuously washed with an oscillating shower [5].

From here the stock is further de-watered by use of a belt press before being transferred to storage tanks where acid is added [8]. The stock then moves through into a number of hydro-cyclone cleaners which further remove sludge for disposal. The stock is then transferred into multistage slot screens where any residual grit or sand is removed, and these contaminants are transferred to the sludge tower for disposal. The stock is thickened by the use of rotary or drums thickeners; this equipment is usually a rotating drum arranged horizontally inside a stationary casing and driven by a shaft. The walls of the drum are stainless steel with a removable mesh screen or filter cloth that is tailored for maximum removal of liquid from the stock [5]. The thickened stock is transferred to a storage tank before being further de-watered by the use of a belt press. Dispergers remove any leftover stickies and mix in bleaching agents before the stock is transferred to a storage tower ready to be processed by the paper making machine [8]. At this point the paper stock is transferred to the headbox of the paper making machine, and then follows a very similar process as described in Section 2.3.1.

Figure 7 Stock preparation process flow sheet of a newsprint mill de-inking process 1 [courtesy of Bridgwater paper mill] [88]

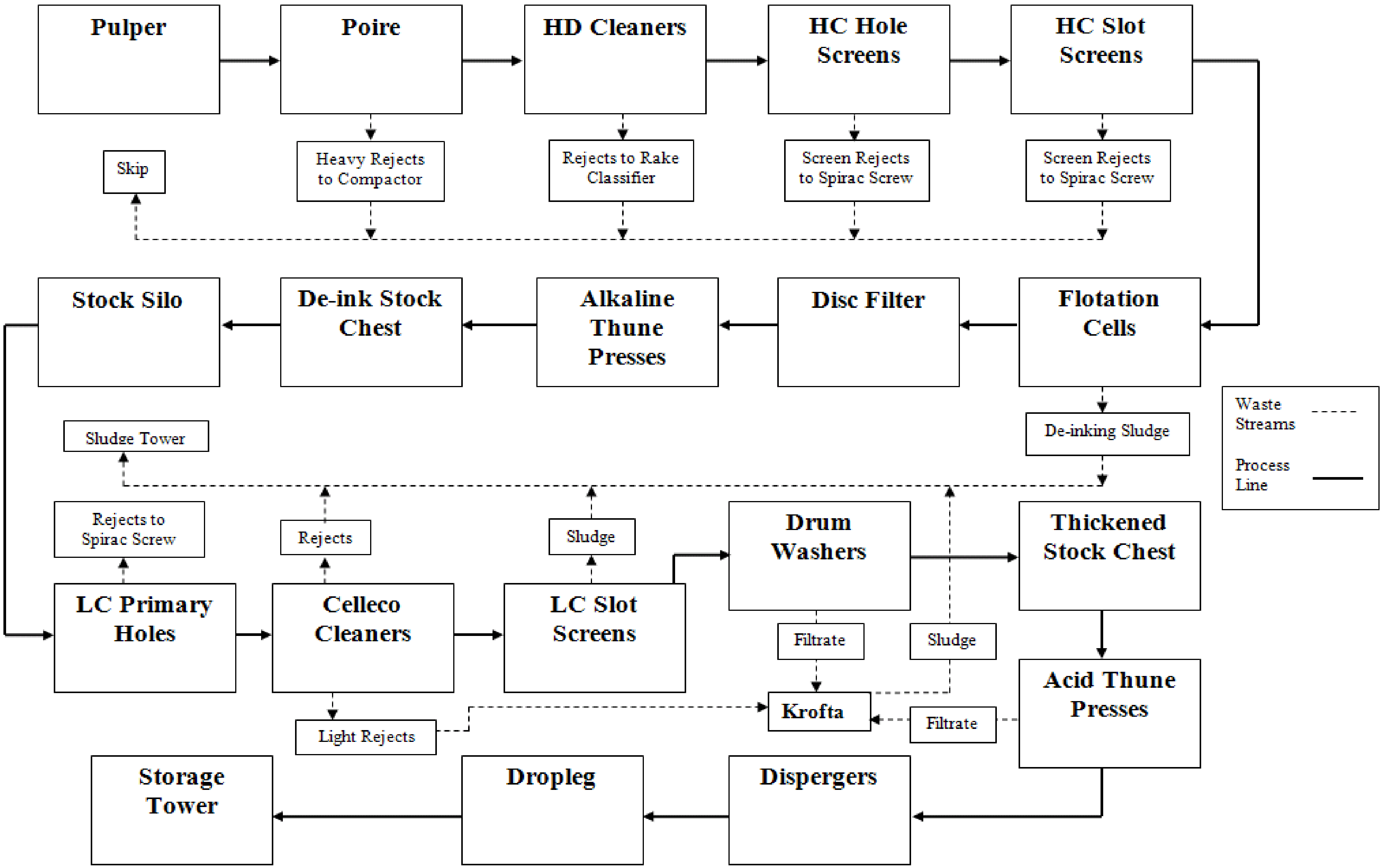


Figure 8 Stock preparation process block diagram of a newsprint mill de-inking process 1

Figure 9 Stock preparation process flow sheet of a newsprint mill de-inking process 2 [courtesy of Bridgwater paper mill] [88]

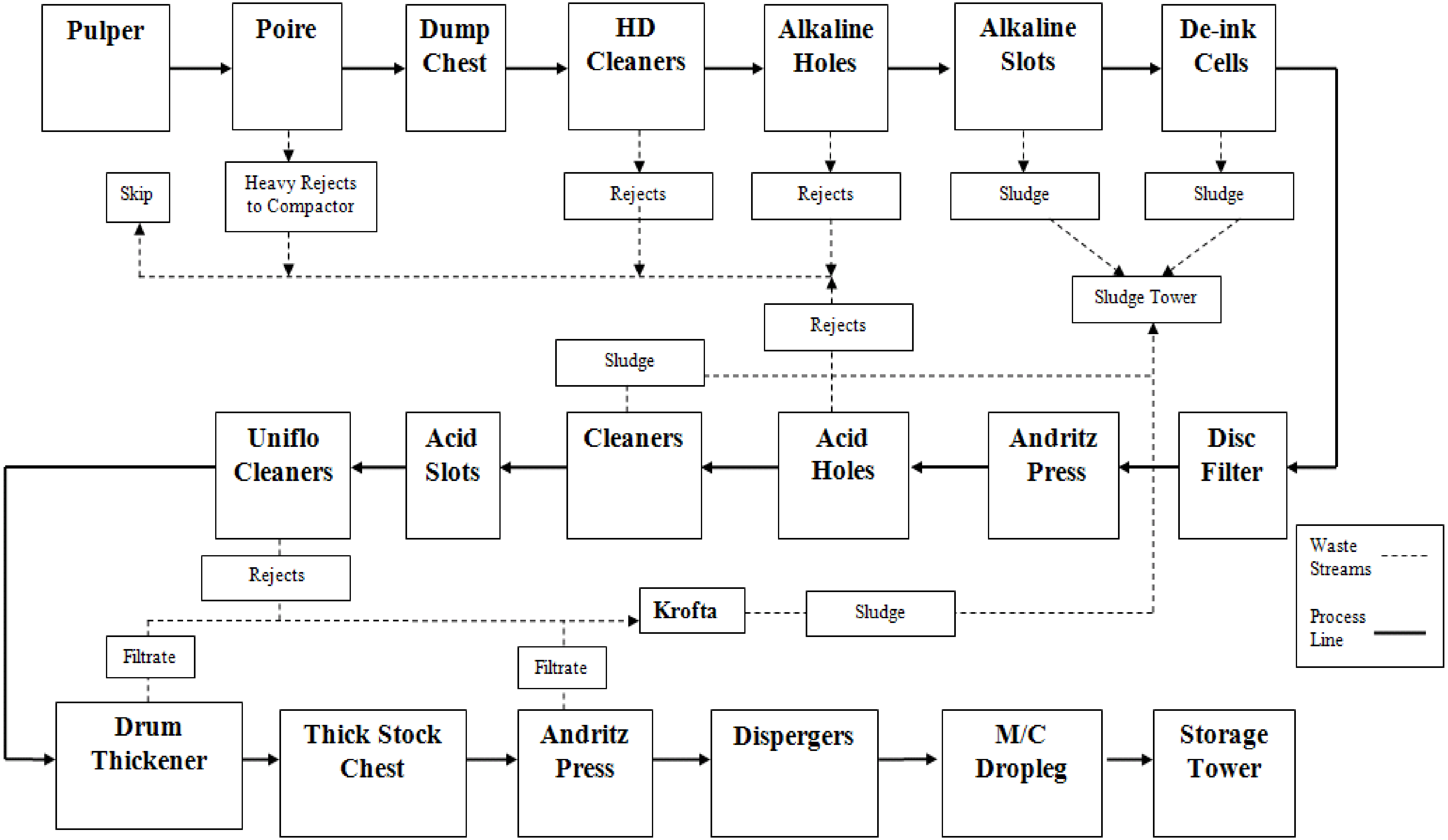


Figure 10 Stock preparation process block diagram of a newsprint mill de-inking process 2

2.3.3 An overview of a tissue mill's paper manufacturing process

The tissue mill explored in this study (KC) produces three types of tissue products, namely hand towels, tissue paper and non-woven cleansing wipes. Over 65,000 tonnes of these products are produced collectively on-site each year and the mill uses 100 % recycled office grade paper as a feedstock [85]. The site is split into three installations: the Coleshill mill which produces the tissue paper; the Delyn mill which produces hand towels; and the Flint mill which produces cleansing wipes [85].

The Coleshill stock preparation plant is composed of a de-ink plant and a virgin pulper system. The de-ink plant converts low grade waste paper into clean, white, low ash pulp for consumption by the tissue machine, using 100 % low grade waste paper as a raw material. The de-ink plant uses pulping, screening, centrifugal cleaning, washing, dispersing, flotation, de-inking and bleaching processes [85]. A dissolved air flotation (DAF) clarifier is utilised to re-circulate internal water [8]. The Coleshill tissue machine produces up to 30,000 tonnes per year of white bathroom tissue [85]. Once passed the forming section the sheet goes through vacuum de-watering and then on to a steam heated Yankee dryer cylinder with gas fired heated hoods. Sheets leaving the Yankee are weighed and monitored by moisture scanners, and then tissue reels are cut on-line to the required width [85]. Waste tissue bales are processed in a tissue waste baling compactor for use in the Delyn Mill.

The Coleshill process is shown in Figure 11. The process begins with the raw feedstock being brought in and stored in a warehouse. The feedstock at this point is in the form of bales and these are de-wired and then loaded on to the pulping conveyor. Pulping is done on either a continuous or batch mode [6]. Bales are loaded and weighed, and information passes to the process operators on-site who calculate the correct weight of paper, water and process chemicals to be added into the pulper. A high consistency pulper mixes the waste paper and process water into a slurry and a surfactant chemical is added to help the flotation section [6]. The de-inking plant can process up to 180 tonnes/day of waste paper at an average process yield of 55 %.

The de-fibred pulp moves through the de-trashing screen removing coarse contaminants such as plastic and wet strength paper. The rejects are rinsed to recover fibre and then put onto a drainage conveyor. Free draining water is recovered and the de-watered rejects are transferred into a skip for disposal [85].

Coarse cleaning and screening equipment removes staples, paperclips, stones and other hard materials which could damage downstream equipment [5]. There are two centrifugal cleaners in parallel and hard materials pass through a rejects trap at the base of the cleaner. There is a constant counter current stream of water which separates good fibre from heavy contaminants. Losses from this cleaning and screening are around 0.5 % [85]. Accepts from the cleaners pass onto a coarse pressure screen and rejects from this screen move on to a fibre recovery screen that is automatically purged at set times [85].

Flotation is utilised for the removal of fine ink and dirt particles from the stock. Air micro-bubbles are mixed with the incoming stock, attaching themselves to ink particles assisted by the action of the surfactant added at the pulper stage [6]. These bubbles float to the surface with the ink particles. An extraction fan takes the ink off the surface of the flotation cell [6].

A vortex separator is used to separate entrained air from the reject sludge and the sludge moves from a barometric leg into a small transfer tank. The sludge is pumped to sludge tanks before de-watering on sludge presses. The higher the ash content of the stock, the higher the percentage loss and the higher the ash/filler content of the sludge [8].

Fine screens, consisting of 3 stages to reduce fibre loss, are used to remove sticky material with particles greater than 0.20 mm [85]. Rejects from the final screen (2 % loss) go through a vibrating gravity strainer and water is recovered. De-watered rejects move into a primary sludge tank to mix them with other reject streams and pump them into sludge presses [6]. Accepts from the fine screens go through centrifugal cleaners and a number of stages ensure fibre loss is minimised. Rejects pass through the same de-watering process as those from the fine screens.

Printing, writing and coated papers contain around 10-35 % of mineral fillers which give the paper the correct opacity and surface absorbency [8]. Washing removes the particle size range which covers filler. Washers function to remove filler and fines and thicken. The washing process is ideally suited to thickening from 1 % to 10 % consistency. The washed stock sticks to the solid roll to be filtrated into an accepts screw conveyor. Filtrate from the washers moves to the primary Dissolved Air Flotation (DAF) clarifier [8].

Screening, cleaning and flotation processes remove ink particles following separation from fibres. The separation process is not 100 % efficient therefore stock at high consistency and elevated temperature is passed between the rotating plates of a disperger [6]. Steam is applied to assist fibre dispersion and to heat the stock before disperging. A high temperature softens sticky material and removes all ink particles. The dispersed contaminants are removed from the stock by a final washing stage. A wire press thickener and steam heated screw are used to reach the required consistency and to control temperature.

Waste paper containing dyed paper has additional colour removed by an optimised sodium hydrosulphite bleaching stage [85]. Bleaching equipment is composed of a medium consistency pump with vacuum degassing and an up flow bleach tower. Fresh hydrosulphite solution is dosed to the suction of the medium consistency pump then it mixes with the pulp. The chemical used is a blend of chelants, buffers and sodium hydrosulphite, which increases efficiency due to decomposition and removes sulphur dioxide emissions [85]. Dry powder is delivered in sealed flowbins that are protected from contact with water. The bleaching reaction takes place in de-aerated stock at a high temperature to increase efficiency and limit chemical usage [6]. The thermal energy for this process is obtained from the energy of tissue machine hood exhaust gases.

The virgin pulper system operates at approximately 6 % consistency with a batch weight of around 1.0 tonne. Stock from the pulper passes through a high density centrifugal cleaner removing oversized contaminants and is fed into the final stock chest. Stock is pumped from the final stock chest and passes through two refiners. Refined stock then flows to the tissue machine thick stock headbox, and from this point follows a similar process as described in Section 2.3.1.

Hand towels are produced at the Delyn installation, and the overall process follows very similar processing steps to that of the Coleshill mill. However there exists some differences; for example it is not necessary to clean de-ink or bleach the stock to the same standards as the Coleshill stock. The resulting product is a much lower grade of tissue paper than that produced at the Coleshill mill.

Cleansing wipes are produced at the Flint mill and are composed of disposable non-woven fabrics soaked in a gentle cleansing solution [5,6,8]. The material used in cleansing wipes is made by a process that presses a single sheet of material from a mass of separate fibres composed of cotton and rayon, as well as plastic resins like polyester, polyethylene, and polypropylene and is designed to be disposable, absorbent yet durable.

Water is the main ingredient of cleansing wipes and serves as a carrier and diluent for the other ingredients [85]. Cleansing wipes also contain, mild detergents mixed with moisturizing agents, fragrance, preservatives and oils. Humectants are added to cleansing wipes to prevent premature drying. Preservatives, such as methyl and propyl paraben, are added to avoid microbial growth. Packaging, such as thermo-molded plastic tubs, keep the cloths free from contamination.

The process to produce cleansing wipes can either be a dry laid process or a wet laid process. The dry laid process is used to make non-woven fabrics from plastic resins. Plastic polypropylene pellets are melted and pushed through tiny holes, by air pressure. As the fibre cools, it condenses to form a sheet and hot metal rollers flatten and bond it. A wet laid process is used for softer cloths. The fibres are formed into liquid slurries using chemicals and water. The paste is pressed, dried and cut into flat sheets by rollers.

Treatment of the non-woven cloth begins as it is fed from storage rolls onto coating machinery where cleansing solution is applied. The fabric is then run through a cleansing solution and folded into small pouches by automated equipment. During this packing process a liquid feed mechanism, injects moisturizing liquid into the towelette and a heat sealer closes the pouch.

The cleansing wipes undergo a series of quality check points during the manufacturing process to ensure no leakage, faulty closure and to conform to tear-strength requirements and to carry out tests on the cleansing solution. Cleansing wipes which do not meet these stringent quality control checks are rejected from the process as co-form rejects (dry and wet wipes) [85].

Figure 11 A block diagram of the Coleshill process [courtesy of KC mill] [85]

2.4 Secondary fibre paper mill waste streams

The wastes which are of greatest concern to secondary fibre paper mills due to their quantities and difficulty of disposal are reject materials (all mills) and de-inking sludge (newsprint and tissue mills). These waste streams constitute the focus of this study.

Other types of wastes produced at secondary fibre mills include primary sludge, the underflow residue from the primary wastewater treatment plant clarifier, and secondary sludge which evolves from secondary waste water treatment basins [6]. Both tend to contain a significant fraction of fibrous, organic (i.e. combustible) content and are relatively easy to de-water [6]. Also both primary and secondary sludge are produced in much smaller quantities than de-inking sludge and reject wastes.

The method of treating wastewater at a mill is usually by anaerobic digestion. This process produces methane rich gases which can be used as a substitute for natural gas, to be burnt in a boiler to raise process steam or used as a fuel for on-site CHP systems [6]. This is already an established technology at the SSK mill and works well to produce energy rich biogases. For these reasons, wastewater treatment sludge was not considered in this study.

2.4.1. Reject wastes from a newsprint mill

When recovered paper feedstock is brought into a mill, it often contains large amounts of other waste, such as plastic, metal, textiles and glass [86]. These waste fractions are rejected from the process by screening at various stages through the pulping process (see Section 2.3.2). The wet reject material, which can have a moisture content in excess of 70 %, is separated from the paper pulp and often placed in large skips or metal containers to de-water before being transported to landfill sites, (although some paper mills choose to further de-water their rejects mechanically using a screw press or compactors).

The rejects ejected at the beginning of the process are usually coarse heavier materials such as clothing, PET bottles and other larger objects [86]. The rejects gradually become smaller and lighter as the process moves further downstream, until in the later stages the rejects are composed of no more than grit and sand. The rejects from different process stages are kept in separate skips within the process, but ultimately are combined together in one mass for landfill disposal [86]. The composition of these rejects varies widely, both with time and also depending on the mill's specific process. A newsprint mill's reject waste will contain large amounts of textiles, metal, glass, (hard plastics) from CDs and bottles, and (soft plastics) from plastic wrap and film, and other general household waste [3, 86]. Approximately 5 % of a typical newsprint mill's waste will be comprised of these materials and at the AN mill this can equate to as much as 5,000 dry tonnes per year (10,000 wet tonnes) [2]. Generally reject material coming from a newsprint mill is broadly comparable to municipal solid waste but with a rather higher fibre content, and it is the extreme heterogeneous nature of the waste along with its day-to-day variability which imposes the requirement for costly pre-processing if the material is to be used as a fuel. Furthermore to prevent dioxin formation, plastics in the form of PVC within these rejects may require either initial removal from the feedstock or further treating downstream before any subsequent thermal conversion process, and implementing this extra pre-treatment step to the process would further increase capital costs.

The cost of landfilling these waste fractions is increasing year upon year and as a result many paper mills are finding this disposal practice increasingly unsustainable. Furthermore in some mainland European countries such as Germany and the Netherlands, landfilling of these wastes has already been prohibited [3], and a similar course may be taken in the UK at some point in the future. Reject material produced from the AN mill is shown in Figure 12 and typical constituents are given in Table 1.

Figure 12 Light rejects generated from the AN mill

Table 1 Typical constituents found in reject waste from a newsprint mill

Typical constituents of reject material	
PET & HDPE	Paper
Compact Discs	Textiles
Glass	Staples
Steel	Paper Clips
Aluminium Cans	Bottles
Bags	Packaging

2.4.2. Reject waste from board and tissue mills

Rejects generated from board mills and tissue mills are broadly similar, but differ from those from a newsprint mill. This is due to the paper grade feedstock which is used at board and tissue mills. Board mills only use OCC as the feedstock, and tissue mills use recycled mixed office grade paper [84,85]; therefore it is less likely that substantial amounts of contaminants will find their way into the processes. As at a newsprint mill, rejects ejected from board and tissue mills are coarse at the start of the process and gradually get lighter further downstream. Unlike at a newsprint mill, some lighter fractions such as grit and sand are land-spread to reduce disposal costs, and so these fractions are kept separate to the coarser material.

In general landfilled pulper rejects produced at board and tissue mills are comprised mainly of fibres and plastics. At the SSK mill, the composition over a year averages 70 % plastics (mainly soft plastic wrap and film) and 30 % fibres [84], whereas at the KC mill the proportion of plastics (soft and hard) is rather higher [85]. The moisture content is high (up to 60 wt %) and the ash content is relatively small (10-15 wt %). The SSK mill produces approximately 7,000 dry tonnes per year (17,000 wet tonnes) of total rejects [84], whereas pulper rejects from the KC mill are approximately 1,800 dry tonnes per year (4,500 wet tonnes)[85].

Due to the high plastics content, the rejects produced at both mills have a high gross heating value similar to low grade coal (18-28 MJ/kg dry basis) [6]. Other rejects produced further downstream contain some metals such as staples and paper clips, as well as small amounts of glass. However these quantities are significantly lower in comparison to the amounts produced at newsprint mills. Figure 13 shows plastic and fibre rejects ejected from the SSK mill.

Figure 13 Board mill rejects ejected from the SSK mill

2.4.3 Co-form rejects (dry and wet wipes)

Co-form rejects are derived from non-woven mills only and are essentially the rejected non-woven materials used to make cleansing wipes (often referred to as baby wipes). Cleansing wipes must meet stringent quality control checks to ensure no leakage, faulty closure and to conform to tear-strength requirements before they can be sold, and the rejected co-form material is essentially the wipes which do not meet these standards, and are therefore discarded from the manufacturing process.

Dry co-form rejects refers to the cleansing wipes before moisture and other antibacterial reagents are added, and are composed of approximately 30 wt% polypropylene and 70 wt % wood pulp fibres [85]. Wet co-form rejects refers to the cleansing wipes after water and other cleansing ingredients have been added, and they usually have a moisture content of approximately 70 wt % [85]. At the KC mill the quantity of this particular waste stream is very small and equates to around 500 dry tonnes per year [85]. Figure 14 and 15 shows wet and dry co-form rejects respectively.


26/07/2011

Figure 14 Wet co-form rejects produced from the KC mill


26/07/2011

Figure 15 Dry co-form reject reels produced from the KC mill

2.4.4. De-inking sludge waste from newsprint and tissue mills

De-inking sludge refers to the float or scum that is evolved from the air flotation process used to remove inks and dyes from the recycled paper fibres [6], and approximately 1 million tonnes of de-inking sludge is produced in the UK each year [9]. De-inking sludge is only produced at paper mills which have de-inking processes such as newsprint mills and tissue mills. Board mills produce no de-inking sludge as the de-inking process is not required for the manufacture of packaging grade paper.

De-inking sludge consists of short length fibres, inks, dyes, fillers and other pigment based contaminants [6]. The de-inking process is usually carried out by flotation, which uses relatively small amounts of chemicals or water, or by mechanical agitation with alkali and surface active chemicals. A traditional bleaching process may also be required to further increase pulp quality and brightness and these bleaching reagents may also end up in the de-inking sludge [6]. Recycled paper entering a mill can contain a large fraction of inorganic substances including dyes and fillers such as kaolin (Al_2O_3, SiO_2, H_2O), talc ($Mg_3Si_4O_{10}(OH)_2$), calcium carbonate ($CaCO_3$) and clays which are added to improve printability, smoothness, opacity and appearance of the finished paper product [10], and significant amounts of these also end up in the sludge. Newsprint and tissue paper tend to contain lesser amounts of fillers compared to (for example) a high quality glossy magazine, due to the paper generally being at a lower quality grade.

De-inking sludge has a high moisture content in the range 35-60 % [10]. It has a low typical gross heating value (HHV) of 6-7 MJ/kg on a dry basis [6], largely because of the high ash content which can range anywhere between 45-65 % depending on the quality of the fibre initially brought into the mill. Thus when the sludge is dried and then subsequently burnt a significant amount of solid ash residue is left behind. The main constituents of this ash are aluminium and calcium metal oxides derived from the fillers used in the de-inking process [10]. Other metals are also present in the ppm range.

Currently the main sludge disposal methods are combustion (direct co-firing with natural gas to raise process steam) and land-spreading. These are both costly, and environmentally undesirable due to the potential for land contamination and the use of fossil natural gas. Some mills also dispose of their sludge by selling it to companies for use as cattle bedding, and in some cases, ash residues from the combustion of sludge is sold to the cement industry as a concrete additive. Figure 16 shows de-watered de-inking sludge as it is ejected from a newsprint mill's process.

Figure 16 De-watered de-inking sludge from a newsprint mill

2.5 Energy requirements of the paper mills

Each of the paper mills explored in this work requires large quantities of energy in the form of both steam and electricity to run the paper manufacturing process. Steam is required mainly by the paper machine to dry paper sheets and electricity is required on-site to run drives, pumps, motors, computers and other machinery. As the cost for importing both electricity from the national grid and natural gas from energy suppliers increases, many mills have installed their own energy systems on-site to facilitate in providing some of the energy they require.

2.5.1 Current energy system, AN mill

The AN mill currently requires 44 MW of electrical energy (MWe) to run the entire mill [86]. The majority of this energy is supplied by a gas turbine which burns natural gas. A proportion of this energy (approximately 2.3 MWe) is supplied by a steam turbine using steam raised from the exhaust gases from a sludge combustor [86]. The on-site sludge combustor is a bubbling fluidised bed and co-fires approximately 117,000 tonnes per year of wet de-inking sludge with natural gas (as the sludge on its own would not sustain combustion) [86]. The exhaust gases are routed directly to a superheated steam boiler which raises steam for the steam turbine. Some of the steam produced is routed to the paper machine to dry paper sheet [86].

2.5.2 Current energy system, SSK mill

The SSK mill currently requires approximately 30.5 MW of heat (MWth) and 8.4 MWe of electrical energy to run the paper mill [84]. The vast majority of this energy is produced on-site using natural gas as a fuel. Approximately 48.5 MWth of natural gas is imported in total. A further 0.9 MWe is imported from the national grid [84].

The mill has two natural gas fired CHP turbines (CHP 1 and CHP 2) which generate a total of 7.7 MWe. The mill also has two natural gas fired boilers (Boiler 2 and 3) to raise a combined total of approximately 23.9 MWth of steam at a pressure of 15 bar gauge and temperature of 200 ^{o}C [84]. A third stand-alone boiler (Boiler 1) is fired mostly on biogas produced from the anaerobic digestion of waste process water which creates between 300-550 m^3/h of gas composed of 70 % CH_4 and 30 % CO_2 by volume (approximately 2 MWth) [84]. The current system will change in the near future as the vast majority of the biogas will be routed instead to a CHP engine to create 890 kWe with heat from the exhaust gases raising steam for the mill in a waste heat boiler [84].

2.5.3 Current energy system, KC mill

The energy consumed at the KC mill is mostly electricity required to power machinery and heat to dry paper sheets, and most of this energy is used at the Delyn and Coleshill Installations [85]. No electricity or heat is generated on-site. Approximately 5 MWe and 2.6 MWe are imported from the national grid by the Coleshill and Delyn mills respectively [85]. Natural gas is also imported (9.3 MWth and 6.7 MWth by Coleshill and Delyn mills respectively) and is used as a heat source to fire two steam boilers one at each site which generate steam for the dryers [85].

2.6 Advanced thermal conversion techniques

There are three thermal conversion techniques which could be used to process paper industry wastes into useful energy products; these are combustion, gasification and pyrolysis. Both gasification and pyrolysis are advanced thermal conversion techniques, well suited to the particular types and quantities of wastes produced at each of the secondary fibre mills. In each process de-inking sludge, reject wastes and blends of the two can be easily converted into gaseous or liquid fuels that can be used in a gas turbine or engine based CHP system. This configuration would yield higher electrical efficiencies than direct combustion with a steam cycle [3, 11, 12, 13]. Furthermore these systems are preferred as pollutants can easily be controlled to low levels [3, 14] and each unit can easily be scaled to suit individual mill requirements. Lastly both gasification and pyrolysis processes being advanced thermal conversion technologies receive extra government support by way of incentives as opposed to traditional combustion systems. For example pyrolysis and gasification processes are entitled to receive higher amounts of renewable obligation certificates (ROCs) [15] than traditional combustion systems.

2.6.1 Gasification

Gasification systems gained their popularity in World War Two when fuel shortages were widespread and people required an alternative and cheap means of generating energy. Wood-fired downdraft gasifiers were often used to produce a combustible gas for powering vehicles and electrical generators for buildings. When Middle Eastern crude oil became abundant after World War Two, gasification processes were displaced. However in recent years, increasing concerns have been raised over the role of fossil fuels in climate change as well as over their rate of depletion, and now many energy producers are revisiting the gasification of biomass and wastes in the quest for a renewable, cheap and reliable source of fuel. Many researchers have explored methods to make the process more efficient and reliable by tackling issues such as tar formation and temperature variations which can be problematic in gasifiers.

Although gasification is not a new technology, its application in the pulp and paper industry is considered to be in the early stages of development. Gasification in this work is the defined as the partial combustion of paper industry waste feedstocks into a fuel gas called a product gas, composed of mainly CO, CO_2 H_2 CH_4, H_2O and N_2, that can be used in heat, power or CHP applications. The partial combustion of the feedstock is achieved using air as the oxidising agent.

There are three main types of reactor configuration which exist for the gasification of biomass and wastes. These are: the fluidised bed gasifier, the entrained flow gasifier and the fixed bed downdraft gasifier, [16]. In a fluidised bed gasifier feedstock is injected into either a bubbling or circulating bed of sand or a mixture of char and another inorganic heating or catalytic media such as dolomite. Rapid heating rates achieve gasification temperatures almost instantaneously. The oxidising agent provides the fluidising medium as well as taking part in the gasification reactions; this can either be air, oxygen enriched air, CO_2, steam or a combination of these. In an entrained flow gasifier gasification reactions take place in a cloud of suspended particles. This type of gasifier usually operates at high temperatures in excess of 1000 ^{o}C which means higher throughputs can be achieved and thus this type of gasifier is usually suited to larger scales > 50MWth. However the present study focuses on the fixed bed downdraft gasifier. It is the preferred configuration for small-scale distributed power generation at inputs < 5MWth [16], which broadly matches the waste stream tonnages being considered here. Furthermore, the literature review revealed that this option has not been widely researched despite advantages of simplicity and cost. The characteristics of each reactor are summarised in Table 2, and the features of the fixed bed downdraft gasifier are then discussed in detail.

Table 2 The characteristics of the three main biomass gasifier types [17]

The gasification steps involved in fixed bed downdraft gasification are drying, de-volatilisation, oxidation and reduction [17]. These are depicted in Figure 17. The first step of the process involves the biomass drying at temperatures between 70-105 ^{o}C [16]. The moisture released from the biomass as water vapour passes through the gasifier and contributes towards important reduction reactions (see Equations 3 and 4). The feedstock continues to move through the gasifier and is further heated to temperatures between 250-550 ^{o}C in which range the feedstock de-volatilises [16]. The volatile components move through the gasifier into the oxidation and reduction zones where they further react with the oxidising agent and char passing through the unit [18]. Partial combustion of the volatiles takes place at the oxidation zone and temperatures within this region are increased to between 700-1100 ^{o}C [17]. The drying, de-volatilisation and reduction steps of gasification are all endothermic and require significant heat input and this is supplied auto-thermally by the exothermic oxidation reactions [18]. The oxidation reactions occurring may be represented by Equation 1, the oxidation of carbon, although the volatiles will in practice be oxygenated hydrocarbons and water vapour will also be formed.

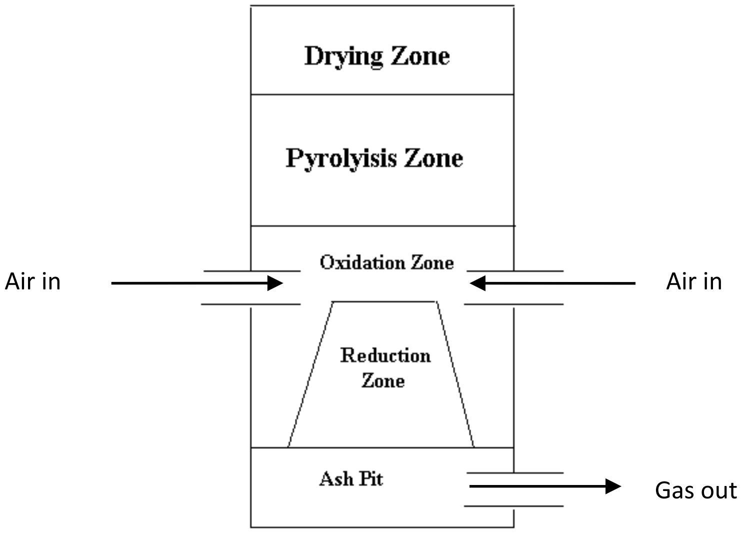


Figure 17 A diagram of fixed bed downdraft gasification

Oxidation reaction

$$\mathrm{C} + \mathrm{O_2} \leftrightarrow \mathrm{CO_2} \qquad \Delta \mathrm{Hr} = 393.8\ \mathrm{KJ/(mol.k)} \qquad (1)$$

The reduction zone of a gasifier is a region where most of the gasification reactions occur [16]. Here the remaining volatiles and the products of the oxidation zone react with each other and with the solid char in a complex set of reactions which may be simplified to Equations 2-5. Heats of reaction in Equations 1-5 are at 298K).

Heterogeneous water gas reaction

$$\mathrm{C} + \mathrm{CO_2} \leftrightarrow 2\mathrm{CO} \qquad \Delta \mathrm{Hr} = 172.6\ \mathrm{KJ/(mol.k)} \qquad (2)$$

$$\mathrm{C} + \mathrm{H_2O} \leftrightarrow \mathrm{H_2} + \mathrm{CO} \qquad \Delta \mathrm{Hr} = 131.4\ \mathrm{KJ/(mol.k)} \qquad (3)$$

Homogenous water gas shift reaction

$$\mathrm{CO} + \mathrm{H_2O} \leftrightarrow \mathrm{H_2} + \mathrm{CO} \qquad \Delta \mathrm{Hr} = -41.2\ \mathrm{KJ/(mol.k)} \qquad (4)$$

Methanation reaction

$$\mathrm{C} + 2\mathrm{H_2} \leftrightarrow \mathrm{CH_4} \qquad \Delta \mathrm{Hr} = -74.9\ \mathrm{KJ/(mol.k)} \qquad (5)$$

The final product gas formed during the reduction reactions is composed mainly of CO, CO_2 H_2 CH_4, H_2O and N_2, with small amounts of condensable organic vapours (tar) which are essentially the long chain hydrocarbons and other volatiles which did not crack in the reduction reactions. In a conventional downdraft gasifier tars can range typically between 1-5 g/ Nm^3 [19] of product gas depending on the conditions and type of feedstock, although modifications such as throatless gasifiers can reduce the tar content down to as low as 50-250 mg/ Nm^3 [19]. Once cleaned the product gas can either be used in an IC engine or gas turbine to produce electricity and heat; this is a proven and well established technology [18].

The downdraft gasifier does however have some drawbacks. For example high ash containing feedstocks may cause clogging or ash softening within the narrow throat section of the gasifier and this requires the need for constant agitation of the gasifier's bottom ash grate. Processing such feedstocks may not be feasible without significant gasifier modification [17]. Furthermore fixed bed downdraft gasification requires a high level of pre-processing as the feedstock should be in a homogenous relatively large particulate form such as pellets, chips or briquettes (see Section 2.7); this is an extra step in the process and could be costly especially when dealing with feedstocks with high variability such as paper industry rejects.

A typical gasification application for paper industry waste feedstock could be a CHP plant with an electrical output of approximately 500 kWe (comprising of two 250 kWe gasifier units). Around 3,000 dry tonnes per year of processed rejects would be required to feed such a system, or 12,000 dry tonnes per year of sludge due to its much lower heating value. Modules could be added to accept higher throughputs. The complete system would consist of feedstock pre-processing (conveying, drying, pelletising and feeding), gasification (gasifier, char and ash removal, gas conditioning and cooling, gas storage) power generation (single fuel gas engine, modified diesel engine or gas turbine) coupled to a waste heat boiler.

2.6.2 Pyrolysis

Pyrolysis is defined in this work as the thermal decomposition of organic matter in the complete absence of an oxidising agent. The pyrolysis process consists of a set of complex chemical reactions involving the formation of radicals. These reactions are endothermic and so require a heat input. As the temperature increases chemical bonds in the feedstock are broken and unstable radicals form, which then react with each other to form more stable low molecular weight molecules [20].

Pyrolysis takes place in the temperature range 280-850 ^{o}C [20] depending on the nature of the feedstock, the desired products and the particular pyrolysis process employed. Similarly pressures can range from hypobaric to atmospheric to elevated pressure [18].

In general pyrolysis processes can be broken down into three main types: slow, intermediate or fast pyrolysis. The main differences between the three processes are the heating rates and solid residence times involved, the reaction temperatures, and the relative yields of solids, liquids and gaseous products formed. The three processes are summarised in Table 3.

Table 3 Typical product yields (dry feed basis) obtained by different modes of pyrolysis of wood [18]

Fast pyrolysis involves very high heating rates at a reaction temperature of around 500 ^{o}C with very short solid/vapour residence times followed by rapid quenching in order to achieve maximum liquid yields (“bio-oil”) [18]. Slow pyrolysis is the opposite with slow heating rates, long solids residence times and low reaction temperatures, and it favours the production of high yields of solid char. Even though fast pyrolysis has the ability to produce high yields of pyrolysis oil [18], this liquid can contain large amounts of high molecular weight compounds as the vapour residence times are very short, not allowing sufficient time for these compounds to be sufficiently cracked [20]. This can lead to highly viscous bio oils which may cause storage and handling problems, phase separation and the need for further liquid upgrading.

Slow pyrolysis requires lesser amounts of heat energy to achieve the desired pyrolysis temperatures; however much longer residence times are required [18]. The product distribution favours high solid yields, with little or no pyrolysis liquids formed. Having a process tailored towards the production of maximum quantities of solids is not suitable for feedstocks with high ash contents and relatively small amounts of fixed carbon such as de-inking sludge.

Intermediate pyrolysis is between the two with moderate reaction temperatures and solids residence times, and produces intermediate amounts of solids, liquid and gaseous products. Figure 18 shows the working principle. Intermediate pyrolysis offers advantages over both fast pyrolysis and slow pyrolysis in that it has the potential to produce relatively low molecular weight pyrolysis oil [20].

There are several feasible configurations for integrating an intermediate pyrolysis process into a CHP system, but the preferred route here would be to condense the pyrolysis liquids and then subsequently utilise both these and the permanent gases (which also have a calorific content) in a dual fuel diesel engine, with the remaining solid char combusted to create additional heat for the process. Paper mill residues are well suited to be processed in this way and at industrial scale the whole system would consist of waste pre-treatment (sorting, sizing and drying), then intermediate pyrolysis to generate pyrolysis oil, char and permanent gases, then a dual fuel engine in parallel with a char combustor.

Figure 18 The working principle of intermediate pyrolysis [20].

2.6.3 Combined heat and power (CHP)

As has been previously stated, paper mills require both heat for drying paper sheets and electricity for drives, motors and machinery. Heat and power are often supplied independently, typically by burning natural gas and by importing electricity from the grid.

With a CHP system the waste heat produced in the generation of electricity can be used to provide local and process heat in the form of hot water and steam [21]. The viability of CHP depends on many factors and the full economic analysis can be complex.

Electricity generation in a CHP unit usually comes from a gas or steam turbine or an IC engine connected to an electrical generator. For the present study an IC engine would be most suitable as it can be coupled easily to both a fixed bed downdraft gasifier and a pyrolysis reactor to run on both liquid bio-oil and product gas.

CHP systems incorporating IC engines recover heat from the engine exhaust with a waste heat boiler, and also from the engine cooling system. In this way a large proportion of the fuel supplied is used to provide both heat and power. Generating energy in this way is usually much more efficient than independent generation of electricity and heat.

Figure 19 gives illustrative data for a CHP system with 100 units of energy input. In this case 40 units of heat and 40 units of power are produced. The CHP unit therefore has a heat to power ratio of 1:1. To produce the same amount of heat conventionally would require 150 units of primary energy.

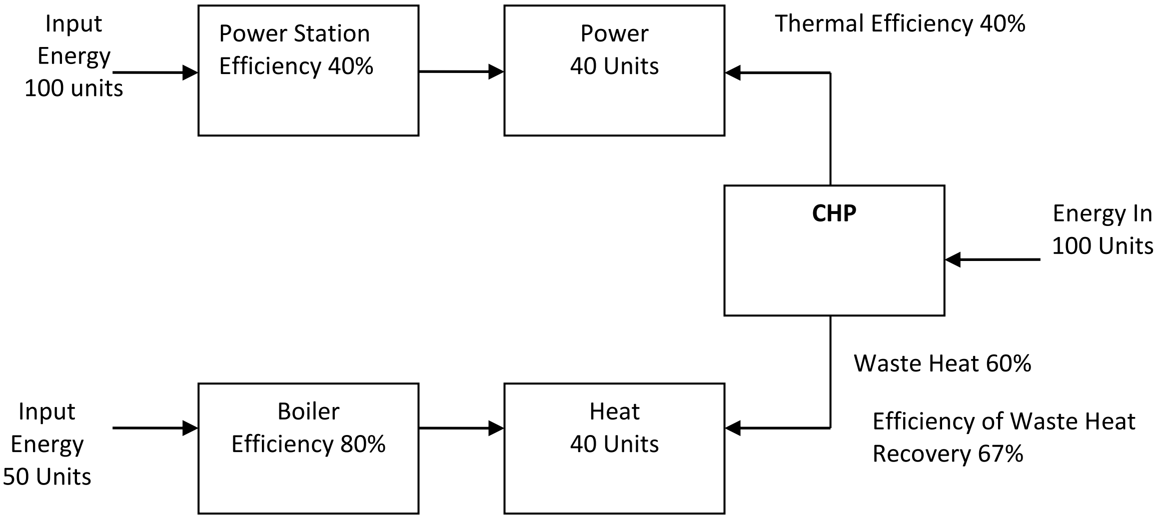


Figure 19 CHP vs conventional heat and power supply.

CHP unit efficiency = 80%

Conventional approach efficiency = 80/150 or 53%

The heat to power ratio is significant. At high values where considerably more heat is required than electrical power, the energy savings offered by CHP are reduced. The ideal application of CHP is where the end user has a constant demand for heat and power. This situation occurs often on a large scale. There are also many instances in industry and commerce where a CHP unit can be designed to meet a base heat load with any excess power being exported to the supply network [21].

The implementation of a CHP system in paper mills is not uncommon and several mills are currently operating using CHP from direct natural gas combustion, from the combustion of gas generated from the anaerobic digestion of waste water, or from the co-firing of solid wastes with natural gas, to drive either a steam turbine in a Rankine cycle, or a gas turbine or IC engine with the exhaust gases utilised to raise steam.

2.7 Feedstock pre-treatment

Wastes produced by secondary fibre paper mills cannot be used directly as a feedstock for thermal conversion, either by gasification or pyrolysis, without some degree of pre-treatment. However, it is important to minimize the necessary pre-treatment steps for economic reasons. All the waste streams being considered here have high moisture content at the point of extraction. High moisture containing feedstocks give rise to high water content product gases or liquids, with consequently reduced calorific value. Similarly high temperatures necessary for these thermal conversion processes cannot be sustained with high moisture containing feedstocks. Therefore, moisture must be reduced significantly in order to obtain high thermal efficiencies in all thermal conversion systems. A maximum moisture content of 15 wt% is recommended for gasification and pyrolysis processes [22]. To reach this level of moisture, evaporative drying is required. The drying of various paper waste streams can be achieved in many different dryer types, including fluidised beds, rotary drum dryers, or paddle dryers. Drying of paper industry wastes like de-inking sludge is a well demonstrated technology and is described in detail elsewhere [6, 23, 24].

The preferred dryer for waste streams such as de-inking sludge is a rotary drum dryer [6]. There is a wide variety of configurations for rotary dryers, but the preferred configuration is the heating of agitated material in a continuous rotating drum that serves the purpose of improved heat transfer efficiency by exposing more surface area of the material to the heating medium [6]. This type of dryer is extremely effective in reducing the moisture content of de-inking sludge, and moisture contents as low as 3 wt% have previously been achieved in this type of dryer. At full scale, the rotary dryer would be directly heated using exhaust gases from the CHP process (either engine exhaust gases or exhaust gases from a combustor downstream). The hot gases would come into direct contact with the wet material, driving off moisture.

Once the feedstocks (de-inking sludge and rejects) are dry they both require pelleting before lab scale gasification and pyrolysis processes. De-inking sludge pelletises readily and this is thought to be because of the clay like properties within the sludge which improves the binding characteristics. Larger sized briquettes of paper industry rejects and sludge wastes, which are better suited for industrial sized gasification processes, have also been demonstrated by a Danish based briquetting company and another UK based company Salvetch Ltd as shown in Figure 20.

Figure 20 Paper Industry waste briquettes produced by a Danish based company.

Rejects generated from paper mills are extremely heterogeneous therefore to avoid blockages or damage to the gasifier the reject stream must undergo some sorting in order to remove glass, metals, stones and other large object. Chlorine containing species must also be removed to prevent corrosion in the gasifier and in a downstream engine, also to eliminate the emissions of toxic dioxins produced from chlorine containing compounds. The US defines waste containing more than 1000 ppm chlorine as a hazardous waste, and therefore subject to strict dioxin control [89].

Processing wastes in this way has been demonstrated successfully by larger mills operating in Europe. One such mill is Smurfit Kappa's Roermond paper mill. Their 'Rofire' facility, built by Siemens, sorts, dry's and pelletises board mill rejects for use as a fuel. The end product is a refuse derived fuel (RDF) type pellet with a calorific value of 22-24 MJ/kg, an ash content of less than 7 %, and a size of 8 mm diameter by 15-20 mm length (Figure 21). The pellets produced at this site have a moisture content of less than 5 %, a total volatile content of 80 %, a chlorine content of less than 2 % and an overall density of 450 kg/m^3 [25].

Figure 21 Rofire (RDF) Fuel Pellets.

Another Dutch based company operates a “Foxfire” process which produces fluff rejects for use as an energy fuel for the cement industry. The process takes reject material from paper mills and processes it in the following way. Firstly an initial pre-mixing and pre-shredding of the rejects is carried out and the pre-shredding mainly removes steel and stones, the material then goes through a de-watering press, after de-watering there is further shredding, the material then passes through a transportation screw which has a magnet to remove steel objects, a secondary high frequency magnet removes non-ferrous metals, separated metals from the process are then sold on as scrap and this helps generate further income to the process. The fibrous material then passes into a hopper which acts as a buffer before a further drying stage. The final end product produced is in the form of reject fluff with a calorific value of 22-23 MJ/kg, a plastic content between 40-60 % and each component is approximately 10-20 mm long. High levels of chlorine are separated from the reject stream before further processing and this is carried out by near-IR-technology, however this is expensive [26]. The entire Foxfire process costs 2 million Euros to install and requires 70,000 tonnes of feedstock annually.

3.0 PREVIOUS WORK

3.1 Chapter summary and scope of review

This chapter analyses the current state of the art for the thermal conversion of paper industry wastes and similar feedstocks for energy production. The scope of the review is defined as the assessment of peer reviewed journals of previous research carried out on the gasification and pyrolysis of paper industry wastes (sludge and rejects) and similar compositional wastes. The overall objective is to assess what research has already been achieved within this field and to identify where there are areas for either improvement or further work.

The scope of this chapter also includes gasification and pyrolysis experiments of municipal solid waste or refuse derived fuels (RDF). These waste compositions are similar to rejects produced from paper mills (see Section 2.4) and so are considered relevant to this work. Other thermal conversion technologies or feedstock types are beyond the scope of this review. A summary and critical analysis of the finding from the review are presented in Section 3.3.

3.2 Literature review

Indirect steam gasification of paper mill sludge waste [10]

A study on the indirect steam gasification of paper mill wastes, mainly plastic rejects and short length fibres from a recycled fibre mill has been carried out by Durai-S et al, [10]. In this work the authors performed gasification experiments at 950 ^{o}C, with a reaction time between 4-10 hours in a continuous steady state experiment, using a fluidised bed reactor fitted with an internal steam heated coil. This coil was claimed to give improved heat transfer and an even heat distribution throughout the fluidised bed. The fluidised bed consisted of $CaCO_3$ (limestone) particles, used for the reduction of dioxin formation in the product gas. It was suggested that this was achieved because calcium material has the effect of absorbing HCl released during gasification.

The fluidising gas used was superheated steam generated from the downstream cooling of product gases. The particular feedstock used in the study consisted mainly of plastic reject material and short fibres expelled from a clarifier within a paper mill. The results of the study showed that the fluidised bed gasification performed well while handling high moisture, high plastic and high ash containing feedstocks. It was reported that dioxin formation of the product gases was within emission limits. The product gas was composed of approximately 80 % combustible gases, and had a calorific value of 13.7-15.2 MJ/Nm^3. This was the highest energy containing product gas produced from all reviewed studies of this type of feedstock.

The authors state that the main advantage of using fluidised bed gasification of paper wastes is that it does not require pelletisation and high levels of drying of the feed, and thus results in improved economics of the process through the reduction of feedstock pre-treatment costs. However the low density of fibrous material in the fluidised bed caused high levels of entrainment and this increased the amount of tar and char produced. High levels of tar in the product gas can severely affect the performance of an engine or gas turbine being used for electricity production. Closer control of gas superficial velocities were proposed as the solution; however there was no further testing of this.

One advantage of this process is that as superheated steam needed for the process is raised from the subsequent gas cooling, there is no need for an external heat input.

Energy and materials recovery from recycled paper sludge [27]

Frederick et al, [27] carried out two studies, one on the pyrolysis and the other on gasification of primary sludge generated from a recycled fibre paper mill. In the experiments the authors focused on producing a fuel gas for energy and reusing the residual ash residue as an admixture mineral for Portland concrete.

The authors performed the gasification experiments in a batch furnace over a temperature range of 500-900 ^{o}C, using air as the gasifying medium with equivalence ratio of 0.58 and a reaction time of 1 hour. Higher temperatures above 900 ^{o}C were reported to be effective in the calcination of kaolinite and $MgCO_3$ compounds and so reduced the overall yields of ash inorganic residues after gasification to approximately 45 % of the total dry sludge feed input. The ash was later tested for its properties as a concrete admixture mineral and was found to be slightly outside the required specification; however the authors reported that slight alterations to the temperature of the gasification reaction would remedy this, although this was not demonstrated.

The authors carried out a material and energy balance of the total gasification system and the thermal efficiency of the process, defined as the heating value of the product gas plus the energy available from steam generation divided by the total energy input at a reference temperature of 25 ^{o}C, was 69.8 %. It was stated that there was enough residual heat from product gas cooling to provide heat energy for both drying the sludge and pre-heating the air used for gasification. The product gas contained 17.1 vol% hydrogen and 5.4 vol% carbon monoxide the remaining composition of the gas was not stated. The gas produced had a gross heating value of 2.64 MJ/Nm^3 (dry basis) and the authors discussed that using oxygen-enriched air for the gasification reaction would help quite considerably to increase this value as it reduces the N_2 present. However using a gasifying agent with high oxygen purity would in turn incur higher operating costs of the gasification process.

For the pyrolysis reactions the authors [27] focused on producing a fuel gas for energy and then subsequently gasified the residual char solids to increase the overall efficiency of the process. Pyrolysis of the sludge was performed in a laminar entrained flow reactor (LEFR) at high heating rates and at temperatures of between 500 and 900 ^{o}C. The optimal pyrolysis temperature for conversion into useful hydrocarbon fuel gases was reported as 900 ^{o}C. This gave the smallest amount of residual char remaining and it also gave a conversion of approximately 60 % of carbon in the solid feed into a useful fuel gas, of which the main components were CO (35% v/v) and C_2H_4O (13% v/v). Performing pyrolysis at these higher temperatures was stated to result in the successful calcination of $CaCO_3$ and $MgCO_{3,}$ which makes up of a large fraction of the solid sludge initial composition. Negligible levels of NOx were formed at higher pyrolysis temperatures and less than 1 % of residual metals from the feed were found in the post cyclone filter. The char produced from the pyrolysis runs were subsequently gasified and it was reported that the residual fixed carbon in the char gasified readily. However no further data was collected on product gas composition or the total remaining ash yields.

Syngas yield during pyrolysis and steam gasification of paper [28]

Ahmed et al, [28] carried out two similar studies to that of Frederick et al, [22] on the pyrolysis and steam gasification of waste paper obtained from municipal solid waste. In this study both pyrolysis and steam gasification was performed in a bench scale batch reactor over a temperature range of 600-1000 °C. Pyrolysis conditions were achieved by purging the reactor with nitrogen and the gasification experiments were carried out with steam also using nitrogen as a trace gas.

The authors studied the product gas yield, composition and the energy thermal efficiency with time. The main findings of the study were that material destruction, hydrogen yield and energy yield are increased with gasification as compared to pyrolysis. The increase in hydrogen yields was attributed to the use of steam within the reactions, which promoted the forward water gas shift reaction. The effect of this could be seen by the gradual reduction in CO formation in the product gases as the hydrogen yield increased. This was due to the reduction reactions of char favouring the production of H_2 rather than CO. There was a direct relationship between the overall product gas yield and temperature. Temperatures above 900 °C produced the highest yields of product gas.

The superiority of the gasification conditions was said to be the result of residual char gasification reactions; they were found to be more effective at gasification temperatures than at pyrolysis temperatures and started at around 700 °C. High activation energies of the gasification process had the effect of increasing the overall gasification time. These activation energies were reported to be in the range 250-440 kJ/mol for gasification and 50-300 kJ/mol for pyrolysis. The activation energy of char was reported to be 149 kJ/mol.

Fixed bed gasification of steam treated food waste (FW) and municipal solid waste (MSW) [29]

Fixed bed gasification trials of municipal solid waste were carried out by Malik et al [29]. The study focused on small scale pilot plant fixed bed gasification trials of dried and pelleted wastes for the production of product gas for use in a small scale engine for electricity production. Pre-treatment involved the initial sorting of the wastes to remove pieces of glass, metals and other large objects. The remaining material consisting of mainly paper fibres and plastic were oven dried and pelletised for gasification. The characterisation of this feedstock revealed that it contained a high ash content in the region of 50 wt% on a dry basis and an initial high moisture content of 51 wt% on an as received basis. 32 wt% of the feedstock was paper and cardboard material. Ultimate analysis revealed a carbon content of 36 wt% and an overall gross heating value of around 10 MJ/kg (dry basis). Total yield of combustible gases from MSW gasification amounted to approximately 20 % however it is not known whether this was by mass or volume. The whole product gas had a gross calorific value of approximately 3 MJ/kg (dry basis) and the main combustible gas formed was CO. The authors noted that there was a significant time delay between the production of the gases to when they were analysed. The method used for gas storage was by inflatable balloons and this may have resulted in a significant loss of gas especially hydrogen which has the tendency to leak from gas balloons. This therefore places doubt on the accuracy of the gas analysis. The gas was reported to have fuelled a 2 kWe gas engine successfully with no reported problems.

The authors concluded that the use of MSW as a fuel for fixed bed gasification was possible; however the high ash content of the feedstock would make the process problematic at a large scale.

Co-gasification of biomass mixed with plastic wastes [30]

A study on the co-fired gasification of biomass with plastic wastes was carried out by Pinto .F et al, [30]. Steam gasification experiments were performed in a fluidised bed gasifier at temperatures up to 900 °C. The feedstock used was pine wood mixed with varying levels of plastic polyethylene material. It was reported that the use of dolomite within the fluidised bed acted as a strong catalyst for the reduction of tar formation. Furthermore results showed that using gasification temperatures above 900 °C helped to further reduce these tars by thermal cracking. High temperature also decreased the formation of hydrocarbon gases which in turn reduced the overall calorific value of the gases produced. Co-firing high ash containing plastics with wood improved gasification reactions compared with only plastic waste gasification, as the low ash content in the wood helped counterbalance negative effects of high ash containing materials. The authors gave two of the main problems with the gasification of plastics as feeding difficulties and clogging from deposits of black carbon powder which agglomerate within cooler areas of the gasifier such as the cyclone, although the pelletising of the plastics was found to overcome this problem.

Overall the results from the study showed that co-firing plastic wastes with biomass was achievable with plastic wastes up to 60 % (w/w). The presence of polyethylene favoured the formation of H_2 and decreased the formation of CO. The effect of steam/waste ratios used in the experiments was found to have no significant effect on the reactions, although it was suggested that ratios below 0.6 were not advisable due to loss of fluidisation in the bed. The highest value of gas yield found was 1.96 Nl/g dry ash free of which the main gases were H_2 (50% v/v) and CO (30% v/v). A gas gross heating value of 18.3 kJ/Nl was obtained when 60 % of polyethylene was used in the waste mixture.

Gasification of biomass wastes and residues for electricity production [31]

Faaij .A et al, [31] developed a model using the ASPEN ^{plus} process simulation package for the gasification of biomass wastes and residues for electricity production. The system selected for the study looked at the atmospheric circulating fluidised bed gasification of wastes coupled to a combined cycle gas turbine to generate electricity. The exact feedstock modelled was wood, grass, organic, domestic wastes and sludge. The compositions of the product gas were derived by laboratory scale experiments and the results served as the basis for subsequent model calculations. It was stated that high ash containing fuels such as sludge and domestic wastes can only be used to a limited extent and must be mixed with cleaner fuels such as wood. The ash limits proposed for high ash containing materials were 10-20 %. Overall the fuels studied produced a net calculated efficiency to electricity production of approximately 35-41 %. Estimated capital cost for the full implementation of such a plant was between 1500-2300 ECU per kWe installed. The operating costs of the full system including logistics and the purchase of raw materials, which were in some cases negative, and varied from –6.7 to 8.5 ECUct/kWh. Negative values imply a gate fee i.e. the plant is paid to take waste feedstocks. The environmental performance of the system was reported to meet the strict emission standards for waste incineration of all feedstocks in the Netherlands.

Overall the authors concluded that processing wastes in this way does not entail fundamental technical and economic barriers and is a promising technology for the future treatment of wastes.

Waste management for pulp and paper production in the European Union [3]

A detailed review of available technologies for processing sludge wastes was recently carried out by Monte .M.C et al, [3]. The study highlighted that due to increased taxes, landfills are quickly being phased out as a final destination for wastes in Europe. The current preferred method adopted by paper mills for processing paper wastes is by incineration coupled to an energy recovery boiler to raise process steam, in turn to run either a steam turbine for electric generation or to be used directly in the paper making process for drying paper sheets. However raising steam by combustion only for electricity production is an inefficient process especially at small scales particularly for smaller mills, which generate small quantities of wastes. It also requires a large capital investment and problems regarding incineration emissions and boiler corrosion are of concern. It was noted by the authors that other options such as pyrolysis, gasification, land-spreading, composting and ash reuse for building materials are being applied, although pyrolysis and gasification are still in the early research and development stages and much further work is needed for their optimisation. Emphasis was placed on the uncertainty of the environmental impact of such processes as well as the economics associated with them. The authors concluded that there remains much scope for further work on thermal treatment methods like pyrolysis and gasification.

Characterisation and pyrolysis behaviour of different paper mill waste materials [32]

The characterisation and pyrolysis behaviour of different paper mill sludge waste materials was studied by Mendez et al, [32]. In this study the authors used a variety of characterisation techniques to study the pyrolysis behaviour of wastes generated from both a virgin pulp mill and a recycled fibre mill. The main objective of the study was to characterise eight different paper mill wastes all from different origins using SEM, FTIR, DRX, and TGA analysis techniques. It was stated that the pyrolysis of paper sludge obtained from a recycled paper mill contained much higher levels of $CaCO_3$ and clays than that from a virgin pulp mill. It was stated that recycled fibre paper mills produce sludge wastes with elevated chemical stability of organic matter and this is the result of a high content of cellulosic fibres in the waste. However constant successive recycling of these fibres by paper mills leads to their gradual degradation. This was proven by XRD analysis. TGA analysis indicated that the presence of mineral matter and the degradation of cellulosic fibres significantly influence the pyrolysis behaviour. Furthermore weight loss of waste materials from recycled paper mills continues at higher temperatures greater than 500 ^{o}C. This was stated as being the result of the dehydration of kaolinite and the dehydration of carbonates and these results correspond well to similar studies performed by Frederick et al, [27].

Thermal processing of paper sludge and characterisation of its pyrolysis products [33]

The thermal processing of paper sludge and characterisation of its pyrolysis products was carried out by Strezov et al, [33]. In this study the authors investigated the properties of paper sludge during slow pyrolysis over a temperature range of 100-700 ^{o}C with a heating rate of 10 ^{o}C/min. The results showed a direct relationship between temperature and gas yield. It was reported that the highest gas yields were of CO and CO_2 and these amounted to almost 25 % of the paper sludge dry weight loss at 500 ^{o}C. Only small amounts of H_2, C_2H_4 and C_2H_6 evolved at this temperature equating to around 1 % of the total dry mass loss. When pyrolysis runs were performed at higher temperatures the yields of these gases also increased although this was never greater than 2 wt%. Significant amounts of condensable organic vapours were produced at 500 ^{o}C and this amounted to approximately 40 wt%. These yields decreased when the pyrolysis runs were carried out at higher temperatures. The oils formed were further characterised by liquid GC-MS and were found to be mainly organic acids with the major contribution being linoleic acid, 2,4-decadienal acid and oleic acid.

The authors further reported that the char solid residue produced at 500 ^{o}C amounted to approximately 36 wt% with a proximate analysis of 56 % ash, 12 % residual volatile matter and 32 % fixed carbon. Had the experiments been performed at pyrolysis temperatures above 750 ^{o}C the ash content may have been reduced by the decomposition of carbonates releasing CO_2. The char residue had an overall gross heating value of 13.3 MJ/kg and the total energy balance of the product gas formed during the process had a gross heating value only slightly higher than the thermal energy input required to pyrolysis the dried paper sludge.

Materials from paper industry waste materials and their use in CU (11) removal from water [34]

Mendez .A et al, [34] carried out a study on the absorbent materials from paper industry wastes and their use in removing Cu (11) from water. During this study the authors carried out characterisation by TGA pyrolysis of de-inking sludge. The pyrolysis run was carried out at a temperature of 650 ^{o}C. The results showed that there were two significant decomposition rates; the first was at a temperature of 348 ^{o}C which corresponded to the decomposition of cellulose and dehydration of clays. The second significant rate of mass loss was at 787 ^{o}C and this was attributed to the decomposition of carbonates. This corresponds well to the findings from similar studies within the literature Frederick et al, [27] Mendez et al, [32]. The char yield from the pyrolysis runs was approximately 60 wt%. The authors also explored the effect of heating rates on the overall char yield. The two heating rates explored were 3 ^{o}C/min and 10 ^{o}C/min. The effect observed was that the char yield decreased when the heating rate increased, although only by 2 wt %. Nevertheless these results demonstrated that varying pyrolysis heating rates does have an effect on the overall char yield. In light of this it was proposed that if fast pyrolysis was performed at heating rates of up to 1000 ^{o}C/s then the observed effect may be that the char yield decreases even further. This would remove the need for post thermal treatment of the residual char by alternative thermal conversion techniques such as gasification and it would also have the effect of producing lower quantities of solid residues for disposal, thus improving the economics of a proposed paper waste to energy system.

High efficiency electricity and products from biomass and waste; Experimental results of proof of principle of staged gasification and fuel cells [35]

Oudhuis .A et al [35] of ECN biomass and ECN fuel cell technology studied the pyrolysis followed by gasification of willow biomass and pelletised paper industry reject waste generated from a secondary fibre board mill (Rofire pellets), in order to evaluate their potential use as fuels for electricity production. This system also integrated the use of a solid oxide fuel cell (SOFC).The tests were carried out using an ECN two stage gasifier “Pyromat” and the connected fuel cell was a downscaled Sulzer HEXIS 1 kW_e SOFC stack. The full experimental set up consisted of a feed hopper, horizontal screw pyrolysis reactor, a secondary oxygen blown (tar cracker) gasifier, a series of cleaners for upgrading producer gas quality, consisting of a wet scrubber, dust filter, condenser and filters both (ZnO and carbon) and finally a solid fuel oxide stack (SOFC) which was a 1 kWe stack containing 5 cells of 60-100 cm^2 each with state of the art electrolyte supported with 3 plate interconnects.

The operating conditions were set up to test both Rofire and willow feedstocks at a feed rate of 3 kg/h. The process was carried out at atmospheric pressure, pyrolysis temperature of 550 ^oC, tar cracker temperature of 1300 ^oC and a SOFC utilisation of 80 %. The results from this study showed successful operation for several days on both willow and Rofire feedstocks. $H_{2,}$ CO and CO_2 were by far the largest fraction of gases formed, 36.7 vol%, 33.3 vol% and 28.3 vol% respectively. Other gases were also found such as CH_4 and N_2 but these were in much smaller volumes giving a combined total of approximately 2 vol%.

The electrical efficiency of the SOFC was calculated and found to be between 35-41 %; the authors outlined that at full scale the electrical efficiency would be comparable to a coal fired power plant although it is not clear if this would include the utilisation of char generated from the process. Some negative aspects were encountered such as soot formation in the product gas, slight degradation of the SOFC over time and decreased electrical efficiencies with the addition of 50 vol% steam. Overall the objective to prove in principle that the use of secondary fibre paper mill rejects for electricity generation by pyrolysis was achieved although further work is required in order to optimise process efficiencies with the utilisation of char generated from the process.

Stoichiometric, mass, energy and exergy balance analysis of counter current fixed-bed gasification of post-consumer residues [36]

Rao .M et al [36] studied the air blown counter current (updraft) fixed bed gasification of municipal solid waste refuse derived fuel (RDF) pellets. The mass and energy results obtained were compared with similar results obtained from the gasification of biomass and other waste feedstocks.

The experimental set up was comprised of a throatless fixed bed counter current air blown gasifier with relevant tubing for air in and gas out as well as suitable sampling probes for gas and tar analysis. The experimental procedure consisted of filling the gasifier with a known amount of charcoal and feedstock, igniting the gasifier and allowing steady state to be achieved. This was determined by a constant gasification temperature. Once steady state had been reached samples of gas and tar were extracted from the gasifier to be analysed. Residual char and ash was removed for further analysis after the reaction was terminated. Results showed a product gas of 25.5 vol% CO, 18.5 vol% H_2 49.7 vol% N_2 5.3 vol% CO_2 and 1 vol % CH_4. The tar yield was observed to be within the range 30-150 g/Nm^3 however the authors did not give a specific figure.

Overall the authors concluded that gasification can be adapted to a wide variety of solid fuels, as the cold gas efficiency with the RDF pellets reached over 73 %. The authors reported some problems with ash as it was too high within RDF pellets (approximately 24 wt%). The sensible heat produced by RDF pellets accounted for 6-7 % of the total energy and 2-3 % of the total exergy. This indicated the total energy output of the fuel was mostly in the chemical form with little wasted as sensible heat and so it was determined that RDF pellets can be a competitive fuel for thermal energy conversion through gasification.

Recovery of plastic wastes from dumpsites as refuse-derived fuel and its utilization in small gasification system [37]

Chiemchaisri. C et al [37] investigated the use of refuse derived fuels sourced from Nonthaburi in Bangkok, Thailand as a gasification feedstock. Waste samples were collected and characterised for their physical and chemical properties. The waste was composed of mainly a mixture of paper, plastics, metals and rubber. Polyethylene was found to be the most abundant plastic and the plastics had a volatile content of 72-92 wt% and the moisture content was also very low, 0.85-3.46 wt%. After analysis the plastics were found to contain 0.65-1.26 wt% chlorine and 0.22-0.36 wt% sulphur with a gross heating value of 27.5-38.5 MJ/kg. The feedstock was pre-treated which involved the manual removal of glass rubber and large stones. The feedstock was then further screened by a trammel screen to remove soil and earth. The recovered fraction of feedstock was shredded into small pieces of 10 mm. The shredded feedstock, composed mainly of plastics, was mixed with a cassava root stem and briquetted. Different mixtures of the plastic wastes and the cassava root stem were produced at varying weighted ratios. These briquettes were then left to dry for 5 days at ambient conditions. The optimum blend of RDF briquette was found to be a weight ratio of cassava root to plastic wastes of 0.8 or higher. The optimum plastic content in the briquettes was 55.56 wt% with a gross heating value of 26 MJ/kg. The dried RDF briquettes were then utilised in a small scale downdraft gasifier operating under atmospheric pressure with gas flow and temperature being monitored. Air was supplied to the gasifier with a varying flow rate of between 30 m^3/h and 50 m^3/h. The gas produced was sampled and analysed for its composition in a gas chromatograph and cold gas efficiency was also obtained. Gasification temperatures averaged 504 ^{o}C with a product gas composed mainly of CO_2, CO, CH_4 and H_2. The heating value of the gas obtained was 1.49 MJ/Nm^3 on a dry basis and the cold gas efficiency was 32 %. When the air input was increased to 50 m^3/h the gasification temperature increased to 604 ^{o}C and the CO and CH_4 content in the gas increased. The calorific value of the gas also increased to 1.76 MJ/m^3 on a dry basis with a cold gas efficiency of 66 %.

Gasification characteristics of combustible wastes in a 5 ton/day fixed bed gasifier [38]

Lee. S et al [38] studied the gasification behaviour of biomass and plastic wastes in an industrial scale fixed bed gasifier. This study compared the gasification characteristics of wood chips and residue plastic fuel (RPF). The authors analysed the effects of bed temperature (500-1000 ^{o}C), oxygen concentration (0-21 vol%) and solid/oxygen ratio (0-15) on product gas composition calorific value and the cold gas efficiency. Each run was carried out over duration of 10 hours. The feed rate and oxygen flow rate were used to control the bed temperatures. The industrial waste feedstock was composed of 35 % biomass and 65 % plastics with a gross heating value ranging between 18 MJ/kg and 26 MJ/Kg (dry basis). The wood chips and the RPF were gasified separately and their results were compared. Due to the low moisture content and higher calorific value in the RPF, the bed temperature in RPF gasification was more stable than for wood chip gasification. The bed temperatures of the RPF gasification experiments were higher than wood chip gasification due to the lower moisture content and higher calorific value of the plastic feedstock.

The product gas was composed of 33.9 % CO, 26.1 % H_2, 10.7 % CH_4, and 29.2 % CO_2 v/v% with a calorific value of 11 MJ/Kg (dry basis) for RPF gasification and 34.4 % CO, 10.7 % H_2, 6 % CH_4, and 48.9 % CO_2 (v/v%) with a calorific value of 8 MJ/Kg (dry basis)for wood chip gasification. The RPF gasification showed a solid/oxygen ratio of 7 produced the highest calorific value gas at 13 MJ/kg (dry basis) while wood chips gasification solid oxygen ratio of 15 produced the most valuable gas with a calorific value of 9.5 MJ/Kg (dry basis).

Gasification of biomass/high density polyethylene mixtures in a downdraft gasifier [39]

Bacaicoa .G et al [39] studied the feasibility of operating a gasification plant using biomass/high density polyethylene mixtures and investigated the characteristics of the gas produced.

Gasification was performed in a downdraft gasifier with a nominal feed rate of 50 kg/h. During the experiment, steady state was determined when the temperatures in the oxidation zone and reduction zone were relatively constant. Air flow rate into the gasifier determined bed temperatures and therefore the reaction rate and the solid feed rate. Air flow rate variation was the only variable in the experiment. During the experiment, different mixtures of biomass/HDPE were used. The results showed that the consumption of the mixture of wood chips and HDPE increased as the air flow rate increased. The overall gas produced from the gasification increased with air flow rate.

Gasification of HDPE showed that CH_4 and C_2H_4 increased with HDPE ratios but other hydrocarbon species such as C_2H_2 and C_2H_6 concentration were not affected. CH_4 levels increased due to the HDPE cracking reactions. The CO produced when 10 % HDPE was gasified with 90 wt% wood was reported to be approximately twice as much as that of pure wood gasification. As the air flow rate and concentration of HDPE increased, the composition of CO increased and CO_2 decreased. The CO/CO_2 ratio showed a clear decrease when the feed was composed of 15 % HDPE and 85 % wood. It was suggested that HDPE favoured the Boudouard reaction against the water gas shift reaction due to higher carbon or hydrocarbon radicals generated from cracking. As the temperature increased in the reduction zone, more CO was produced while the H_2 levels stayed relatively constant suggesting that the Boudouard reaction was favoured as opposed to the water gas shift reaction at higher temperatures.

The heating value of the gases produced increased with feed rate and the heating value increased even further when the ratio of HDPE to wood increased. HDPE feed increased the concentration of CH_4, C_2H_6, C_2H_4 and C_2H_2 in the product gas. The heating values of the product gas from the 10 % HDPE and the 15-17 % mixture were 5.6-6.1 MJ/Mm^3 and 6.1-6.9 MJ/Nm^3 (dry basis) respectively.

Prediction of Producer gas quality for two stage gasification of selected waste feedstocks [40]

This paper focused on the development of a thermodynamic model to examine the appropriate conditions for the gasification of landfill gas, waste oil, refuse derived fuel (RDF), and municipal solid waste (MSW). The MSW utilised was from two different locations with MSW 1 obtained from low-income Asian countries and MSW 2 obtained from high-income countries. The gasification conditions assumed were those of a two stage gasifier with a nominal value of 15 tonnes per hour. The first stage was a free fall reactor and the second was a fixed bed re-former where gases formed from stage one flowed upwards through the second stage for reforming reactions to occur. The product gas was assumed free of light and heavy condensable hydrocarbons (tars).

The gasification took place in either the presence of oxygen or oxygen-steam mixtures. The oxygen or steam-oxygen mixture injection into the system was controlled by the model to ensure ratios between waste and oxidising agent in each stage was optimum to achieve desired temperature. The model assumed that the heat loss across the reactor was 10 % which is typical for industrial sized gasifiers of this type. The results obtained during gasification suggested that the oxygen volume per kg of waste should be maintained at an optimum value to ensure that the highest percentage of CO and H_2 was achieved. The injection of higher ratios of oxygen results in lower CO and H_2 percentage. The model also showed that CO and H_2 percentage was inversely proportional to the ratio of oxygen required for complete combustion of the feedstock.

The gasification of MSW 1 showed very low CO and H_2 which was assumed to be negligible. However, when MSW 1 was gasified with landfill gas the H_2 and CO percentage increased significantly to 19 vol % collectively.

Comparing the product gas power potential obtained from MSW 1 and MSW 2, MSW 2 was a better option due to its higher energy content. MSW 2 from high income countries was reported to contain significantly higher ratios of plastics and paper compared to MSW 1 which increased its energy content.

Energy and resource utilisation of de-inking sludge pyrolysis [41]

Rui Lou et al [41] investigated the pyrolysis characteristics of de-inking sludge for potential energy generation. The main aim of this study was to thermally characterise the properties of the pyrolysis products produced from de-inking sludge and also to investigate the application of de-inking sludge through a pyrolysis technique as an energy resource. The de-inking sludge utilised for this study was air dried, ground and oven dried at 105 ^{o}C for 6 hours. The particles were maintained at 40-60 mesh.

Three different techniques of pyrolysis were applied in this study. Thermo-gravimetric analysis (TGA) at 800 ^{o}C using nitrogen, Py-GC/MS at 800 ^{o}C under a helium atmosphere, and pyrolysis in a static tubular furnace using nitrogen. The experimental results revealed that de-inking sludge contained 41.46 % ash with a low gross heating value of 10.35 MJ/kg (dry basis). The proximate analysis of the de-inking sludge and its calorific value was compared to other feedstocks and its combined volatiles and fixed carbon content were very similar to maize straw.

From the TGA runs it was found that the pyrolysis process of de-inking sludge is made up of 3 steps; evaporation, devolatisation and decomposition of residues. The evaporation occurred at 71 ^{o}C, pyrolysis peaked at 361 ^{o}C and the combustion of fixed carbon and decomposition of calcium carbonate was observed to peak at 772 ^{o}C. When the Py-GC/MS was used for the experiment, it was found that different temperature ranges favoured different pyrolysis products. Benzene and its derivatives peaked at 85.23 % when the temperature was 800 ^{o}C.

When experiments were carried out in a tubular furnace a three phase product mixture consisting of incondensable gas (29.78 wt%), condensed liquid (24.41 wt%) and solid (45.81 wt%) were found to be the products. The gaseous products were composed of 31.6 % CO, 21.53 % CO_2, 19.54 % CH_4, 9.62 % C_2H_4, 17.71 % H_2 by volume. The decomposition of $CaCO_3$ contributed to the CO_2 in the gaseous products. The solid product obtained was composed of mainly inorganic material, mainly $CaCO_3$, and the pyrolysis oil produced was made up of mainly benzene, toluene, ethylbenzene and xylene. The authors concluded that the oil and gas produced can be used as an energy fuel and the $CaCO_3$ can be reused as paper filler.

3.3 Summary of literature review

From the review of the literature it has been identified that there still exists a requirement to investigate in detail a CHP system based on advanced thermal conversion techniques for the conversion of secondary fibre paper mill de-inking sludge and rejects. Very limited research has been carried out on the gasification of these specific waste streams, and the work which has been done has entirely focused on fluidised bed technologies with no work at all on fixed bed downdraft gasification. This can be explained by the problems for downdraft gasifiers of feedstocks with very high ash content such as de-inking sludge, and also feedstocks with high plastics content such as rejects which can lead to agglomeration above the throat [39]. One approach would be to co-fire these materials with conventional biomass feeds. Some workers have looked at this for general waste plastics, but there has been no attempt to take this approach specifically with paper industry wastes [39].

The review of the literature has also identified the pyrolysis of de-inking sludge as an area to be further investigated. Some studies have looked at the pyrolysis of secondary fibre paper mill rejects for energy generation such as the work carried out by ECN [35]; other studies have investigated the characteristics of the products formed from de-inking sludge pyrolysis such as the work of Rui Lou et al [41]. The authors used laboratory analytical techniques, mainly TGA and py-GC-MS, to analyse the pyrolysis products formed from de-inking sludge, and concluded this had potential for energy production but that considerable further work would be required at larger scale. Since then no further attempt has been made to up-scale or further investigate the findings from this study. Furthermore there has been no published literature which investigates the use of de-inking sludge in intermediate pyrolysis systems and there also exists no literature on the use of de-inking sludge pyrolysis products as a fuel source for engines. Therefore there remains much scope for further research in these areas.

4.0 PROJECT AIM, OBJECTIVES AND PLAN

4.1 Project aim

The aim of this work was to investigate the possibility of utilising the wastes produced by secondary fibre mills as an energy source in a site-based CHP system based on the advanced thermal conversion technologies pyrolysis and gasification.

4.2 Project objectives

1) Develop methods to condition selected paper industry wastes to enable thermal conversion. This stage required detailed characterisation of wastes in terms of moisture content, and degree of contamination with non-calorific substances e.g. metals and ash. This method required the initial drying and conditioning of residues.

2) Through trials, quantify the energy recovered through both fixed bed downdraft gasification and intermediate pyrolysis of selected wastes (de-inking sludge and rejects). This stage examined the characteristics and performance of each process in detail including the yield and quality of the gaseous, liquid and solid products.

3) Identify the advantages and disadvantages of each process in order to evaluate the optimal process route for each participating mill.

4) Carry out industrial implementation studies for full scale schemes to be integrated at each of the three participating mill sites. This involved detailed design specifications including full mass/energy balances, and includes consideration of the use of waste heat from each site, waste conditioning strategies (sorting, sizing, drying and pelleting), and options for residue management.

5) For each of the three participating sites, carry out an economic evaluation of the proposed scheme, and also a quantification of environmental benefits in terms of both waste reduction and sustainable energy development.

4.3 Project plan

Four different types of waste generated from the three secondary fibre paper mills have been explored in this work. These are de-inking sludge and pulper rejects generated from AN mill, de-inking sludge, pulper rejects and co-form rejects (dry and wet wipes) generated from KC mill, and pulper rejects generated from SSK mill.

The experimental work plan began with the pre-treatment of these selected waste streams. Each feedstock required some degree of pre-treatment as previously described in Section 2.7.

To achieve the pre-treatment 500-1000 kg of each feedstock was acquired from each participating mill. The feedstocks as received required extra conditioning such as drying, sorting and sizing in order to be processed by subsequent thermal conversion processes. In order to achieve this external companies were approached (Salvetch Ltd and Envirosystems Ltd).

Envirosystems Ltd carried out the drying of de-inking sludge, which were then transported to Aston University for pelletisation. Salvetch Ltd carried out the drying, sorting, shredding and pelletising of the pulper rejects and co-form rejects. After conditioning of the wastes subsequent characterisation was carried out at Aston University.

After pre-treatment, the feedstocks were characterised by a full proximate, ultimate and HHV analysis to determine the thermo-chemical properties. A thorough analysis of the characterisation results helped to better understand the thermal behaviour and energy products from each feedstock when subjected to thermal conversion processes either by pyrolysis or gasification.

To achieve the second objective of thermally converting the waste residues, access to suitable test facilities was necessary. Aston University has a 10 kg/h laboratory-scale downdraft gasifier which was the main test vehicle for the gasification studies. Aston also has a 20 kg/h intermediate pyrolysis reactor (the “Pyroformer”) and this was the main test vehicle for the pyrolysis studies.

Co-fired blends of pulper rejects, de-inking sludge and wood chips were selected for gasification experiments. Rejects are ideal candidates for gasification trials as they contain low ash contents and high calorific values which would provide sufficient energy for the exothermic oxidation reactions to occur. Rejects also contain high volatile contents from the plastics present; when cracked these volatiles form product gases with elevated heating values. However rejects often suffer from agglomeration caused by the melting of plastics; therefore it was necessary to blend rejects in lower proportions with wood chips and de-inking sludge to resolve this issue. AN rejects were tested first; this was followed by SSK rejects and then finally KC rejects. The rejects were blended in varying proportions with wood chips and de-inking sludge to determine the optimum gasifiable blend for each mill.

De-inking sludge was selected for pyrolysis experiments. De-inking sludge was better suited to be processed by pyrolysis as it contains very high ash levels. High ash containing fuels perform poorly in downdraft gasifiers: firstly as suitable gasification temperatures cannot be reached without the support of a higher calorific value support fuel; and secondly due to the need for constant ash removal. On the other hand, the pyrolysis process utilises an augur type reactor known as the “Pyroformer” [46]. This reactor has the ability to continuously remove the high ash content of the de-inking sludge. Likewise as pyrolysis processes are endothermic the unit is heated indirectly by electrical heating bands therefore the need for a higher calorific support fuel is removed.

The pyrolysis products were characterised and the pyrolysis oil produced was blended with biodiesel in varying proportions and subsequently tested in a diesel lister engine to determine the performance, emissions and combustion characteristics of the new fuel.

On completion of the experimental work an assessment of each process was carried out in order to determine the optimal process route for scale up at each of the participating mills this satisfied the requirements of the third objective.

To achieve the forth objective of carrying out an industrial implementation study of the proposed processes for each of the candidate mills, a detailed site survey at each mill was required. This stage included individual site visits to assess current on-site energy systems and to gather relevant data in terms of flow sheets mass and energy balances. Once all data was acquired a specific process tailored to meet each of the mills current energy demands was designed. The system proposed aimed to minimise costs by best utilising the infrastructure currently available on-site and adapting each process to fit accordingly.

To achieve the final objective a full economic appraisal was performed on each of the systems proposed in the industrial implementation studies. The aim was to evaluate each process in terms of capital investment, operating costs and payback time to justify whether each installation would be economically feasible. This stage of work also established what the environmental benefits would be if such installations were implemented at each site.

5.0 EXPERIMENTAL METHODS

5.1 Chapter summary

This chapter describes the experimental methods used in this study. It begins by describing the types and quantities of feedstocks received. It then leads on to explain in detail how these feedstocks are pre-treated, characterised and thermally converted by fixed bed downdraft gasification and intermediate pyrolysis, detailing the procedure and equipment to be used. It also gives details on the methods and equipment to be used to analyse the products from each experimental run.

5.2 Feedstocks

5.2.1 De-inking sludge

Approximately 700 kg of de-inking sludge was received from AN mill and approximately 150 kg of de-inking sludge was received from KC mill. The de-inking sludge was de-watered by screw presses at each mill before being supplied. De-inking sludge as received contained an initial moisture content of approximately 35 wt% for AN mill and 45 wt% for KC mill and was of a grey flaky granular form.

5.2.2 Pulper rejects

Approximately 1 tonne of pulper rejects was acquired from each participating mill. The material was extracted directly from each mill's pulper waste compactors, and as received contained a very high moisture content averaging 55 wt%.

5.2.3 Co-form rejects (dry and wet wipes)

Approximately 500 kg of co-form rejects (both dry and wet wipes) was acquired from KC mill. The dry wipes were delivered in reels and the wet wipes were delivered within their plastic packaging. The moisture content of the wet wipes was approximately 70 wt% as received and that of the dry wipes was < 10 wt%.

5.2.4 Mixed wood chips

Approximately 500 kg of mixed wood chips was obtained from a local midland based wood fuel supplier (Midlands Wood Fuels Ltd) and was composed of mixed UK forest wood (mainly spruce) with an as received moisture content of approximately 26 wt%. After chipping the wood was of approximate dimensions 15-40 mm length by 10-30 mm width and 1-5 mm thickness.

5.3 Feedstock pre-treatment methods

5.3.1 De-inking sludge pre-treatment

De-inking sludge feedstock as received contained a moisture content of approximately 35 wt% from AN mill and 45 wt% from KC mill. It was further dried to a moisture content of < 3 wt% by an external company Envirosystems ltd using a rotary drum drier at an inlet gas drying temperature of 1000 ^{o}C [87]. The process was carefully controlled to prevent de-volatilisation of the de-inking sludge. The dried sludge was in a dry flaky form, suitable for pelletisation. This was achieved using a roll and die 9PK-200 7.5 kWe motorised pelletiser with total capacity of 100-150 kg/hr throughput. The pellets formed were of dimensions 6 mm diameter by 15 mm length (Figure 22).

Figure 22 Dried de-inking sludge pellets from AN mill.

5.3.2 Rejects pre-treatment

Reject material as received contained a very high moisture content averaging 55 wt%. The pre-treatment of rejects was carried out by an external company, Salvetch ltd.

Before pre-treatment, Salvetch characterised each feedstock to quantify the general compositions as received. AN mill rejects consisted of 18 wt% plastics (mainly hard CDs, bottles etc.), 64 wt% fibres and 18 wt% metals. The SSK pellets consisted of 15 wt% plastics (mainly soft film, and wrap), 77 wt%

fibres with the remaining 8 wt% being metals. The KC mill rejects were stated as 20 wt% plastics (soft and hard), 74 wt% fibres and 6 wt% metals. The KC co-form rejects were composed of 30 wt% plastics (polypropylene) and 70 wt% fibres.

The pre-treatment began with the initial manual sorting of the material to remove non-ferrous metals such as aluminium cans, glass bottles, stones and other large objects. The rejects were then further sorted on lines with overband metal detection to remove other ferrous metals such as staples and paper clips. The residual material consisting of mainly mixed plastics and fibres was then size reduced using an industrial shredder, and then hot air dried for moisture reduction to < 20 wt%. The rejects were then pelletised using an industrial pelletiser with a 6 mm die and a compression ratio of 9:1. The pellets were subsequently dried further to a moisture level of approximately 5-8 wt%, and given a 'consolidation' re-pelleting to insure their integrity.

The final product was approximately 500 kg of pelletised pulper rejects from each mill and 250 kg of co-form rejects from KC mill. The final composition of the pellets were approximated to 22 % plastics (mainly hard) and 78 wt% fibres for the AN mill rejects, 16 wt% plastics (mainly soft) and 84 wt% fibres for the SSK mill, 21 wt% plastics (mainly soft) and 79 wt% fibres for the KC mill rejects and 30 wt% plastics (polypropylene) and 70 wt% fibres for KC co-form pellets. The pellets had a size range of 6 mm diameter by 15-20 mm length and each had an overall bulk density of 494 kg/m^3. An example of the reject pellets is shown in Figure 23.

Figure 23 SSK mill reject pellets

5.3.3 Wood chip pre-treatment

Wood chips as received contained a moisture content of 26 wt%. They were oven dried for a period of 12 hours at 70 ^{o}C in a Funditor tray drying oven with a maximum capacity of 20 kg. The average moisture content after drying was approximately 9 wt%. The wood was further sieved using a 2 mm mesh to remove fines. No further pre-treatment was necessary before gasification experiments.

5.4 Feedstock characterisation

The feedstocks were characterised for proximate and ultimate analysis and their gross heating values.

5.4.1 Moisture content (all feedstocks)

All moisture contents of the feedstocks were determined using a moisture analyser (Sartorius MA35) with a programmed temperature of 105 ^{o}C. Total moisture content was determined gravimetrically by measuring the total weight loss of solid sample with increasing temperature until no further weight loss was measured at the programmed temperature.

5.4.2 Proximate analysis (de-inking sludge)

De-inking sludge was characterised by a proximate analysis to determine the moisture, volatiles, fixed carbon and ash present. This was carried out by Thermo Gravimetric Analysis (TGA) in a Perkin Elmer Pyris 1 TGA device with auto sampler. Before analysis all samples were prepared in accordance with ASTM E1756-01. Approximately 5 mg of dried de-inking sludge was loaded into a tarred crucible and pyrolysis of the sample was carried out under an inert atmosphere of N_2 with a temperature programme of:

- Heating from ambient to 50 °C at heating rate 5 °C/min
- Hold for 5 minutes at 50 °C
- Heating from 50 °C to 105 °C at heating rate 5 °C/min
- Hold for 5 minutes at 105 °C
- Heating from 105 °C to 900 °C at heating rate 25 °C/min
- Hold for 15 minutes at 900 °C
- Cooling to ambient at cooling rate 25 °C/min

The weighted moisture content was determined at 105 °C, total fixed carbon was determined as the weight of solids after cooling and the total volatile content was obtained by difference.

The total ash content of de-inking sludge was determined by TGA combustion under the same programme temperature, using a purged atmosphere of air. After cooling the residual weight of ash was determined and recorded as a percentage of the original sample.

Proximate analysis of the reject pellets and wood chips was carried out by Marchwood Scientific Services Ltd. Due to the heterogeneous nature and variability of the materials, a sample size of 200 g was used.

5.4.3 Ultimate analysis (de-inking sludge)

Samples of dried de-inking sludge were analysed externally by Medac Ltd using a Carlo- Erba EA1108 CHNS-O analyser by total oxidation. Elemental compositions (C H N O, S, and Cl) were quantified.

Proximate analysis of the reject pellets and wood chips was carried out by Marchwood Scientific Services Ltd. Again due to the heterogeneous nature and variability of the materials, a sample size of 200 g was used.

5.4.4 Gross heating value (HHV)

The gross heating value in (MJ/Kg) of the dried de-inking sludge was determined using a Parr 6100 bomb calorimeter, and was verified using the unified correlation for fuels (Equation 6) developed by Channiwala et al [42, 43].

$$HHV\ \left(\frac{MJ}{kg}\right) = 0.3491\ (C) + 1.1783\ (H) + 0.1005\ (S) – 0.1034\ (O) – 0.0151\ (N) – 0.0211\ (A) \quad (6)$$

C, H, S, O, N and A (ash) are the mass fractions from the ultimate analysis expressed as percentages.

Gross heating value of the reject pellets and wood chips was measured by Marchwood Scientific Services Ltd. Again due to the heterogeneous nature and variability of the materials, a sample size of 200 g was used.

5.4.5 ASTM Standard test methods for de-inking sludge characterisation

In addition to the proximate, ultimate and ash analysis, de-inking sludge was further characterised to determine the total solids in the sample. This was performed using the American standard test method ASTM E1756-01, and the amount of ash in the sample at a temperature of 575 ^{o}C was determined using ASTM E1755-01.

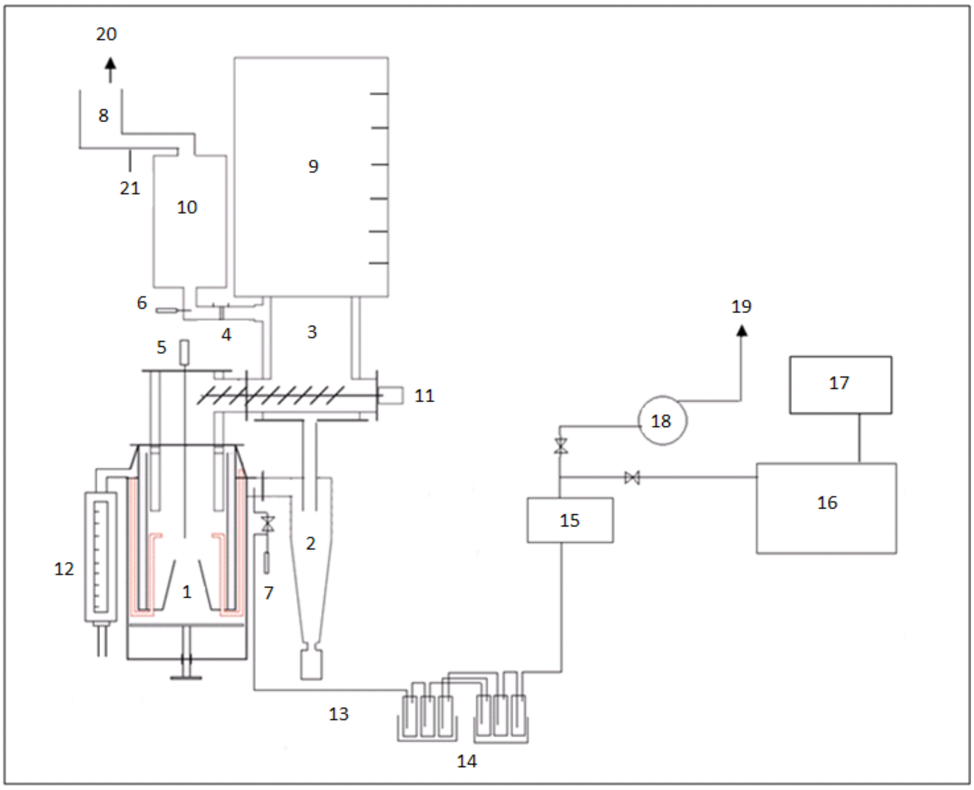
5.4.6 Inductively coupled plasma (ICP) metals analysis

As the majority of de-inking sludge is composed of inorganic ash, samples of de-inking sludge ash which had been produced at a temperature of 575 ^{o}C according to the ASTM E1755-01 standard test method for ash were analysed externally by Medac Ltd, using an ICP metals analysis for detection of all the metal elements present using an acid digestion method.

5.5 Thermal conversion trials

5.5.1 Fixed bed downdraft gasification trials

The gasification of de-inking sludge, rejects and wood chips in this work was carried out using a 10 kg/hr fixed bed downdraft gasifier. The unit, also known as the Gasifier Experimenter's Kit (GEK), was originally designed and manufactured by All Power Labs in the USA [44]. The unit (shown in Figure 24) operates under negative pressure using a downstream venturi air ejector to draw the gasification air through. It is composed of a hopper (9), feed dryer (3), motorised auger (11), gasifier (1), cyclone (2), carbon absorption filter (10) and swirl burner (8).



1 Gasifier, 2 Cyclone, 3 Heat Exchanger Drying Bucket, 4 Orifice Plate, 5 Thermocouple, 6 Thermocouple,7 Thermocouple, 8 Swirl Burner, 9 Calibrated Glass Hopper, 10 Carbon Absorption Filter, 11 Auger ,12 Air Rotameter, 13 Gas Sampling Line, 14 Gas Wash Bottles, 15 Digital Mass Flowmeter,16 Gas Chromatograph, 17 Computor , 18 Gas Suction Pump, 19 Vent, 20 Main Vent, 21 Venturi Ejector

Figure 24 A schematic diagram of the GEK gasification system

In order to determine the mass balance for each test it was necessary to modify the unit and install further instrumentation. A calibrated glass feed hopper was installed to determine feedstock flow rates, a calibrated air rotameter was installed for air inlet flow rates, k-type thermocouples for temperature measurements, pressure transducers linked to a Gasifier Control Unit (GCU), measured differential pressures, a calibrated orifice plate for product gas outlet flow rates and a gas sampling line for tar, water and product gas composition measurements. All electronically recordable data were sent to the (GCU) and logged every second. In all experiments the gasification medium used was pre-heated air and the pre-heat was derived from a heat exchanger jacket between the hot product gases leaving the reactor and ambient air entering the reactor.

Before each experiment the gasifier was cleaned to remove tar fouling, ash and char before being reassembled. At the start of each run the bed of the gasifier was filled with approximately 2-3 kg of fresh wood charcoal and the hopper filled with the prepared feedstock of known weight. The unit was then sealed gas tight to ensure no air leaks and this was confirmed by performing a cold run before each experiment. The experiments were initiated by opening the venturi ejector valve, opening an ignition port on the gasifier and using a propane burner to light the gasifier bed. After ignition was achieved feedstock was fed into the gasifier from the hopper and the reactor was then left to reach gasification temperatures of approximately 800-1000 ^{o}C at the oxidation zone, and once gasification was within this temperature range the flare was ignited. The nature of the design of the GEK unit is such that temperature is controlled by altering the air flow rate entering the gasifier. Therefore at start up the air inlet flow rate was maintained at 10 m^3/hr, however as the experiment proceeded the air flow rate was either slightly increased or decreased to stabilise fluctuations in gasification temperature. Each run lasted approximately 3-5 hours and depending on the material used consumed between 10 and 25 kg of feedstock. The feed rate was determined by multiplying the average bulk density of the feedstock by the reduction of hopper volume.

5.5.1.1 Tar analysis

Tar was quantified by a tar sampling system developed by CEN [45] in which a sample of product gas was removed from the gasifier and routed through a series (“train”) of gas washer bottles that condense the tars under low temperatures and by the use of a propan-2-ol (iso-propanol) extraction solvent. The remaining clean product gas was then routed through an Aalborg DFM digital mass flow meter and then directly into a Hewlet Packard HP-5890 GC-TCD for detection and quantification.

The tar train was composed of six interconnected bottles. Bottles, one, two and four were put in a hot bath at 50 °C while bottles three, five and six were put into a cold bath utilising dry ice. All the bottles apart from the sixth were filled with iso-propanol to act as the scrubbing agent. All the bottles apart from the first were filled with glass beads to improve the scrubbing efficiency. Figure 25 shows the schematic of a tar train setup

Figure 25 Producer gas tar cleaning system [45]

A rotary evaporator was then used to separate the tar/iso-propanol mixture using a STUART model RE-300 rotary evaporator with round bottom flask. Rotary evaporation was conducted to analyse the tar content present in the sample of product gas. The iso-propanol/tar/water mixture collected from the tar sampling system was weighed and placed into a round bottom flask also of known weight. The flask was connected to the rotary evaporator with the sample flask partially immersed in a water bath at 50 °C. The collection flask was connected to a shell and tube water cooled condenser which operated under vacuum. The sample was simultaneously rotated and heated at a constant speed of 40 rpm, evaporating the water and iso-propanol which condensed and collected in the collection flask, leaving only the tar remaining in the sample flask. Figure 26 shows rotary evaporator set up.

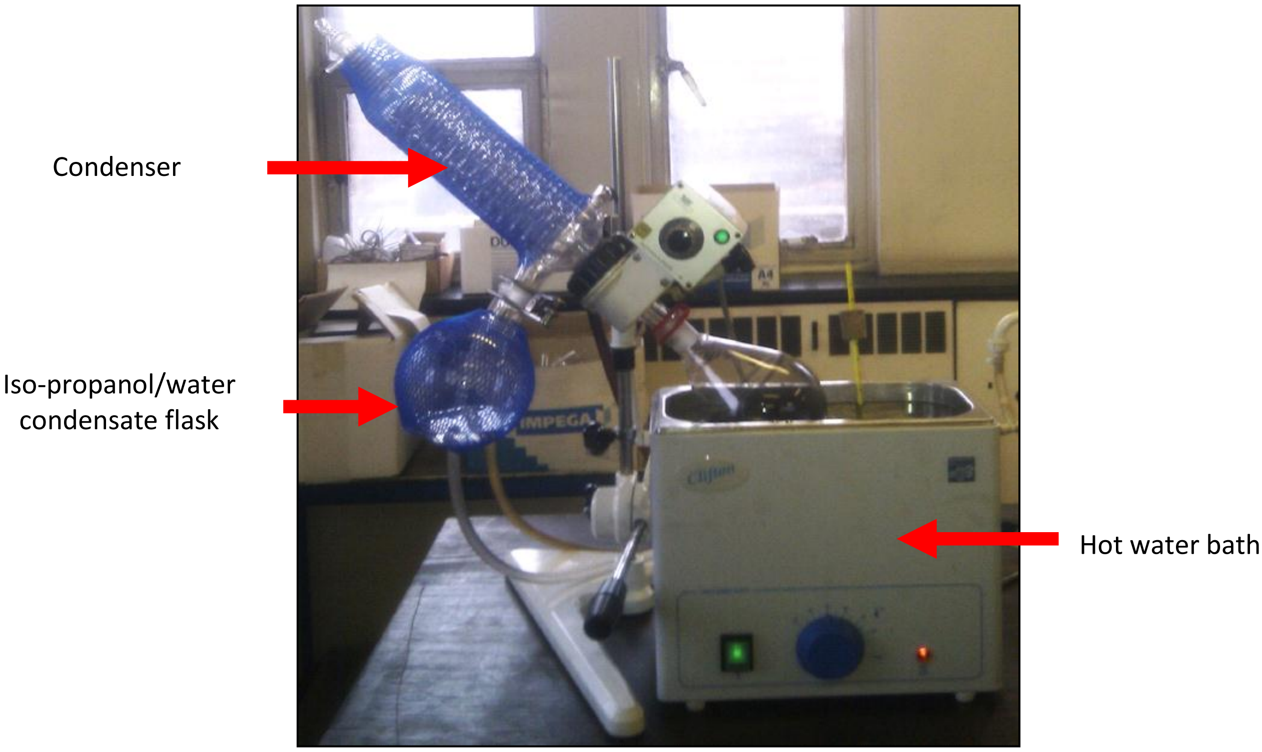


Figure 26 Rotary evaporation system

The separation was monitored by visually observing the water/iso-propanol mixture collected in the condensate flask. Once the separation was complete, the rate at which water/iso-propanol solution was being collected reduced and eventually stopped. The flask that contained the initial mixture only contained tar at this point. Tar appeared to form a dark highly viscous layer on the flask surface. When the separation was complete, the rotary evaporator was de-pressurised. The tar laden flask was taken and its weight was measured. The difference in weight between the empty flask and the tar laden flask was the total amount of tar collected in the product gas which passed through the tar train.

5.5.1.2 Water condensate analysis

Water condensate after the extraction of tar was determined by a V20-Compact volumetric Karl-Fischer titration unit using a Hydranal composite 5K titrant. Karl Fischer titration was carried out on the tar free iso-propanol to determine the water content of the product gas which passed through the tar train. The mass of the iso-propanol water mixture was weighed before the titration. The result from the titration was used to scale up the water produced per cubic metre of product gas in the gasifer. Figure 27 shows the Karl Fischer titration set up that was used to analyse the results.

Figure 27 Karl Fischer titration setup

Before using the equipment, it was ensured that the glass beaker was clean by rinsing with fresh solvent (Hydranal or Methanol) and draining. The empty beaker was then filled with the fresh solvent until the sensor was submerged. The sensor is a double platinum pin indicator that detects the end point of the titration. It is connected to an on board microprocessor which calculates the water content in the sample.

The solvent was titrated with an iodine based titrant to standardise the water present. A clean syringe was used to inject 2 g of the isopropanol/water sample and the weight was manually input into the system before starting the sample titration. The titration was monitored until the run was complete. Once complete, the water content in the sample was displayed on the screen.

5.5.1.3 Product gas analysis

Gas analysis was carried out using a Gas Chromatograph Thermal Conductivity Detector (GC-TCD) in a Hewlet Packard HP-5890 series II device with 60/80 Carboxen 1000 column. The column was calibrated to detect H_2, CO, CO_2, N_2 and CH_4. Oven temperature was pre-programmed to an initial temperature of 35 ^oC and ramped to 225 ^oC at a rate of 20 ^oC/min. Helium with a flow rate of 30 ml/min was used as the carrier gas. Approximately 150 micro litres of clean product gas was extracted from the gas sampling system using a gas tight syringe and injected directly into the GC for analysis.

5.5.2 Intermediate pyrolysis trials

The pyrolysis of de-inking sludge was carried out in the “Pyroformer”, a 20 kg/hr intermediate pyrolysis unit. The pyroformer [46] was originally manufactured in Germany before being installed and commissioned at the European Bioenergy Research Institute (EBRI) at Aston University. The installation is shown in Figure 28.

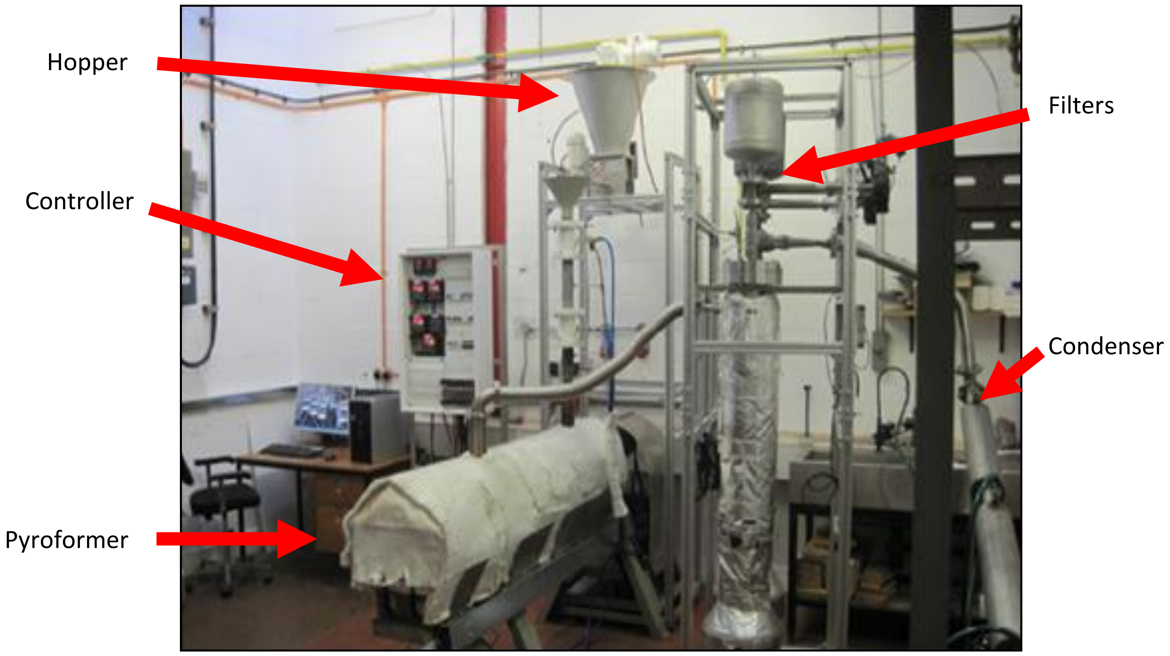
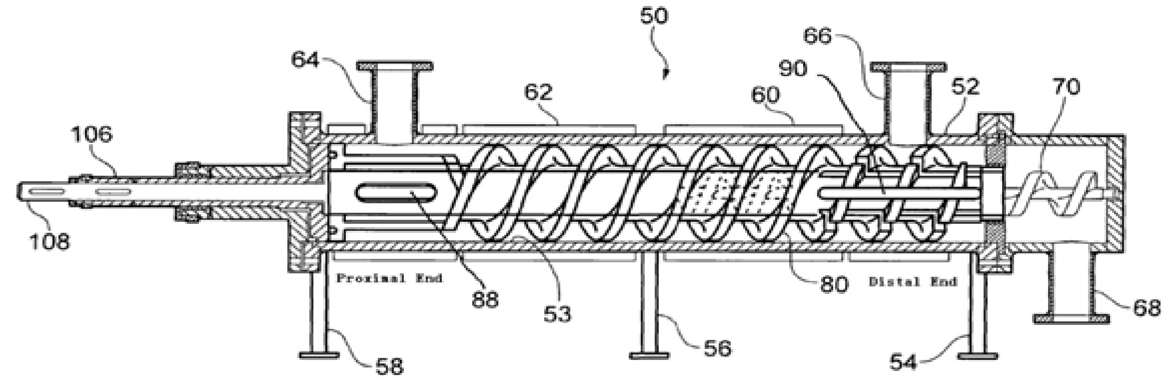


Figure 28 Pyroformer intermediate pyrolysis system Aston University

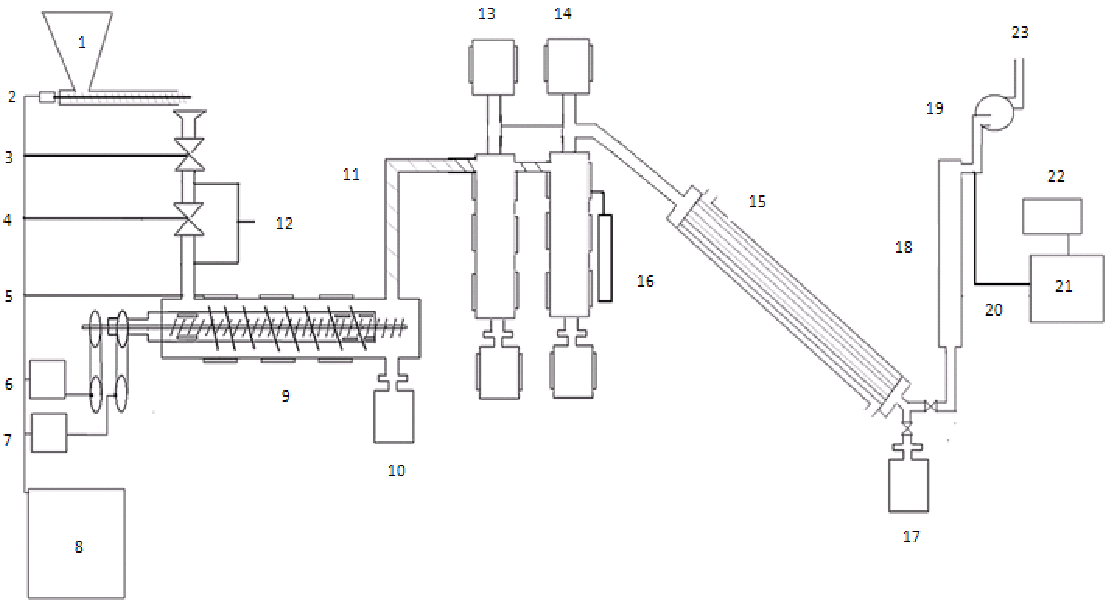
The pyroformer is essentially an auger pyrolysis reactor with two counter-rotating co-axial screws which can process a feed in an inert atmosphere of nitrogen at a rate of up to 20 kg/hr (Figure 29). The reactor is heated externally by electrical heating bands and the feed moves through the screw conveyor system whilst being heated to the specified pyrolysis temperature. Some solid residue (char and ash) is recycled within the reactor, the remainder drops out at the downstream end of the reactor, and the pyrolysis vapours and gases exit through an outlet pipe. The recycling of char increases the char to feedstock ratio in the reaction zone which promotes catalytic cracking of the primary vapours to lower molecular weight vapours and permanent gases. It also serves to recycle heat within the reactor and increase the heating rate experienced by the feedstock. The design is well suited to de-inking sludge as it can process feeds with very high ash contents.



106 External Screw Shaft, 108 Internal Screw Shaft, 64 Feed Inlet Pipe, 62,60,52,50 Electrical Heating Bands, 58,56,54 Supports, 88 External Screw Slot, 90 External Screw Slot, 80 External Screw, 70 Internal Screw, 68 Solid Drop Out Pipe, 66 Gas Outlet Pipe

Figure 29 Internal mechanics of the pyroformer reactor

Before each run the pyroformer was heated to its set point temperature of 450 °C. Then, after purging with N_2, the unit was fed with pelletised feedstock at a rate of approximately 15 kg/hr from a feed hopper. A steady state was established after approximately 60 minutes. For removal of entrained solid particulates the hot organic vapours and permanent gases were first passed through hot gas filter candles which pre-cleaned by purging with N_2 and maintained at 450 °C and then routed to a shell and tube water cooled condenser where the great majority of the pyrolysis vapours were condensed. Remaining vapours and the permanent gases were then passed to an electrostatic precipitator for aerosol knock out and then to a GC-TCD for detection. Approximately 3 litres of pyrolysis oil was collected from each run. Once saturated the solids within the pyroformer began to drop out of the char exit port indicated as 68 on Figure 29. These solids were collected in a char collection pot, and at steady state the rate at which the char is collected becomes constant. A schematic of the process is shown in Figure 30.



1 Feed Hopper, 2 Auger, 3 Actuating Value, 4 Actuating Valve, 5 Electric Heating Bands, 6 Electric Motor, 7 Electric Motor, 8 Main Control Board, 9 Pyrolysis Reactor, 10 Char Collection Vessel, 11 Heated Line, 12 N_2 Purge Line, 13 Heated Filter Candle, 14 Heated Filter Candle, 15 Shell and Tube Condenser, 16 Filter Candle Control Board, 17 Pyrolysis Oil Collection Vessel, 18 Electrostatic Precipitator, 19 Gas Suction Pump, 20 Gas Sampling Line, 21 Gas Chromatograph, 22 Computer, 23 Extraction Vent

Figure 30 Schematic diagram of the intermediate pyrolysis system

5.5.2.1 Pyrolysis products analysis

After production, the pyrolysis oil, gases and solids were characterised to determine their physical and chemical compositions. Successful utilization of pyrolysis oil for its application as a liquid fuel requires sound understanding of its characteristic properties. The physical and chemical properties of pure pyrolysis oil are known to affect the combustion and performance characteristics of an engine as well as posing storage and handling difficulties. Samples of de-inking sludge pyrolysis oil produced were characterised to determine water content, viscosity, gross heating value, elemental composition, flash point, density, pH, acid number, copper corrosion, ICP metals, ash, carbon residue, and liquid compounds.

5.5.2.2 Water content

Pyrolysis oil contains water which is comprised partly of the moisture present in the feedstock and partly of water created in the pyrolysis reactions (“reaction water”). A high water content gives rise to low heating value, low ignition temperature, immiscibility with hydrocarbon fuels and phase separation problems. But on the other hand, it slows down the polymerization rate of the oil which delays its ageing and favours low viscosity thereby improved atomization and combustion in an engine [47, 48, 49, 50]. The water content of all pyrolysis oil samples was determined by a Mettler Toledo V20-Compact volumetric Karl-Fischer titration unit using a Hydranal composite 5K titrant and was carried out in accordance with the ASTM-E203 standard test procedure.

5.5.2.3 Kinematic viscosity

Kinematic viscosity is a measure of resistance to shearing forces or angular deformation during fluid motion [50]. It is one of the most important properties for fuel performance in an engine. For pyrolysis oil, viscosity mainly depends upon the type of feedstock and the pyrolysis technique used to produce the oil. A higher viscosity in pyrolysis oil is generally helpful in preventing internal leakages in the fuel-line/pump which is indirectly related to an increase in brake power. However if the viscosity is too high it may cause blockages and restrict fuel flow. Pre-heating the fuel before injection can help to reduce viscosity thereby improving atomization and combustion. The presence of surplus water content or alcohols in the oil reduces viscosity and increases oil stability by slowing down the rate of ageing [51]. Kinematic viscosity of all pyrolysis oil samples in this work was measured by Cannon-Fenske routine viscometer, and was measured in accordance with the ASTM-D445 standard test procedure.

5.5.2.4 Gross heating value (HHV)

The HHV of the pyrolysis oil and char samples was determined using a Parr 6100 bomb calorimeter and verified using the Channiwala equation as previously described in Section 5.4.4

5.5.2.5 Ultimate analysis

The elemental composition of fuels is an important parameter to measure as the amount of carbon and hydrogen present directly relates to the energy content. Also higher oxygen to carbon ratios in pyrolysis oil tends to lower the energy density and miscibility of the oil with conventional fuels. It is also accelerates the polymerization reactions which leads to oil instability. Also the amount of sulphur and nitrogen in the fuel indicates the degree to which SO_X and NO_X emissions will ultimately be released upon its combustion. These compounds present in pyrolysis oil can also cause engine corrosion and wear as they are strongly acidic. For determination of the elemental compositions (C H N O and S), samples of the pyrolysis oil and pyroformer solid residues were analysed externally by Medac Ltd using a Carlo- Erba EA1108 CHNS-O analyser by total oxidation.

5.5.2.6 Flashpoint

Flashpoint is defined as the minimum temperature at which a fuel begins to evolve vapours to form a flammable mixture with air in the presence of an ignition source. Generally, it indicates the volatility and fire hazard associated with storage and handling of a fuel [52]. Flashpoints of the pyrolysis oil were measured using an automated closed cup FP tester according to ASTM-D93 standard test procedure.

5.5.2.7 Density

Density is an important factor for calculating the fuel consumption rate. For pyrolysis oil, density is generally high due to the presence of high molecular weight compounds. The density of the pyrolysis oil was measured at 22 ^{o}C using a glass hydrometer of range (0.8—1.0gm/cu.cm) according to ASTM-D4809 standard test procedure.

5.5.2.8 pH

The substantial presence of corrosive acids in pyrolysis oil such as acetic and carboxylic acids is responsible for very high acid numbers (between 25—100mg KOH/g) and low pH in the range 2-3.

High acidity poses severe corrosion problems in engines and storage vessels. The presence of a high water content in pyrolysis oil tends to reduce acidity to a certain extent [49, 50,]. Here, pH of the pyrolysis oil was determined using Fisher-brand pH meter calibrated with standard buffer solutions according to ASTM-D7544 standard test procedure.

5.5.2.9 Acid number

Acid number of the pyrolysis oil was measured potentiometrically with an alcoholic potassium hydroxide titrant, using a Mettler Toledo G20 Compact titrator device in accordance with ASTM-664-04 standard test procedure.

5.5.2.10 Copper corrosion, ICP-metals and cetane index

Copper corrosion of the pyrolysis oil, as the name implies, gives an indication of how corrosive the oil is to copper, ICP metals analysis identifies the metal elements present in the pyrolysis oil ash and the cetane index is an indication of the readiness of a fuel to ignite in a compression-ignition engine. For most pyrolysis oils, a typical cetane index is around 5-6 although some authors report values in the range of 13-14 [49, 51] Samples of pyrolysis oil were measured externally by Alcontrol Laboratories Ltd for determination of copper corrosiveness, ICP-metals and cetane index.

5.5.2.11 Ash content

Ash is a mixture of various alkali metals such as sodium, potassium and calcium which are mainly responsible for corrosion, erosion and deposition in the engine [50] including blocking/corroding of injectors, fuel-line/fuel-pump etc. Ash content of the pyrolysis oil was determined in a Carbolite furnace model AAF-1100 in accordance with the ASTM-D482-03 standard test procedure.

5.5.2.12 Carbon residue

Carbon deposits around engine injectors can often cause clogging and fouling, therefore by knowing how much carbon residue is present in the pyrolysis oil an assessment of whether or not it will cause clogging of the engine can be made. Carbon residue of the pyrolysis oil was determined in accordance with the ASTM-D524-09 standard test procedure.

5.5.2.13 Liquid GCMS

GC-MS was performed to identify the most abundant compounds present within the pyrolysis oil. Samples of the oil were diluted in ethanol in a 1:10 ratio. GC-MS analysis was conducted using a Hewlett Packard HP 5890 Series II Gas Chromatograph with an automatic injector and auto sampler with a DB 1706 non-polar capillary column of 60m x 0.25 mm x 0.25 µm. Helium was used as the carrier gas at a constant flow rate of 1.5 ml/min. The initial oven temperature was 40 ^{o}C and ramped up to 290 ^{o}C. Identification of the GC-MS peaks was based on the comparison between the Mass finder library and Wiley library.

5.5.2.14 Permanent gas analysis

Permanent gas samples were taken after condensing of the oils and injected into a GC-TCD at 25 minute intervals using a gas tight syringe. Gas analysis was carried out using a Gas Chromatograph Thermal Conductivity Detector (GC-TCD) in a Hewlett Packard HP-5890 Series II device with a 60/80 Carboxen 1000 column. Oven temperature was pre-programmed to an initial temperature of 35 ^{o}C and ramped to 225 ^{o}C at a rate of 20 ^{o}C/min and Helium with a flow rate of 30 ml/min was used as the carrier gas.

5.6 Pyrolysis oil/biodiesel blending and characterisation

Biodiesel used in this study was obtained from Brittania Oils Ltd (Birmingham, UK) and was produced from waste cooking oils. Blends with pyrolysis oil were prepared by mixing in an agitated tank without the addition of any surfactants and then allowed to settle for a period of 24 hours. The blends were then filtered using a 1 µm sock filter to remove fine solid particulates and dust prior to engine testing. Once prepared the blends were characterised according to (Sections 5.5.2.2-5.5.2.13).

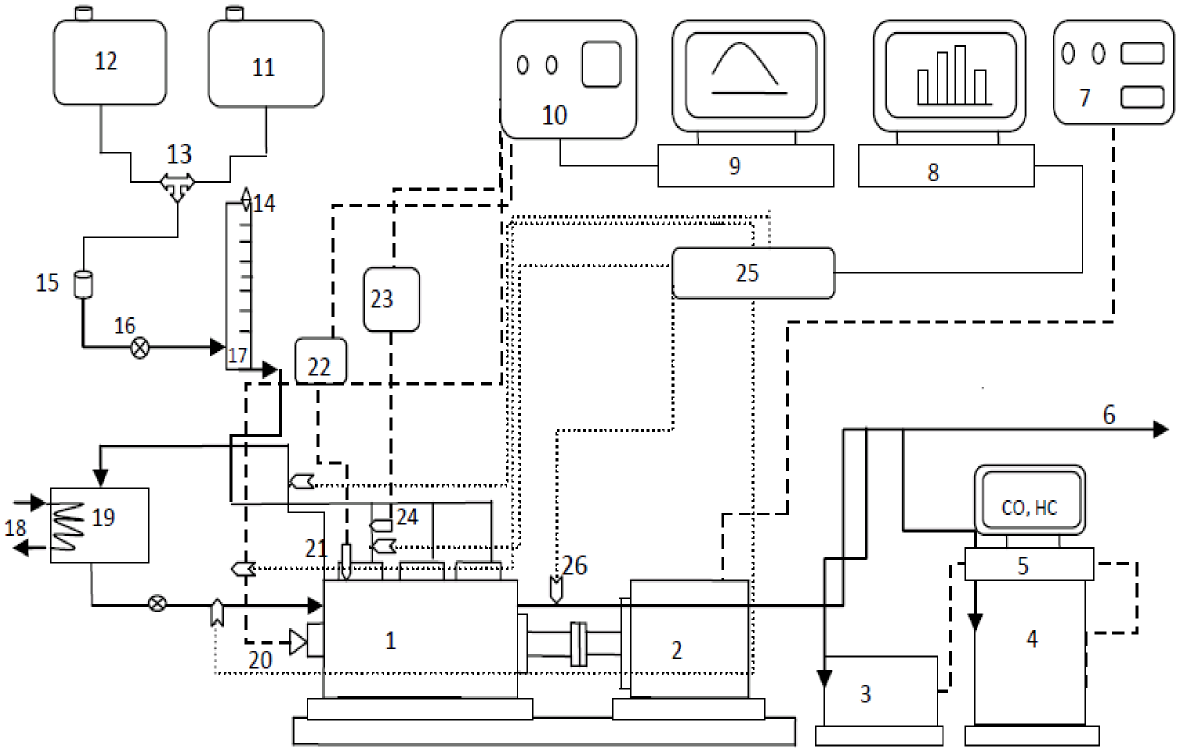
5.7 Pyrolysis oil/biodiesel engine testing

Table 4 shows the specifications of the indirect injection water cooled Compression Ignition (CI) engine selected for the study. The experimental arrangement is shown in Figure 31.

Table 4: Engine Specifications [65]

Manufacturer	Lister Petter (UK)
Model/type	LPWS Bio3 water-cooled
No. of cylinders	3
Bore/stroke (mm)	86/80
Rated speed (rpm)	1500
Continuous power at rated speed (kW)	9.9
Overload power at rated speed (kW)	10.9
Type of fuel injection	Indirect injection. Self-vent fuel system with individual fuel-injection pumps
Fuel pump injection timing	20° BTDC
Cylinder capacity (litre)	1.395
Compression ratio	1:22
Minimum full load speed (rpm)	1500
Continuous power fuel consumption at 1500 rpm	3.19 litres/hr (fossil diesel)
Glow plug	Combustion-chamber glow plugs
Exhaust gas flow	41.4 litres/sec at full loads at 1500 rpm
Jacket water flow at full load	33 litres/min (at 1500 rpm)
Maximum engine jacket water temperature (°C)	99 - 102

A dual-fuel supply system was designed so that the engine can be started and warmed up with Fossil Diesel (FD) or Biodiesel (BD) before switching to the De-inking Sludge Pyrolysis Oil/ Biodiesel (DSPO-BD) blend after about 10 minutes (Figure 31). At the end of each test run the engine was reverted back to FD operation in order to remove the remaining DSPO-BD blend from the fuel supply and engine injection system. Additional filters were connected into the fuel supply system. The fuel supply tanks were placed at 3 m height for reliable fuel flow and to overcome the pressure losses in the additional fuel filters. Stainless steel piping and valves were used to avoid any corrosion or erosion. The radiator was by-passed by a header tank type HX (Bowman UK) supplied with tap water. The radiator itself and the fan were kept in place to enable comparisons of the engine performance with the rated figures provided by the manufacturer and for cooling the engine accessories. An eddy current dynamometer (Froude Hofmann AG80HS) was used to measure and adjust the engine load and speed (Figure 31). The torque and speed accuracies of the dynamometer were ±0.4 Nm and ±1 rpm respectively. A five-gas emission analyser (Bosch BEA 850) and smoke opacity meter (Bosch RTM 430) were used to analyse the exhaust gas components and to measure the smoke intensity respectively. A graduated cylinder and stop watch were used to measure the fuel consumption rate (Figure 31). K-type thermocouples were used to measure the temperatures of the exhaust gas, fuel inlet, lubrication oil and engine jacket water. A LabVIEW® data acquisition system was used to log the temperatures at the various locations. The engine was operated at different loads at a constant speed of 1500 rpm. A pressure sensor (Kistler 6125C11) and charge amplifier (Kistler 5064B11) were used to measure pressure in the cylinder nearest to the radiator end of the engine. Another pressure sensor (Kistler 4065A500A0) and amplifier (Kistler 4618A0) were used to measure the fuel line injection pressure. This sensor was installed on the fuel delivery line of the same cylinder. An optical encoder (Kistler 2614A) was installed for detection of the crank angle position. The amplifiers and the encoder electronics were connected to the 'KiBox' (Kistler, model 2893AK8) for data logging, which was connected to a PC through an ethernet port. KiBoxCockpit software was used to measure and analyse the various engine combustion parameters.



1 Engine, 2 Dynamometer, 3 Smoke meter, 4 Exhaust analyser, 5 Exhaust data acquisition, 6 Exhaust gas discharge, 7 Dynamometer controller, 8 NI data acquisition for temperature, 9-10 Kistler combustion analyser, 11 Jatropha/Karanj tank, 12 Diesel tank, 13 3 way valve, 14 Vent screw, 15 Additional fuel filter, 16 Valve, 17 Fuel measurement, 18 Cold water flow to HX, 19 HX to cool jacket water, 20 Crank angle encoder, 21 Cylinder pressure transducer, 22,23 Amplifier, 24 Injection pressure sensor, 25 NI DAQ, 26 Thermocouple

Figure 31 Schematic diagram of the engine test rig

6.0 RESULTS AND DISCUSSION

6.1 Chapter summary

This chapter describes the feedstock characterisation results, and then presents all the results from all the thermal conversion trials carried out. Downdraft gasification trials focused on the blending of wood chips with rejects and de-inking sludge from all mills to create a product gas. Pyrolysis trials focused on the intermediate pyrolysis of de-inking sludge from the AN and KC mills to create a pyrolysis oil which can be used as a fuel. The pyrolysis oils which were created from these trials were subsequently blended with biodiesel, characterised, and tested in a diesel engine to determine their feasibility as a fuel for CHP applications at these mills. There currently exists no available literature on these particular processes to thermally convert these particular wastes, which is where the novelty of this work rests. The chapter concludes by identifying the optimum process route for implementation at each of the candidate mills.

6.2 Feedstock characterisation

All feedstocks acquired were characterised by a full proximate, ultimate and HHV analysis.

6.2.1 De-inking sludge characterisation

In general de-inking sludge has a high ash content and a low calorific value, and the sludge produced at the AN mill is very similar in terms of composition to that of the KC mill. Before proximate and ultimate analysis of de-inking sludge, samples from each mill were first analysed to determine the initial moisture content using a Sartorius MA35 moisture analyser. The moisture content of the KC mill's de-inking sludge was found to be 48 wt% and that of the AN mill's de-inking sludge 35 wt%.

Table 5 shows the proximate and ultimate analysis and HHV of the de-inking sludges. It is observed from Table 5 that the de-inking sludges have very low HHVs of 6.4 MJ/Kg for the AN mill and 7.0 MJ/Kg for the KC mill and very high ash contents of 51.6 wt% for the AN mill and 43.6 wt% for the KC mill, which would necessitate a continuous ash removal system in an industrial scale fixed bed downdraft gasifier. The ultimate analysis revealed very similar carbon, hydrogen, and oxygen contents between the two mills. Sulphur and chlorine contents were very low. De-inking sludge from the KC mill contained a slightly higher nitrogen content which may lead to higher NOx emissions upon its combustion.

Table 5 Proximate and ultimate analysis and HHV of de-inking sludge

	AN Mill De-inking Sludge	**KC Mill De-inking Sludge**
Proximate Analysis wt% (dry basis)		
Moisture	1	1.3
Volatiles	46.3	55.1
Fixed Carbon	1.1	<0.1
Ash	51.6	43.6
HHV (MJ/kg)	6.4	7.0
Ultimate Analysis wt% (dry basis)		
Carbon	21.1	21.7
Hydrogen	2.3	2.8
Oxygen*	24.7	29.8
Nitrogen	0.3	2.1
Sulphur	<0.1	<0.1
Chlorine	<0.1	<0.1

* Obtained by difference

The HHVs shown in Table 5, which were determined using an oxygen bomb calorimeter, were also calculated using the Channiwalla equation (see Section 5.4.4) [42]. The results were found to be 7.5 MJ/kg for the KC mill sludge and 6.4 MJ/Kg for the AN mill sludge, which are in reasonable agreement with the values of Table 5.

Thermal gravimetric analysis (TGA) was carried out using a Perkin Elmer Pyris 1 TGA device in order to assess the thermal degradation behaviour of de-inking sludge. For moisture, volatiles and char, an inert atmosphere of nitrogen was used, with a heating rate of 25 °C/min and a programmed maximum temperature of 900 °C (TGA pyrolysis). To determine ash, an atmosphere of air was used, again with a heating rate of 25 °C/min and a maximum programme temperature of

900 °C (TGA combustion). The TGA pyrolysis and combustion weight loss and derivative weight loss (DTG) profiles, for both the KC mill and AN mill de-inking sludges are shown in Figures 32-39. In general the two sludges show similar behaviour, whether exposed either to air or an inert atmosphere of nitrogen. Weight loss is seen between 200 °C and 450 °C, and then again between 700 °C and 800 °C. The rate of weight loss between 200 °C and 450 °C peaks at approximately 350-400 °C, and this is the direct result of the decomposition of the cellulosic fibres and other de-inking minerals. As a result of the increased amount of volatiles present within the KC de-inking sludge, the rate of weight loss is observed to be higher for the KC mill sludge (Figures 34 and 35) than for the AN mill sludge (Figures 38 and 39). The rate of weight loss between 700 °C and 800 °C peaks at approximately 800 °C and this is largely the result of the thermal decomposition of $CaCO_3$ present within the ash. Similar peak rates are observed for the two sludges. Above 900 °C, approximately 50 wt% of the original dry solid sample is present as inorganic metal oxide compounds, so no further weight loss is experienced.

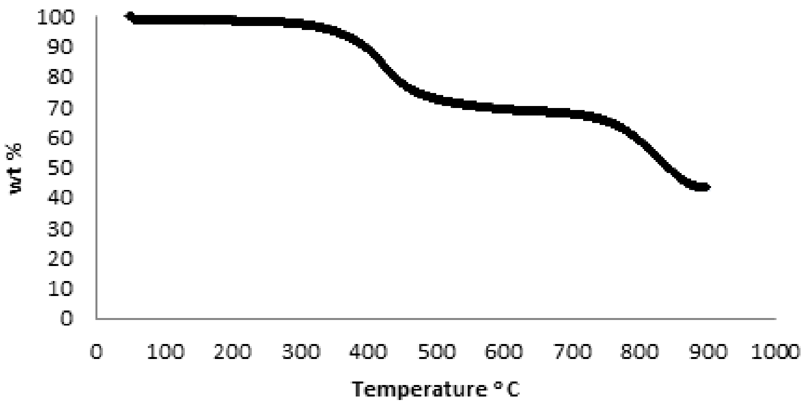


Figure 32 De-inking sludge TGA combustion profile (KC mill)

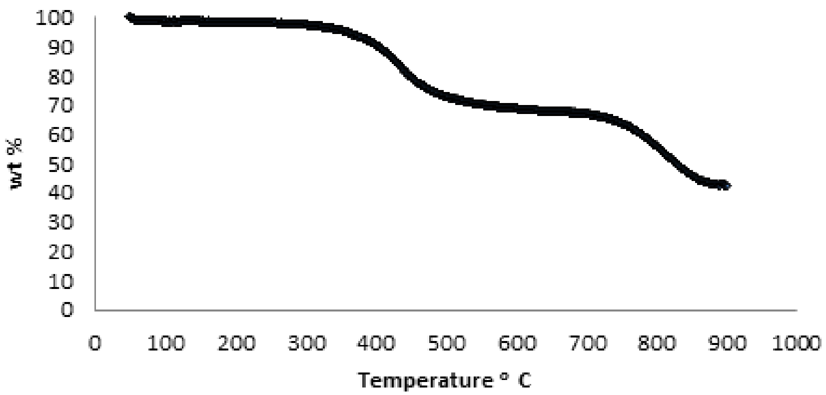


Figure 33 De-inking sludge TGA pyrolysis profile (KC mill)

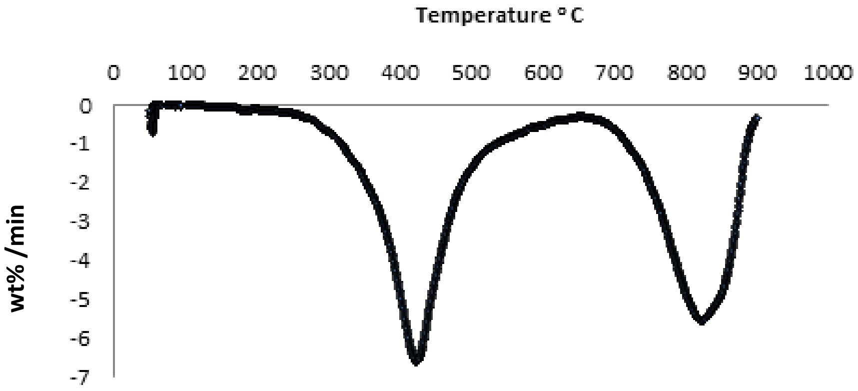


Figure 34 De-inking sludge DTG combustion profile (KC mill)

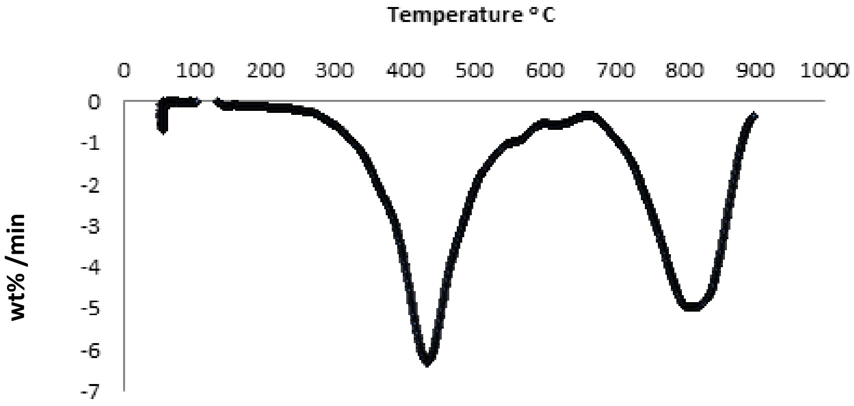


Figure 35 De-inking sludge DTG pyrolysis profile (KC mill)

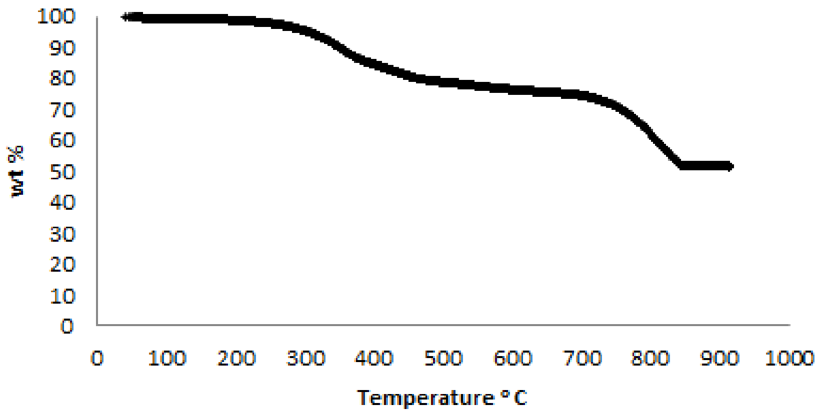


Figure 36 De-inking sludge TGA combustion profile (AN mill)

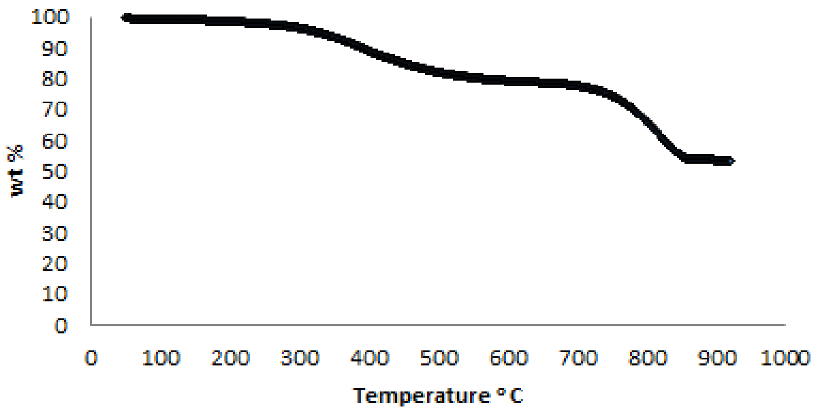


Figure 37 De-inking sludge TGA pyrolysis profile (AN mill)

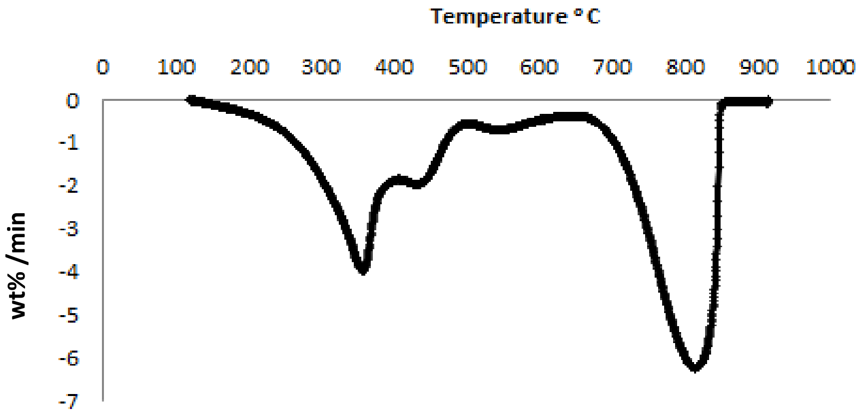


Figure 38 De-inking sludge DTG combustion profile (AN mill)

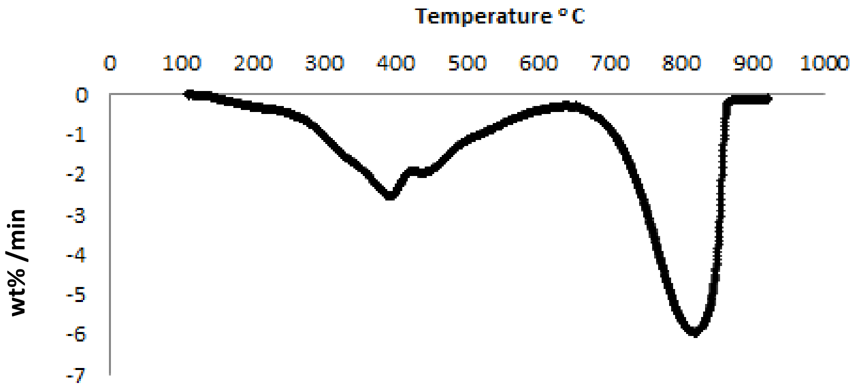


Figure 39 De-inking sludge DTG pyrolysis profile (AN mill)

Further characterisation tests were carried out using ASTM standards in order to validate moisture content (as received) and to determine ash content at a lower ash temperature of 575 °C. The results of these tests are presented in Tables 6-7.

Table 6 ASTM E1756-01 Standard test method for determination of total solids in biomass

Sample (As Received Basis)	**Run**	**Moisture Free Solid wt%**
AN Mill De-inking Sludge	1	64.8
AN Mill De-inking Sludge	2	65.2
KC Mill De-inking Sludge	1	52.4
KC Mill De-inking Sludge	2	51.6

Table 7 ASTM E1755-01 Standard test method for ash in biomass

Sample (Dry Basis)	**Run**	**Ash Content wt%**
AN Mill De-inking Sludge	1	74.7
AN Mill De-inking Sludge	2	74.4
KC Mill De-inking Sludge	1	64.0
KC Mill De-inking Sludge	2	61.9

The ASTM tests show that de-inking sludge produced by the KC mill contains a slightly higher moisture content as received than that produced from the AN mill. This depends on the degree to which the sludge is de-watered on-site at each mill. Although both mills use screw presses to de-water their sludge, the AN mill's screw press system is slightly more efficient. The results also show that AN mill's de-inking sludge contains a higher ash content at 575 °C than that of the KC mill. This is the result of the AN mill's initial feedstock containing larger amounts of filler and other inorganic substances which appear in the de-inking sludge ash.

The ash remaining after combustion at 575 °C was analysed for metals composition. The metal oxides present within both de-inking sludge ashes were found to be predominantly aluminium and calcium, with other metal elements present in far smaller quantities. The most common metal oxide present in the ash at 575 ^{o}C was $CaCO_3$, representing over 60 wt%. A higher ashing temperature (>750 ^{o}C) would result in the calcination of $CaCO_3$ to release CO_2 and form lime (CaO). Therefore it is suspected that de-inking sludge ash may share similar properties to that of calcined limestone or dolomite which would make its use as a solid medium in a fluidised bed gasifier of interest, with catalytic properties for the cracking of tars at elevated temperatures (>800 ^{o}C). Moreover the physical appearance of de-inking sludge ash is of a granular powder with a high surface area which fluidises readily and is well suited for use as a catalyst. Aluminium, commonly present in ash as Al_2O_3, was also abundant at approximately 20-30 wt%. Other metal oxides were present in smaller quantities and these included silicon oxides, titanium oxides, and magnesium oxides.

It has also been shown that the mineral ash-forming content of de-inking sludge can be further reduced before gasification or combustion by as much as 65 wt % by acid pre-treatment (acid washing); both HCl, and H_2SO_4 have been shown to work well [53]. Acid washing of biomass for ash-forming mineral removal is well documented [54,55,56] and its application to pre-treat de-inking sludge is also feasible in principle, however the effect on the gasification or pyrolysis products is unknown and requires further work.

The extremely high ash content along with the very low calorific value of de-inking sludge found from the characterisation results indicates that de-inking sludge is not a suitable fuel for downdraft gasification and would almost certainly need to be blended with higher calorific fuels to provide sufficient energy to sustain gasification. The principal attraction to a mill of processing this waste fraction by either gasification or pyrolysis would come from the reduction of disposal costs rather than from the conversion to a fuel. Carrying out the direct pyrolysis of the sludge in the Pyroformer would be a more suitable thermal conversion option. The Pyroformer has the ability to handle high ash content feedstocks because of its augur-screw design.

6.2.2 Reject, co-form and wood chip characterisation

Due to the heterogeneous nature and large particulate size of the rejects, co-form and wood chips, samples of these feedstocks were sent after pre-treatment to Marchwood Scientific Services Ltd for a full proximate and ultimate analysis the results are shown in Table 8.

Table 8 Proximate, ultimate analysis and HHV of rejects, co-form and wood chips

	Mixed Wood Chips	**KC Mill Pulper Reject Pellets**	**KC Mill Dry Co-form Reject Pellets**	**KC Mill Wet Co-form Reject Pellets**	**AN Mill Pulper Reject Pellets**	**SSK Mill Pulper Reject Pellets**
Proximate Analysis wt% (dry basis)						
Moisture	3.7	2.0	2.5	9.5	<0.2	7.6
Volatiles	86.6	82.2	82.7	79.5	83.4	73.5
Fixed Carbon	9.4	9.1	7.7	7.4	7.2	10.1
Ash	0.3	6.7	4.1	3.6	9.4	8.8
HHV (MJ/Kg)	15.4	24.8	20.4	20.4	18.3	22.9
Ultimate Analysis wt% (dry basis)						
Carbon	45.6	70.5	58.9	60.4	60.9	53.3
Hydrogen	5.8	8.3	6.1	6.4	3.4	7.5
Oxygen*	48.0	13.9	30.2	28.8	23.5	29.6
Nitrogen	0.3	0.5	0.4	0.6	0.4	0.5
Sulphur	<0.1	<0.1	0.2	0.2	0.4	<0.1
Chlorine	<0.1	0.1	0.1	<0.1	2	0.3

Notable from Table 8 is the similarity between the dry and wet co-form material with the only significant difference being higher moisture content of the wet co-form which is 7 % higher than the dry co-form material, this is due to the difficulty of removing water during the drying process.

The pulper reject fuel pellets all have a higher HHV (averaging 22 MJ/kg) than the wood chips (15.4 MJ/kg); this is due to the presence of plastics within the pellets which averaged 20 wt% in total.

The rejects also have a much lower ash content compared to de-inking sludge, but the ash is significantly higher than for wood chips. From Table 8 the total volatile fraction and fixed carbon content of the pulper reject pellets is similar to that of wood chips. Chlorine content of the rejects is observed to be higher than that of wood chips and this is thought to be as a result of residual PVC material in the plastic pellets.

From the proximate and ultimate analysis results the use of rejects for thermal conversion by downdraft gasification would in principle not be problematic as the ash content is relatively low and the calorific value of the pellets high, which means sustained gasification of the fuel would be achievable. However the concern for gasification of the rejects lies with the level of plastics present within the pellets. Feeds with high plastics content in downdraft gasifiers can lead to agglomeration above the throat as the slow heating rate and compression of materials within the gasifier cause plastics to melt and stick together [39]. This has led to the approach adopted here to blend the plastics with wood chips, thus reducing the tendency for the plastics to bind together. Blending in this way also results in less chlorine containing species in the product gas. Chlorine compounds when combined with oxygen can form harmful dioxins if present in sufficient quantities, and these are extremely expensive to remove from exhaust gases.

6.3 Fixed bed downdraft gasification results and discussion

6.3.1 Gasification trial sequence

Table 9 shows the gasification trials which were carried out, using the GEK downdraft gasifier (see Section 5.5.1 for a description of this unit).

Table 9 Gasification trial sequence

Trial Number	Feedstock Blend (wt%)
1	20% AN Pulper Reject Pellets, 80% Wood Chips
2	50% AN Pulper Reject Pellets, 50% Wood Chips
3	30% AN Pulper Reject Pellets, 70% Wood Chips
4	10% AN Pulper Reject Pellets, 10% De-inking Sludge, 80% Wood Chips
5	15% AN Pulper Reject Pellets, 20% De-inking Sludge, 65% Wood Chips
6	10% AN Pulper Reject Pellets, 40% De-inking Sludge, 50% Wood Chips
7	20% SSK Pulper Reject Pellets, 80% Wood Chips
8	50% SSK Pulper Reject Pellets, 50% Wood Chips
9	70% SSK Pulper Reject Pellets, 30% Wood Chips
10	80% SSK Pulper Reject Pellets, 20% Wood Chips
11	100% SSK Pulper Reject Pellets
12	41% KC Pulper Reject Pellets, 15% Wet Co-form, 11% Dry Co-form, 33% Wood Chips

In the following Sections, each trial is described. Key results are presented as these sections proceed, but a full summary of results and associated discussion can be found in the Summary Section, 6.3.6.

6.3.2 Commissioning tests

The GEK gasifier was initially commissioned by carrying out several runs using mixed wood chip as the feedstock. This was done to optimise the operating procedure, establish the optimal running conditions and determine baseline performance. The first run resulted in significant tar formation as heat up of the gasifer was initiated without a char bed in place. Having a char bed in the reduction zone is fundamental as this is where the heterogeneous gasification reactions take place as hot gases from the pyrolysis and oxidation zones above react with carbon in the char [57]. The main reactions occurring in this zone include the Bouduard reaction, the water gas shift reactions and the methane production reactions. At high temperatures CO_2 produced from combustion in the oxidation zone reacts with carbon in the char and promotes the forward Boudouard reaction forming CO. Similarly CO and H_2 are produced when carbon in the char reacts with steam in the gas;

carbon in the char also reacts with some of the H_2 produced to form CH_4 [57]. Modifications to the second run included putting in place a charcoal bed and an active carbon filter to absorb condensable organic vapours downstream of the gasifier. This run proved to be successful, a consistent flare was achieved and a product gas with an average calorific value of 5 MJ/Nm^3 was formed. Tar measured by the tar sampling system was found to be 2 g/Nm^3 which was significantly less than the first run. Air inlet flow rate averaged 10 m^3/h and gasification temperatures in the oxidation zone of the gasifier averaged 1000 ^oC. This run established the operating basis for all subsequent trials.

6.3.3 *Gasification of AN mill rejects*

Initial trials attempted to gasify AN mill reject pellets on their own without the use of wood chips as a co-fired blend. However the gasifier suffered from agglomeration problems caused by melting of plastics. Figure 40 illustrates the extent of agglomeration.

Figure 40 Agglomeration caused by melting of plastics (6 inches diameter)

The focus of subsequent trials was to determine to what extent the pellets could be co-fired with wood chips before agglomeration occurred.

6.3.3.1 Trial 1: 20 wt% AN mill rejects, 80 wt% mixed wood chips.

The first co-fired trial was carried out using a blend of 20 wt% rejects with 80 wt% mixed wood chips. The reduction zone and throat of the gasifier was initially loaded with approximately 3.2 kg of charcoal. Above this region the gasifier was loaded with approximately 2.7 kg of dried mixed wood chips to ensure successful start-up and the achievement of steady state conditions. The gasifier hopper was loaded with a blend of 20 wt% reject pellets and 80 wt% wood chips which were pre-mixed together beforehand. The total loading in the hopper of the mixed feedstock was approximately 13.5 kg. The gasifier bed was ignited and the gasifier started well and steady state was soon established on wood. The flare was subsequently ignited and this was observed to be strong and stable throughout the duration of the experiment which lasted approximately 4 hours. Oxidation zone temperature averaged 1076 ^{o}C during steady state conditions. The gas outlet temperature averaged 67 ^{o}C at steady state conditions. The feed rate was determined by multiplying the average bulk density of the feedstock by the reduction in hopper volume over a given period of time and this was calculated to be approximately 3.7 kg/h of the mixed feedstock. An average air inlet flow rate of 6.71 m^3/h and an average gas outflow of 13.6 m^3/h were measured.

During steady state gasification approximately 232 litres of gas was withdrawn from the gasifier and routed through the tar sampling system. 4 product gas samples were taken at 30 minute intervals. As a result of a small air in leakage through the tar sampling system, oxygen was detected by the GC, therefore the results were corrected and normalised by eliminating the proportion of air in the sample to determine the true amount of product gas formed. Table 10 shows the normalised product gas GC results.

Table 10 Normalised product gas compositions Trial 1: 20 wt% AN mill rejects 80 wt% wood chips

Product Gas	Sample 1 vol%	Sample 2 vol%	Sample 3 vol%	Sample 4 vol%	Average vol%
N2	42.0	43.4	50.7	43.9	45.0
H2	19.0	17.7	11.4	16.5	16.2
CO	26.3	25.9	21.2	24.2	24.4
CO2	10.7	10.9	13.8	12.3	11.9
CH4	1.9	1.9	2.8	2.9	2.4

The average gas composition shown in Table 10 has a HHV of 6.3 MJ/Nm^3. The slightly higher N_2 content of sample 3 is believed to be the result of a higher intake of air into the gasifier, air flow rate varied slightly in order to control gasification temperatures.

The gasification experiment continued without incident for a total duration of 4 hours. No evidence of agglomeration of plastics in the reject feedstock was found, and this is believed to be due to the presence of non-plastics-containing particles between the plastics-containing particles, preventing them from contacting each other and thus binding. It was also observed that the ash grate did need to be agitated more often than with 100 % wood chips in order to dislodge some build-up of material. The tar content in the product gas was calculated to be 3.78 g/Nm^3, a figure higher than that achieved with 100 % wood chips (2 g/Nm^3).

The water content of the product gas was 16.7 g/Nm^3. High amounts of water vapour in the product gas reduce its calorific value, therefore it is important to reduce the level if possible. Further evaporative drying of the feedstock before gasification would achieve this, although it is important that the water content is not too low as some water vapour is required for the important water gas shit reactions occurring to produce H_2; also low water containing feedstocks can result in excessive gasification temperatures. This work determined that moisture content ranging from 10-15 wt% was found to be acceptable with using the GEK. The mass balance from the run is shown in Table 11.

Table 11 Mass balance Trial 1: 20 wt% AN mill rejects 80 wt% wood chips

	In (kg/h)			**Out (kg/h)**					**Closure**
Feedstock Blend (wt %)	**Air**	**Feed**	**Tot**	**Ash**	**Gas**	**H_2O**	**Tar**	**Tot**	**%**
20% AN Mill Reject Pellets, 80% Wood Chips	7.39	3.70	11.09	0.08	9.26	0.17	0.04	9.55	86

Table 11 shows that for every kg of feed input approximately 2.5 kg of product gas was formed. This is similar to what would be expected with downdraft gasifiers running on wood (about 2.6 kg) [57]. The product gas yielded an energy content of 18 kWth and the mass balance closed at 86 %. The source of error may have been due to a combination of factors including the inability to measure residual char in the gasifier. A more accurate closure could have been obtained by placing the gasifer on a weighing scale to measure total mass before and after each experiment, however due to the size of the unit it was impractical to implement this strategy.

In general as equipment is scaled up it becomes increasingly difficult to obtain accurate closures as there is increased tendency for errors to occur. A major source of error often experienced is from air entering the gasifier through small gaps between the fittings and lose seals, although it was attempted to minimise this by performing a cold run to identify any gas leaks before each hot run.

6.3.3.2 Trial 2: 50 wt% AN mill rejects, 50 wt% mixed wood chips.

The second gasification trial attempted to gasify 50 wt% reject pellets with 50 wt% mixed wood chips. The hopper was loaded with a total of 17.5 kg of the blended feedstock. The gasifier bed was ignited and the gasifier started well and steady state was soon established on wood. The flare was ignited and this was observed to be stable for approximately 60 minutes and during this period the gasification temperature was maintained within the range of between 800^{o}C to 1200^{o}C.

Approximately 15 minutes after the start-up period a significant decrease in reaction temperature was observed, from 1200 ^{o}C to < 600 ^{o}C. During this period the flare could not be sustained and a significant reduction in feed rate indicated that there was blockage occurring within the reactor. Blockage was also evident from a significantly reduced pressure drop measured across the bed (this was recorded to be around 600 Pa, whereas under normal operation it is >1200 Pa) and the associated significant drop in air flow rate from 10 m^3/h to less than 2 m^3/h. After a further 5 minutes had elapsed no more feedstock could be fed into the gasifier from the hopper. At this point it was apparent that no further gasification of this blend would be achieved. The reaction was terminated for further inspection after cool down. Upon opening the gasifier plastic pellet agglomeration was apparent. This is shown in Figure 41

Figure 41 Agglomeration caused by melting of plastics gasification Trial 2

This experiment demonstrated that 50 wt% reject pellets cannot be successfully gasified with wood chips as a result of high levels of plastics within the pellet.

6.3.3.3 Trial 3: 30 wt% AN mill rejects, 70 wt% mixed wood chips.

The third gasification run was carried out using a lower blend ratio of reject pellets, 30 wt% rejects with 70 wt% mixed wood chips. The gasifier hopper was loaded with 12.8 kg of the prepared feedstock. The gasifier bed was ignited and the gasifier started well. The flare was ignited and this was observed to be stable for approximately 2 hours; during this period the gasification temperature at the oxidation zone was between 800^oC and 1200^oC. The feed rate was calculated to be 4.2 kg/h. After approximately 2 hours the flare could not be sustained and a significant reduction in feed rate indicated that again there was blockage within the reactor. This was supported again by reductions in the bed pressure drop and air flow rate. Once again gasification was terminated for further inspection. Plastic pellet agglomeration (Figure 42) was again evident, although the extent of the agglomeration appeared to be much less than that in Trial 2.

Figure 42 Agglomeration caused by melting of plastics gasification Trial 3

From Trials 1-3, the optimal ratio of AN mill rejects in a blend with wood chips while avoiding bed agglomeration was found to be approximately 20 wt%.

6.3.3.4 Trial 4: 10 wt% AN mill rejects, 10% AN mill de-inking sludge, 80 wt% mixed wood chips.

The next series of trials introduced de-inking sludge into the reject/wood chip mix while keeping the ratio of rejects to wood chips at or below approximately 20 wt%.Trial 4 blended 10 wt% rejects with 10 wt% de-inking sludge pellets and 80 wt% wood chips. The gasifier hopper was loaded with a total of 12.5 kg of the mixed feedstock. The gasifier bed was ignited and the gasifier started well and steady state was soon established on wood.

The flare was ignited and this was observed to be strong and stable and this lasted throughout the duration of the experiment which was approximately 3 hours. Gasification temperatures within the oxidation zone of the gasifier averaged 1000 °C. The gas outlet temperature was measured after cyclone separation and averaged 62 °C at steady state conditions. Pressure difference across the gasification bed averaged 950 Pa, and the differential pressure measured between the active carbon filter outlet and atmosphere averaged 1100 Pa. The feed rate was calculated to be approximately

4.2 kg/h. The flow rate of air into the gasifier averaged 5 Nm^3/h and product gas flow rate out of the gasifier averaged 12 Nm^3/h.

During steady state operation a sample of approximately 237 litres of gas was withdrawn from the gasifier and routed through the tar sampling system. 4 gas injections were taken at timed 30 minute intervals. Table 12 shows the normalised GC results.

Table 12 Normalised product gas compositions Trial 4: 10 wt% AN mill rejects, 10% AN mill de-inking sludge, 80 wt% mixed wood chips

Product Gas	Sample 1 vol%	Sample 2 vol%	Sample 3 vol%	Sample 4 vol%	Average vol%
N2	40.5	46.4	48.1	54.1	47.3
H2	18.7	16.2	12.0	10.7	14.4
CO	26.1	25.0	24.9	21.4	24.4
CO2	12.7	9.9	12.2	12.3	11.8
CH4	2.0	2.3	2.8	1.5	2.2

The average composition corresponds to an HHV of 6 MJ/Nm^3. Gasification continued without incident throughout the experiment which lasted 3 hours. After the experiment there was no evidence of agglomeration in the bed. The tar level in the product gas was 2.15 g/ Nm^3, and the water content 11 g/Nm^3. The mass balance from the run is shown in Table 13.

Table 13 Mass balance Trial 4: 10 wt% AN mill rejects 10 wt% de-inking sludge 80 wt% wood chips

	In (kg/h)			Out (kg/h)					Closure
Feedstock Blend (wt%)	Air	Feed	Tot	Ash	Gas	H_2O	Tar	Tot	%
10% AN Mill Reject Pellets, 10% De-inking Sludge 80% Wood Chips	11.28	4.22	15.50	0.27	13.46	0.14	0.03	13.90	90

Table 13 shows that for every kg of feed input approximately 3.18 kg of product gas was formed. This is slightly higher than that which would be expected with downdraft gasifiers running on wood

[57], where about 2.6 kg would typically be formed. This product gas had an energy content of 6.28 kWth and the mass balance closed at 90 %. As this run was successful the next run aimed to increase the proportion of de-inking sludge in the blend while keeping the ratio of rejects to wood chips at or below 20 wt%.

6.3.3.5 Trial 5: 15 wt% AN mill rejects, 20% AN mill de-inking sludge, 65 wt% mixed wood chips.

The blend used in the fifth trial was 15 wt% rejects, 20 wt% de-inking sludge and 65 wt% wood chips. The gasifier hopper was loaded with 15.57 kg of the blended feedstock. The gasifier bed was ignited and the gasifier started well and steady state was soon established on wood. The flare was ignited and this was observed to be strong and stable throughout the duration of the experiment. Gasifer temperatures averaged 1075 ^{o}C in the oxidation zone. The gas outlet temperature measured after the cyclone averaged 74 ^{o}C at steady state conditions. Pressure difference across the gasification bed averaged 670 Pa this was much lower than that observed with previous runs; the cause may have been due to a less restrictive bed within the gasifier. The feed rate was approximately 6.24 kg/h, the flow rate of both air into the gasifier averaged 8.13 m^3/h and gas outflow averaged 19 Nm^3/h.

During steady state operation a sample of approximately 186 litres of gas was withdrawn from the gasifier and routed through the tar sampling system. 3 gas samples were taken in total and the results from the 3 manual injections are presented in Table 14.

Table 14 Normalised product gas compositions Trial 5: 15 wt% AN mill rejects 20 wt% AN mill de-inking sludge, 65 wt% wood chips

Product Gas	**Sample 1 vol%**	**Sample 2 vol%**	**Sample 3 vol%**	**Average vol%**
N2	44.4	50.9	47.0	47.5
H2	16.4	13.4	15.3	15.0
CO	27.6	23.6	22.9	24.7
CO2	10.9	11.6	13.0	11.9
CH4	0.6	0.4	1.7	0.9

The average composition corresponds to an HHV of 4.2 MJ/Nm^3. Tar content of the product gas was 4.8 g/Nm^3. The water content was 15.6 g/Nm^3. The mass balance from the run is shown in Table 15.

Table 15 Mass balance Trial 5: 15 wt% AN mill rejects, 20 wt% AN mill de-inking sludge, 65 wt% wood chips

	In (kg/h)			Out (kg/h)					Closure
Feedstock Blend (wt%)	**Air**	**Feed**	**Tot**	**Ash**	**Gas**	**H_2O**	**Tar**	**Tot**	**%**
15% AN Mill Reject Pellets, 20% AN Mill De-inking Sludge, 65% Wood Chips	9.21	5.27	14.48	0.63	17.34	0.30	0.09	18.37	127

Table 15 shows that for every kg of feed input approximately 3.3 kg of product gas was formed. The product gas had an energy content of 7.5 kWth. As this run was successful the proportion of de-inking sludge was increased in the next run.

6.3.3.6 Trial 6: 10 wt% AN mill rejects, 40% AN mill de-inking sludge, 50 wt% mixed wood chips.

The sixth gasification trial was carried out using a blend of 10 wt% rejects with 40 wt% de-inking sludge and 50 wt% wood chips. The gasifier hopper was loaded with 19 kg of the blended feedstock. The flare was ignited and this was observed to be strong and stable and lasted throughout most of the 4 hour experiment. Gasification temperatures within the oxidation zone of the gasifier averaged 1052 °C. The gas outlet temperature measured after the cyclone averaged 95 °C at steady state conditions. Pressure difference across the bed averaged 880 Pa. The differential pressure was also measured between the carbon filter outlet and atmosphere and this averaged 1600 Pa. Therefore total pressure drop between the gasification bed and filter was 720 Pa which was significantly higher than observed with other successful runs. The feed rate was measured to be approximately 10 kg/h. The average flow rate of air into the gasifier was 9.4 Nm^3/h and average gas outflow was 21 Nm^3/h.

During steady state operation a sample of approximately 198 litres of gas was withdrawn from the gasifier and routed through the tar sampling system. 3 gas samples were taken at 30 minute intervals; the results from the 3 manual injections are given in Table 16.

Table 16 Normalised product gas compositions Trial 6: 10 wt% AN mill rejects 40 wt% AN mill de-inking sludge and 50 wt% wood chips

Product Gas	**Sample 1 vol %**	**Sample 2 vol %**	**Sample 3 vol %**	**Average vol%**
N_2	37.0	48.4	63.6	49.7
H_2	16.8	11.8	5.9	11.5
CO	27.9	22.9	14.5	21.8
CO_2	15.0	15.6	15.6	15.4
CH_4	3.2	1.2	0.4	1.6

The average composition corresponds to a HHV of 5 MJ/Nm^3. This experiment did not require continous agitation, however as the experiment progressed the reaction seemed to slow down until it eventually self terminated. This was the result of a build up of de-inking sludge ash within the ash grate and reduction zone of the gasifier. This was confirmed when the gasification bed was inspected after cooling. Figure 43 shows de-inking sludge ash build up in the reduction zone. It is identified by its distinctive white CaO colouring.

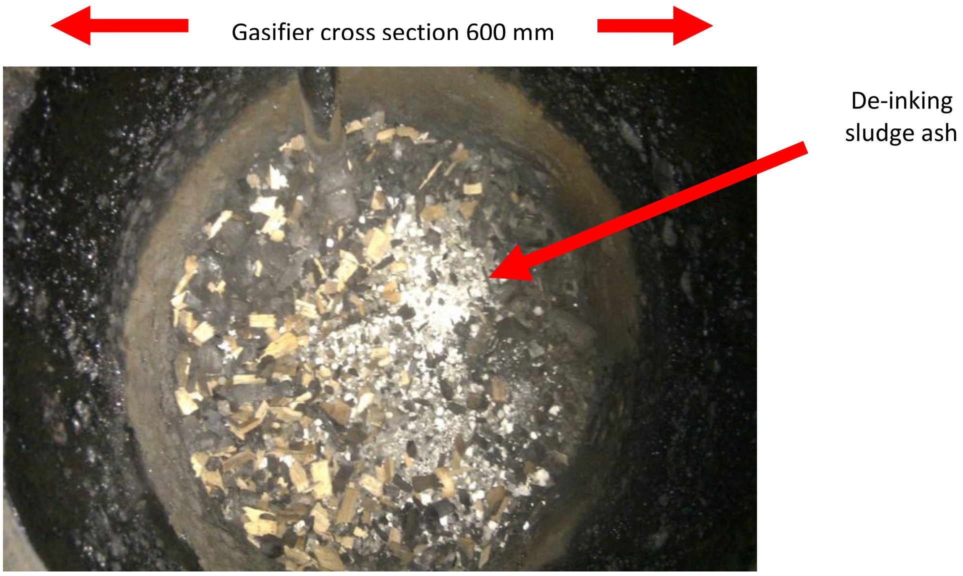


Figure 43 De-inking sludge ash in reduction zone

Significant ash build up in the reduction zone increases gasification bed density; as ash forms the gasifier bed becomes more compact and this restricts gas flow through the reactor. When char in in the reduction zone is replaced with ash reduced reduction reactions occur which are vital to sustain gasification. The significant build up of de-inking sludge ash in the reduction zone was believed to be the principal cause of the experiment self terminating. For succesful gasification of higher blends, it would be essential to have a continuous ash removal system.

From this experiment it was also evident that de-inking sludge pellets were not holding their form in the feeding process, but were crumbling instead. This was mainly found around the screw feeder of the hopper which may have also contributed to the high pressure drop in the gasifier bed (Figure 44).

Figure 44 Crumbled AN mill de-inking sludge pellets

Stronger pellets would limit the formation of fines. Pellet strength studies have been carried out as previously disscussed [37] but are beyond the scope of this work.

Approximately 198 litres of gas was sampled in the tar sampling system.The tar content in the product gas was 1.9 g/Nm^3. The water content was 15.52 g/Nm^3. The mass balance from the run is shown in Table 17.

Table 17 Mass balance Trial 6: 10 wt% AN mill rejects 40 wt% de-inking sludge 50 wt% wood chips

	In (kg/h)			Out (kg/h)					Closure
Feedstock Blend (wt%)	**Air**	**Feed**	**Tot**	**Ash**	**Gas**	**H_2O**	**Tar**	**Tot**	**%**
10% AN Mill Reject Pellets, 40% AN Mill De-inking Sludge, 50% Wood Chips	11.28	10.26	21.54	2.25	19.02	0.33	0.04	21.64	100

Table 17 shows that for every kg of feed input approximately 1.85 kg of product gas was formed. The product gas had an energy content of 6.3 kWth.

This experiment determined that the maximum amount of de-inking sludge which can be blended with wood chips and rejects in a downdraft gasifier is greater than 20 % but less than 40 wt%. High de-inking sludge fractions result in significant ash build up in the gasifier thus leading to the reactions self terminating.

The results overall which were obtained from the AN mill reject gasification experiments show that two out of the six experiments failed due to bed agglomeration, and these utilised reject portions of 30 wt% or greater, indicating (together with observations) that the cause of the agglomeration was the high proportion of plastics within the mixture. As stated in Section 2.4 the AN mill accepts recycled paper with a tendency to contain a broader range of plastics than (for example) the SSK mill. These plastics are extremely heterogeneous and vary significantly over time, which makes their quantification difficult. However, in general they are often composed of a combination of plastic film wrap (soft plastics) and (hard plastics) from CDs, plastic bottles etc. It is the latter which is believed to be the principle cause of agglomeration witnessed here; as the gasification of rejects which contained mainly soft plastics had significantly reduced tendency to agglomerate (see Section 6.3.4).

All the failed experiments demonstrated similar behaviour; they started well, but as time elapsed the feeding rate reduced drastically with inconsistent flare quality. Whenever this was observed, the runs would self-terminate. The main cause of the failure was established when the gasifier was opened and the state of the gasification bed inspected. In all these experiments, agglomeration appeared to occur mainly above the throat of the gasifier. It was also found that the reduction zone of the gasifier was virtually empty with no char, indicating that all the residual char in the bed had combusted and had not been replaced with fresh char moving down. Char is the main source of carbon in the reduction zone and plays an important role in gasification reactions. Therefore the absence of char in this zone meant the important reduction reactions could not occur, leading to less CO and H_2 formation from water gas shift reactions and Boudouard reactions and less methane formation from methanation reactions. Furthermore as fresh material could not flow freely through the gasifier important oxidation reactions of the volatiles could not occur. The absence of combustible gases was the cause of the inconsistent flare.

The optimum wood-rejects mixture was observed to be 20 wt% pellets and 80 wt% wood chips. It is thought that the reduced amount of plastics-rich reject pellets present in the feedstock resulted in better separation between and isolation of agglomerating particles and therefore fewer tendencies to fuse together.

The experiments that utilised de-inking sludge as part of the feedstock were all successful. However, there were some issues that that were experienced during the experiments such as the crumbling of the de-inking sludge pellets and reduced gasification performance caused by the high levels of ash present in the de-inking sludge.

6.3.4 *Gasification of SSK mill feedstocks*

6.3.4.1 Trial 7: 20 wt% SSK mill rejects, 80 wt% mixed wood chips.

The gasification of SSK rejects began with the testing of 20 wt% rejects with 80 wt% wood chips. The gasifier hopper was loaded with a total of 17 kg of the prepared feedstock. The gasifier bed was ignited and the gasifier started well and steady state was soon established. The flare was ignited and this was observed to be stable for the duration of the experiment. Gasification oxidation zone temperature was kept within the range 800°C-1200°C. The feed rate was calculated to be approximately 5 kg/h. The average flow rate of air into the gasifier was 10 m^3/h and of product gas out of the gasifier 17 m^3/h. During steady state operation a sample of approximately 281 litres of gas was withdrawn from the gasifier and routed through the tar sampling system, and a sample of approximately 100 micro litres of gas was manually injected into the GC device. On this occasion only one gas sample could be measured due to a technical error with the GC. The result is presented in Table 18

Table 18 Normalised product gas compositions Trial 7: 20 wt% SSK mill rejects, 80 wt% wood chips

Product Gas	Sample 1 vol %
N_2	51.5
H_2	11.0
CO	19.1
CO_2	16.1
CH_4	2.3

The composition shown in Table 18 has an HHV of 4.9 MJ/Nm^3. The gasification experiment continued without incident for a total duration of 3 hours and there was no evidence of agglomeration in the bed. Tar content of the product gas was 2 g/Nm^3 and water content 21 g/Nm^3. The mass balance from the run is shown in Table 19.

Table 19 Mass balance Trial 7: 20 wt% SSK mill rejects 80 wt% wood chips

	In (kg/h)			Out (kg/h)					Closure
Feedstock Blend (wt%)	**Air**	**Feed**	**Tot**	**Ash**	**Gas**	**H_2O**	**Tar**	**Tot**	**%**
20% SSK Mill Reject Pellets, 80% Wood Chips	7.33	4.63	11.96	0.09	11.65	0.27	0.03	12.05	101

Table 19 shows that for every kg of feed input approximately 2.5 kg of product gas was formed. The product gas had an energy content of 5.3 kWth and the closure was 101 %.

6.3.4.2 Trial 8: 50 wt% SSK mill rejects, 50 wt% mixed wood chips.

The second SSK run increased the SSK reject blend to 50 wt% with 50 wt% wood chips. The hopper was initially loaded with 18 kg of feedstock. The gasifier bed was ignited and the gasifier started well and steady state was soon established. The flare was ignited and this was observed to be stable for the duration of the experiment. Gasification oxidation zone temperature averaged 1000 ^{o}C at steady state conditions. The feed rate was measured to be 7 kg/h. The flow rates of air into the gasifier and product gas out of the gasifier were 8.4 Nm^3/h and 24 Nm^3/h respectively. During steady state operation a sample of approximately 381 litres of gas was withdrawn from the gasifier and routed through a tar sampling system, and 4 samples of approximately 100 micro litres of gas were manually injected into the GC device. Results are given in Table 20.

Table 20 Normalised product gas compositions Trial 8: 50 wt% SSK mill rejects, 50 wt% mixed wood chips.

Product Gas	Sample 1 vol %	Sample 2 vol %	Sample 3 vol %	Sample 4 vol %	Average vol%
N_2	16.1	46.4	23.3	44.5	38.1
H_2	24.9	16.2	21.7	15.2	17.7
CO	47.4	25.1	47.4	32.5	35.0
CO_2	6.5	10.0	4.8	6.1	6.9
CH_4	5.2	2.3	2.7	1.5	2.2

The average gas composition has an HHV of 8 MJ/Nm^3. The gasification experiment continued without incident for a total duration of 4 hours with the flare observed to be strong and consistent. There was no evidence of agglomeration in the bed; this contrasts with Trial 2 using AN mill reject pellets, and is thought to be due to the lower levels of hard plastics present in SSK rejects, as this feedstock was mainly comprised of paper fibres and plastic polyethylene wrap. Upon inspection of the gasifier bed after the run it was evident that some ash slagging was present; however this could easily be rectified with careful control of gasification temperature as well as a continuous ash removal system. Fluctuating air flow rate during the run may have been the cause of the lower N_2 experienced in sample 1, Table 20. Tar content of the product gas was 0.89 g/Nm^3. The low level may be the result of increased temperatures experienced in the gasifier as a result of increased level of plastics. Water content of the product gas was 6.43g / Nm^3. The mass balance from the run is shown in Table 21.

Table 21 Mass balance Trial 8: 50 wt% SSK mill rejects, 50 wt% wood chips

	In (kg/h)			Out (kg/h)					Closure
Feedstock Blend (wt%)	**Air**	**Feed**	**Tot**	**Ash**	**Gas**	**H_2O**	**Tar**	**Tot**	**%**
50% SSK Mill Reject Pellets, 50% Wood Chips	10.94	7.14	18.08	0.32	19.26	0.16	0.02	19.77	109

Table 21 shows that for every kg of feed input approximately 2.7 kg of product gas was formed. The product gas had an energy content of 10 kWth and the closure was 109 %.

6.3.4.3 Trial 9: 70 wt% SSK mill rejects, 30 wt% mixed wood chips.

The third SSK gasification test was carried out using a blend of 70 wt% SSK mill pellets with 30 wt% wood chips. The gasifier hopper was loaded with a total of 22 kg of the prepared feedstock. The gasifier bed was ignited and the gasifier started well and steady state was soon established. The flare was ignited and this was observed to be stable throughout the duration of the experiment. Gasification oxidation zone temperatures measured on this occasion were in excess of 1000 ^{O}C. The feed rate was 5.7 kg/h. The flow rates of air into the gasifier and product gas out of the gasifier were approximately 10.8 Nm^3/h and 29 Nm^3/h respectively.

During steady state operation a sample of approximately 396 litres of gas was withdrawn from the gasifier and routed through a tar sampling system 3 samples of approximately 100 micro litres of gas was manually injected into the GC device. The results from the manual injection are given in Table 22.

Table 22 Normalised product gas compositions Trial 9: 70 wt% SSK mill rejects, 30 wt% mixed wood chips.

Product Gas	Sample 1 vol %	Sample 2 vol %	Sample 3 vol %	Average vol%
N_2	55.8	44.6	50.9	50.4
H_2	13.9	21.2	14.8	16.6
CO	23.3	25.9	24.3	24.5
CO_2	5.8	6.2	8.6	6.9
CH_4	1.2	1.9	1.3	1.5

The average gas composition has an HHV of 6 MJ/Nm^3. The gasification experiment continued without fault for a total duration of 3 hours there was no evidence of agglomeration of plastics in the reject waste pellet feedstock and the flare was observed to be strong and consistent. The tar content of the product gas was 4.4 g/Nm^3 and the water content 70 g/ Nm^3. The mass balance from the run is shown in Table 23.

Table 23 Mass balance Trial 9: 70 wt% mill rejects, 30 wt% wood chips

	In (kg/h)			Out (kg/h)					Closure
Feedstock Blend (wt%)	**Air**	**Feed**	**Tot**	**Ash**	**Gas**	**H_2O**	**Tar**	**Tot**	**%**
70% SSK Mill Reject Pellets, 30% Wood Chips	13.32	5.74	19.06	0.36	22.79	2.06	0.13	25.35	133

Table 23 shows that for every kg of feed input approximately 4 kg of product gas was formed. The product gas had an energy content of 12 kWth. The closure of the mass balance for this run was very poor, due to a damaged air inlet pipe which occurred toward the end of the run after all measurements had been taken. The pipe allowed a significant amount of air into the gasifier which resulted in erroneous flow readings. This pipe was subsequently repaired immediately after the run.

6.3.4.4 Trial 10: 80 wt% SSK mill rejects, 20 wt% mixed wood chips.

The fourth SSK gasification run used a blend of 80 wt% SSK mill rejects with 20wt% wood chips. The gasifier hopper was loaded with 25 kg of the prepared feedstock. The gasifier bed was ignited and the gasifier started well and steady state was soon established. The flare was ignited and this was observed to be stable for the duration of the experiment. Gasification temperatures in the oxidation zone remained in excess of 1000 ^{o}C. The feed rate was measured to be 6 kg/h. The flow rate of air into the gasifier was 7 Nm^3/h and product gas out was 15.5 Nm^3/h. During steady state operation a sample of approximately 326 litres of gas was withdrawn from the gasifier and routed through a tar sampling system. 4 samples of gas were extracted and injected into the GC. The results from the manual injections are presented in Table 24.

Table 24 Normalised product gas compositions Trial 10: 80 wt% SSK mill rejects, 20 wt% mixed wood chips.

Product Gas	Sample 1 vol %	Sample 2 vol %	Sample 3 vol %	Sample 4 vol %	Average vol%
N_2	41.9	44.5	38.9	44.6	42.5
H_2	17.8	15.9	14.2	17.0	16.2
CO	23.4	23.7	24.6	21.6	23.3
CO_2	11.7	10.9	16.2	12.0	5.2
CH_4	5.1	4.9	6.0	4.7	12.7

The average composition had an HHV of 7.3 MJ/Nm^3. The gasification experiment continued without incident for a total duration of 3.5 hours and the flare was observed to be strong and consistent. There was no evidence of agglomeration in the bed, although the bed did need to be agitated regularly. The tar content of the product gas was 5.9 g/Nm^3 and water content 21 g/Nm^3. The mass balance from the run is shown in Table 25.

Table 25 Mass balance Trial 10: SSK 80 wt% rejects 20 wt% wood chips gasification

	In (kg/h)			Out (kg/h)					Closure
Feedstock Blend (wt%)	**Air**	**Feed**	**Tot**	**Ash**	**Gas**	**H_2O**	**Tar**	**Tot**	**%**
80% SSK Mill Reject Pellets, 20% Wood Chips	8.43	5.62	14.02	0.4	13.5	0.33	0.09	14.32	102

Table 25 shows that for every kg of feed input approximately 2.4 kg of product gas was formed. The product gas had an energy content of 7.4 kWth and the closure was 102 %.

6.3.4.5 Trial 11: 100 wt% SSK mill rejects.

The fifth SSK gasification run attempted to gasify 100 % SSK reject pellets without the use of wood chips as a co-fired blend. However this run was unsuccessful due to bed agglomeration. Upon inspection it was evident that the melting of plastics present in the SSK pellets was the principle cause of the agglomeration (Figure 45).

Figure 45 SSK gasification Trial 11 agglomeration

Hence the maximum proportion of SSK mill reject pellets in a blend with wood chips that was successfully tested was 80%.

6.3.5 *Gasification of KC mill feedstocks*

Only 900 and 500 tonnes per year of rejects and co-form respectively are produced at the whole Kimberly Clark Flint site. These are small amounts, and at full scale a high fraction of wood chips would be required to fuel a full scale 250 kg/h downdraft gasifier on the reject stream alone. The wood would obviously need to be purchased by the mill, and initial estimates indicate that it would not be economically feasible for KC to implement such a system for the reject stream. The only configuration which was believed to be feasible was to blend the rejects and co-form dry and wet wipes together in the proportions that they arise from the KC mill, and then co-fire this blend with wood in a proportion that corresponded to the full utilisation of the KC waste streams over a year in a 250 kg/h gasifier. The blend that was tested to achieve this was 41 wt% pulper reject pellets, 15 wt% wet co-form, 11 wt% dry co-form, 33 wt% wood chips. However major bed agglomeration was encountered and it was decided that no further testing of this waste stream would be carried out.

6.3.6 Summary of gasification results

Table 26 shows the feedstock blends tested in gasification trials and whether the trials were successful.

Table 26 Feedstock blends tested and gasification performance

Trial Number	Feedstock Blend (wt%)	Status
1	20% AN Pulper Reject Pellets, 80% Wood Chips	Successful
2	50% AN Pulper Reject Pellets, 50% Wood Chips	Unsuccessful
3	30% AN Pulper Reject Pellets, 70% Wood Chips	Unsuccessful
4	10% AN Pulper Reject Pellets, 10% De-inking Sludge, 80% Wood Chips	Successful
5	15% AN Pulper Reject Pellets, 20% De-inking Sludge, 65% Wood Chips	Successful
6	10% AN Pulper Reject Pellets, 40% De-inking Sludge, 50% Wood Chips	Successful
7	20% SSK Pulper Reject Pellets, 80% Wood Chips	Successful
8	50% SSK Pulper Reject Pellets, 50% Wood Chips	Successful
9	70% SSK Pulper Reject Pellets, 30% Wood Chips	Successful
10	80% SSK Pulper Reject Pellets, 20% Wood Chips	Successful
11	100% SSK Pulper Reject Pellets	Unsuccessful
12	41% KC Pulper Reject Pellets, 15% Wet Co-form, 11% Dry Co-form, 33% Wood Chips	Unsuccessful

From Table 26 it is observed that the unsuccessful trials were in most cases from the testing of AN mill and KC mill rejects and co-form pellets, and this is thought to be largely due to the levels of hard plastics present within the pellets which caused agglomeration and blockages within the gasifier. The most successful trials were with SSK mill rejects. Initial trials attempted to gasify the reject pellets without the use of wood chips as a co-fired blend. However the gasifier suffered from agglomeration problems caused by melting of plastics. In all cases where it occurred, agglomeration was found to be mainly within the pyrolysis zone of the gasifier at moderate temperature levels; as the plastics reach this zone they become soft and extremely sticky causing bridging and binding above the gasifier throat, and this causes reduced flow within the gasifier unit and leads ultimately to unsuccessful gasification. One of the most important factors when using the downdraft gasifier is the ability for feedstock to freely move through the unit by gravity. Note this problem would be reduced in a fluidised bed gasifier, where the heating rate is much higher and particles entering the gasifier reach full reactor temperature almost instantaneously.

The focus of subsequent trials was to determine to what extent the pellets could be co-fired with wood chips before agglomeration occurred. Trials 1, 2 and 3 focused on the gasification of AN rejects with wood chips, however the maximum blend which could be achieved in these cases was only 20 wt % rejects with 80 wt % wood chips. Trials 4, 5 and 6 focused on introducing AN de-inking sludge to the blend whilst keeping the ratio of rejects to wood chips at or below approximately 20 wt%. The de-inking sludge blend was then gradually increased to determine the maximum blend of rejects and de-inking sludge which could be co-fired with wood chips. The maximum blend which was achievable in this mix was found to be below 40 wt% of de-inking sludge; at higher percentages the ash content of the gasifier bed rose to levels which were unacceptable for gasification, with limited carbon content and excessive pressure drop.

The most successful trials were the blending of the SSK reject pellets with wood chips (Trials 7, 8, 9 and 10). The most successful of these was the blending of 80 wt % SSK pellets with 20 wt% wood chips. It is the composition of this particular feedstock which is thought to be the key parameter for its successful gasification. It contained a lower proportion of hard plastics than the other rejects tested.

In Trial 12 which gasified KC mill feedsocks with wood, the level of plastics present was again too high for successful gasification, causing major agglomeration.

Table 27 shows the composition of the product gases from the successful trials. The air equivalence ratio is included. This is defined as the actual volume flow rate of air consumed (AAI) divided by the actual feed mass flow rate multiplied by the stoichiometric volume of air required to combust 1 kg of feed [58] and is calculated by Equation 7.

$$ER = \frac{AAI}{(AFI \times SA)} \qquad (7)$$

AAI = Actual air volume flow rate consumed (m^3/h)

AFI = Actual feed mass flow rate consumed (kg/h)

SA = Stoichiometric air required to combust 1 kg feed (m^3/kg)

Each successful run produced a low/medium product gas heating value and when mixed with air and ignited a strong, consistent flare was achieved.

Table 27 Product gas volume compositions of successful gasification trials (vol%)

Trial N°	**Feedstock Blend (Wt %)**	**H_2**	**N_2**	**CO**	**CH_4**	**CO_2**	**Gas H.V** (MJ/Nm^3)	**Air Equivalence Ratio**
1	20% AN Mill Reject Pellets, 80% Wood Chips,	16.2	45.0	24.4	2.4	11.9	6.3	0.36
4	10% AN Mill Reject Pellets, 10% AN Mill De-inking Sludge, 80% Wood Chips,	14.4	47.3	24.3	2.2	11.8	6	0.53
5	15% AN Mill Reject Pellets 20% AN Mill De-inking Sludge, 65% Wood Chips,	15.0	47.5	24.7	0.9	11.9	4.2	0.36
6	10% AN AN Mill Reject Pellets 40% AN Mill De-inking Sludge, 50% Wood Chips	11.5	49.7	21.8	1.6	15.4	5	0.27
7	20% SSK Mill Reject Pellets, 80% Wood Chips	11.0	51.5	19.1	2.3	16.1	4.9	0.28
8	50% SSK Mill Reject Pellets, 50% Wood Chips	17.7	38.1	35.0	2.2	6.9	8	0.24
9	70% SSK Mill Reject Pellets, 30% Wood Chips	16.6	50.4	24.5	1.5	6.9	6	0.34
10	80% SSK Mill Reject Pellets, 20% Wood Chips	16.2	42.5	23.3	5.2	12.7	7.3	0.22

Trials 8 and 10 with SSK mill rejects produced the highest calorific value gases overall (8 and 7.3 MJ/Nm^3 respectively) with generally elevated levels of H_2 and other combustible gases as well as lower amounts of N_2. This can be explained by the fact that both these runs used a low air equivalence ratio in comparison to all other runs which enhanced the product gas heating value.

In all trials high levels of N_2 were present in the product gas as a result of using air as the oxidising medium. Using oxygen enriched air as the oxidising medium would reduce the level of N_2 present and thus would increase the calorific value of the gases produced, although a cost would be associated with the enrichment. However oxygen rich air may lead to an excessive rise in gasification temperatures which could cause ash slagging and damage to gasifier components.

CO_2 produced in all runs did not exceed 17 v/v% and a proportion of the CO_2 produced is considered to be carbon neutral as it is derived from wood chips and paper fibres which originate from wood pulp (a carbon neutral source of biomass). It is also observed from Table 27 that the effect of increasing the de-inking sludge content in the feed was to increase level of CO_2 in the product gas. This is thought to be the result of calcination reactions of $CaCO_3$ present within de-inking sludge ash which occur above 700 ^{o}C to form CaO and CO_2. Increasing the level of CO_2 in the product gas has a diluting effect and reduces the overall gas calorific value. Therefore to maximise product gas heating value the de-inking sludge content should be kept to a minimum.

Table 28 shows both the tar and water content of the product gas from each successful gasification trial. After each run the gasifier was disassembled and some traces of tar deposits in outlet piping and especially around the venturi ejector were found.

Table 28 Tar and water condensate

Trial N°	Feedstock Blend (wt%)	Tar (g/Nm^3)	Water (g/Nm^3)
1	20% AN Reject Pellets, 80% Wood Chips	3.78	16.70
4	10% AN Mill Reject Pellets, 10% AN Mill De-inking Sludge, 80% Wood Chips	2.15	11.00
5	15% AN Mill Reject Pellets, 20% AN Mill De-inking Sludge, 65% Wood Chips	4.80	15.60
6	10% AN Mill Reject Pellets, 40% AN Mill De-inking Sludge, 50% Wood Chips	1.90	15.50
7	20% SSK Mill Reject Pellets, 80% Wood Chips	2.00	21.00
8	50% SSK Mill Reject Pellets, 50% Wood Chips	0.89	6.40
9	70% SSK Mill Reject Pellets, 30% Wood Chips	4.40	70.20
10	80% SSK Mill Reject Pellets, 20% Wood Chips	5.80	21.00

Water condensate formation in outlet piping was found to be minimal, which was due to the low moisture content of the feedstock. The tar content ranged from 1.9-4.8 g/Nm^3 for the AN mill tests and 0.89-5.8 g/Nm^3 for the SSK mill tests, and the water content ranged from 11-16.7 g/Nm^3 for the AN mill and 6.43-70.2 g/Nm^{3}for the SSK mill. These tar contents are in many cases higher than those observed from wood gasification, which is typically 1-2 g/Nm^3 in this type of gasifier [59]. However at full scale careful control of gasification temperatures along with the use of downstream tar clean up equipment such as scrubbers, filters or tar crackers has been shown to reduce the amount of tar in the product gas to acceptable levels for use in an engine [57]. In this work tar removal was achieved using a carbon absorption filter, but tar levels downstream of the filter were not measured. Table 29 shows the mass balance closures for each successful gasification trial.

Table 29 Mass balance and closures of each successful gasification trial.

Feedstock Blend (wt%)	In (kg/h)			Out (kg/h)					Closure %
	Air	Feed	Tot	Ash	Gas	H2O	Tar	Tot	
20% AN Mill Reject Pellets, 80% Wood Chips	7.39	3.70	11.09	0.08	9.26	0.17	0.04	9.55	86
10% AN Mill Reject Pellets, 10% AN Mill De-inking Sludge, 80% Wood Chips	11.28	4.22	15.50	0.27	13.46	0.14	0.03	13.90	90
15% AN Mill Reject Pellets, 20% AN Mill De-inking Sludge, 65% Wood Chips	9.21	5.27	14.48	0.63	17.34	0.30	0.09	18.37	127
10% AN Mill Reject Pellets, 40% AN Mill De-inking Sludge, 50% Wood Chips	11.28	10.26	21.54	2.25	19.02	0.33	0.04	21.64	100
20% SSK Mill Reject Pellets, 80% Wood Chips	7.33	4.63	11.96	0.09	11.65	0.27	0.03	12.05	101
50% SSK Mill Reject Pellets, 50% Wood Chips	10.94	7.14	18.08	0.32	19.26	0.16	0.02	19.77	109
70% SSK Mill Reject Pellets, 30% Wood Chips	13.32	5.74	19.06	0.36	22.79	2.06	0.13	25.35	133
80% SSK Mill Reject Pellets, 20% Wood Chips	8.43	5.62	14.02	0.4	13.5	0.33	0.09	14.32	102

The closures from the mass balance presented in Table 29 were in most cases within 10 %, The general applicability of observations from the present work depends on the degree to which it can be assumed that the performance of the GEK gasifier is representative of full scale. This is not clear. The design of the GEK is based on the Imbert concept which is common to most successful downdraft gasifier designs, and effort has been made in the design to limit thermal losses by using recuperative heat exchange. There is therefore no obvious reason to suppose that the temperature time history seen by a feedstock particle will change significantly on scale up, and the behaviour of a particle in response to a given temperature time history should also be unaffected (it is the same material). However, the important issue is whether the tendency of the softening plastics within the particle to cause agglomeration with neighbouring particles and form a blockage remains the same. This is more likely to be a function of porosity and packing geometry which might well change with scale, but the present work has not allowed this to be explored.

If despite the above concerns the performance of the GEK gasifier is taken as representative of full scale operation for any fixed bed downdraft gasifier design, then it can be concluded that the use of fixed bed downdraft gasification to convert paper industry wastes would be practical only for reject wastes produced from the SSK mill, and a small amount of wood would need to be co-fired. The levels of hard plastics present in AN mill and KC mill rejects prevent successful gasification above about 20 wt% blends with wood. From a paper mill's perspective it may not be economically attractive to buy large quantities of mixed wood chips even if such material is renewable and therefore eligible for renewable obligation certificates (ROC's), as price can be high and availability problematic. Reduction of the plastics content of AN mill and KC mill rejects by pre-treatment might be an option, but the plastics content is very high in these streams and there may not be enough residual fibrous material left to justify the gasification route. The SSK mill rejects on the other hand contain mainly soft plastics wrap and film, and their partial removal by pre-treatment might be attractive in that the need for co-firing with wood may be removed. Assuming that the results obtained from this work are scalable then rejects could be pre-treated on-site at a paper mill and used as a fuel in a 250 kg/hr fixed bed gasifier, which in principle could create enough product gas to power a 250 kWe gas engine. The exhaust gases from the engine could then be re-used for drying the feedstock. Multiple downdraft gasifier units in parallel could potentially be installed for higher tonnages of rejects. It may also be an economically attractive option to produce fibrous reject pellets for sale as a gasification fuel to existing wood gasifier plants. However, the success in this study does not guarantee the successful operation at industrial usage, (several thousand hours of continuous operation at 250 kW).

The inclusion of de-inking sludge to the mix of rejects and mixed wood chips was observed to have little or no effect on reducing agglomeration problems caused by plastics. The level of ash present within de-inking sludge restricted its use It is thought that higher blends of this feedstock maybe possible by using a fluidised bed gasifier. Alternatively a more attractive option for paper mills would be to process de-inking sludge by pyrolysis. This has been proven to yield high energy pyrolysis oils which can be used in combustors, gasifiers, boilers and engines for CHP generation [60]. One advantage of processing de-inking sludge by pyrolysis as opposed to gasification is that no co-firing or support fuel is required.

6.4 Intermediate pyrolysis results and discussion

In total two pyrolysis runs were carried out using the Pyroformer to investigate the intermediate pyrolysis of de-inking sludge. The first (Trial 1) used AN mill de-inking sludge, and the second (Trial 2) used KC mill de-inking sludge.

To begin each run the Pyroformer was first purged with N_2 and started by switching on the electric augur screws. The inner screw was set to move the feed in the forwards direction at 16 revolutions per minute and the outer screw was set to recycle char in the reverse direction at 8 revolutions per minute. The heating bands on the Pyroformer were then switched on and heated to a set point temperature of 450 ^{o}C. The candle filters were switched on and maintained at 450 ^{o}C. This was followed by opening the water supply to the condenser and also switching on the electrostatic precipitator.

After all equipment was operating at the set point conditions, feeding commenced by loading the hopper with approximately 15 kg of de-inking sludge pellets, switching on the screw feeder and switching on the actuating valves which regulated inert conditions within the unit. The de-inking sludge pellets were fed into the unit at a mass flow rate of 15 Kg/h. Steady state conditions were measured as the time taken for the pyrolysis outlet gas temperature to stabilise, and this took approximately 60 minutes. Each run lasted for a total duration of approximately 6 hours, composed of a 3 hour heat up period, 1 hour to reach steady state and 2 hours running at steady state. Each run consumed a total of 45 kg of feedstock (30 kg at steady state). Approximately 3 litres of steady state pyrolysis oil was collected from each run. The two trials well proceeded without technical error. Table 30 shows the run conditions and mass yields from each trial.

Table 30 Intermediate pyrolysis steady state run conditions and product yields

Steady State	**Intermediate Pyrolysis Trial 1 and Trial 2**
Feed Consumption	30 kg
Feed Rate	15 kg/h
Run Duration (steady state)	2 hours
Pyrolysis Temperature	450 °C
Heating Rate (Approximate)	90 °C/min
Solid Residence Time	Minutes
Gas Residence Time	Seconds

Obtained by difference *

A typical mass balance of this process at steady state produces 9.3 wt% condensable organic vapours (pyrolysis oil), 0.9 wt% aqueous phase, 10.8 wt% permanent gases with the remaining 79 wt% being the solid inert residues, mainly calcium based. This was consistent with both runs.

6.4.1 Pyrolysis oil analysis

In each case, after condensing of the organic vapours formed and removal of an easily separated aqueous phase (by gravity settling), the pyrolysis oil was found to be of a high quality with a very high calorific value (close to 40 MJ/kg, comparable to biodiesel). The pyrolysis oil was characterised for its energy content and ultimate analysis; these are presented in Table 31. “AN DSPO” denotes AN mill De-inking Sludge Pyrolysis Oil, “KC DSPO” KC mill De-inking Sludge Pyrolysis Oil. Results are also shown for fossil diesel FD and for biodiesel BD for comparison.

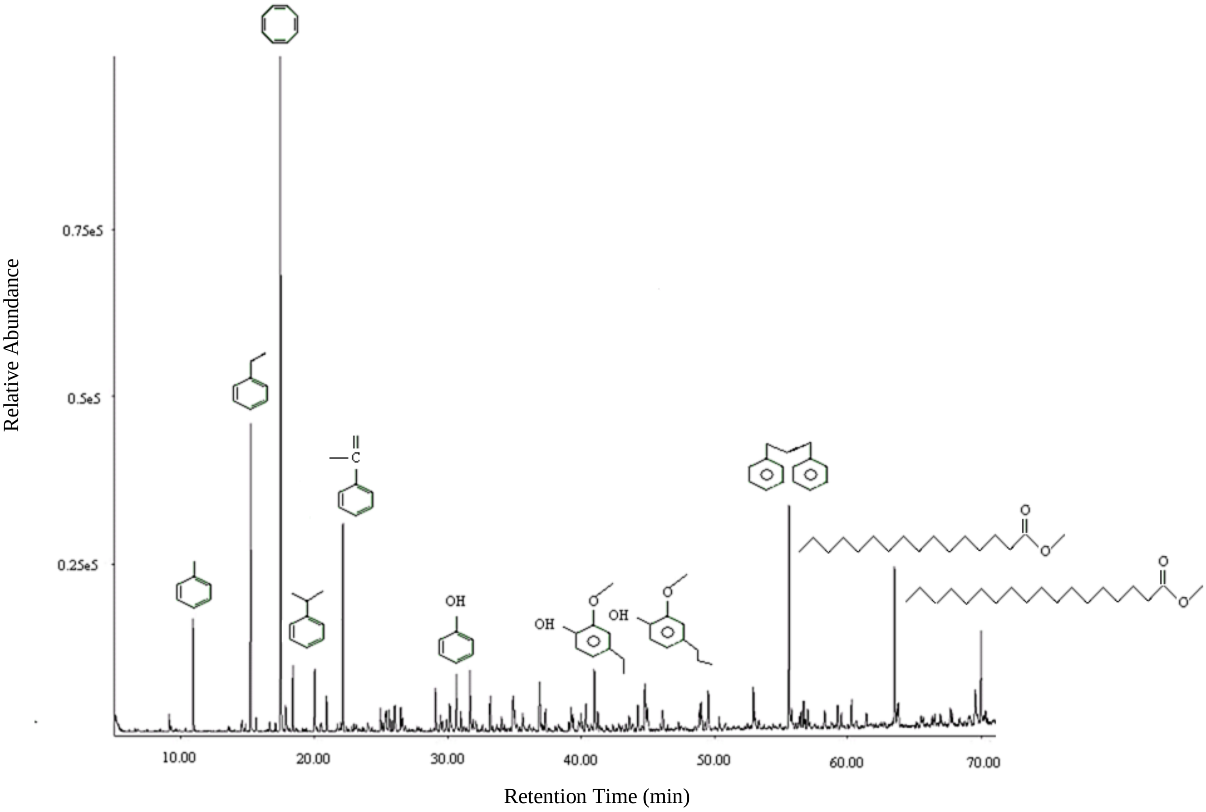
Table 31 Ultimate analysis and HHV of AN mill and KC mill de-inking sludge pyrolysis oil

	AN DSPO Result wt%	**KC DSPO Result wt%**	**FD Result wt%**	**BD Result wt%**
Carbon	78.71	76.58	85.60	78.86
Hydrogen	10.08	8.38	13.37	12.63
Nitrogen	1.02	1.86	<0.1	<0.1
Sulphur	0.55	0.58	<0.1	0.74
Oxygen	10.08	11.27	1.01	8.36
Ash	<0.02	<0.02	<0.01	<0.01
HHV (MJ/kg)	37	36	45	39

Table 31 shows an average carbon content of approximately 77 wt% for DSPO this is similar to 79 wt% for biodiesel (BD) and about 9 % lower than fossil diesel (FD); hydrogen contents of 8-10 wt% for DSPO are around 30 % lower than the 13 wt% value FD and 28 % lower than BD. Table 31 also shows that the intermediate pyrolysis of de-inking sludge produces significantly reduced oxygen content of the oils from that of the feedstocks, which is extremely favourable for fuel oils as it enhances the stability of the fuel [61]. Oxygen contents of 10-11 wt% for DSPO compare well with 8 wt% for BD but are approximately 7 times higher than FD. Research indicates that modest oxygen content in a fuel oil can have a beneficial effect on emissions from diesel engines as well as on combustion efficiency, since oxygenated fuels require less air for complete combustion [62]. Reduced CO emission from both DSPOs compared to FD may be expected, but increased NOx emission may result from the high nitrogen content of the oils [62]. The oxygen content of DSPO is much lower than fast pyrolysis oil obtained from traditional biomass feedstocks such as wood or straw (typically 30-50%) [63].

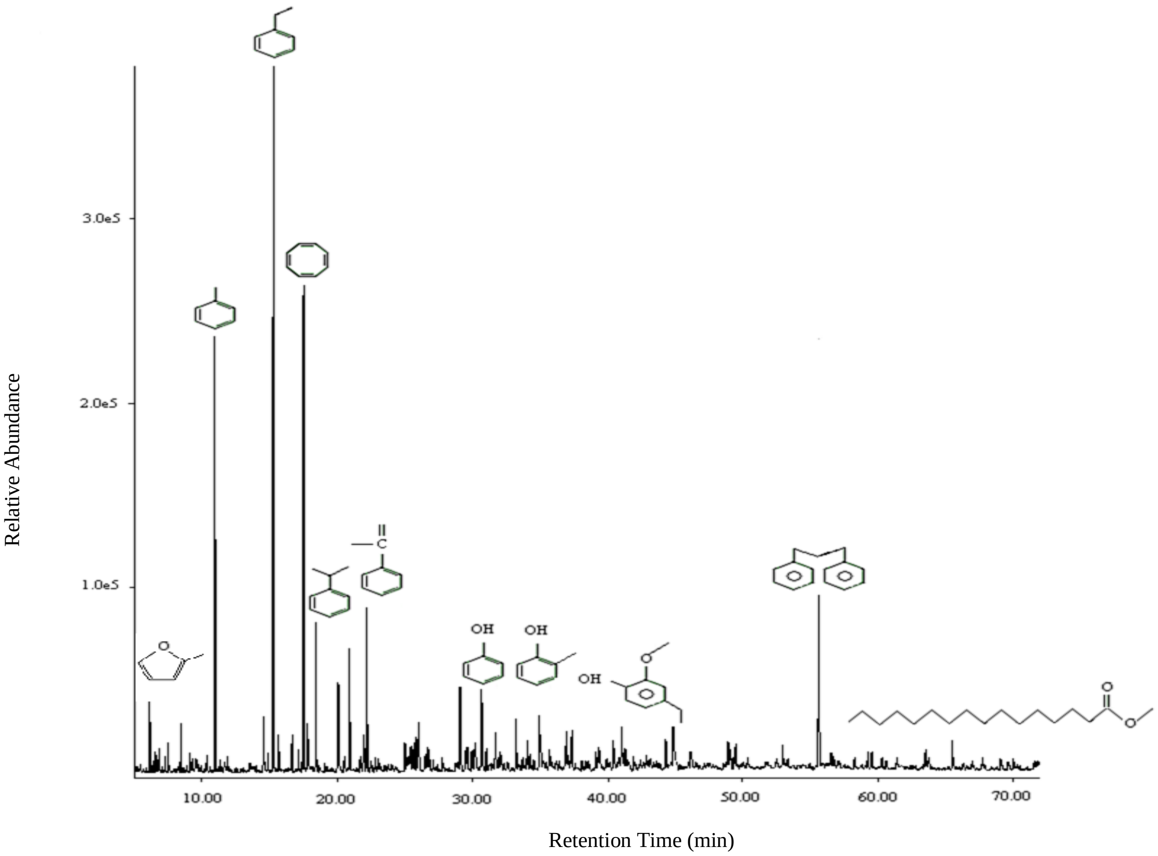
A comparison of the higher heating values of the oils is also shown in Table 31. DSPO has a satisfactory heating value for fuel oils, 37 MJ/kg for AN mill oil and 36 MJ/kg for KC mill oil compared with 45 MJ/kg for FD and 39 MJ/kg for BD. DSPO and biodiesel contain less combustion energy than diesel mainly because of the presence of oxygen. This derives from the presence of hydroxyl, aldehyde, carboxyl, ether and carbolic groups. Ash content of the oils was measured as <0.02 wt% for DSPO. This should be acceptable for furnace and engine applications.

A sample of the oils obtained from both the KC mill and AN mill runs was dissolved in an ethanol solvent and further analysed using liquid GC-MS. The chromatographs of each GC-MS run are presented in Figures 46 and 47, in which the top 11 most abundant chromatograph peaks are identified (more than 200 peaks corresponding to different organic compounds were detected in each mass spectrum).



Peak Number	Retention Time	Chemical Name	Chemical Formula	Area %
1	10.93	Toluene	C_7H_8	4.14
2	15.24	Ethylbenzene	C_8H_{10}	12.62
3	17.47	1,3,5,7 Cyclooctatetraene Styrene	C_8H_8	28.41
4	18.39	1-Methylethyl Benzene	C_9H_{12}	2.68
5	22.14	1-Methylethenyl Benzene	C_9H_{10}	9.10
6	30.63	Phenol	C_6H_6O	3.18
7	40.98	4-Ethyl-2-Methoxy Phenol	$C_9H_{12}O_2$	3.18
8	44.93	2-Methoxy-4-Propyl Phenol	$C_{10}H_{14}O_2$	0.93
9	55.60	1,3-Propanediyl-Bis-Benzene 1,1	$C_{15}H_{16}$	11.91
10	63.50	Hexadecanoic Acid, Methyl Ester	$C_{17}H_{34}O_2$	7.34
11	69.97	Octadecanoic Acid Methyl Ester	$C_{19}H_{38}O_2$	4.81

Figure 46 Liquid GC-MS chromatograph of AN DSPO



Peak Number	Retention Time	Chemical Name	Chemical Formula	Area %
1	6.16	2-Methyl-Furan	C_5H_6O	1.56
2	10.96	Toluene	C_7H_8	12.35
3	15.28	Ethylbenzene	C_8H_{10}	22.99
4	17.51	1,3,5,7 Cyclooctatetraene Styrene	C_8H_8	15.67
5	18.43	1-Methylethyl Benzene	C_9H_{12}	4.47
6	22.18	1-Methylethenyl Benzene	C_9H_{10}	5.50
7	30.63	Phenol	C_6H_6O	3.94
8	33.19	2-Methyl Phenol	C_7H_8O	1.83
9	41.00	4-Ethyl-2-Methoxy Phenol	$C_9H_{12}O_2$	2.00
10	55.60	1,3-Propanediyl-Bis-Benzene1,1	$C_{15}H_{16}$	7.56
11	63.50	Hexadecanoic Acid, Methyl Ester	$C_{17}H_{34}O_2$	0.73

Figure 47 Liquid GC-MS chromatograph of KC DSPO

The area % given in Figures 46 and 47 represents the peak areas of each identified component as a fraction of the whole integrated mass spectrum, thus giving an approximate mass fraction of that component. It is observed that the peaks obtained from KC DSPO and AN DSPO are quite similar. The highest abundance detected from AN DSPO was 1,3,5,7 Cyclooctatetraene Styrene representing an area% of 28.41 % and KC DSPO was Ethylbenzene representing an area % of 22.99 %, collectively these two compounds represented the largest fraction of both oils. The major peaks detected were also found to be consistent with those found from the work of Lou .R et al [41]. Both KC DSPO and AN DSPO also contain fatty acid methyl esters which are the major components of biodiesel. In general the oils are complex organic mixtures consisting of carbon chains ranging from C5-C15. Similar to fossil diesel fuel, they mainly contain paraffins, naphthenes and aromatics. Aromatic hydrocarbons are the most abundant component in the pyrolysis oils. Phenols present in the oils represent an area% of approximately 7 %. This is the cause of the high acidity of the oils. The high aromatics content also gives rise to the low cetane index of the oils, as aromatics have poorer combustibility compared with paraffins and naphthenes [64].

Analysis of the oils to determine their physical properties relevant to use as a fuel was also carried out, with the results shown in Table 32 together with results for diesel and biodiesel. Note that the oils were first separated from the aqueous phase.

Table 32 AN DSPO and KC DSPO fuel-related properties

Physical Properties	**Units**	**AN DSPO Result wt%**	**KC DSPO Result wt%**	**FD Result wt%**	**BD Result wt%**
Water Content	%	4	3	0.05	0.37
pH	-	4.8	4.7	7.01	7.75
Total Acid Number	mgKOH/g	26.0	32.8	0.023	0.489
Copper Corrosion 40 ^{o}C	24 hours	3a	2c	1a	1b
Flash Point	^{o}C	168	160	68	170
Density 22 ^{o}C	Kg/m^3	980	984	832	890
Cetane Index	-	19	23	>40	>47
Kinematic Viscosity 40^{o}C	cSt	12.3	9	3.01	8.20
Conradson Carbon Residue	wt %	3.9	4.6	0.06	<0.01
Lubricity (size of wear scar)	mm	219	215	276	202
Surface Tension 40 ^{o}C	mN/m	27.9	26.7	25.7	31.7

Table 32 shows that DSPO contains a lower water content than fast pyrolysis oil from traditional biomass feedstocks which usually give a water content of approximately 20% [63] this leads to improved combustion characteristics of DSPO. Water is always present in pyrolysis oil. There are two main sources for this water, feedstock water which is unaltered, and reaction water which is produced during the pyrolysis process. Due to the extensive drying of the de-inking sludge before pyrolysis, the oils only contained 3-4 wt% water. This amount does not greatly reduce the oil heating values. Furthermore, modest water content in the oils can reduce the combustion temperature in the cylinders and thereby reduce NOx emission [65]. However, the water content may result in metal electrochemical reaction in the long term, and could decrease the oil stability by allowing the growth of microorganisms if there is any further phase separation [66].

DSPO is acidic as indicated by a pH value of approximately 4.7, however this is less acidic than oils produced by the fast pyrolysis of biomass which usually have a pH value of around 2-3 [63]. However results from acid number titration indicate that DSPO has an average acid number of approximately 30 mgKOH/g; which is significantly higher than 0.49 for BD and 0.02 for FD, and may therefore cause corrosion and consequent damage to fuel system components. As previously described, the high acid number may be partly due to the presence of phenolics in the oils. Further treatment to reduce acid number is strongly recommended. Blending the oil with BD (an alkaline fuel) has been investigated and the results are presented later.

The corrosive effect is related to the acidity and reactivity of the oils. Pictures of copper strips immersed in DSPO at 40 °C in different time phases have been illustrated in Figure 48.

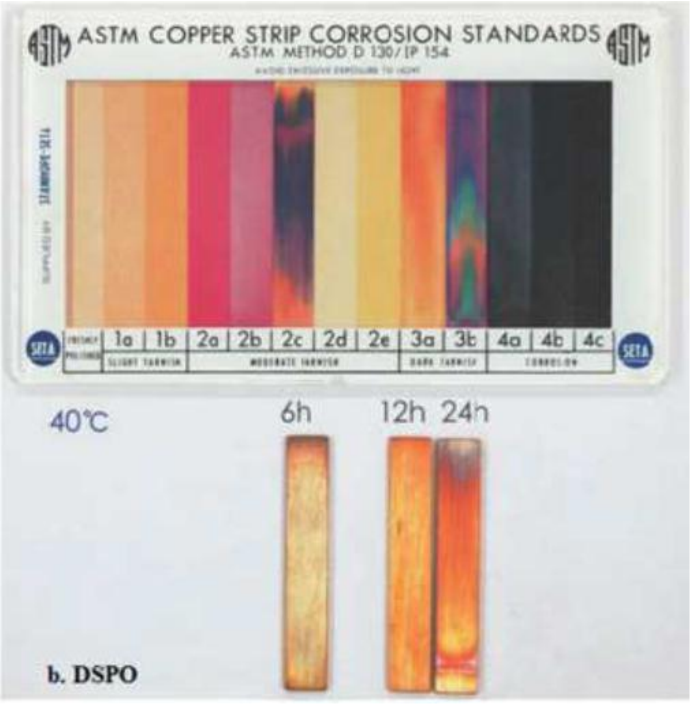


Figure 48 Copper Strips AN DSPO

It is observed that the class of corrosiveness of DSPO to copper corrosion strips are 3a for AN DSPO (24h/40 °C) and 2c for KC DSPO (24h/40 °C). This means DSPO has a relatively low ability to corrode copper.

DSPO has a high flash point, averaging 164 °C, which is comparable to BD at 170 °C and much higher than FD at 68 ^{o}C. DSPO is therefore considered to be sufficiently safe for transportation and storage.

It can be seen from Table 32 that DSPO has a density higher than conventional FD and BD, 980 kg/m^3 and 984 kg/m^3 for AN DSPO and KC DSPO respectively compared to 832 kg/m^3 and 890 kg/m^3 for FD and BD respectively. Fuel density is directly related to the engine performance since the injection system works principally on a volume basis. A greater mass of the higher density pyrolysis oils will be injected, and this compensates to some extent for the low heating value of the oils when compared to BD. However, an increase in fuel density leads to an advance in fuel injection timing and a greater spray cone angle as well as reduced fuel spray penetration in the cylinder [67]. Highly dense fuels may result in a poorer combustion and unstable engine operation.

The Calculated Cetane Index of the oils is presented in Table 32. DSPO shows a greatly reduced CCI compared with FD and BD. The main reason for this is that pyrolysis oils have a high density and significant aromatics content, and for the same carbon number aromatics have the lowest cetane number, followed by naphthenes and paraffins [64]. A low CCI indicates that pyrolysis oils may have difficulty in compression ignition in a diesel engine and may exhibit incomplete fuel combustion and consequent reduced engine thermal efficiency. It is therefore recommended that ignition improvers or blends of pyrolysis oil with biodiesel be used for diesel engine applications.

Kinematic viscosity of DSPO is shown in Table 32. DSPO was measured to be 12 and 9.60 cSt for AN DSPO and KC DSPO respectively. This is similar to 8.2 cSt for BD. High fuel oil viscosity is considered beneficial in lubricating the fuel supply system and thus decreasing mechanical wear, however on the other hand it worsens the flow characteristics of the oil and its atomisation quality which can cause incomplete combustion and engine power loss. The viscosity of the oil is significantly lower than that of biomass derived fast pyrolysis oils which results in improved flow characteristics as well as making the fuel less corrosive to equipment component parts. An ICP metal scan of the metal elements present in the oils was carried out and is presented in Table 33

Table 33 ICP metals present within the AN and KC DSPO.

Metal	AN DSPO (ppm)	KC DSPO (ppm)
P	91	5
Sn	<1	<1
Mo	1	1
B	6	1
Zn	53	16
Pb	2	1
Cd	<1	n/a
Ni	<1	5
Ba	21	1
Mn	1	<1
Fe	4	1
Cr	1	<1
Si	522	160
Al	3	<1
V	<1	<1
Ca	69	10
Cu	3	<1
Ti	<1	<1
Na	12	1
Mg	2	<1
Li	n/a	<1

From Table 33 it is observed that the metal levels present within the oils are low, with Si being the highest fraction measured at 160 and 522 ppm for KC DSPO and AN DSPO respectively.

This work has shown that oils produced from the intermediate pyrolysis of de-inking sludge in the Pyroformer process are appropriate for use as diesel engine fuels in terms of elemental composition and some physical and chemical properties. In some respects the oils are similar to commercial BD.

The oil characterisation proves that the pyrolysis process successfully decomposed a part of waste sludge feedstock into organic compounds in the carbon chain range C5 to C18 with high carbon and hydrogen contents and relatively low oxygen content, giving high energy content similar to BD. The relatively low water content in the oils also contributes to the high energy content. DSPO is found have similar lubricity to FD fuel and a high flash point for storage and transport.

However the nature of the oil composition gives rise to an unfavourably high acidity level, which can lead to metal corrosion. This can be militated against by blending the oil with alkaline fuels such as BD. Cetane index of the oils is calculated to be lower than the BD, which implies the possibility of poor compression ignition in a diesel engine; again blending with a higher cetane fuel such as BD can resolve this. The balance of density, viscosity and surface tension of the oils indicates that the quality of atomisation may be not as good as with FD and BD. This will cause deteriorated combustion and increased soot in engine emissions. The oil is also found to be high in carbon residue content and this may cause clogging in injector nozzles and coke formation in the combustion chamber.

In summary, the pyrolysis oils have some satisfactory characteristics for use as diesel engine fuels, but at the same time there are some characteristics that may cause deteriorated engine performance and reduced engine life. In their unblended form the oils would be most suitable as furnace fuels for boiler or similar applications. The negative effects of most of the characteristics for engines could however be alleviated if they are blended with other better quality fuel oils, since they are theoretically proportional to the fraction of pyrolysis oil in the blend. Preliminary investigations showed that both pyrolysis oils are highly miscible with BD, and therefore further work on confirming the long-term miscibility of the pyrolysis oils in BD and the characteristics of blends of different proportions was carried out. The results are presented in Section 6.5.

6.4.2 Permanent gas analysis

Approximately 11 wt% of the total products from the Pyroformer was permanent gases. Gas analysis using a GC-TCD analyser was performed on these gases, and results are presented in Table 34 (normalised to 100%) together with HHV.

Table 34 Normalised pyrolysis gas composition from the pyrolysis of de-inking sludge

Normalised Pyrolysis Gas Composition	**AN Mill De-inking Sludge vol%**	**KC Mill De-inking Sludge vol%**
H_2	-	1.9
CO	22.7	25.5
CO_2	71.2	66.3
CH_4	6.1	6.3
HHV (MJ/Nm^3)	5.5	6.2

The gas was composed of mainly CO_2 and CO with an HHV of between 5-6 MJ/Nm^3. Very little hydrogen was seen with the majority of the calorific value of the gas attributed to the CH_4 and CO fractions. This energy carrying gas would be suitable for use as a fuel co-fired with pyrolysis oil or char in a dual fuel combustor to meet the heat demands of the Pyroformer or a dryer; it may also be possible to co-fire the gas with a liquid fuel in a dual fuel engine.

6.4.3 Residual solids analysis

The majority of the product formed from the intermediate pyrolysis of de-inking sludge was solid residues which made up approximately 79 wt% of the mass balance. The solids were mostly inorganic with a majority being calcium carbonate and the remainder made up of other metal oxides such as Al_2O_3. The solids were further characterised for elemental composition and energy content. The results are presented in Table 35. They revealed that full organic decomposition did not occur during pyrolysis, some residual carbon and hydrogen remained in the solids and this subsequently resulted in a residual HHV of between 3-5 MJ/kg.

Table 35 Ultimate analysis and HHV of de-inking sludge pyrolysis solid residues

	AN Mill De-inking Sludge Solid Residue Result wt%	KC Mill De-inking Sludge Solid Residue Result wt%
Carbon	17.12	20.02
Hydrogen	1.13	2.19
Nitrogen	0.26	0.38
Sulphur	<0.1	<0.1
Oxygen*	21.49	23.91
Ash	60	53.5
HHV (MJ/Kg)	3.3	4.9

* Result obtained by difference

The results from Table 35 indicate that the solid residues cannot be used directly as a fuel in a solids combustor without blending with a higher calorific value support fuel, as the HHV is too low and the ash content too high. Co-firing with the pyrolysis oil and/or the permanent gases produced could however be successful. Combustion of the solid residues at a temperature > 700 ^{o}C would calcine the $CaCO_3$ fraction into CaO and thus would reduce the overall weight of the solid ash leaving the combustor, reducing ash disposal costs. Ashing of the pyrolysis solid residue was carried out in a furnace at a temperature of 800 ^{o}C, and a further reduction of 40 wt% by weight was achieved due to the release of residual volatiles and the calcination of $CaCO_3$, thus leaving a total remaining inorganic ash content of 60 wt%.

In addition to the ultimate analysis, TGA (combustion) was carried out on the residual solids. The TGA derivative weight loss profiles are presented in Figures 49 and 50 and show the weight loss profiles of AN mill and KC mill de-inking sludge residual solids respectively.

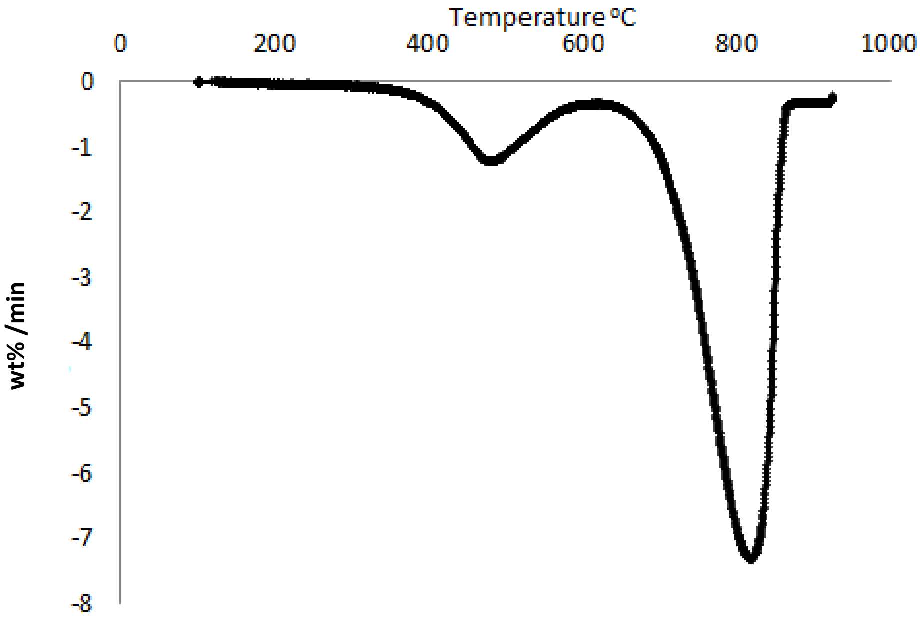


Figure 49 AN mill de-inking sludge pyrolysis solid residue DTG curve

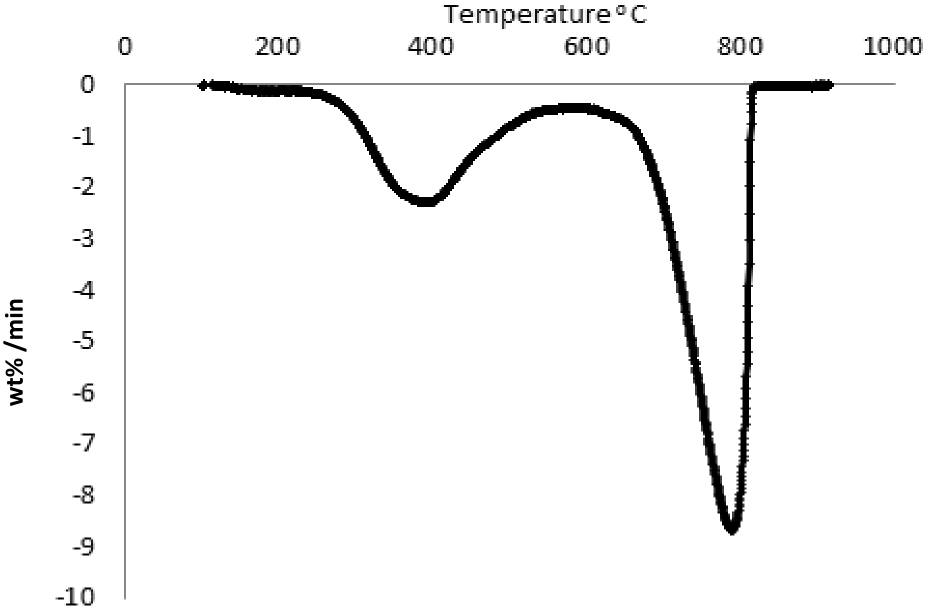


Figure 50 KC mill de-inking sludge pyrolysis solid residue DTG curve

In general the two solid residues show similar behaviour when combustion was carried out in an atmosphere of air. Weight loss is seen between 200 °C and 600 °C, and then again between 600 °C and 800 °C. The rate of weight loss between 200 °C and 600 °C peaks at approximately 400-450 °C, and this is the direct result of the decomposition of the residual cellulosic fibres and other organic components remaining in the solids. The rate of weight loss between 600 °C and 800 °C peaks at approximately 790 °C and this is largely the result of the thermal decomposition of $CaCO_3$ present within the ash. Similar peak rates are observed for the two residues. Above 900 °C, approximately 60 wt% of the original pyrolysis solids sample is present as inorganic metal oxide compounds, so no further weight loss is experienced.

6.4.4 Summary of de-inking sludge pyrolysis trials

This work has developed a method to utilise de-inking sludge waste fractions for the on-site production of combined heat and power (CHP) using the advanced thermal conversion method known as intermediate pyrolysis. The pyrolysis oil produced had an HHV of 36-37 MJ/kg and was composed mainly of aromatics and fatty acid methyl esters. The oil was found to have improved fuel physical properties when compared to pyrolysis oil produced from the fast pyrolysis of biomass, and was similar in some respects to BD. A permanent gas fraction was also produced which contained a HHV of 5-6 mJ/Nm^3 and was mainly composed of CO, CO_2 and a small amount of CH_4. The rest of the processed feedstock remained as solids which contained a small calorific value ranging from 3-5 MJ/Nm^3 and was mainly composed of $CaCO_3$.

6.5 BD-DSPO blending and engine test results and discussion

The following work presents the results of BD-DSPO blending and engine testing and was carried out jointly in a collaborative effort with Dr Hossain. A and Siddiqui .S et al, at Aston University [65]

6.5.1 BD-DSPO blend characterisation

A batch of AN DSPO was directly blended with BD by mixing in an agitated tank and allowing to settle for a period of 24 hours. After blending the mixture was characterised for its fuel chemical and physical properties with the intention of using the blend as a fuel in an indirect injection Lister-type diesel engine. Two blends were prepared in total the first was composed of 20 vol% DSPO with 80 vol% BD and the second was 30 vol% DSPO with 70 vol% BD. Table 36 shows physical and chemical properties of DSPO, BD, FD and DSPO-BD blends

Table 36 Measured properties of DSPO, BD, DSPO − BD blends and FD

Physical and chemical properties	**DSPO**	**BD (B100%)**	**20% DSPO 80% BD**	**30% DSPO 70% BD**	**FD**
Kinematic viscosity (cSt) at 40°C	12.3	8.2	8.91	9.35	3.01
Flash point temperature (°C)	168	170	105	118	68
pH value @ 22°C	4.8	7.75	5.91	5.73	7.01
Acid number (mg KOH/g)	26.0	0.489	6.74	7.66	0.023
Density (kg/m^3) @ 22°C	980	890	906	920	832
Higher heating value (MJ/kg)	37.04	39.29	38.79	38.58	44.67
Lower heating value (MJ/kg)	34.91	36.49	36.22	36.08	41.87
Water content (% wt.)	4.00	0.37	0.94	1.70	0.06
Carbon residue (% wt.)	3.89	<0.01	0.316	0.518	0.059
Ash content (% wt.)	<0.02	<0.01	<0.01	<0.01	<0.01
Carbon (% wt.)	78.71	77.20	77.15	77.34	84.73
Hydrogen (% wt.)	10.08	13.21	12.11	11.80	13.20
Nitrogen (% wt.)	1.02	0.10	<0.10	<0.10	<0.10
Oxygen (% wt.)	10.08	9.39	10.54	10.66	1.40
Sulphur (% wt.)	0.55	<0.10	<0.10	<0.10	<0.10

Table 36 shows that the viscosity and flash point temperature of the DSPO were approximately 4 times and 2.5 times higher than that of FD respectively. On the other hand, the viscosity of BD was about 2.7 times higher than that of FD. Flash point temperatures of the BD and DSPO were almost the same. With regard to LHV, Table 36 shows only small differences between DSPO and BD. In contrast, LHV of DSPO was lower by about 17 % than for FD. Density, acid number and carbon residue values of DSPO were considerably higher than those of FD and BD; in the case DSPO−BD blends these values were decreased significantly. The carbon content in FD was 7–9 % higher than in DSPO and in BD. Nitrogen and sulphur content in the DSPO was much higher than in the BD and in FD. On the other hand, sulphur content was at trace levels in the DSPO-BD blends. In the DSPO, hydrogen content was lower by approximately 31% but oxygen content was higher by about 7 times than that of FD. Ash content was at trace levels in all fuels. Overall by blending the DSPO with BD significantly improved physical fuel properties were achieved in particular reduced acid number, water content, and carbon residue which makes the blend a better fuel for engines reducing its corrosiveness and potential to block filters and clog pipes.

6.5.2 Engine performance and exhaust emissions

Engine performance and emission parameters when operated on DSPO-BD blends were assessed against those of FD and BD (B100) operation. Brake specific fuel consumption (BSFC) and thermal efficiencies results were better for 20 % blends than for 30 % blends (Figure 51, 52 and 53).

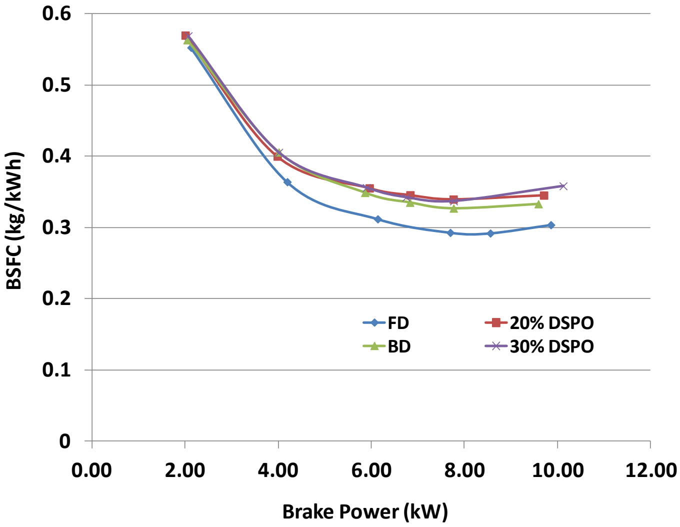


Figure 51 BSFC (wt.) vs. Brake power

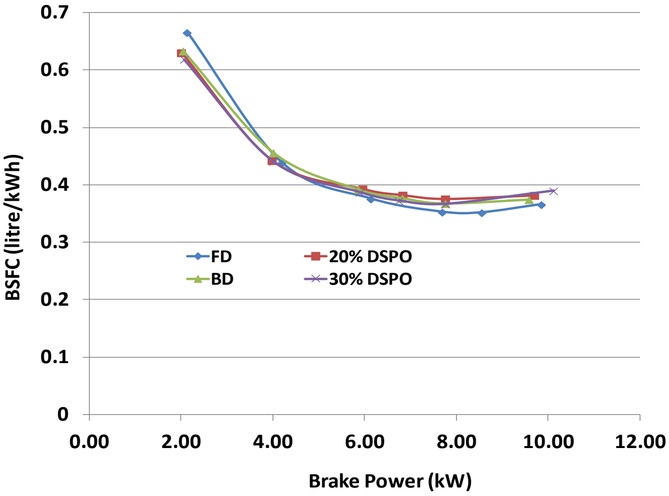


Figure 52 BSFC (vol.) vs. Brake power

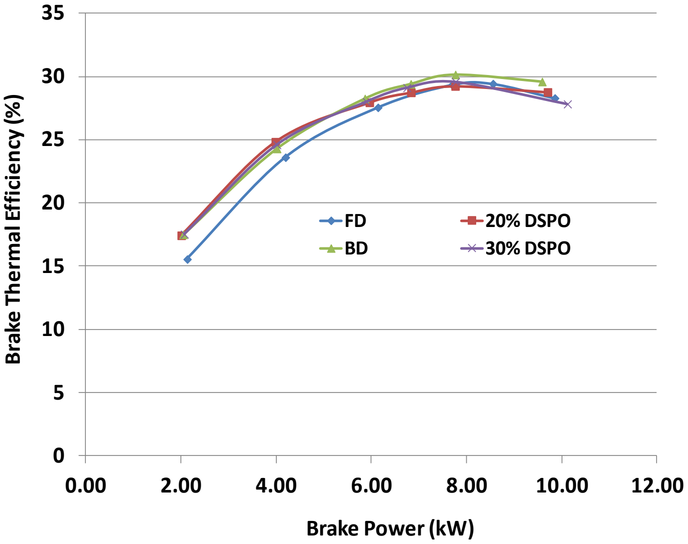


Figure 53 Thermal efficiency vs. Brake power

At full load, BSFC was 14−18 % and 4−8 % higher than FD and BD respectively, when operated on both 20 % and 30 % DSPO-BD blends (Figure 51). As the LHVs of the DSPO blends are less than that of FD or BD, more fuel is needed for the same engine output. Similarly as the viscosities of the DSPO-BD blends are higher than that of FD this leads to less efficient mixing of fuel and air. In contrast, the fuel consumptions were comparable when expressed in volume rather weight basis due to the higher densities the DSPO-BD blends; and this was almost similar when compared to BD in all load conditions, but higher by approximately 6 % than FD at full load (Figure 52). At low loads, the brake thermal efficiency from both blends were close to those from BD and 4 % higher than FD, but at full load efficiencies were about 3−6 % lower than BD and were closer to FD (Figure 53).

Emissions of CO_2 were almost the same for BD and DSPO-BD blends at all loads, but higher by about 4 % than FD at higher load conditions (Figure 54). Although CO_2 emissions appear higher when running on the blends a proportion of the CO_2 can be considered as carbon neutral, whereas it cannot with FD operation.

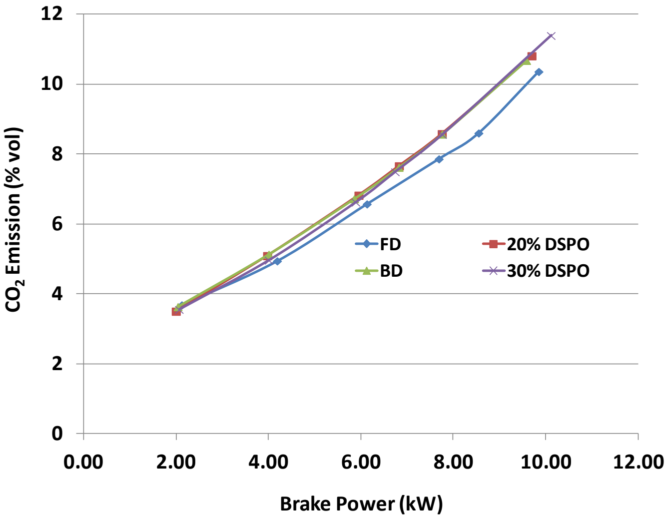


Figure 54 CO_2 emissions vs. brake power

The amount of air intake (i.e. oxygen content) was constant throughout the engine test so, at higher loads the fuel to air ratio decreased, and hence in the case of FD operation, CO emission increased sharply at full load (Figure 55) due to the lack of oxygen content in the mixture. In contrast to this, sharp increases in the CO emissions were not noticed in the case of DSPO-BD blends due to having lower carbon-to-oxygen ratio as compared to FD (Figure 55).

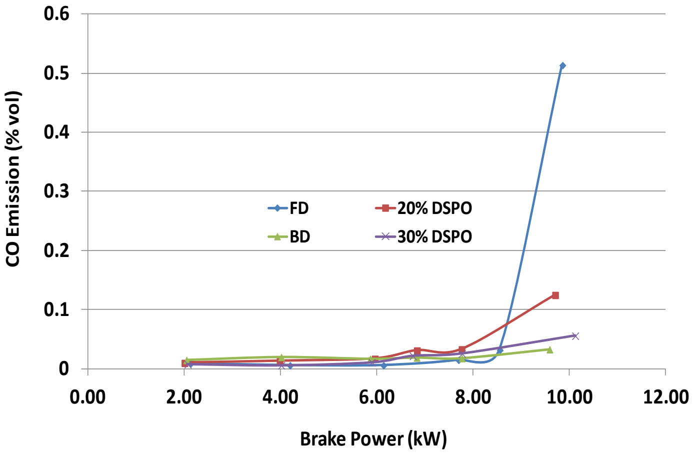


Figure 55 CO emissions vs. brake power

In general, at higher loads CO and CO_2 emissions were higher with DSPO blends when compared with FD, as higher amounts of the DSPO blends were burned for the same engine output.

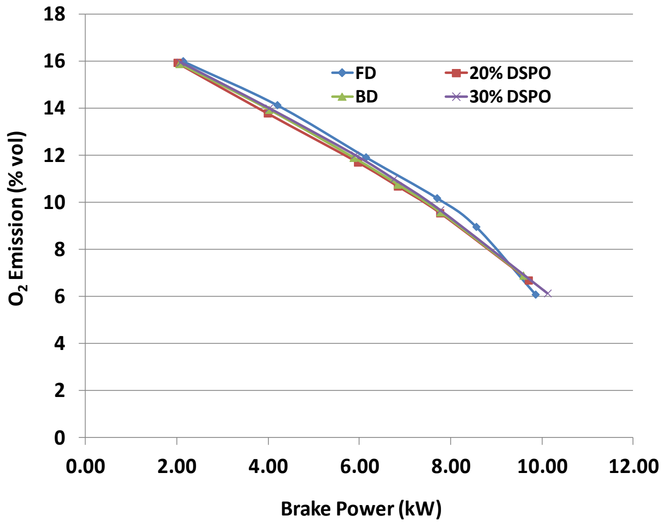


Figure 56 O_2 emissions vs. brake power

Oxygen emissions were almost the same between DSPO blends and BD; but slightly lower for all three fuels when compared with FD (Figure 56).

Higher combustion temperatures in CI engines generally give higher NOx emissions. The presence of higher water content in the DSPO-BD blends lowered the combustion temperature which caused lower NOx emissions (Figure 57). For 20 % and 30 % blends and at full load, NOx emissions were decreased by about 12 % and 6 % respectively compared to FD. In addition, these observations were consistent with the higher density of 30 % DSPO-BD blend the higher the density the more NOx emitted.

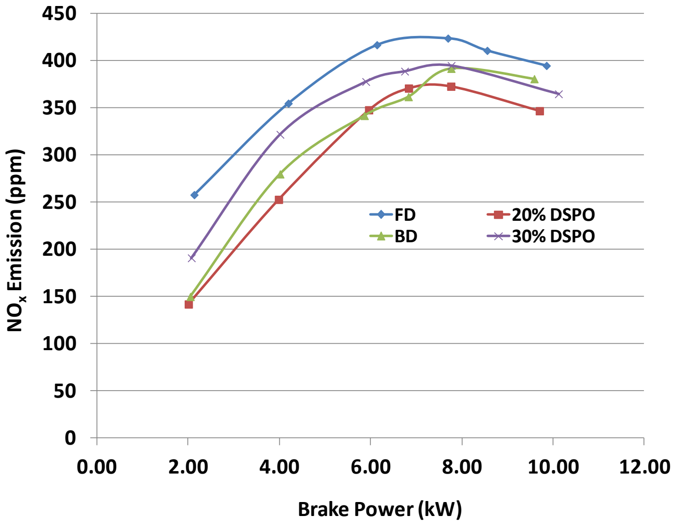


Figure 57 NOx emissions vs. brake power

Exhaust temperature is important for poly-generation applications (e.g. CHP, tri-generation). Little difference was observed in exhaust gas temperatures among the four fuels tested (Figure 58). Smoke levels were similar at low load conditions for all four fuels; but at higher loads, the smoke opacity values of DSPO-BD blends were slightly lower than for FD and BD.

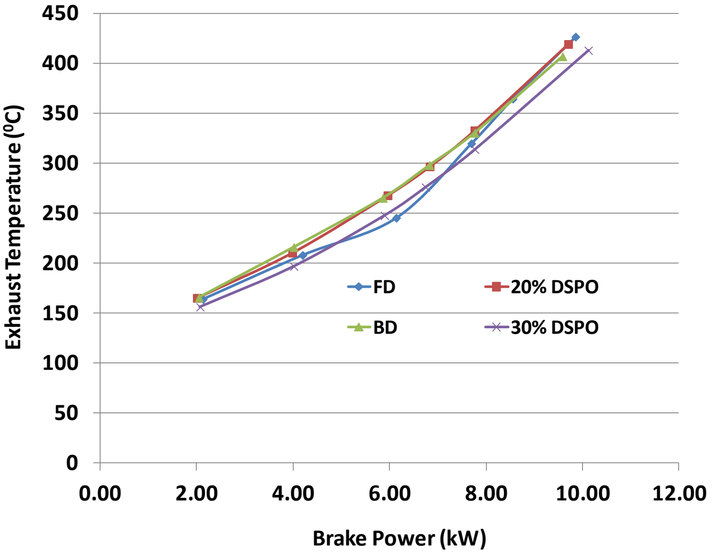


Figure 58 Exhaust gas temperatures vs. brake power

6.5.3 Combustion characteristics

Smooth engine operation was observed with the 20 % DSPO blend; however, the engine experienced minor knocking when operated on the 30 % DSPO blend, as evident from the pressure crank angle diagram (Figure 59 and 60).

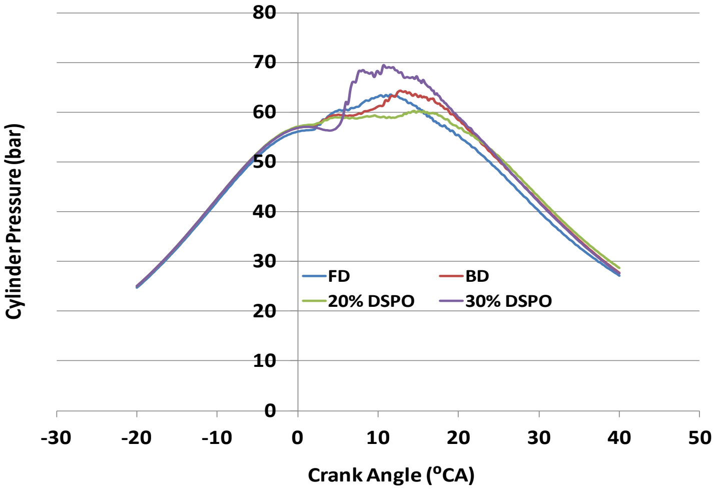


Figure 59 Cylinder pressure vs crank angle at full (100%) load

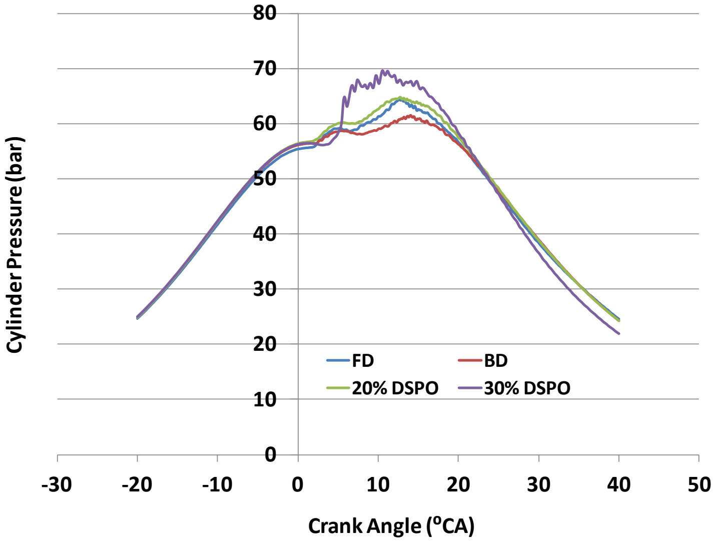


Figure 60 Cylinder pressure vs crank angle at full (70%) load

The low cetane number of the 30 % DSPO blend is believed to have caused this behaviour. Typical cetane number of DSPO was about 20; whereas cetane numbers of FD and typical BD are> 40. Peak cylinder pressures of 30 % DSPO blend were about 6−13 % and 5−6 % higher respectively, when compared to BD and FD operation. The cylinder pressure profiles for 20 % DSPO blend, BD and FD were almost similar, and only minor peak pressure variations were observed.

At low loads, integral heat release from combustion was almost the same for all fuels; and at higher load conditions, the integral heat released by DSPO-BD blends were decreased but the peak burn rates were higher (Figure 61 and 62).

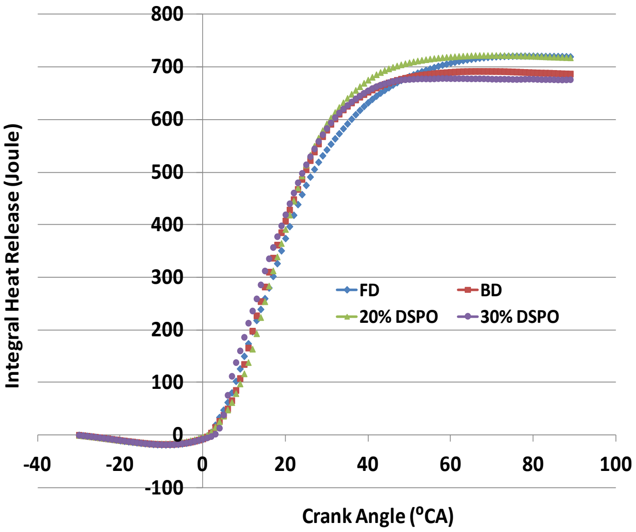


Figure 61 integral heat releases at full (100%) load

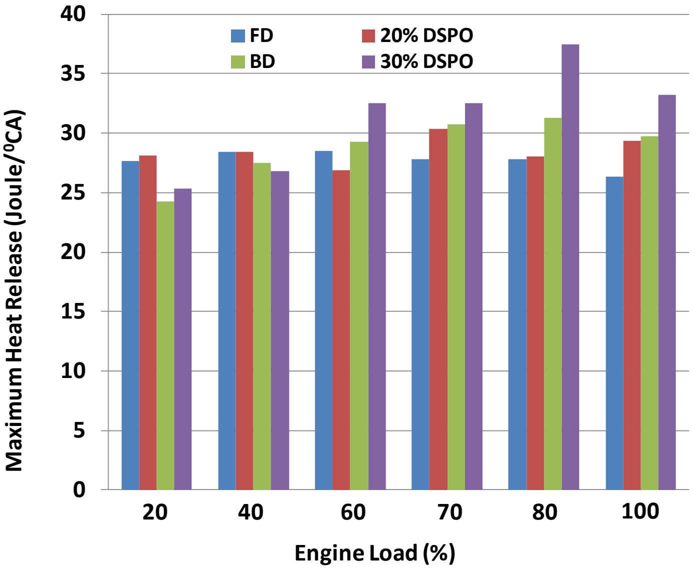


Figure 62 Maximum heat release rate at different loads

At full load, the peak burn rate of the 30 % DSPO blend was about 26 % and 12 % higher than with FD and BD respectively (Figure 62). Higher peak burn rates in the case of DSPO blends may have been caused by long ignition delay and short combustion periods. Total combustion duration is defined as the duration of the crank angle between 5 % and 90 % combustion. Ignition delay is related to the ignition quality (i.e. cetane number) of the fuel. Compression ratio, engine speed, cylinder gas pressure, temperature of the air intake, and quality of fuel spray affects the ignition delay period.

Figure 63 shows that, in the case of DSPO-BD blends the start of combustion was delayed compared to FD at most load conditions.

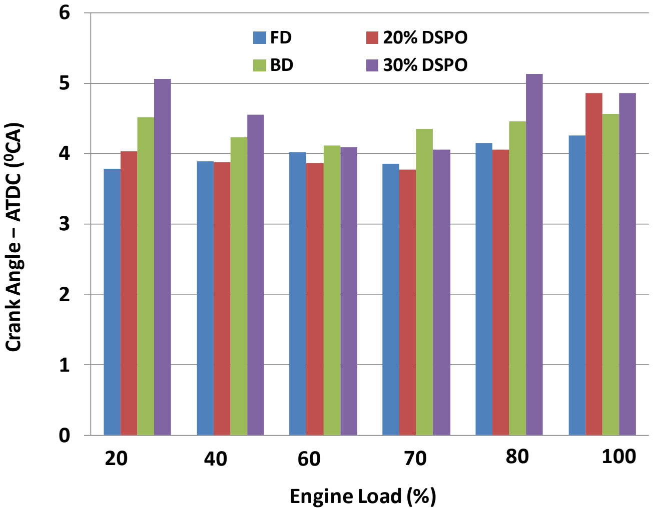


Figure 63 Crank angle positions at 5 % combustion

Total combustion duration increased with engine load for all fuels. In all load conditions, the combustion duration was shorter for the 30 % DSPO blend than for FD and BD. For the 20 % DSPO blend, the duration was shorter than for FD operation only at higher load conditions; no significant trends were observed between 20 % DSPO blend and BD in this respect (Figure 64).

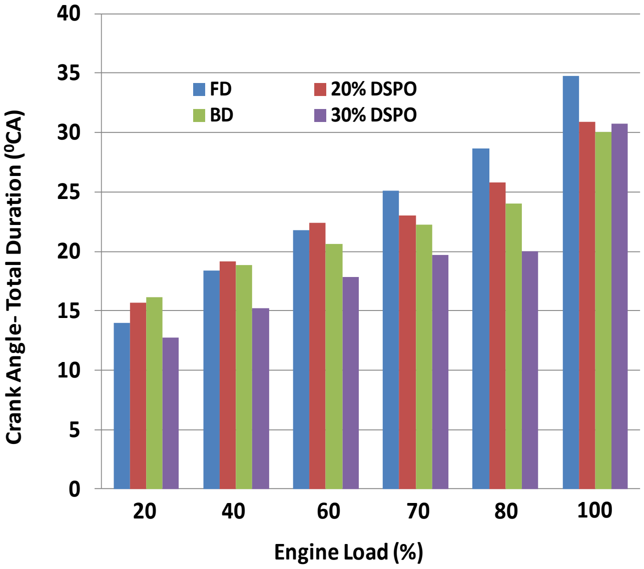


Figure 64 Total combustion duration

The short duration in the case of DSPO operation may be explained by less efficient mixing with inlet air as compared to the FD and BD, leading to a smaller amount of air-fuel mixture effectively available for combustion. At full load, combustion duration of the 30 % DSPO blend operation was almost 12 % lower than for FD (Figure 64). In the case of engine testing with other types of pyrolysis oils [68, 69], long ignition delays and short combustion periods were also reported. Higher cylinder pressure and high heat release rates of pyrolysis oil combustion are also reported in the literature [70, 71, 72]. The fuel injection pressure was higher in the case of DSPO blends; at full load, it was higher by approximately 17 % than FD.

6.5.4 Summary of BD-DSPO blending and engine tests

A three-cylinder indirect-injection CI engine, with nominal output 9.9 kW, has been tested with 20 % and 30 % de-inking sludge pyrolysis oil blended with biodiesel. Performance, emissions and combustion characteristics were compared against FD and BD (B100) operation. The physical and chemical properties of all four fuels were measured. With DSPO blends, full engine power was achieved. Between the 20 % and 30 % blends, there were few differences in the results from the engine tests. However when compared to FD and BD, there were a number of small but significant differences when using DSPO-BD blends.

At full load, the BSFC was increased by 6 % on a volume basis and 14−18 % on a weight basis when compared with FD; whereas, BSFC was only 4−8 % higher on a weight basis when compared to BD operation. At full load, brake thermal efficiencies were about 3−6 % lower than BD but were almost similar to FD. Compared to FD, CO_2 and NOx emissions were increased by 4 % and decreased by 6−12 % respectively. At full load, CO emission of 30 % DSPO blends was almost 10 times lower than FD operation. Compared to FD, peak cylinder pressures were about 5-6 % higher for 30% blend and were almost the same for the 20 % DSPO blend. In the case of DSPO-BD blends, the start of combustion was delayed but the burn rate was high as compared to FD. At full load, the peak burn rates of 30 % DSPO blend were 26 % and 12 % higher than the FD and BD operation respectively. Total combustion duration was decreased for both blends; for 30% blend, at full load the duration was decreased by 12% when compared to FD. The cylinder gas pressure diagrams indicated stable engine operation with 20 % DSPO blend, but the engine experienced some minor knocking in the case of 30 % DSPO blend.

This study concludes that up to 20% DSPO blended with biodiesel can be used successfully without adding any ignition additives or surfactants. After three hours of operation no deterioration in engine condition was observed.

6.6 The optimal process routes

This chapter has investigated in detail the downdraft gasification and intermediate pyrolysis of paper industry wastes produced from AN, SSK and KC mills with the intention of possible scale up and implementation at each site. There currently exists no available literature on these particular processes to thermally convert these particular wastes, which is where the novelty of this work rests.

The optimal gasification route was found to lie with the co-fired blend of SSK rejects with wood chips. This test produced the most promising results in terms of potential scale up at SSK. A maximum blend of 80 wt% SSK rejects with 20 wt% wood chips was discovered to gasify successfully for several hours with no performance issues. A product gas with a calorific value of 7.3 MJ/Nm^3 was formed. The product gas has a promising potential to be used as a fuel in a CHP scheme integrated within the mill. A full industrial implementation study on how this system could potentially be integrated at the SSK mill is presented in Chapter 7.

Other gasified blends using AN and KC rejects with wood chips were not as successful. The agglomeration of hard plastics within the rejects was found to be the principal cause of failure. A blend of 20 wt% rejects with 80 wt% wood chips was discovered to be the maximum gasifiable. At each site this would equate to the mill importing approximately 20,000 dry tonnes per year of dried wood chips to process their annual tonnages of rejects. From a practical and economical point of view it would not be feasible for these mills to import such a large quantity of wood chips. On the basis of this the processing of AN and KC rejects were therefore not explored further.

Introducing de-inking sludge to the gasified blend of AN and KC rejects with wood chips was also problematic at levels above 40 wt% de-inking sludge. Ash levels within the gasifier rose to levels which were unacceptable for gasification. Although a continuous ash removal system may alleviate this, it was beyond the scope of this study to explore the effect further. A more suitable option for the processing of de-inking sludge was discovered to be by intermediate pyrolysis.

The intermediate pyrolysis of de-inking sludge produced from the AN and KC mills was carried out using a Pyroformer reactor. Among the products formed was a high calorfic value pyrolysis oil (36/37 MJ/kg) as well as a permanent gas fraction which also contained a calorific value. The oil produced was found to work well as a 20 vol% blend with biodiesel in a Lister-type diesel engine. the combustion of the fuel in the engine yielded lower NOx and co emissions when compared to fossil diesel, making it a more sustainable engine fuel for potential CHP applications at the AN and KC mills.

Another main advantage of this process is that it potentially removes the need for paper mills to import large quantities of natural gas to co-fire sludge in a combustion unit which is the current practice for sludge reduction at the AN mill. At full scale it is envisaged that the inert solids formed from the pyrolysis of de-inking sludge would be co-fired with the hot pyrolysis vapours which are formed from the process in a combustion unit downstream. The ash product formed could then be sold to cement industries as cement or concrete admixture. Some of the hot exhaust gases from the combustion unit would be required to supply the heat for the drying and pyrolysis processes, but with the majority of exhaust gases being routed to a boiler for electricity or heat generation. A full industrial implementation study on how this system could potentially be integrated at AN and KC mills is presented in Chapter 7.

7.0 INDUSTRIAL IMPLIMENTATION STUDY

7.1 Chapter summary

This chapter investigates how the proposed gasification and pyrolysis processes explored in Chapter 6 would be implemented at each of the case paper mills. Information and calculations are based on detailed discussions with the mills.

The AN mill would integrate a de-inking sludge pyrolysis process into its existing sludge combustor CHP system. This would convert a fraction of its de-inking sludge into pyrolysis vapours which would heat the combustor replacing a substantial fraction of natural gas which is currently used as the heating fuel.

The SSK mill would utilise its entire reject stream as a co-fired blend with wood chips in a downdraft gasification system. The system proposed includes a pre-treatment plant for sorting, sizing and drying of the rejects, and a gasification plant including a product gas cleaning system. The proposed system would generate enough heat energy to power an existing boiler at the mill which currently powered using biogas.

The KC mill would integrate a de-inking sludge pyrolysis process which would utilise all its de-inking sludge to generate pyrolysis vapours which would be combusted in an existing boiler which is currently heated using natural gas. The method of integration, is presented in this chapter, energy flows and detailed calculations from each of the proposed processes are presented in Appendix A, B and C.

7.2 AN mill industrial implementation study

The AN mill currently requires 44 MWe of energy to run the entire paper mill. The majority of this energy is supplied by a gas turbine which burns natural gas. A proportion of this energy, approximately 2.3 MWe, is supplied by a steam turbine which is driven by steam raised using the exhaust gases from a sludge combustor. This system is illustrated in Figure 65. In this study a proposed pyrolysis system to extract de-inking sludge volatiles for CHP generation will be integrated into the existing sludge combustor system. 8000 annual operating hours are assumed.

7.2.1 AN mill current process (sludge combustor) description

The sludge combustor at the AN mill is a bubbling fluidised bed boiler which is maintained at a temperature range of 800-1000 °C. The fluidising medium is the combustion products of natural gas and air, and the inert bed material is sand. There are two locations where air is injected into the unit. The first (primary) is at the base of the unit which fluidises the bed, and the second (secondary) is located in the middle of the unit. This arrangement promotes heat transfer, complete combustion, and reduced emissions through staged combustion.

Approximately 350 tonnes/day of wet de-inking sludge at approximately 33 wt% moisture is conveyed into the bed via screw feeders. The physical limit on the amount of sludge which can be combusted is dictated by the total air available and by the proportion of natural gas needed to sustain combustion. The forced draft fans supplying air to the combustor are currently running at maximum capacity and thus the physical limit on sludge input has been reached, unless larger fans are installed.

Once inside the bed the sludge is combusted at the set point temperature. The exhaust gases pass up through the unit and then back down through a convective pass containing superheater, evaporator and economiser tubes in which superheated steam is generated. The movement of exhaust gases through the unit is encouraged by induced draft fans located downstream of the combustor. Fly ash entrained by the exhaust gases is collected and removed in bag house filters also located downstream of the combustor. Ash retained within the bed is also extracted intermittently and conveyed by screw conveyors into ash skips. A total of approximately 130 tonnes/day of ash is produced by the sludge combustor. The exhaust gases proceed to a stack where they are released to atmosphere at a temperature of approximately 130 °C, and 22 % of the total heat energy of the combined fuels is lost to the atmosphere at this point. This minimum temperature must be maintained to avoid the risk of condensation of water vapour and consequent corrosion of the stack.

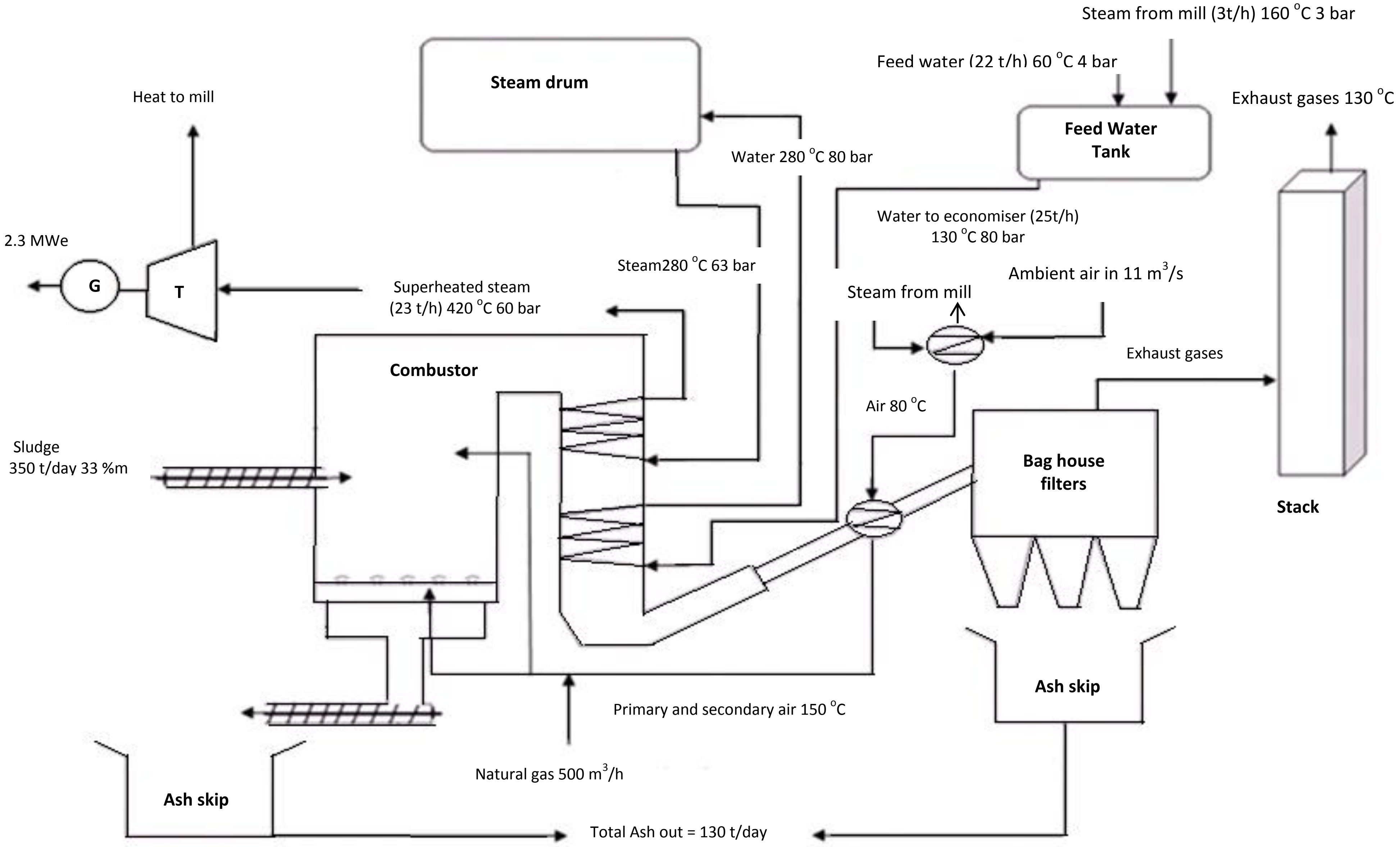


Figure 65 Current CHP system at the AN mill

A total of approximately 11 m^3/s of pre-heated air is supplied to the combustor and the pre-heat is supplied by two heat exchangers. The first is a steam heat exchanger which heats air from ambient to 80 ^oC and the steam for this heat exchanger is supplied from the mill. The air then passes into a second heat exchanger which is heated by exhaust gases from the combustor; here the air temperature rises from 80 ^oC to 150 ^oC. The hot air is then pre-mixed with natural gas before being fired into the combustor.

The water supplied to the to the heat exchangers is initially derived from a hot water well on-site and approximately 22 tonnes/h. of water at 60 ^oC and 4 bar(g) is pre-mixed with 3 tonnes/hr. of steam supplied from the mill at 160 ^oC and 3 bar(g). These streams are combined and pumped into a feed water tank at 80 bar. Oxygen scavengers are also added at this point to reduce corrosion. A total of 25 tonnes/hr of water at 130 ^oC and 80 bar(g) leaves the feed water tank and is pumped into the economiser. Water leaves the economiser at approximately 280 ^oC and 80 bar(g) and is routed into a steam drum located on the roof of the combustor. The water is then circulated through the evaporator tubes where steam is created. The steam leaves the drum and passes to the superheater which is located higher up in the convective pass where the exhaust gases are much hotter. Here the steam is superheated to a temperature of 420 ^oC and 60 bar(g) pressure. Approximately 23 tonnes/hr of this steam is routed to a steam turbine where 2.3 MWe is generated for the paper mill. Of the low pressure steam leaving the steam turbine, around 11 MWth is routed back into the mill for drying of paper sheets, the remainder is used elsewhere in the mill. The maximum steam production capacity of the boiler is approximately 30 tonnes/hr.

The parasitic load of the CHP system is 0.5 MWe (used to drive pumps, compressors feeders etc); therefore net electricity production is approximately 1.8 MWe. A maximum of 5 MWe could potentially be produced by the steam turbine if it operated in fully condensing mode. Figure 66 shows all calculated energy flows of the sludge combustor and the detailed energy balance calculations can be found in Appendix A1.

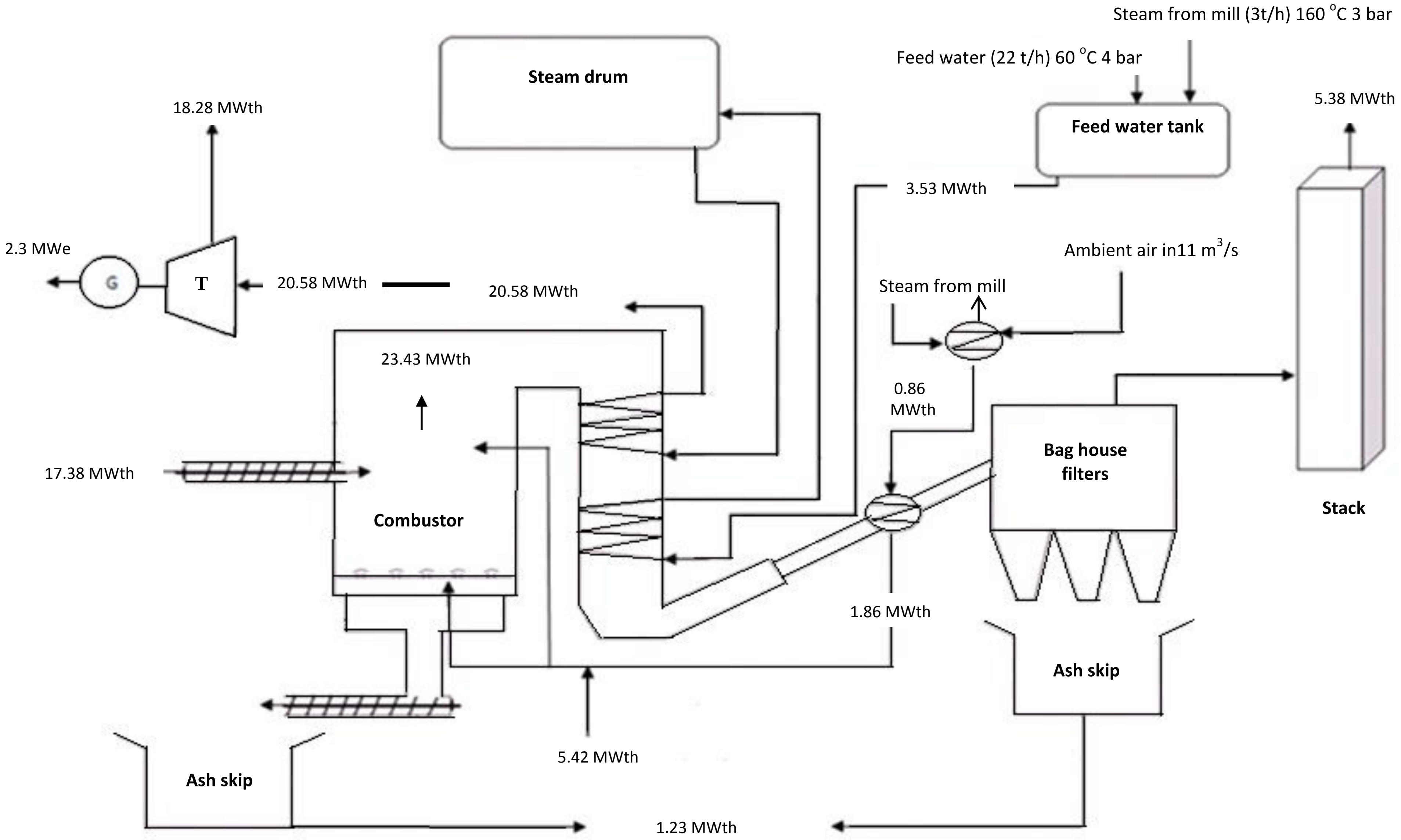


Figure 66 Current AN mill process energy flows

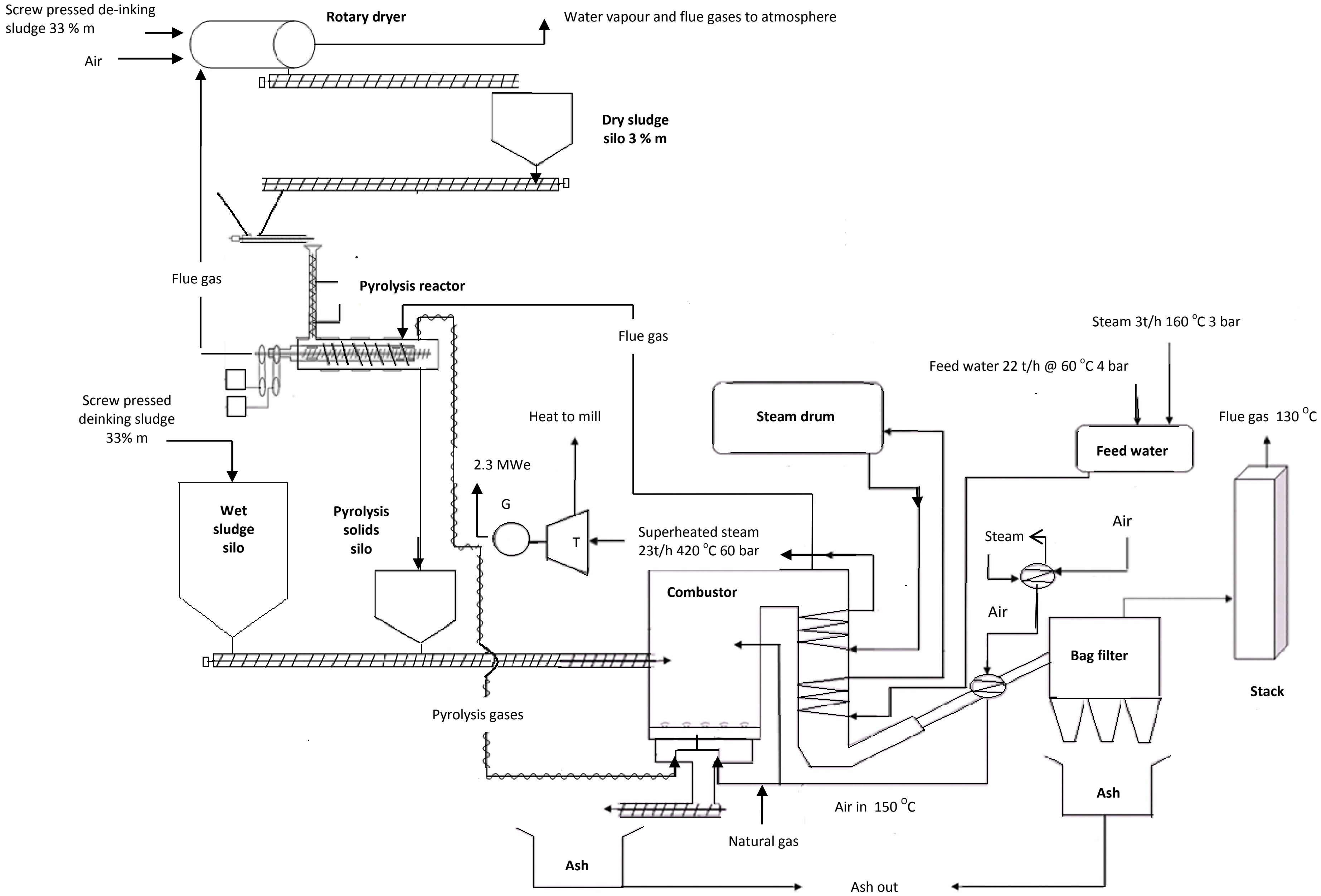


Figure 67 Proposed AN mill process

7.2.2 AN mill proposed process description

The new proposed pyrolysis process illustrated in Figure 67 is based around the current sludge combustion system at the AN mill which would remain largely unchanged, but with the addition of a pyrolysis line to process the remaining de-inking sludge product. The design of the system aims to minimise the amount of modification necessary to the current system, so as to minimise capital cost. It is also believed to be the most feasible configuration in terms of practicality and maintenance.

The new pyrolysis line requires approximately 33,333 wet tonnes per year of screw pressed sludge at 33 wt% moisture to be rotary drum dried to 3 wt% moisture content at a temperature of 100 ^{o}C. The sludge is then transferred to a holding silo using screw conveyors and then fed into the pyrolysis reactor. Pyrolysis of the sludge takes place at 450 ^{o}C and at this temperature volatiles, gas and water vapour are emitted, together with a solid char residue. The uncondensed gaseous products are transferred to the sludge combustor where they substitute a substantial fraction of the existing natural gas demand. The vapours/gases are transferred by heated line to the sludge combustor, into which they are injected via some of the existing natural gas nozzles. It is believed that the pyrolysis vapours/gases, in combination with the remaining natural gas input, have a sufficiently high heating value to maintain satisfactory combustion within the fluidised bed. The char is transferred to a pyrolysis solids silo and then combined with the existing feed of 350 tonnes per day of screw pressed de-inking sludge and conveyed via a screw feeder into the combustor.

At the exit of the combustor, a small fraction of the combustion products are diverted for indirect heating of the pyrolysis unit and subsequently to meet the drying duty. The remaining exhaust gases have an identical energy content to the existing system, and the boiler convective pass is unchanged both in terms of design and duty. It is assumed that the evaporator section also requires no modification and will operate at the same duty.

From the energy balance presented in Appendix A2, the energy from the complete combustion of the pyrolysis products is enough to replace approximately 67 % of natural gas currently used. Only

1.79 MW of natural gas is required by the new system as opposed to the 5.42 MW currently used. Furthermore, all of the sludge is now consumed and the need for land-spreading is eliminated.

It is assumed that all fuels will be combusted in the existing sludge combustor. This would represent a small uprating from the existing duty (23.4 to 24.9 MW) and it is assumed that the combustor can accommodate this. Once the exhaust gases for pyrolysis and drying have been diverted, the remaining exhaust gas energy content (and flow rate to a first approximation) is unchanged, and the boiler convective pass requires no modification. It is assumed that despite the small uprating of the combustor, the evaporator tubes will see a similar heat load, so that the overall steam generation will be unchanged.

The pyrolysis vapours, gases and char almost exactly balance the previous natural gas requirement in terms of energy, although some natural gas must be retained to supply the drying and pyrolysis demands. It is assumed that the total air requirement is unchanged, there will be a small reduction in air requirement due to the change in fuel from natural gas to pyrolysis vapours/gases and char, but a small increase from the uprating of the unit, and these are assumed to balance.

It is further assumed that the combustor exhaust gases required for the pyrolysis and drying heat demand would be extracted from the top of the combustor (i.e. before the convective pass) at 900 °C. These would pass first to the pyrolysis unit where they would cool to around 600°C, and hence to the dryer where they would cool to 100 °C. The gases could be diluted with air at inlet to the dryer if an inlet temperature of 600°C is excessive. Natural gas required by the new system is 3966 Nm^3/day which constitutes a total reduction of natural gas usage by the new system of 67 %.

7.3 SSK mill industrial implementation study

The SSK mill currently requires approximately 244 GWh/y (30.5 MWth) of heat and 67 GWh (8.4 MWe) of electrical energy, to produce an overall quantity of approximately 192,000 tonnes per year of brown paper reels. The vast majority of this energy is produced on-site using natural gas fired gas turbines and natural gas fired boilers. An additional amount of heat energy approximately 98 GWh (12.3 MWth) is produced using a combination of biogas and natural gas in a third boiler. Approximately 388 GWh/y (48.5 MWth) of natural gas is imported in total. A further 7 GWhe/y (0.87 MWe) of electrical energy is imported from the national grid and a small amount of surplus electrical energy is returned back to the national grid from the on-site CHP systems. The current system is illustrated in Figure 68. In this study a proposed fixed bed downdraft gasification system producing a product gas by co-firing rejects and wood chips will be integrated into the existing biogas boiler as a replacement for natural gas and biogas. 8000 annual operating hours are assumed.

7.3.1 SSK mill current process description

Wastewater generated at the SSK mill is processed by an anaerobic process which reduces the majority of the dissolved organic content. Anaerobic treatment has an impressive and proven track record in the paper industry, which is why SSK selected anaerobic technology as the basis for a wastewater treatment solution at their site. The biogas, a valuable by-product of the anaerobic treatment process, is currently used to fuel an on-site boiler [75].

The anaerobic digestion process at the SSK mill has been designed to remove 80 % of the chemical oxygen demand (COD) and further reduce the level of suspended solids in the treated wastewater [75]. Utilising the biogas in one of the existing boilers further reduces the mill’s energy costs, as less natural gas is required for the mill’s steam generation demand, helping in turn to reduce the mill’s carbon footprint.

There are two energy systems which currently exist on-site. The first comprises two natural gas fired gas turbines (CHP 1 and CHP 2) which generate a total of 62 GWhe/y (7.7 MWe), and have an efficiency of approximately 25 %. The second comprises 3 boilers, one of which (Boiler 1) has been modified to allow both biogas from the on-site wastewater treatment plant and natural gas to be used as a fuel. Boiler 2 and Boiler 3 are natural gas fired boilers which generate steam for the mill. Boiler 2 and Boiler 3 raise a combined total of approximately 191 GWh/y (23.9 MWth) of steam at a pressure of 15 bar and temperature of 200 ^{o}C.

The biogas used in Boiler 1 is produced from the anaerobic digestion of waste water which creates between 300-550 Nm^3/h of gas composed of 70 % CH_4 and 30 % CO_2 by volume. This quantity is sufficient to supply approximately 16 GWh/y (2 MWth) of thermal energy to the boiler. This arrangement will however change in the near future as the vast majority of the biogas will be routed instead to a new CHP engine to create 890 kWe, with heat from the exhaust gases passing to a waste heat boiler to raise steam. Therefore Boiler 1 will either have to operate using natural gas or an alternative fuel supply if it is to remain operational.

Approximately 307 GWh/y (38.4 MWth) of natural gas is used by the CHP turbines and boilers, and approximately 81 GWh/y (10.1 MWth) is used in the biogas boiler where it is combined with 16 GWh/y (2 MWth) of biogas and 0.3 GWh/y (0.04 MWth) of gas oil which is used intermittently. A total of 98 GWh/y (12.2 MWth) of steam is raised from this boiler at 15 bar and 200 ^{o}C; the majority of this steam is used to facilitate in the drying of paper sheets in the mill. The water condensate from the steam after it has been used contains residual heat energy; this water is combined with additional make up water and returned to the boilers for re-use. A total of 44 GWh/y (5.5 MWth) of water condensate heat is returned to the boilers.

In total there is approximately 97 GWh/y (12.2 MWth) of available waste heat emitted from the existing CHP systems in exhaust gases after steam has been raised. This heat can potentially be recovered and re-used, for example as a drying medium for the pre-treatment drying of rejects, or to preheat combustion air. Figure 68 shows the current energy system at the SSK mill with energy flows; note steam losses from breaks in the line downstream of the CHP waste heat boilers are not included in the calculation. Detailed energy balance calculations can be found in Appendix B1.

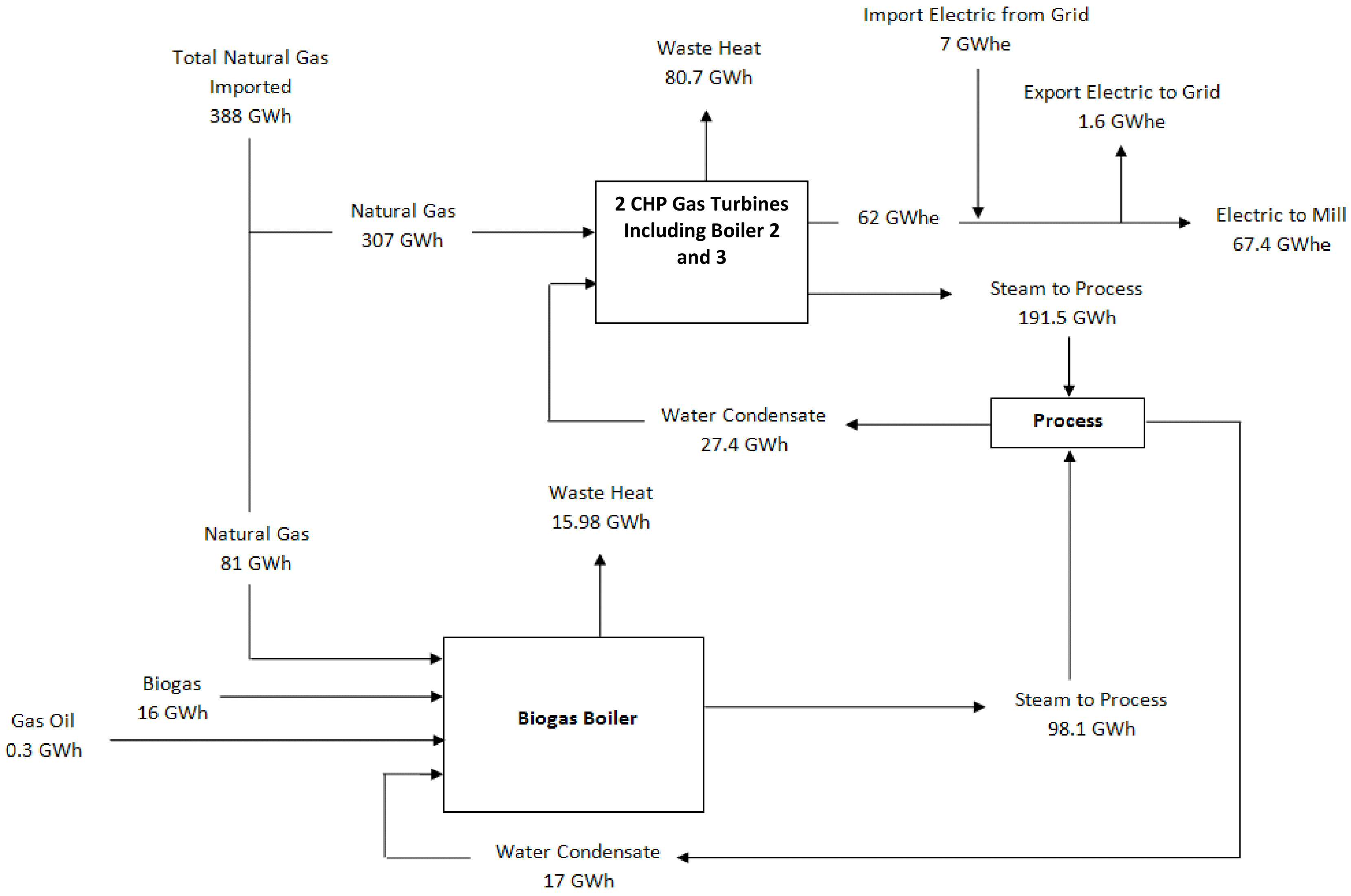


Figure 68 Current CHP system at the SSK mill

7.3.2 SSK mill proposed process description

The rejects created at the SSK mill which will be considered for the proposed gasification system are the general rejects consisting of 30% plastics (film, wrap, bags etc.) and 70% fibre which originates from wood pulp. Government incentives are available for the gasification of renewable feedstocks and to what extent this would be applicable to this reject stream is explored in Chapter 8. Approximately 17,000 wet tonnes /year of this stream is considered for the new process. These rejects have an initial moisture content of 55 wt%.

Drying of the rejects will be necessary before gasification in order to reduce the moisture content down from 55 wt% to approximately 10 wt%, and it is recommended that the heat necessary to do this would come from waste heat of the existing on-site energy systems . It is believed that the lack of space for a gasifier and pre-treatment facility could be an issue at the SSK site although a footprint for the new facility would need to be confirmed.

The proposed plant consists of a pre-treatment facility and gasification plant which are integrated together and linked to the existing biogas boiler as a replacement for biogas and natural gas to raise 12.3 MW of steam for the mill. The intention is that this will be a replacement for biogas, as in the near future the biogas will be routed to a separate biogas engine which will produce approximately 890 KW electrical for the mill.

Before rejects can be used in any thermal conversion process to produce energy they must first be pre-conditioned to remove contaminants, inorganic substances and water. The proposed pre-treatment is based on an existing facility which Smurfit Kappa (SK) has already implemented at the Roermond plant in the Netherlands.

The pre-treatment process begins by transporting 17,000 tonnes per year of general rejects with an initial moisture content of 55 wt% into the pre-treatment facility by tractor/trailers or conveyors. Here the rejects are dumped into initial holding silos for pre-processing. The rejects are then sieved to loosen the initial bulk of material, and spread evenly onto conveyor belts with overband metal detector magnets. As the rejects are transported along the conveyors, manual pickers or operators continuously monitor the mass flow and lift out any oversized objects which may cause damage to downstream equipment, when identified the belt is stopped and these objects are manually removed and placed in skips for landfill. As the rejects are transported along this conveyor ferrous metals are lifted by magnets and are transported to large ferrous metal containers to be sold as scrap. The rejects then pass through eddy currents where non-ferrous metals are repelled and are removed. The remainder of the rejects at this point composed of plastics and fibres fall into a screw compactor where it is de-watered from 55 wt% moisture down to 40 wt% moisture. A flip-flow sieve then separates the larger particles (> 30 mm), which are cut into smaller fragments in a shredder underneath. After re-mixing of the shredded particles with a top up of 20 wt% wood chips the organic mass dries in a rotating drum dryer, where the dry content is increased to 90 wt%. This is achieved by using hot exhaust gases from the existing waste heat boilers downstream. While the heaviest particles remain behind in the dryer, the light fraction is transported by the drying gases and reclaimed in a separator and two cyclones downstream. Finally the reject/wood chip product is pelletised and transported to the gasifier feed silos ready for thermal conversion by gasification. The drying gases from the drum dryer pass through a gas scrubber, where dust and odour are removed and is then discharged to the atmosphere from a stack. The final feedstock is a dry pelletised organic matter with an average calorific value of approximately 21 MJ/kg similar to that produced at the Rofire facility.

The thermal conversion process begins by transporting the reject pellets and wood chips from the feedstock holding silos into multiple downdraft gasifiers which are each designed to process a total capacity of 250 kg/hr of dry feed. These gasifiers are well established units used by companies such as Biomass Engineering Ltd. Feedstock enters the gasifier via screw feeders. The temperature of the feedstock increases as it passes through each zone of the gasifier. The gasification air is supplied to the throat of the gasifier, and enters the oxidation zone. An air equivalence ratio of between 0.2-0.3 is used. The product gas exits the gasifier and is routed into cyclones to remove fly ash and traces of dust particulates and soot entrained in the product gas. The product gases leave the cyclones at a temperature slightly above the condensation point of tar > 500 ^{o}C and then enter into a three stage scrubbing train. The three stage scrubbing system uses chilled waste vegetable oil as the scrubbing medium and serves three purposes. The first is to remove the sensible heat of the product gas, the second is to remove tar from the gas to prevent clogging and fouling in pipes and components downstream and the third is to remove water vapour entrained in the gas. The scrubbing oil is recovered and sent to an oil recovery tank to be recycled. The product gas then passes into a riser where it is elevated to a product gas buffer tank situated on the roof of the building for safety reasons. As the gas rises it cools down further to ambient conditions before being stored in a gas buffer tank. The product gas at this stage is ready to be used in the existing biogas boiler. All of the product gas is routed to the existing biogas boiler where it is premixed with natural gas to produce 12.3 MW thermal steam energy. Figures 69-71 illustrate the proposed process and a detailed energy balance calculations of the proposed process can be found in Appendix B2.

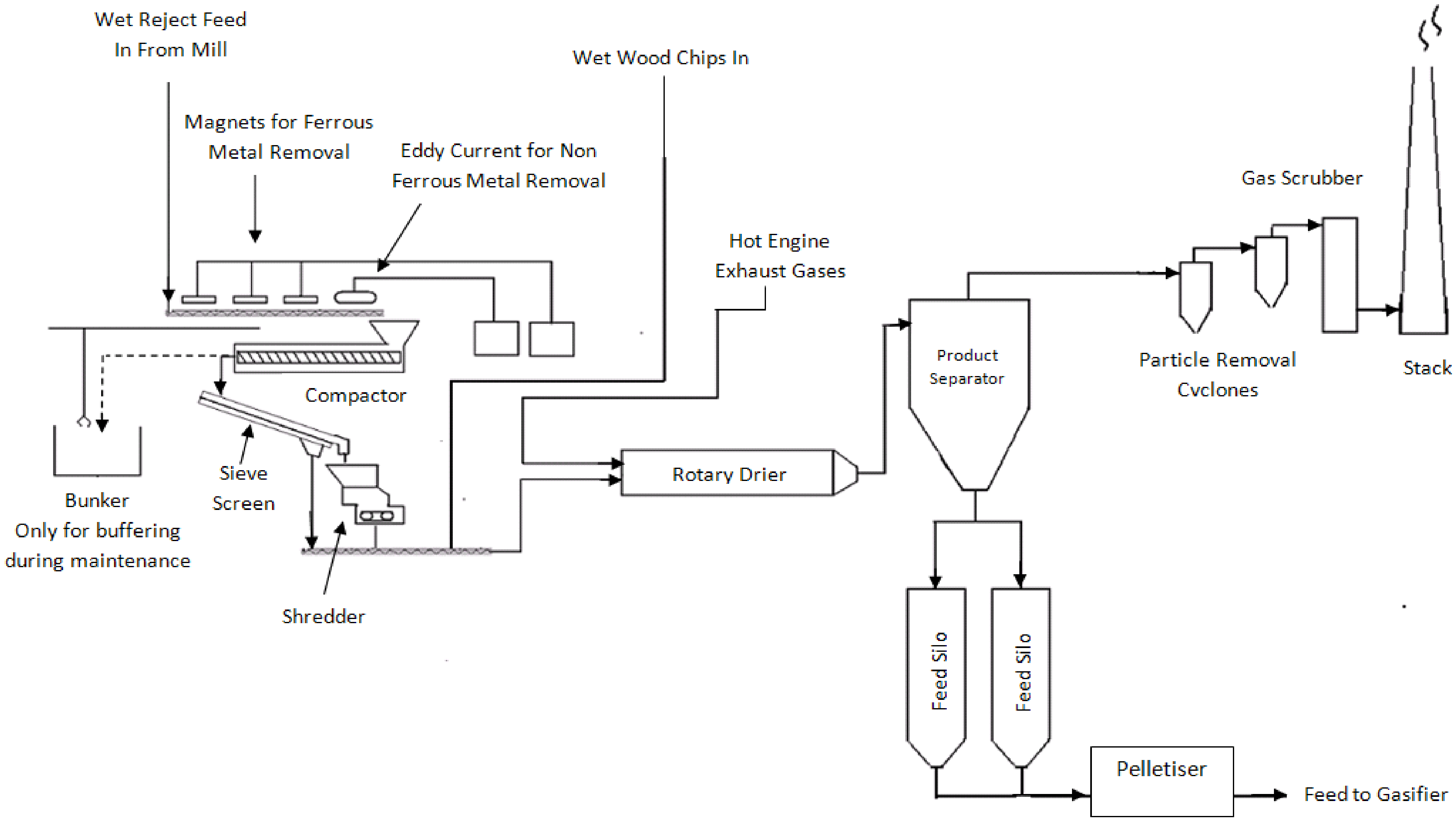


Figure 69 Proposed SSK pre-treatment plant

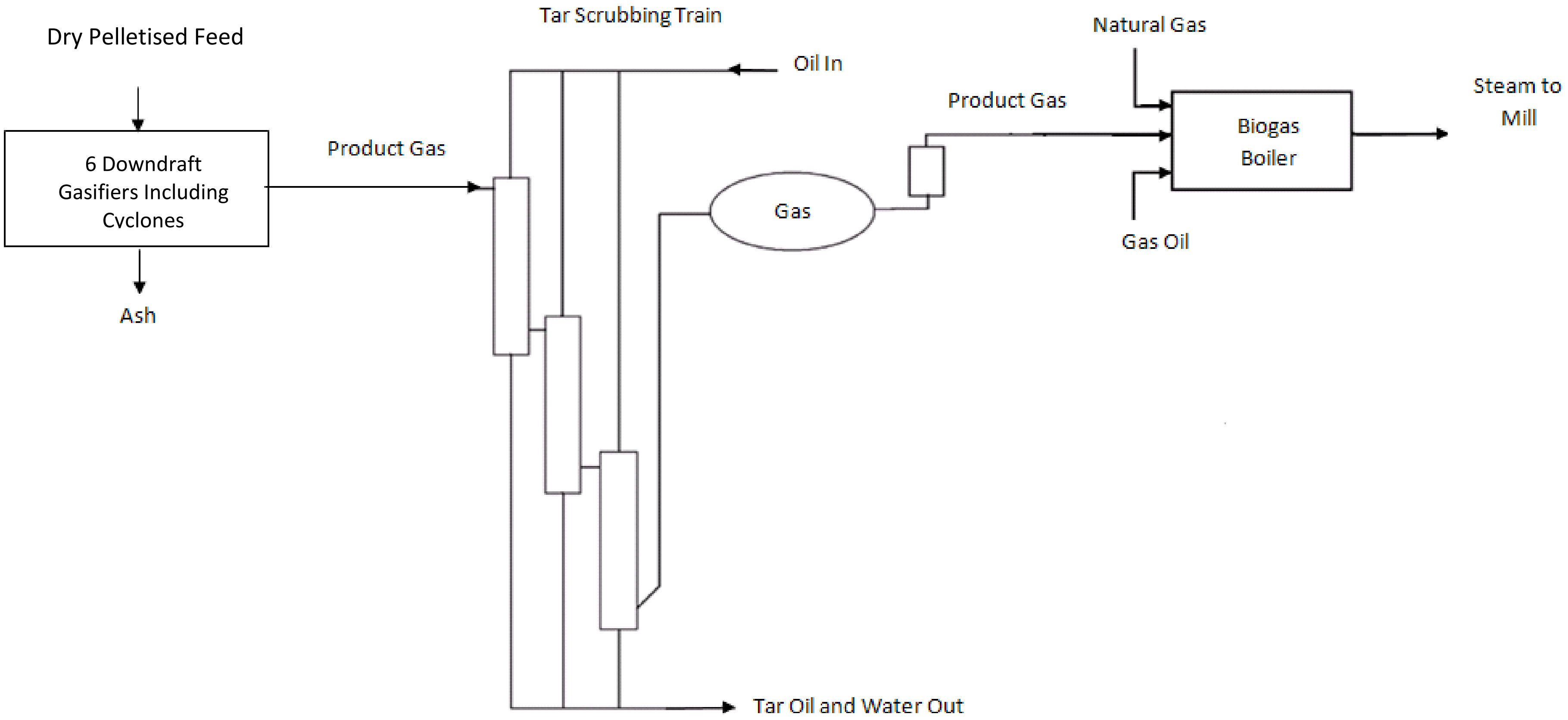


Figure 70 Proposed SSK gasification plant

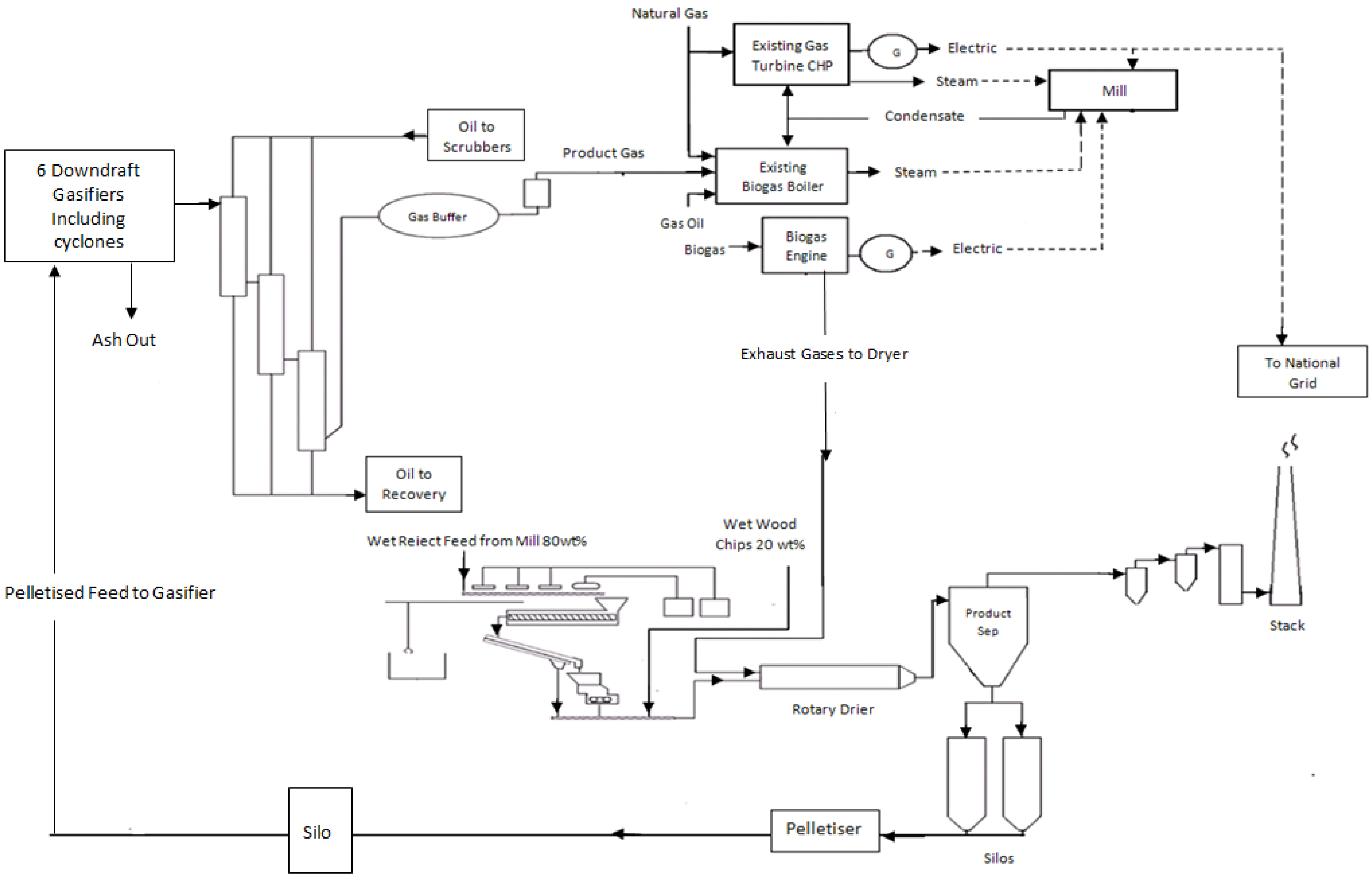


Figure 71 Complete proposed SSK plant

Overall the system is energy deficient by 4.9 MW which will be provided by natural gas combustion. It is assumed that all fuels will be combusted in the existing biogas boiler. The chemical energy supplied by the product gases from the gasifiers is sufficient to replace all of the biogas used by the existing boiler as well as reducing the amount of natural gas used (currently 10.1 MWth) by approximately 48.5 %. However the number of downdraft gasifiers required for this duty is calculated to be 6 in parallel. This may not be feasible to implement at the SSK site as considerable space is required for both the pre-treatment and gasifier installations and there is limited space available at the SSK site to accommodate this. However this process may be feasible at other SK sites which have similar compositional feedstocks and more space for such a large installation.

7.4 KC mill industrial implementation study

There currently exist three installations which operate at the KC site. These are the Flint installation, the Coleshill installation and the Delyn installation. Collectively they produce over 60,000 tonnes per year of tissue paper, cleansing wipes and hand towels. This industrial implementation study has focused exclusively on the Coleshill installation as the potential candidate for a proposed de-inking sludge pyrolysis plant, as Coleshill is the highest consumer of energy at the KC site. Furthermore due to the location of Coleshill and its close proximity to the feedstock, as well as the availability of space surrounding the installation, it is ideally situated to house a pyrolysis reactor and associated equipment.

Energy data provided for this study is based on figures from the year 2000, however it reasonable to assume that these figures are representative of current figures as annual production of tissue products produced at the Coleshill installation has not significantly changed since then.

Energy imported to the KC site consists mostly of natural gas purchased from BP and electricity purchased from British Energy. Supplier contracts are reviewed on an annual basis. The electricity consumption at the Coleshill installation is approximately 5 MWe. The total amount of natural gas imported to the Coleshill installation is 9.3 MWth.

In this study a proposed pyrolysis system to extract de-inking sludge volatiles for heat generation will be integrated into the existing boiler at Coleshill mill (referred to henceforth as the KC mill). The system will provide steam at 14 bar pressure and 210 °C for the mill. 8000 annual operating hours are assumed.

7.4.1 KC mill current process description

The approximate distribution of steam throughout the KC mill is shown in Figure 72. The main energy consumer is a natural gas fired boiler which has a maximum energy output of 8.5 MWth. This is a dual fuel fired unit and has a three pass wet back design with a large shell and heating surface for optimum efficiency. The boiler is rated to produce a maximum of 14 bar steam at a rate of 13,600 kg/hr and operates at an efficiency of 80 %. Since the initial installation of the boiler, an oxygen trim system (automatic control of excess O_2 by adjustment of air supply) has been fitted to the unit to improve operating efficiency. Natural gas is used as the primary fuel with gas oil available as an alternative fuel supply. Due to the availability of natural gas, the gas oil is used very infrequently. Under normal operating conditions the boiler will produce 7.5 tonnes/hr of steam at a temperature of 210 °C and pressure of 14 bar.

Most of the steam is exported from the boiler to the tissue machine where it is used in a Yankee dryer. Condensate from the Yankee passes through a series of separators to recover as much energy as possible. The dryer is also heated by natural gas, via two 3.5 MW dual fuel rotary cup burners which supply hot air to the to the wet and dry end hoods above the Yankee cylinder. The hot air (300-480 °C) is blown with high velocity against the tissue. This is a highly efficient unit which enables the natural gas consumption to be kept to a minimum. Some of the steam from the boiler is also used in a disperger heating screw to directly heat the tissue stock to 90 °C and to heat the disperger dilution water. The final use of the steam is to heat the Yankee cylinder coating tank. Energy balance calculations around the Coleshill boiler are can be found in Appendix C1.

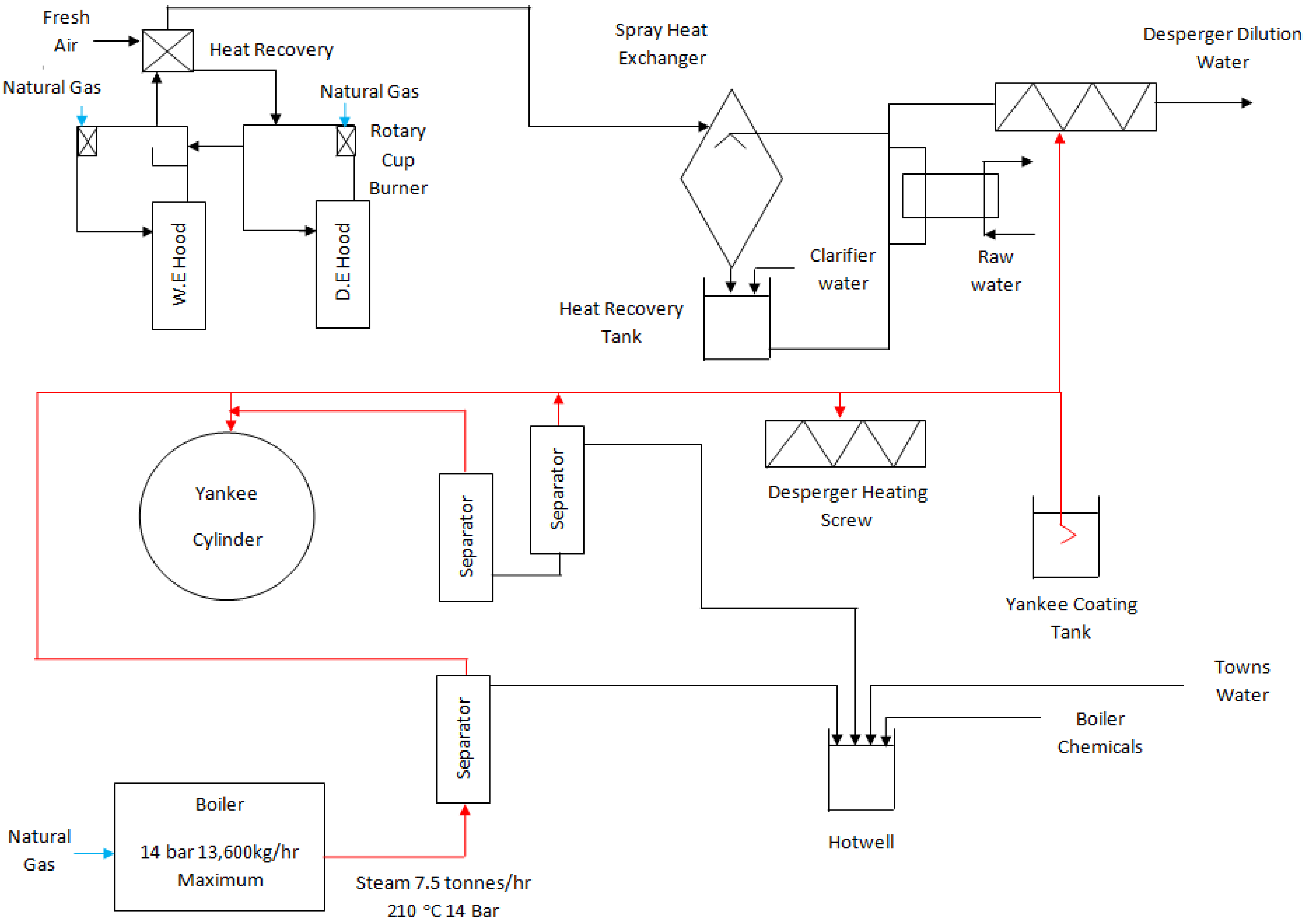


Figure 72 Current process at the KC mill

7.4.2 KC mill proposed process description

The proposed pyrolysis plant is designed to process a maximum of 40,000 wet tonnes per year of de-inking sludge at a moisture content of 45 wt%. The KC mill currently disposes of its de-inking sludge either to land reclamation schemes where it can neutralise acidic soils, as an animal bedding medium or as a boiler fuel to a local paper mill.

The pyrolysis process illustrated in Figure 73 is based around the current boiler. The design of the process aims to minimise the amount of modification necessary to the current system and thus the capital cost. It is also believed to be the most practical arrangement.

It is assumed that the existing boiler will continue to operate with the same energy output, but with 53 % of the natural gas displaced by the use of exhaust gases from the combustion of pyrolysis vapours and solid residues produced upstream by the pyrolysis unit.

The new system requires approximately 40,000 wet tonnes per year of screw pressed sludge at 45 wt% moisture to be rotary drum dried. Once dried to a moisture content of 3 % the sludge is fed into the pyrolysis reactor. Pyrolysis of the sludge takes place at 450 ^{o}C and at this temperature volatiles, gas and water vapour are emitted. All vapours are routed via heated lines to a fluidised bed combustor and are pre-mixed with air to combust the residual pyrolysis solid residues.

The exhaust gases from the combustor are split into two streams. One is routed to the pyrolysis reactor and then to the dryer to provide the energy required by the endothermic pyrolysis and drying processes. The second is routed to the boiler to replace some of the natural gas. The remainder of the process downstream from the boiler remains the same as the existing system. Cost savings to the KC mill by integrating such a system would come from the reduction in sludge disposal costs as well as the reduction in natural gas used. An income could also be gained from the potential sale of de-inking sludge ash to the cement industry as well as from Renewable Heat Incentives (RHIs). Detailed energy balance calculations of the proposed process can be found in Appendix C2.

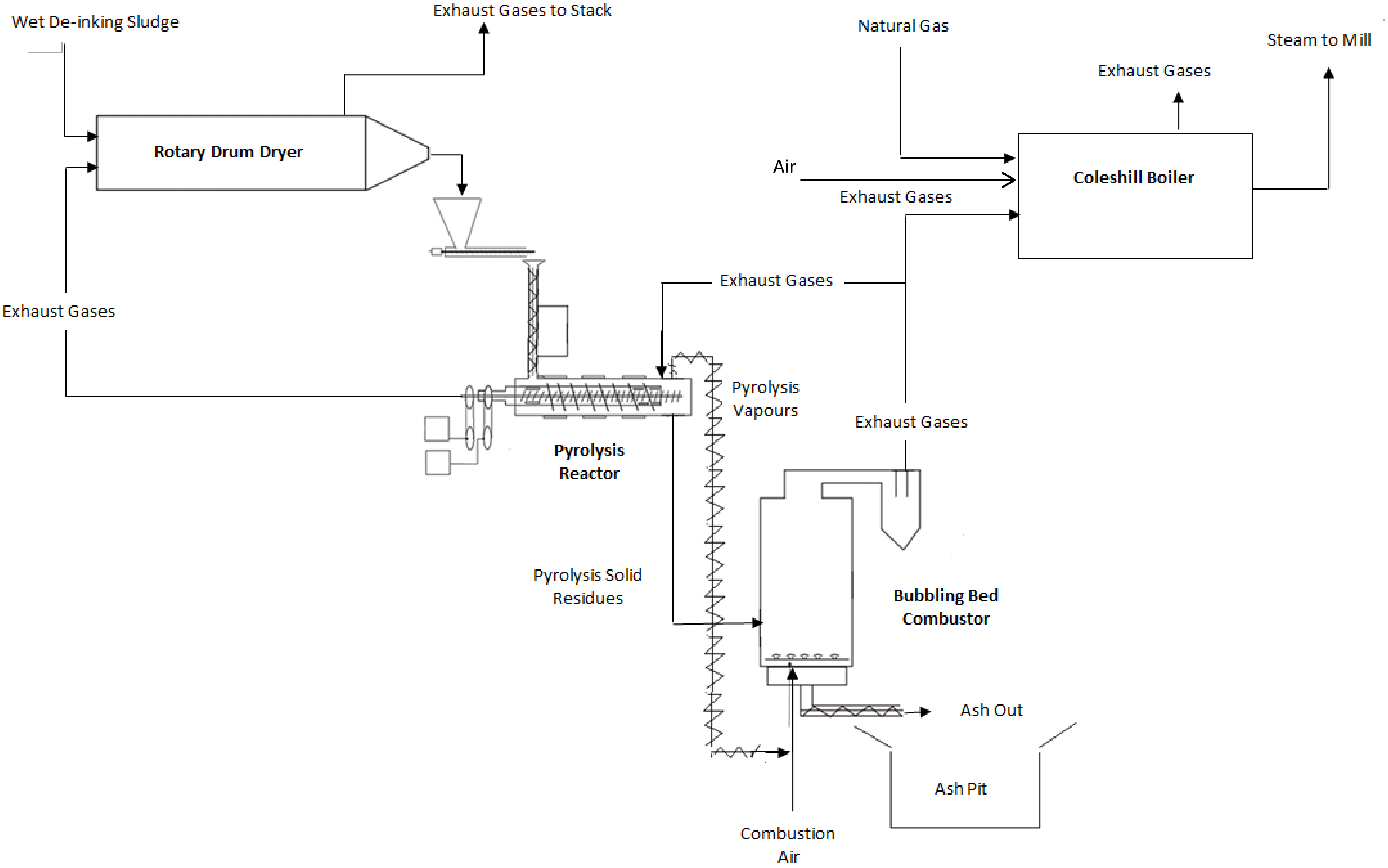


Figure 73 New proposed KC process description

From the energy balance calculations presented in Appendix C2, the new system is overall energy deficient by 3.51 MW. The deficient energy is to be supplied from natural gas combustion which will be fired directly into the Coleshill boiler. Once the exhaust gases for pyrolysis and drying have been diverted from the combustor exit, the remaining exhaust gas is routed to the boiler which requires no modification. It is assumed that the evaporator tubes will experience virtually the same thermal environment, so that the overall steam generation will be unchanged.

It is further assumed that the combustor exhaust gases required for the pyrolysis and drying heat demand would be extracted from the top of the combustor at 900 °C. These would pass first to the pyrolysis unit where they would cool to around 600 °C, and hence to the dryer where they would cool to 100 °C. The gases could be diluted with air at inlet to the dryer if an inlet temperature of 600°C is excessive. Natural gas required by the new system is 324 Nm^3/hr which constitutes a total reduction of natural gas usage by the new system of 53 %.

8.0 ECONOMIC EVALUATION

8.1 Chapter summary

This chapter is based on the findings from the industrial implementation study presented in Chapter 7. Detailed economic calculations are carried out to determine the total capital investment, operating costs and payback time if each mill were to implement the proposed processes. Figures presented in the study are based on cost data acquired from equipment suppliers and cost information gained from discussions with the mill. Where cost data are not available or limited then cost estimation techniques are used to obtain a best estimate. Calculations are largely based around sensitive cost data provided by the mills. Due to the confidentiality of this information this section of the thesis only provides a summary of the main findings from the study.

8.2 Economic evaluation criteria

The base year for the study is 2012; this being the year that all cost data was collected. Therefore all costs are expressed as 2012 GB£ unless otherwise specified. An inflation rate of 3 % is assumed and a total project life is taken to be 25 years.

All costs are presented as preliminary estimates only and are calculated within typically +/– 30 % of the true cost. A more precise economic appraisal would require a thorough analysis and critique into the specific aspects of the plant design which is beyond the scope of this study.

This economic evaluation has essentially been broken down into four parts:

(i) Total Capital Cost (TCC), that is to say the total capital investment required to install all equipment necessary up to the point that the plant is handed over to the mill and is operational under normal conditions,

(ii) Operating Costs (OC) which includes all the costs incurred for operating the plant on a day to day basis such as labour, energy, maintenance and repair costs,

(iii) Income or Savings gained from the operation of the proposed process, and

(iv) Payback Time, this is defined as the total length of time required to recover the Total Capital Cost (TCC) of the plant.

For the purpose of this study the following conversion rates have been used:

1.22 Euros = 1 GB£

1.6 US $ = 1 GB£

Some cost parameters are common for all mills. Where this has been the case it has been stated.

8.2.1 Total Capital Cost (TCC)

The Total Capital Cost (TCC) is defined as the sum of the Fixed Capital (FC) and Working Capital (WC). FC is the total cost of the plant ready for start-up and includes the design, engineering, construction and supervision costs, all items of equipment including their installation, all piping, instrumentation and control systems, building and structural costs and auxiliary facilities including utilities, land, civil and engineering works. FC costs are a once only cost that is not recoverable at the end of the project life other than via scrap value.

WC is the additional investment needed over and above the FC cost to start the plant up and to finance a period of operation. This cost includes start up fees, the purchase of initial catalysts, raw materials and intermediates in the process. WC is recoverable at the end of the projects life and can usually vary from between 5-30% of FC cost [76].

TCC in this study is based on the detailed factorial method for estimating the cost of chemical process plants [76]. It considers the purchase cost of each major piece of plant equipment and factors in all other costs as estimated increments of these items.

The calculation of TCC begins with the purchase cost of each piece of equipment (PCE). It is assumed that all plant items are bought new. Increments are then added for erection, instrumentation, piping and ducting, associated electrical equipment, structures and buildings and civil works to give the total physical plant costs (PPC). Engineering, design, contractor’s fees and a contingency allowance are then added to give the Total Fixed Capital (FC) cost and finally the working capital (WC) cost of 10 % is added to give TCC.

8.2.2 Operating Costs (OC)

The operating costs incurred by the mills are essentially the day to day running costs of the new facility and are comprised of both fixed and variable operating costs. The fixed costs includes maintenance, labour, supervision plant overheads, capital charges, rates and local taxes, insurance, licence fees and royalty payments and the variable costs includes the purchase of raw materials, operating materials and utilities. The operating costs are given on an annual basis.

8.2.3 Payback Time

The Payback Time is calculated as the total time taken to recover TCC, from launch of the project.

8.3 AN mill economic evaluation

The new AN mill plant described in Section 7.2.2 will process a total of 33,333 wet tonnes per year of de-inking sludge and will aim to integrate a drying and pyrolysis reactor with the existing sludge combustion system for the production of sustainable CHP. The plant will utilise the full amount of de-inking sludge produced at the mill which will result in a reduction in its current natural gas usage by approximately 67 %. .

8.3.1 Cost of current AN mill system

The AN mill currently produces around 150,000 tonnes/year of de-inking sludge at 33% moisture. Approximately 116,667 tonnes/year of de-inking sludge is burnt in a sludge combustor using natural gas, with the great majority of the remaining 33,333 tonnes/year sent for land-spreading. 75% of the ash produced, 32,500 tonnes/year, is sold to cement and concrete block manufacturing companies. The remainder of the ash, approximately 10,800 tonnes/year, is sent to an alternative disposal site. The total cost for purchase of natural gas and disposal of sludge is approximately £1,000,000 per year.

8.3.2 AN mill Total Capital Cost (TCC)

For the proposed process at the AN mill a relatively low working capital would be required as it is a simple process which does not require the use of catalysts or the purchase of raw materials; therefore WC is taken as 10 % of the FC cost.

The two major plant items required by the mill are a dryer and a pyrolysis reactor. The mill also requires auxiliary equipment to facilitate in the process such as pumps, fans, conveyors, silos, instrument control systems, valves etc. The incremental factors used to calculate the cost of these minor plant items are given in Table 37.

8.3.2.1 Dryer cost

The preferred dryer to carry out the drying duty would be a rotary drum dryer. This has been proven to dry sludge very efficiently down to a moisture content of < 5 wt% and is currently the type preferred by companies such as Envirosystems ltd who dry de-inking sludge for use as cattle bedding. To estimate the purchase cost of the rotary drum dryer companies were approached to obtain a quote. The dryer would be delivered new to the mill gates and the cost of purchase does not include other costs such as erecting or commissioning.

From the general enquiries a UK based supplier TCR was able to give a quote for a rotary drum dryer which met the required specification. Their dryer would utilise combustion exhaust gases to directly dry a maximum capacity of 5 tonnes per hour of sludge from an initial moisture content of approximately 33 wt% down to < 5 wt%. The dryer is suitable for de-inking sludge applications and would be sufficient to evaporate approximately 1 tonne per hour of water vapour with an exhaust gas exit temperature of approximately 100 ^{o}C. The specification and quote for the dryer as provided by TCR is given below:

Triple Stage Rotary Dryer complete with cyclone, fan and control panel but no provision for any scrubbers that may be required.

Dryer Specification

Dryer length	6000 mm
Diameter	2000 mm
Weight	8,000kg
In feed diameter	400 mm
Rotation speed	5 – 11 rpm
Drum drive power	10 kW
Fan	30kW
Cyclone Diameter	1600 mm
Height	4550 mm

The control panel contains starters, overloads, isolator, emergency stops etc.

PCE dryer = £ 110,000

This was taken to be a representative of a typical PCE for the rotary drum dryer and was therefore used as the bases in calculating the TCC as shown in Table 37.

8.3.2.2 Pyrolysis reactor cost

The "Pyroformer" pyrolysis reactor unit would be delivered new to the mill and would be manufactured by WGD based in Germany. Although it was not possible to obtain an accurate quote for a 3 tonne per hour Pyroformer, some cost information was available for a 500 kg per hour unit, and a rapid capital cost estimating technique was therefore used to relate the capital cost of the reactor with capacity by equation 8.

$C2 = C1\ (S2/S1)^{n}$ (8) [76]

where:

C_2 is the capital cost of the reactor with capacity S_2

C_1 is the capital cost of the reactor with capacity S_1

The value for the index n is commonly taken as 0.6 [76]. The capital cost of a 500 kg per hour unit delivered as new has been stated to be approximately £400,000 [77]

Therefore:

C_2 = £400,000 (3/0.5) $^{0.6}$

C2 = £ 1,180,000

PCE pyrolysis reactor = £ 1,180,000

The PCE costs for the dryer and pyrolysis reactor obtained above were used as the basis in calculating the TCC as shown in Table 37. Factors were estimated according to published data [76].

Table 37 AN mill Total Capital Cost (TCC)

Item	Factor	£ Cost
Equipment Total Purchase Cost (PCE)		£ 1,290,000.00
F0	1*PCE	£ 1,290,000.00
F1 Equipment Erection	0.5*PCE	£ 645,000.00
F2 Piping	0.2*PCE	£ 258,000.00
F3 Instrumentation	0.1*PCE	£ 129,000.00
F4 Electrical	0.1*PCE	£ 129,000.00
F5 Building, Process	0.05*PCE	£ 64,500.00
F6 Utilities	0.25*PCE	£ 322,500.00
F7 Storage	0.25*PCE	£ 322,500.00
F8 Site Development	0.05*PCE	£ 64,500.00
F9 Ancillary Buildings	0.3*PCE	£ 387,000.00
Total	**2.8**	£ 3,612,000.00
Total Physical Plant Cost (PPC)	2.8* PCE	£ 3,612,000.00
	1*PPC	£ 3,612,000.00
Design and Engineering	0.2*PPC	£ 722,400.00
Contractors Fees	0.05*PPC	£ 180,600.00
Contingency	0.1*PPC	£ 361,200.00
Total	1.35	£ 4,876,200.00
Fixed Capital Cost (FC)	1.35*PPC	**£ 4,876,200.00**
Working Capital Cost (WC)	0.1*FC	**£ 487,620.00**
TOTAL CAPITAL COST (TCC)		**£ 5,363,820.00**

8.3.3 AN mill Operating Costs (OC)

The operating costs incurred by the AN mill are essentially the day to day running costs of the new facility and are comprised of both fixed and variable operating costs.

8.3.3.1 Raw materials

These are the essential major materials required to run the process. As the main raw material used in the process is de-inking sludge there is no associated cost.

8.3.3.2 Utilities (services)

This term includes power, steam, compressed air, cooling and process water and effluent treatment. As the majority of the electrical and heat power required to drive the process will be generated by the process itself and little compressed air is required by the process, this cost will be minimal. A small amount of nitrogen is required to purge the pyrolysis reactor intermittently and the cost for this has been accounted for here. A typical value of 9 p/Nm^3 of nitrogen [78] is taken for this expense, it is estimated that a total amount of 16,000 Nm^3 of nitrogen per year will be used to purge the system equating to a total annual nitrogen cost of £1,440.

Although the implementation of the new pyrolysis system would significantly reduce the amount of natural gas required by the sludge combustor, the system would still require approximately 1.3 million Nm^3 per year of natural gas to operate. The total annual cost of utilities is assumed to be £323,240 per year.

8.3.3.3 Maintenance

This item will include the cost of maintenance labour and the maintenance materials required, including equipment spares needed. An initial estimate of 1 % [76] of the FC cost is taken as the annual maintenance costs which is shared between the cost of labour and materials required. Total annual maintenance cost is £48,762.

8.3.3.4 Miscellaneous materials

Included in these costs are items which are not listed under raw materials and covers items such as safety clothing for operators during maintenance, instrument charts, gaskets and other consumables. As a rough guide the cost of miscellaneous materials is taken as 5 % [76] of the total maintenance costs and equates to £2,438.10

8.3.3.5 Operating labour

The mill's current labour force is considered sufficient to run the process with no additional operators required for the day to day running of the process.

8.3.3.6 Ash disposal

An additional throughput of ash will be produced by the sludge combustor from the new system. The mill's ash disposal costs will therefore increase. The mill may be able to offload a certain amount to the cement industry as a concrete additive. This study however has assumed that none of the surplus ash produced by the new system will be sold and therefore would need to be disposed of. A total of £218,000 per year in ash disposal costs is estimated.

8.3.3.7 Supervision

The mill's current labour force is considered sufficient to supervise the process with no additional supervisors required.

8.3.3.8 Plant overheads

Included under this heading are all the general costs associated with operating the plant not included under the other headings; such as, general management, plant security, medical, canteen general clerical staff and safety. All these costs are assumed to be covered under the mill's current expenditure therefore no additional plant overheads are included for the new system.

8.3.3.9 Capital charges

The investment required for building the new installation has been assumed to come from the mill's retained profits and therefore not subject to a capital charge.

8.3.3.10 Local taxes

Local taxes are assumed to be unchanged.

8.3.3.11 Insurance

The annual cost of insuring the new installation is assumed to be covered under the mill's existing insurance policy and therefore not subject to an additional charge.

8.3.3.12 Royalties and licence fees

The new process would be subject to a royalty fee paid to the process creator. An annual cost of 1 % FC is estimated. Total annual royalty cost is therefore £48,762.

8.3.4 Income from ROCs

The AN mill will receive an income by incorporating the new pyrolysis process. De-inking sludge is assumed to contain a maximum of 30 % renewable fibres which originated from wood pulp; the new pyrolysis process would therefore be entitled to receive a revenue generated from Renewable Obligation Certificates (ROCs). According to the Department of Energy and Climate Change, the use of advanced thermal conversion technologies such as pyrolysis to process wastes in this way are entitled to receive up to 2 ROC's per MWh of electricity which is produced, subject to an accreditation by Ofgem (UK government energy regulator) [79]. The calculation is based on the proportion of electricity which is generated from the renewable fraction of the feedstock. An estimation of the amount of ROC's the mill would receive from the new process is as follows.

It is estimated that on a dry basis a maximum of 30 wt% of de-inking sludge originates from renewable wood pulp fibres, approximately 50 wt% is composed of inorganic ash and the remaining 20 wt% is composed of other non-renewable components.

Assuming 8000 operating hours, total electricity produced by the AN mill = 2.3 MWe = 18,400 MWh

Proportion of renewable electricity produced by the AN mill = 18,400 MWh * 0.3 = 5520 MWh

Total number of ROCs issued = 2 per MWh = 5520 * 2 = 11040

Current price per ROC = £39

Total annual revenue from ROC's = 39 * 11040 = £ 430,560

8.3.5 Savings

There are two forms of savings which will be made by the AN mill from implementing the new process. The first is a 67 % saving in the amount of natural gas which is currently used and the second is a saving on sludge disposal costs. A breakdown of the total saving made by the new system in comparison to the old system is presented in Table 38.

Table 38 A breakdown of total savings made by the proposed AN mill process

	New System	**Current System**
Variable Costs (per annum)		
Utilities	-£323,240.00	confidential
Maintenance	-£48,762.00	
Miscellaneous Materials	-£2,438.10	
Fixed Costs (per annum)		
Sludge Disposal		confidential
Ash Disposal Costs	-£218,000.00	confidential
Royalties and Licence Fees	-£48,762.00	
Total	-£641,202.10	confidential
Income		
ROCs	£430,560.00	
Ash Sales	confidential	confidential
Total Old Expense		-£1,073,000.00
Total New Income	£114,358	
Net Saving Per Year	**£1,187,358**	

8.3.6 Payback Time

The Payback Time is calculated as the total time take to recover TCC, from launch of the plant. The time for commissioning is not included in the calculation..

£5,363,820 / £1,187,358 = 4.5 years

8.3.7 AN mill economic evaluation summary

It is estimated that a total capital investment of approximately £ 5.4 million is required to integrate a new drying and sludge pyrolysis system with the AN mill's current sludge combustor system. The new system is designed to process a total of 33,333 wet tonnes per year of de-inking sludge. After all expenses the AN mill would make a net saving of approximately £ 1.2 million per year by implementing the new system. A payback time of 4.5 years is therefore estimated to recover the total capital investment of the project.

8.4 SSK mill economic evaluation

The new SSK mill plant described in Section 7.3.2 will process a total of 17,000 wet tonnes per year of rejects and will aim to integrate a pre-treatment and gasification plant with the existing biogas boiler for the production of 12.3 MW of steam for use in the mill. The plant will utilise the full amount of rejects produced at the mill, thereby reducing its current natural gas usage by 48.5 %.

This economic evaluation gives a broad outline of the investment and cash flow for the mill if it were to implement the proposed process.

8.4.1 Cost of current SSK mill system

The SSK mill currently produces around 17,000 wet tonnes/year of rejects at 55 % moisture which is disposed of by landfill. The SSK mill also imports natural gas for use in its existing biogas boiler. The total cost for purchase of natural gas for this boiler and disposal of rejects is approximately £2,700,000 per year.

8.4.2 SSK mill Total Capital Cost (TCC)

For the proposed process at the SSK mill a relatively low working capital would be required as it is a simple process which does not require the use of catalysts or the purchase of raw materials; therefore WC is taken as 10 % of the FC cost [76].

The two major installations required by the mill are a pre-treatment plant and a gasification plant. The mill also requires auxiliary equipment to facilitate in the process such as pumps, fans, conveyors, silos, instrument control systems, valves etc. The incremental factors used to calculate the cost of these minor plant items are given in Table 39.

8.4.2.1 Pre-treatment plant cost

The pre-treatment facility is modelled around SSK's existing Roemond facility based in the Netherlands which processes approximately 30,000 tonnes per year of rejects into fuel pellets using a very similar process to that described in Section 7.3.2. The PPC cost for this plant was £5.5 Million [80]. Therefore a rapid capital cost estimating technique was used to relate the capital cost of the new facility with capacity by equation 8.

C_2 = £5,500,000 (17,000/30,000) $^{0.6}$

C2 = £3,911,643

Total Physical Plant Cost of pre-treatment facility (PPC)= £ 3,911,643

8.4.2.2 Gasification plant cost

The gasification plant cost was obtained from data available in the public literature [81] for existing downdraft gasification plants. These give the physical plant cost of a 250 kg/hr gasifer installed as £ 3125 per KW.

Therefore it is assumed :

PPC per gasifier = £3125 * 250 =£ 781,250

Total Physical Plant Cost of gasification facility (PPC)=£ 781,250 * 6 = 4,687,500

The typical PPC cost for the pre-treatment and gasification plant was therefore used as the basis in calculating the TCC as shown in Table 39 [76].

Table 39 SSK mill Total Capital Cost (TCC)

Item	Factor	£ Cost
Total Physical Plant Cost (PPC)		£ 8,599,143
	1*PPC	£ 8,599,143
Design and Engineering	0.2*PPC	£ 1,719,829
Contractors Fees	0.05*PPC	£ 429,957
Contingency	0.1*PPC	£ 859,914
Total	1.35	£ 11,608,843
Fixed Capital Cost (FC)	1.35*PPC	**£ 11,608,843**
Working Capital Cost (WC)	0.1*FC	**£ 1,160,884**
TOTAL CAPITAL COST (TCC)		**£ 12,769,727**

8.4.3 SSK mill Operating Costs (OC)

The operating costs are given on an annual basis

8.4.3.1 Raw materials

The main raw material to be purchased by the process is 4,250 tonnes per year wood chips at an estimated cost of £60 per wet tonne [82]. The total cost of raw materials is therefore estimated to be £255,000 per year.

8.4.3.2 Utilities (services)

Although the implementation of the new gasification system would significantly reduce the amount of natural gas required by the biogas boiler, the system would still require approximately 3.6 million Nm^3 per year of natural gas to operate. The total annual cost of utilities is calculated to be £780,000 per year.

8.4.3.3 Maintenance

An initial estimate of 1 % [76] of the FC cost is taken as the annual maintenance costs which is shared between the cost of labour and materials required. Total annual maintenance cost is £116,088.

8.4.3.4 Miscellaneous materials

As a rough guide the cost of miscellaneous materials is taken as 5 % [76] of the total maintenance costs and equates to £5,804

8.4.3.5 Operating labour

The mill's current labour force is considered sufficient to run the process with no additional operators required for the day to day running of the process.

8.4.3.6 Ash disposal

A small amount of ash will be produced by the new system which would need to be disposed of. A total of £65,000 per year in ash disposal costs is estimated.

8.4.3.7 Supervision

The mill's current labour force is considered sufficient to supervise the process with no additional supervisors required.

8.4.3.8 Plant overheads

No additional plant overheads are included for the new system.

8.4.3.9 Capital charges

The investment required for building the new installation has been assumed to come from the mill's retained profits and therefore not subject to a capital charge.

8.4.3.10 Local taxes

Local taxes are assumed to be unchanged.

8.4.3.11 Insurance

The annual cost of insuring the new installation is assumed to be covered under the mill's existing insurance policy and therefore not subject to an additional charge.

8.4.3.12 Royalties and licence fees

The new process would be subject to a royalty fee paid to the process creator. An annual cost of 1 % FC is estimated. Total annual royalty cost is therefore £116,088.

8.4.4 Income from Renewable Heat Incentives RHI's

As there is no electricity produced by the process ROC's are not applicable to this case. However the SSK mill could receive an income from Renewable Het Incentives RHI's. Rejects are assumed to contain a maximum of 70 % renewable fibres which originated from wood pulp; in addition wood chips used in the process would be entitled to receive 100 % benefit. According to the Department of Energy and Climate Change, the use of advanced thermal conversion technologies such as gasification to process renewable wastes are entitled to receive up to 1p/kwh of heat which is produced, subject to an accreditation by Ofgem (UK government energy regulator) [79]. The calculation is based on the proportion of heat which is generated from the renewable fraction of the feedstock. An estimation of the amount of RHI's the mill would receive from the new process is as follows.

It is estimated that on a dry basis a maximum of 85 wt% of the total feedstock originates from renewable wood pulp fibres.

Assuming 8000 operating hours, total heat produced by the new system = 7.4 MWth = 59,200 MWh

Proportion of renewable heat produced by the SSK mill = 59,200 MWh * 0.85 = 50,320 MWh

Current price per RHI = 1 p/kWh

Total annual revenue from RHI’s = 0.01 * 50,320,000 = £ 503,200

8.4.5 Savings

There are two forms of savings which will be made by the SSK mill from implementing the new process. The first is a 50 % saving in the amount of natural gas which is currently used in the biogas boiler and the second is a saving on reject disposal costs. A breakdown of the total saving made by the new system in comparison to the old system is presented in Table 40.

Table 40 A breakdown of total savings made by the proposed SSK mill process

	New System	**Current System**
Variable Costs (per annum)		
Raw Materials	-£255,000.00	
Utilities	-£780,000.00	confidential
Maintenance	-£116,088.00	
Miscellaneous Materials	-£5,804.00	
Fixed Costs (per annum)		
Reject Disposal		confidential
Ash Disposal Costs	-£65,000.00	confidential
Royalties and Licence Fees	-£116,088.00	
Total	-£1,337,980.00	confidential
Income		
RHIs	£503,200.00	
Total Old Expense		-£2,674,000.00
Total New Expense	-£834,780	
Net Saving Per Year	**£1,839,220**	

8.4.6 Payback time

The Payback Time is calculated as the total time take to recover TCC, from, from launch of the plant. The time for commissioning is not included in the calculation.

£ 12,769,727 / £ 1,839,220 = 7 years

8.4.7 SSK Mill economic evaluation summary

It is estimated that a total capital investment of approximately £ 13 million is required to integrate a new reject pre-treatment and gasification system with the SSK mill's current biogas boiler system. The new system is designed to process a total of 17,000 wet tonnes per year of rejects. After all expenses the SSK mill would make a net saving of approximately £ 1.8 million per year by implementing the new system. A payback time of 7 years is therefore estimated to recover the total capital investment of the project.

8.5 KC mill economic evaluation

The new KC mill plant described in Section 7.4.2 will process a total of 40,000 wet tonnes per year of de-inking sludge and will aim to integrate a drying and pyrolysis reactor with a combustor, the heat energy created will fuel the existing boiler. The plant will utilise the full amount of de-inking sludge produced at the mill, thereby reducing its current natural gas usage by 53 %.

8.5.1 Cost of current KC mill system

The KC mill currently produces around 40,000 tonnes/year of de-inking sludge at 45 wt% moisture which is mainly disposed of by land-spreading. The mill also imports natural gas to fuel the existing boiler. The total cost for purchase of natural gas and disposal of sludge is approximately £1,600,000 per year.

8.5.2 KC mill Total Capital Cost (TCC)

For the proposed process at the KC mill a relatively low working capital would be required as it is a simple process which does not require the use of catalysts or the purchase of raw materials; therefore WC is taken as 10 % of the FC cost [76].

The three major plant items required by the mill are a dryer, pyrolysis reactor and fluidised bed combustor. The mill also requires auxiliary equipment to facilitate in the process such as pumps, fans, conveyors, silos, instrument control systems, valves etc. The incremental factors used to calculate the cost of these minor plant items are given in Table 41.

8.5.2.1 Dryer cost

From the general enquiries a UK based supplier TCR was able to give a quote for a rotary drum dryer which met the required specification. Their dryer would utilise combustion exhaust gases to directly dry a maximum capacity of 5 tonnes per hour of sludge from its initial moisture content down to < 5 wt%. This capacity satisfies the demand of both the AN mill as well as the KC mill. The dryer is suitable for de-inking sludge applications, and is sufficient to evaporate approximately 1 tonne per hour of water vapour with an exhaust gas exit temperature of approximately 100 ^{o}C. The specification and quote for the dryer as provided by TCR is given in Section 8.3.2.1:

This was taken to be a representative of a typical PCE for the rotary drum dryer and was therefore used as the bases in calculating the TCC as shown in Table 41

8.5.2.2 Pyrolysis reactor cost

The “Pyroformer” pyrolysis reactor unit would be delivered new to the mill and would be manufactured by WGD based in Germany. Although it was not possible to obtain an accurate quote for a 3 tonne per hour Pyroformer, some cost information was available for a 500 kg per hour unit [77], and a rapid capital cost estimating technique was therefore used to relate the capital cost of the reactor with capacity by equation 8.

C_2 = £400,000 $(3/0.5)^{0.6}$

C2 = £ 1,180,000

PCE pyrolysis reactor = £ 1,180,000

8.5.2.3 Fluidised bed combustor

A fluidised bed combustor is required to further reduce the sludge weight after the pyrolysis reactor. It is indented that the pyrolysis vapours are used to calcine the $CaCO_3$ solid residues releasing further thermal energy which would be available to the boiler. A bubbling fluidised bed combustor was chosen as it is a robust unit which is very suitable for processing sludge materials and is used widely in the paper industry to combust sludge.

Although it was not possible to obtain an accurate quote for a 2.2 tonne per hour combustor some cost information was available for a 4.2 tonne per hour unit [83]; the estimating technique given by equation 8 was then used to obtain the capital cost at the smaller scale. The capital cost of a 4.2 tonne per hour unit delivered as new was taken to be approximately £1,250,000

Therefore

C_2 = £1,250,000 $(2.2/4.2)^{0.6}$

C2 = £ 848,035

PCE Combustor = £ 848,035

The PCE costs for the dryer, pyrolysis reactor and combustor obtained above were used as the basis in calculating the TCC as shown in Table 41.

Table 41 SSK mill Total Capital Cost (TCC)

Item	Factor	£ Cost
Equipment Total Purchase Cost (PCE)		£ 2,138,035.00
F0	1*PCE	£ 2,138,035.00
F1 Equipment Erection	0.5*PCE	£ 1,069,017.00
F2 Piping	0.2*PCE	£ 427,607.00
F3 Instrumentation	0.1*PCE	£ 213,803.00
F4 Electrical	0.1*PCE	£ 213,803.00
F5 Building, Process	0.05*PCE	£ 106,902.00
F6 Utilities	0.25*PCE	£ 534,509.00
F7 Storage	0.25*PCE	£ 534,509.00
F8 Site Development	0.05*PCE	£ 106,902.00
F9 Ancillary Buildings	0.3*PCE	£ 641,410.00
Total	**2.8**	£ 5,986,497.00
Total Physical Plant Cost (PPC)	2.8* PCE	£ 5,986,497.00
	1*PPC	£ 5,986,497.00
Design and Engineering	0.2*PPC	£ 1,197,299.00
Contractors Fees	0.05*PPC	£ 299,325.00
Contingency	0.1*PPC	£ 598,650.00
Total	1.35	£ 8,081,771.00
Fixed Capital Cost (FC)	1.35*PPC	**£ 8,081,771.00**
Working Capital Cost (WC)	0.1*FC	**£ 808,177.00**
TOTAL CAPITAL COST (TCC)		**£ 8,889,948.00**

8.5.3 KC mill Operating Costs (OC)

The operating costs are given on an annual basis.

8.5.3.1 Raw materials

As the main raw material used in the process is de-inking sludge there is no associated cost.

8.5.3.2 Utilities (services)

A typical value of 9 p/Nm^3 of nitrogen is taken for this expense [78], it is estimated that a total amount of 16,000 Nm^3 of nitrogen per year will be used to purge the system equating to a total annual nitrogen cost of £1,440.

Although the implementation of the new pyrolysis system would significantly reduce the amount of natural gas required by the Coleshill boiler, the system would still require approximately 2.6 million Nm^3 per year of natural gas to operate. The total annual cost of utilities is calculated to be £563,040 per year.

8.5.3.3 Maintenance

An initial estimate of 1 % [76] of the FC cost is taken as the annual maintenance costs which is shared between the cost of labour and materials required. Total annual maintenance cost is £80,818.

8.5.3.4 Miscellaneous materials

As a rough guide the cost of miscellaneous materials is taken as 5 % [76] of the total maintenance costs and equates to £4,041

8.5.3.5 Operating labour

The mill’s current labour force is considered sufficient to run the process with no additional operators required for the day to day running of the process.

8.5.3.6 Ash disposal

A throughput of ash will be produced by the combustor. The mill may be able to offload a certain amount to the cement industry as a concrete additive. This study however has assumed that none of the ash produced by the new system will be sold and therefore would need to be disposed of. A total of £106,560 per year in ash disposal costs is estimated.

8.5.3.7 Supervision

The mill's current labour force is considered sufficient to supervise the process with no additional supervisors required.

8.5.3.8 Plant overheads

All these costs are assumed to be covered under the mill's current expenditure therefore no additional plant overheads are included for the new system.

8.5.3.9 Capital charges

The investment required for building the new installation has been assumed to come from the mill's retained profits and therefore not subject to a capital charge.

8.5.3.10 Local taxes

Local taxes are assumed to be unchanged.

8.5.3.11 Insurance

The annual cost of insuring the new installation is assumed to be covered under the mill's existing insurance policy and therefore not subject to an additional charge.

8.5.3.12 Royalties and licence fees

The new process would be subject to a royalty fee paid to the process creator. An annual cost of 1 % FC is estimated. Total annual royalty cost is therefore £80,818.

8.5.4 Income from Renewable Heat Incentives (RHI's)

The KC mill will receive an income by incorporating the new pyrolysis process. De-inking sludge is assumed to contain a maximum of 30 % renewable fibres which originated from wood pulp; the new pyrolysis process would therefore be entitled to receive a revenue generated from Renewable Heat

Incentives (RHI's). According to the Department of Energy and Climate Change, the use of advanced thermal conversion technologies such as pyrolysis to process wastes in this way are entitled to receive a tariff of up to 1p/kWh of heat which is produced, subject to an accreditation by Ofgem (UK government energy regulator) [79]. The calculation is based on the proportion of heat which is generated from the renewable fraction of the feedstock. An estimation of the amount of RHI's the mill would receive from the new process is as follows.

It is estimated that on a dry basis a maximum of 30 wt% of de-inking sludge originates from renewable wood pulp fibres, approximately 50 wt% is composed of inorganic ash and the remaining 20 wt% is composed of other non-renewable components.

Assuming 8000 operating hours, total heat produced by the new process = 48,640 MWh

Proportion of renewable heat produced by the KC mill = 48,640 MWh * 0.3 = 14,592 MWh

Current price per kw/h = 1p

Total annual revenue from RHI's = £0.01 * 14,592,000 = £ 145,920

8.5.5 Savings

There are two forms of savings which will be made by the KC mill from implementing the new process. The first is a 53 % saving in the amount of natural gas which is currently used and the second is a saving on sludge disposal costs. A breakdown of the total saving made by the new system in comparison to the old system is presented in Table 42.

Table 42 A breakdown of total savings made by the proposed KC mill process

	New System	**Current System**
Variable Costs (per annum)		
Utilities	-£563,040.00	confidential
Maintenance	-£80,818.00	
Miscellaneous Materials	-£4,041.00	
Fixed Costs (per annum)		
Sludge Disposal	n/a	confidential
Ash Disposal Costs	-£106,560.00	-
Royalties and Licence Fees	-£80,818.00	
Total	-£835,277.00	confidential
Income		
RHIs	£145,920.00	
Total Old Expense		-£1,600,000.00
Total New Expense	-£689,357.00	
Net Saving Per Year	**£910,643.00**	

8.5.6 Payback time

The Payback Time is calculated as the total time take to recover TCC, from launch of the plant. The time for commissioning is not included in the calculation. .

£8,889,948 / £910,643 = 9.8 years

8.5.7 KC mill economic evaluation summary

It is estimated that a total capital investment of approximately £8.8 million is required to integrate a new drying, pyrolysis and combustion system with the KC mill's current Coleshill boiler system. The new system is designed to process a total of 40,000 wet tonnes per year of de-inking sludge. After all expenses the KC mill would make a net annual saving of approximately £910,000 per year by implementing the new system. A payback time of 9.8 years is therefore estimated to recover the total capital investment of the project.

8.6 Overall economic evaluation discussion

The economic analysis of each process presented in this chapter is based only on best estimates that were made for the investment required and cash flows. The actual cash flow would be affected by any changes made to the operating costs. For example in the case of the SSK mill the price of raw materials (wood chips) may increase and in the case of the AN and KC mills, ash disposal costs may increase. Similarly income gained from each project will also differ if for instance each mill produces less wastes or renewable incentives legislated by government changes.

For capital cost estimates the degree of change would be directly related to changes in plant design, optimisation of the process may reduce costs by reducing the number of plant items required, likewise additional plant items may be required if the configuration of the design is changed.

Overall there are several different methods which can be used to determine the viability of a process. The viability of each process in this work was assessed by calculating the payback time for the project. This gives an indication of the measure of time to recover the total capital investment. Usually most companies would consider a project to be feasible if it can achieve a payback time within 2-5 years of operating [76]. Therefore on the basis of this, the economic evaluation concludes that the de-inking sludge process proposed for the AN mill would be feasible with a calculated payback time of 4.5 years, however the gasification process proposed for the SSK mill (payback time 7 years) and de-inking sludge process proposed for the KC mill (payback time 9.8 years) would not. Process optimisation is recommended for SSK and KC if they seek to reduce the payback time further.

9.0 CONCLUSION

The two main wastes generated from secondary fibre paper mills are rejects (composed mainly of plastics and fibres) and de-inking sludge, both of which are evolved from the pulping process during paper manufacture. The current practice for the disposal of these wastes is either by land-spreading, incineration (using natural gas) or landfilling, each of which presents environmental problems. Furthermore as the cost for disposing these wastes increases year upon year many secondary fibre paper mills are finding it increasingly difficult to remain profitable.

This work has investigated both the fixed bed downdraft gasification and intermediate pyrolysis of rejects and de-inking sludge derived from three secondary fibre paper mills, Aylesford Newsprint (AN) newsprint mill, Smurfit Kappa (SSK) board mill, and Kimberly Clark (KC) tissue mill. Overall the main objectives of the study have been satisfied by the following points:

- Each waste stream acquired (rejects and de-inking sludge) were successfully pre-conditioned and characterised before thermal conversion. Pre-treatment required the drying, sorting and pelletisation of each feedstock before thermal conversion. Each feedstock was subsequently characterised to determine the water content, proximate and ultimate compositions and heating value.

- Through trials the energy through both fixed bed downdraft gasification and intermediate pyrolysis of the selected wastes and waste blends were quantified. The gasification trials used a co-fired blend of wood chips, de-inking sludge and rejects as the feedstock. Multiple gasification trials were performed, and the results showed that the most promising trials were those carried out using reject waste pellets produced from the SSK mill, where as much as 80 wt % of the rejected pellets could be successfully co-fired with wood chips. The product gas composition was 16.24 % H_2, 23.34 % CO, 12.71 % CO_2 5.21 % CH_4 and 42.49 % N_2 (v/v %) with a heating value of 7.3 MJ/Nm^3 After the removal of tar and water condensate the product gas was considered to be of sufficient quality to power a gas engine. The limiting factor for other feedstocks and blends was the agglomeration of plastics present within the feed causing blockage in the gasifier.

The intermediate pyrolysis trial carried out by this work utilised de-inking sludge from the AN and KC mill as a feedstock in a newly patented pyroformer reactor. Among the products formed was a pyrolysis oil which contained a gross heating value of 36-37 MJ/kg and was composed of mainly phenolic compounds and fatty acid methyl esters. The oil produced was found to have improved fuel physical properties when compared to traditional pyrolysis oil produced from the fast pyrolysis of biomass, and was comparable to biodiesel. Permanent gases were also produced with a composition of approximately 24% CO, 6% CH4 and 70% CO2 (v/v%). After pyrolysis the solid char residue also contained a small residual calorific value, and was largely composed of mainly calcium based inert metal oxides.

The pyrolysis oil produced from this work was found to be directly miscible with biodiesel when blended. Blended ratios of 20 % with 80 % (v/v) and 30 % (v/v) with 70 % (v/v) pyrolysis oil to biodiesel respectively were prepared by directly mixing and filtering the two fuels. The blended fuels were subsequently tested in an IDI Lister Petter diesel engine to compare the performance, combustion and emission characteristics with fossil diesel and biodiesel.

Full engine power was achieved with both blends, and very little difference in engine performance and emission results were observed between 20 % and 30 % blends. At full engine load, the brake specific fuel consumption on a volume basis was around 6 % higher for the blends when compared to fossil diesel. The brake thermal efficiencies were about 3–6 % lower than biodiesel and were similar to fossil diesel. Exhaust gas emissions of the blends contained 4 % higher CO_2 and 6–12 % lower NOx, as compared to fossil diesel. At full load, CO emissions of the blends were decreased by 5–10 times. The cylinder gas pressure diagram showed stable engine operation with the 20 % blend, but indicated minor knocking with 30 % blend. Peak cylinder pressure of the 30 % blend was about 5–6 % higher compared to fossil diesel. At full load, the peak burn rate of combustion from the 30 % blend was about 26 % and 12 % higher than fossil diesel and biodiesel respectively. In comparison to fossil diesel the combustion duration was decreased for both blends; for 30 % blend at full load, the duration was almost 12 % lower. The study concluded that up to 20 % blend of de-inking sludge pyrolysis oil with biodiesel can be used in an indirect injection CI engine without adding any ignition additives or surfactants.

- It was concluded that the optimal application for gasification technology is at paper mills which manufacture brown paper for the corrugated board industry, using their rejects stream. Some importing of wood chips as a co-fired fuel may be necessary, although it may be possible to eliminate this by pre-sorting the rejects to remove some of the plastics content. Other rejects tested from a newsprint and tissue mill were not suitable for gasification largely because of the agglomeration of hard plastics present.

  The optimal application for intermediate pyrolysis was found to be the processing of de-inking sludge in the Pyroformer. The main advantage of this is that the unit can tolerate feedstocks with high ash contents and low calorific values thereby removing the need for paper mills to import large quantities of natural gas to co-fire sludge in a combustion unit.

- The industrial implementation study presented in Chapter 7 further investigated how the proposed downdraft gasification system could be integrated on-site at the SSK mill. The optimal process route was found to be 6 downdraft gasifiers in parallel which would co-fire the total amount of rejects produced on-site with wood chips. This would yield enough product gas to fuel an existing on-site boiler, thereby offsetting approximately 50 % of the natural gas which is currently used. The environmental and economic benefits of implementing such a system would be; a 50 % reduction in natural gas usage, approximately 90 % reduction in rejects sent to landfill and a reduction of CO_2 emissions as approximately

70 wt% of the SSK rejects come from a carbon neutral source of biomass (providing the plant is re grown sustainably).

At full scale the industrial implementation studies presented in Chapter 7 revealed that the optimal process route to pyrolyse de-inking sludge at the AN and KC mills would be to co-fire the pyrolysis solid residues with the pyrolysis vapours in a combustion unit downstream. The ash product formed could then be sold to cement industries as cement or concrete admixture. Some of the hot exhaust gases from the combustion unit would be required to run the drying and pyrolysis process itself with the majority of exhaust gases being routed to either a gas turbine or boiler for electricity generation. Processing de-inking sludge in this way would significantly reduce the natural gas requirement of the AN and KC mills by as much as 67 %, as well as significantly reducing the amount of de-inking sludge sent to landfill. As the process utilises advanced thermal conversion methods it would also be entitled to receive benefits by way of ROC's for the AN mill and RHI's for the KC mill. The pyrolysis of de-inking sludge is overall a sustainable solution which would significantly reduce this waste fraction whilst simultaneously generating useful CHP energy at the paper making sites.

- An economic evaluation of each proposed processes was carried out in Chapter 8, to determine the total plant capital costs, operating costs and payback time if each mill were to invest in the proposed processes. The total payback time for the SSK mill was estimated to be 7 years, the AN mill would see a payback time of 4.5 years and 9.8 years would be the return for the KC mill.

Overall this project has laid the foundation for the future development of gasification and pyrolysis technologies at secondary fibre paper mills. It is hoped that as a consequence of this research each paper mill is now in a position to better understand the potential of each of their waste streams (rejects and de-inking sludge) for use as a sustainable on-site fuel in a CHP system either by gasification or pyrolysis. Furthermore it is hoped that each mill is now in a better position to make an accurate assessment on whether the environmental and economic gain from each of the proposed technologies outweigh the necessary capital investment required to implement each process.

10.0 FURTHER WORK

The following work is recommended to take this project further:

- To expand on research investigating the intermediate pyrolysis of de-inking sludge for CHP generation. This research should mainly focus on the up-scaling of the pyrolysis process outlined by this work. A range of operating conditions, reaction temperatures, solid residence times and screw configurations should be explored. Detailed analysis of results and process optimisation should be developed, including optimisation of the reactor design (either single screw one pass design or co-axial design), and recirculation of waste heat from exhaust gases for reuse in the process as a drying medium and as heating medium for the reactors.

- To expand on research investigating the gasification of rejects for CHP generation at secondary fibre mills. This research should mainly focus around the use of fluidised bed gasifiers to process rejects because of the reduced vulnerability to agglomeration blockages; the work should include the coupling of the gasifier to a CHP engine for the testing and analysis of this feedstock. Detailed analysis of results and process optimisation should be further developed. The research should aim to investigate ways of optimising reactor design with recirculation of waste heat from exhaust gases for reuse in the process as a drying medium.

- To test de-inking sludge ash for CO_2 capture and to investigate its use as a tar cracking catalyst. From this work it is perceived that de-inking sludge may share similar catalytic properties to that of dolomite for use as a tar crack medium in fluidised bed gasifiers. It is proposed that this research further investigates this. The research should aim to characterise and test the solid de-inking sludge residues produced from both pyrolysis and de-inking sludge combustion as a tar cracking catalyst and heat transfer medium in a fluidised bed gasifier. Tar levels should be quantified, and compared to both sand and dolomite. Similarly CO_2 levels in product gases should be quantified and compared to both sand and dolomite. Reducing CO_2 in the product gases would serve to increase gas calorific value whilst simultaneously capturing the CO_2 in a solid form as $CaCO_3$ in the ash. This will reduce CO_2 emissions to atmosphere. Product gas quality will also be improved if the effect of de-inking sludge residues can serve to decrease tar formation in the gasifier.

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APPENDIX A1

(AN mill current process energy balance)

This energy balance has been derived for the AN mill's current sludge combustion system explained in Section 7.2.1 and presented by Figure 66. All calculations are based on data acquired during discussions with the mill.

Sludge combustor

Operating temperature = 900 °C, ambient temperature = 20 °C

Input = 350 tonnes/day of sludge at 33% moisture = 4.05 kg/s

Higher Heating Value (HHV) of sludge (dry basis) as measured by Aston University = 6.4 MJ/kg

HHV of wet sludge at 33% moisture = 0.67 * 6.4 = 4.29 MJ/kg

Chemical energy of wet sludge into combustor = 4.05 *4.29 = **17.38 MW**

Approximately 12,000 Nm^3/ day of natural gas is required to combust sludge = 0.139 Nm^3/s

HHV of natural gas = 39.0 MJ/Nm^3 [74]

Chemical energy into combustor from natural gas = 0.139 * 39.0 = **5.42 MW**

Air enters combustor at 11.0 Nm^3/s

Mass flow rate of air entering combustor = 11.0 * 1.29 kg/m^3 = 14.2 kg/s

Sensible heat input to air from primary heat exchanger = 14.2 * 1.01 * (80-20) = **0.86 MW**

Sensible heat input to air from secondary heat exchanger = 14.2 * 1.01 * (150-80) = **1.00 MW**

Sensible heat of air entering the combustor = 0.86 + 1.00 MW = **1.86 MW**

Ash content of sludge (dry basis) as measured by Aston University = 52.1 wt%

Total ash leaving combustor = 4.05 * 0.67 * 0.521 = 1.40 kg/s

Major component of ash (over 60 wt%) is CaO, MW = 56.08 g/mol

Cp of CaO = 10.00 + 0.00484T – 108000/T2 = cal/deg mol [74]

T = (900 + 20)/2 + 273 = 733 K

$10.00 + (0.00484 * 733) – (108000 / 733^2) =13.35$ cal/deg mol

Conversion: 13.35 * 4.187 =55.90 J/kmol, 55.90 / 56.08 g/mol = 1.00 kJ/kg K

Sensible heat of ash leaving combustor = 1.40 * 1.00 * (900-20) = **1.23 MW**

None of the sensible heat energy from the hot ash is recovered.

Total energy out of combustor from exhaust gases = 17.38 + 5.42 + 1.86 – 1.23 = **23.43 MW**

Boiler section

The mill uses the heat of the exhaust gases from the combustor to raise 23 tonnes/hr (6.39 kg/s) of steam in a boiler at 420 °C and 60 bar(g)

Feed water enters economiser at 80 bar(g) pressure and 130 °C. Specific enthalpy = 0.552 MJ/kg.

Energy of water fed into economiser = 6.39 * 0.552 = **3.53 MW**

Water leaves the economiser at 281 °C and 80 bar(g) pressure and enters a steam drum from which the water passes to the evaporator tubes. The steam then passes back to the drum and on to the superheater. Specific enthalpy of superheated steam at 420 °C, 60 bar(g) = 3.226 MJ/kg

Energy of superheated steam = 6.38 *3.226 = **20.58 MW**

Total energy extracted from exhaust gases = 20.58 – 3.53 = **17.05 MW**

Energy lost from exhaust gases to stack = 23.43 - 17.05 - 1.00 = **5.38 MW** or approximately 22 % of the total energy input.

APPENDIX A2

(AN mill proposed process energy balance)

In total the AN mill produces approximately 150,000 wet tonnes per year of de-inking sludge at 33 wt% moisture. Approximately 350 tonnes per operating day is sent to the sludge combustor, for 8000 operating hours per year (116,667 wet tonnes/year). The remaining 33,333 wet tonnes per year are sent for either land-spreading or cattle bedding.

The following calculations are based on the proposed pyrolysis system explained in Section 7.2.2 and illustrated by Figure 67. And is based on the mill continuing to supply the sludge combustor with 116,667 wet tonnes per year of sludge. However the remaining 33,333 wet tonnes per year will now be processed by pyrolysis, with the residual pyrolysis solids also sent to the combustor.

Energy required for drying

To dry the wet sludge from 33 wt% moisture to the 3 wt% moisture content required for pyrolysis, rotary drum dryers will be used.

33,333 t/y at 33 % moisture = 4.167 t/hr of wet sludge to dryer, comprising 2.792 t/hr of dry solids and 1.375 t/hr of water.

After drying to 3% moisture, there are 2.792 t/hr of dry solids and 0.086 t/hr of water.

Hence water to be evaporated = 1.375 - 0.086 = 1.289 t/hr = 0.358 kg/s, remaining water = 0.024 kg/s, and solids = 0.776 kg/s.

Temperature at exit of rotary dryer = 100 °C

Energy required in the dryer is that to evaporate the water (2.592 MJ/kg) plus that to raise the remaining water to 100 °C (0.335 MJ/kg) plus that to heat the solids to 100 °C (assuming a Cp value of 1.2 kJ/kgK).

Total energy = (0.358 * 2.592) + (0.024 * 0.335) + (0.776 * 0.0012 * 80) = **1.01 MW**

Energy balance around pyrolysis unit

Sludge Input

Once sludge has been dried to 3 wt% moisture solid it enters the pyrolysis unit operating at 450 ^{o}C, at a total flow rate of 0.024 + 0.776 = 0.800 kg/s

3 wt% moisture content sludge has a HHV of 6.21 MJ/kg

Chemical energy into pyrolysis unit = 0.800 * 6.21 = **4.97 MW**

Condensable Organic Vapours (Volatiles)

0.800 kg/s of sludge yields 9.3 wt% or 0.0744 kg/s of condensable organic vapours and 0.9 wt% water = 37 MJ/kg (see section 6.4.1)

Chemical energy in volatiles = 0.0744 *37 = **2.75 MW**

Sensible heat of volatiles $Q= \Sigma mCp\Delta T$

Sensible heat of major volatile components at 508 K:

Toluene = 1.887 kJ/kg K

Ethyl benzene = 1.976 kJ/kg K

Phenol = 1.742 kJ/kg K

Average = 1.87 kJ/kg K = 0.00187 MJ/kg K

Sensible heat of volatiles =0.0744 * 0.00187 * (450-20) = **0.060 MW**

Total energy of condensable volatiles = 0.060 + 2.75 = 2.81 MW

Total water vapour produced = 0.009 * 0.800 =0.0072 kg/s

Sensible heat of water vapour at Cp = 1.98 kJ/kg K = 0.0072 * 0.00198 *430 = 6.2 KW = **0.006 MW**

Char (Solid Residues)

79 wt% of the feed entering the pyrolysis unit remains as solid char residue.

Amount of solid residue leaving pyrolysis unit = 0.79 * 0.800 = 0.632 kg/s

Composition of char = 17.12% C, 1.13% H, 0.26% N, < 0.1% S, 21.39% O, 60% ash

HHV of char as measured by Aston University = 3.26 MJ/kg

Chemical energy in char = 0.632 * 3.26 = **2.06 MW**

Major component of char (over 60 wt%) =$CaCO_3$

Cp of $CaCO_3$ = 19.68 + 0.01189T - 307600/T^2 = cal/deg mol [74]

T = 508 K

19.68 + (0.01189 * 508) – (307600 / 508^2) =24.53 cal /deg mol

Conversion: 24.53 * 4.187 =102.69 J/kmol, 102.69 /100 g/mol = 1.02 kJ/kg K

Sensible heat in char = 0.632 * 1.02 * (450-20) = **0.28 MW**

Total energy from char = 0.28 + 2.06 = **2.34 MW**

Permanent Gas

Amount of gas produced from pyrolysis unit is obtained by difference = 10.8 wt%

Amount of gas leaving pyrolysis unit = 0.108 * 0.800 = 0.0864 kg/s

Composition of gas discounting N_2 from purge which would be negligible on a full scale unit =22.69%

CO, 6.1% CH_4, 71.2% CO_2

HHV of CH_4 = 41.2 MJ/Nm^3

HHV of CO = 13.1 MJ/Nm^3

Total HHV of gas = (41.2 MJ/M^3 * 0.061 CH_4) + (13.1 MJ/M^3 * 0.2269 CO) = 5.48 MJ/Nm^3

Average density of gas at STP = 1.61 kg/Nm^3

Total volume of gas produced = 0.0864 / 1.61 = 0.0537 Nm^3/s

Chemical energy in gas = 0.0537 * 5.48 = **0.29 MW**

Sensible heat of permanent gases:

Gas densities at 450°C

CO = 0.47 kg/m^3, mass in 1 m^3 =0.227 * 0.47 =0.108 kg

CH4 = 0.27 kg/m^3, mass in 1 m^3 = 0.061 *0.27 = 0.0162 kg

CO2 = 0.47 kg/m^3, mass in 1 m^3 = 0.712* 0.47 = 0.334 kg

Total mass in 1 m^3 = 0.4582 Kg/m^3

Mass Fractions: CO =23.5%, CH4 = 3.5%, CO2 = 73%

Average Cp Values

CO = 1.064 kJ/kg K, CH_4= 0.74 kJ/kg K, CO_2 = 1.014 kJ/kg K

Average Cp = (1.064 * 0.235) + (0.74 * 0.035) + (1.014 * 0.73) = 1.016 kJ/kg. K

Sensible energy of gas = 0.0864 * 0.00102 * (450-20) = **0.038 MW**

Total energy of gas = 0.038 + 0.29 = 0.33 MW

Energy required for heating pyrolysis unit

Total indirect heat input required for the pyrolysis unit (as determined by Aston University) is approximately 10% of chemical energy of the feed **=** 0.10 *4.97 = **0.50 MW**

Energy balance around Combustor

Solids produced from the pyrolysis unit will be added to 350 t/day of sludge to the combustor.

Natural gas will be displaced by the vapours and gases produced from the pyrolysis unit.

Char from pyrolysis unit to combustor = **2.34 MW**

Vapours/gases from pyrolysis unit to combustor = 2.81 MW + 0.33 MW + 0.0062 = **3.14 MW**

Chemical energy of wet sludge into combustor (from earlier) = **17.38 MW**

Sensible heat of ash out of the combustor

Total ash out (60% of pyrolysis solids is ash) = (0.632 * 0.6) + 1.40 = 1.78 kg/s

Sensible heat of solid residue (major component over 60 wt%) =CaO

Cp of CaO = $10.00 + 0.00484T - 108000/T^2$ = cal/deg mol [74]

T = 733 K

$10.00 + (0.00484 * 733) – (108000 / 733^2)$ =13.35 cal/deg mol

Conversion: 13.35 * 4.187 = 55.90 J/kmol, 55.90 / 56.08 g/mol = 1.00 kJ/kg K

Sensible heat from solid residue leaving the reactor = 1.78 * 0.00100 * (900-20) = **1.57 MW**

None of the sensible heat energy from the solid residue is recovered back into the system

Assuming air requirement is unchanged (see below), sensible heat of air entering the combustor = **1.86 MW**

Energy Out of Combustor from Exhaust Gases = 17.38 + 3.14 + 2.34 + 1.86 – 1.57 = **23.15 MW**

System energy requirement for steam raising (assumed unchanged) = **23.43 MW**

To which must be added:

Energy for drying = **1.01 MW**

Energy for pyrolysis = **0.50 MW**

Total system energy requirement = 23.43 + 1.01 + 0.50 = 24.94 MW

System is energy deficient by = 24.94 - 23.15 = 1.79 MW

APPENDIX B1

(SSK mill current process energy balance)

This energy balance has been derived for the SSK mill's current on-site energy system explained in Section 7.3.1 and presented in Figure 68 and all calculations are based on data acquired during discussions with the mill.

Total natural gas imported

Total natural gas imported = 388 GWh/y = **48.5 MW**

HHV of natural gas = 39 MJ/Nm^3 [74]

Natural gas flow rate = 48.5 / 39 = 1.24 Nm^3/s = 4477 Nm^3/h

Natural gas turbines including boilers 2 and 3 (CHP)

Natural gas to CHP= 307 GWh/y = **38.37 MW**

Natural gas flow rate to CHP= 38.37 / 39 = 0.984 Nm^3/s = 3542.3 Nm^3/h

Electricity from gas turbines = 62 GWhe/y = **7.75 MWe**

Imported electricity from grid = 7 GWhe/y = **0.87 MWe**

Exported electricity to grid = 1.6 GWhe/y = **0.20 MWe**

Net electrical energy to mill = 62 + 7 -1.6 = 67.4 GWhe/y = **8.42 MWe**

Heat energy tot boilers 2 and 3= 307 - 62 = 245 GWh/y = **30.62 MW**

Water condensate returned to boilers 2 and 3= 192,899 t/y = 6.70 kg/s

Specific enthalpy of water condensate = 0.51 MJ/kg

Energy of water condensate to boilers 2 and 3= 6.70 * 0.51 = 27.4 GWh/y = **3.42 MW**

Flow rate of steam from boilers 2 and 3 = 247,307 t/year = 30.9 t/h = 8.59 kg/s

Specific enthalpy of steam at 15 bar and 200 ^{O}C = 2.788 MJ/kg

Total energy of steam from boilers 2 and 3= 8.59 * 2.788 = 191.5 GWh/y = **23.94 MW**

Waste heat by difference = 30.62 MW + 3.42 MW – 23.94 MW = **10.10 MW**

Biogas boiler

Natural gas to boiler = 81 GWh/y = **10.12 MW**

Natural gas flow rate to boiler = 10.12 / 39 = 0.259 Nm^3/s = 932.4 Nm^3/h

Flow rate of gas oil to boiler = 25.7 t/y = 8.9×10^{-4} kg/s

HHV of gas oil = 43.22 MJ/kg

Total energy of gas oil to boiler = 43.22 * 8.9×10^{-4} = 0.30 GWh/y = **0.04 MW**

Composition of biogas 70 vol% CH_4 30 vol% CO_2

HHV of biogas = 26 MJ/Nm^3

Flow rate of biogas = 277 Nm^3/h = 0.077 Nm^3/s

Chemical energy from biogas = 0.077 * 26 = 16 GWh/y = **2.00 MW**

Water condensate returned to boiler = 98,769 t/y = 3.43 kg/s

Specific enthalpy of water condensate = 0.61 MJ/kg

Energy of water condensate to boiler = 3.43 * 0.61 = =17 GWh/y = **2.12 MW**

Flow rate of steam from boiler = 126,627 t/y = 15.83 t/h = 4.40 kg/s

Specific enthalpy of steam at 15 bar gauge and 200 °C = 2.788 MJ/kg

Total energy of steam from boiler = 4.40 * 2.788 = 98.1 GWh/y = **12.26 MW**

Waste heat by difference = 10.12 + 0.04 + 2.00 MW + 2.12 MW – 12.26 MW = **2.02 MW**

APPENDIX B2

(SSK mill proposed process energy balance)

The following energy balance are based on the proposed gasification system explained in Section 7.3.2 and illustrated by Figures 69-71. It has been created according to the following design specifications:

Total Plant Capacity = 21,250 wet tonnes per year comprising of 17,000 wet tonnes per year of rejects (13,000 wet tonnes per year of general rejects and 4,000 wet tonnes per year of ragger rope rejects) and 4,250 wet tonnes per year of wood chips.

Operating hours per year = 8000

Total Plant Life = 25 years

Assumptions

All reactions reach equilibrium, therefore there is no residual solid carbon.

Gasifier walls are heavily lagged using rock wool therefore wall heat losses to the surroundings by conduction are considered negligible.

All streams entering at ambient conditions are calculated at a reference temperature of 20 °C.

All operating conditions as per laboratory scale unless stated.

Sensible heat from the system is not recovered

Energy required for drying

To dry the total 21,250 tonnes per year of wet feedstock from approximately 55 wt% moisture to the 10 wt% moisture content required for gasification, rotary drum dryers will be used.

21,250 t/y at 55 % moisture = 2.656 t/hr of wet feedstock to dryer, comprising 1.195 t/hr of dry solids and 1.461 t/hr of water.

After drying to 10% moisture, there are 1.195 t/hr of dry solids and 0.133 t/hr of water.

Hence water to be evaporated = 1.461 – 0.133 = 1.328 t/hr = 0.369 kg/s, remaining water = 0.037 kg/s, and solids = 0.332 kg/s.

Temperature at exit of rotary dryer = 100 ^{o}C

Energy required in the dryer is that to evaporate the water (2.592 MJ/kg) plus that to raise the remaining water to 100 ^{o}C (0.335 MJ/kg) plus that to heat the solids to 100 ^{o}C (assuming a Cp value of 1.2 kJ/kgK).

Total energy = (0.369 * 2.592) + (0.037 * 0.335) + (0.332 * 0.0012 * 80) = **1.00 MW**

It is assumed that the energy required for drying the feedstock will be provided by SSK's existing on-site energy systems

Energy balance around gasifier

Feed Input

Once feed has been dried to 10 wt% moisture solid it is pelletised and enters the gasifiers

at a total flow rate of 0.037 + 0.332 = 0.369 kg/s

Average HHV of feedstock = 21.4 MJ/Kg

Chemical energy of feedstock into the gasifiers = 0.369 * 21.4 = **7.90 MW**

Standard imbert type downdraft gasifiers at industrial scale are typically offered at a unit size of 250 kg/hr of feedstock or 250 kWe by Biomass Engineering Ltd; therefore it is proposed that approximately 6 gasifiers in parallel are installed to process the total capacity at the SSK site.

Chemical energy of product gas from gasifier

From lab scale energy balance calculations 1 kg of feed yields approximately 2.76 Nm^3 of product gas with an average calorific value of 7.35 MJ/Nm^3 (see Section 6.3.4.4)

Volumetric flow rate of product gas out of gasifier = 0.365 * 2.76 = 1.01 Nm^3/s

Chemical energy of product gas from gasifier = 1.01 * 7.35 = **7.4 MW**

No further heat recovery from sensible heat, water condensate or the ash will be recovered by the system and therefore are not considered in this energy balance. The net available energy from the product gas would therefore be **7.4 MWth.**

Energy required by biogas boiler

Energy required by existing biogas boiler = **12.3 MWth**

Energy provided by product gas = **7.4 MWth**

System is energy deficient by = 12.3 – 7.4 = 4.9 MW

APPENDIX C1

(KC mill current process energy balance)

This energy balance has been derived for the KC mill's current Coleshill boiler explained in Section 7.4.1 and presented by Figure 72, all calculations are based on data acquired during discussions with the mill.

Energy required for raising steam 14 bar pressure 210 °C

Flow rate of steam out of boiler = 7.5 tonnes/hr = 2.08 kg/s

Specific enthalpy of steam at 14 bar pressure 210 °C = 2.83 MJ/kg

Energy required for raising steam = 2.83 MJ/kg * 2.08 kg/s = **5.89 MW**

Boiler running with efficiency of 80 %

Total thermal energy input to boiler from natural gas therefore = **7.36 MW**

Calorific Value of Natural gas = 39 MJ/Nm^3

Flow rate of natural gas to boiler = 7.36 MW / 39 MJ/Nm^3 = 0.189 Nm^3/s = 679 Nm^3/hr

APPENDIX C2

(KC mill proposed process energy balance)

In total the KC mill produces approximately 40,000 wet tonnes per year of de-inking sludge at 45 wt% moisture. All of the de-inking sludge is disposed of.

The following calculations are based on the proposed pyrolysis system explained in Section 7.4.2 and illustrated by Figure73. And are based on the mill processing 40,000 wet tonnes per year by pyrolysis, with the residual pyrolysis solids sent to a fluidised bed combustor.

Operating hours per year = 8000

Total Plant Life = 25 years

Energy required for drying

To dry the wet sludge from 45 wt% moisture to the 3 wt% moisture content required for pyrolysis, rotary drum dryers will be used.

40,000 t/y at 45 % moisture = 5 t/hr of wet sludge to dryer, comprising 2.75 t/hr of dry solids and 2.25 t/hr of water.

After drying to 3% moisture, there are 2.75 t/hr of dry solids and 0.085 t/hr of water.

Hence water to be evaporated = 2.25 - 0.085 = 2.165 t/hr = 0.601 kg/s, remaining water = 0.024 kg/s, and solids = 0.764 kg/s.

Temperature at exit of rotary dryer = 100 ^{o}C

Energy required in the dryer is that to evaporate the water (2.592 MJ/kg) plus that to raise the remaining water to 100 ^{o}C (0.335 MJ/kg) plus that to heat the solids to 100 ^{o}C (assuming a Cp value of 1.2 kJ/kgK).

Total energy = (0.602 * 2.592) + (0.024 * 0.335) + (0.764 * 0.0012 * 80) = **1.64 MW**

Energy balance around pyrolysis unit

Sludge Input

Once sludge has been dried to 3 wt% moisture solid it enters the pyrolysis unit operating at 450 ^{o}C, at a total flow rate of 0.024 + 0.764 = 0.788 kg/s

Dry KC sludge has a HHV of 7 MJ/kg (see Section 6.2.1)

Chemical energy into pyrolysis unit = 0.788 * 7 = **5.52 MW**

Condensable Organic Vapours (Volatiles)

0.788 kg/s of sludge yields 9.3 wt% or 0.073 kg/s of condensable organic vapours with an HHV of 37 MJ/kg, and 0.9 wt% water (see Section 6.4)

Chemical energy in volatiles = 0.073 *37 = **2.71 MW**

Sensible heat of volatiles Q= ΣmCpΔT

Sensible heat of major volatile components at 508 K:

Toluene = 1.887 kJ/kg K

Ethyl benzene = 1.976 kJ/kg K

Phenol = 1.742 kJ/kg K

Average (assuming equal fractions) = 1.87 kJ/kg K = 0.00187 MJ/kg K

Sensible heat of volatiles = 0.073 * 0.00187 * (450-20) = **0.059 MW**

Total energy of condensable volatiles = 0.06 + 2.71 = **2.77 MW**

Total water vapour produced = 0.009 * 0.788 =0.0071 kg/s

Sensible heat of water vapour with Cp of 1.98 kJ/kg K = 0.0071 * 0.00198 *430 = 6.05 KW = **0.006 MW**

Char (Solid Residues)

79 wt% of the feed entering the pyrolysis unit remains as solid char residue.

Amount of solid residue leaving pyrolysis unit = 0.79 * 0.788 = 0.623 kg/s

Composition of char = 20.02% C, 2.19% H, 0.38% N, < 0.1% S, 53.5% ASH, 23.91% O

HHV of char as measured by Aston University = 4.8 MJ/kg

Chemical energy in char = 0.623 * 4.8 = **2.99 MW**

Major component of char (over 60 wt%) =$CaCO_3$

Cp of $CaCO_3$ = 19.68 + 0.01189T - 307600/T^2 = cal/deg mol [74]

T = 508 K

19.68 + (0.01189 * 508) – (307600 / 508^2) =24.53 cal /deg mol

Conversion: 24.53 * 4.187 =102.69 J/kmol, 102.69 /100 g/mol = 1.02 kJ/kg K

Sensible heat in char = 0.623 * 1.02 * (450-20) = **0.27 MW**

Total energy from char = 0.27 + 2.99 = **3.26 MW**

Permanent Gas

Amount of gas produced from pyrolysis unit is obtained by difference = 10.8 wt%

Amount of gas leaving pyrolysis unit = 0.108 * 0.788 = 0.085 kg/s

Composition of gas discounting N_2 from purge which would be negligible on a full scale unit

= 1.9% H2, 25.5 % CO, 6.3% CH4, 66.3 % CO2

HHV of CH_4 = 41.2 MJ/Nm^3

HHV of CO = 13.1 MJ/Nm^3

HHV of H_2 = 13.2 MJ/Nm^3

Total HHV of gas = (41.2 * 0.063 CH_4) + (13.1 * 0.255 CO)+ (13.2 * 0.019) = 6.19 MJ/Nm^3

Average density of gas at STP = 1.61 kg/Nm^3

Total volume of gas produced = 0.085 / 1.61 = 0.0528 Nm^3/s

Chemical energy in gas = 0.0528 * 6.19 = **0.33 MW**

Sensible heat of permanent gases:

Gas densities at 450°C

H2 = 0.082 kg/m^3, mass in 1 m^3 =0.019 * 0.082 =0.0016 kg

CO = 0.47 kg/m^3, mass in 1 m^3 =0.255 * 0.47 =0.119 kg

CH4 = 0.27 kg/m^3, mass in 1 m^3 = 0.063 *0.27 = 0.017 kg

CO2 = 0.47 kg/m^3, mass in 1 m^3 = 0.663* 0.47 = 0.311 kg

Total mass in 1 m^3 = 0.449 kg/m^3

Mass Fractions: H2 =0.36% CO =26.5%, CH4 = 3.8%, CO2 = 69.3%

Average Cp Values

$H2$ = 14.5 KJ/kg K CO = 1.064 kJ/kg K, CH_4= 0.74 kJ/kg K, CO_2 = 1.014 kJ/kg K

Average Cp = (14.5 * 0.0036) + (1.064 * 0.265) + (0.74 * 0.038) + (1.014 * 0.693) = 1.065 kJ/kg K

Sensible energy of gas = 0.085 * 0.00106 * (450-20) = **0.039 MW**

Total energy of gas = 0.039 + 0.33 = **0.37 MW**

Energy required for heating pyrolysis unit

Total indirect heat input required for the pyrolysis unit (as determined by Aston University) is approximately 10 % of chemical energy of the feed **=** 0.10 *5.52 = **0.55 MW**

Energy balance around Combustor

Solids produced from the pyrolysis unit will be routed to the combustor and combined with vapours leaving the pyrolysis reactor.

Char from pyrolysis unit to combustor = **3.26 MW**

Vapours/gases from pyrolysis unit to combustor = 2.77 MW + 0.37 MW + 0.006 = **3.15 MW**

Sensible heat of ash out of the combustor

Total ash out (60% of pyrolysis solids is ash) = (0.623 * 0.6) = 0.37 kg/s

Sensible heat of solid residue (major component over 60 wt%) =CaO

Cp of CaO = $10.00 + 0.00484T - 108000/T^2$ = cal/deg mol [ref perrys]

T = 733 K

$10.00 + (0.00484 * 733) – (108000 / 733^2)$ =13.35 cal/deg mol

Conversion: 13.35 * 4.187 = 55.90 J/kmol, 55.90 / 56.08 g/mol = 1.00 kJ/kg K

Sensible heat from solid residue leaving the reactor = 0.37 * 0.00100 * (900-20) = **0.32 MW**

None of the sensible heat energy from the solid residue is recovered back into the system

Energy Out of Combustor from Exhaust Gases = 3.26 + 3.15 – 0.32 = **6.09 MW**

System energy requirement for steam raising (assumed unchanged) = **7.4 MW**

To which must be added:

Energy for drying = **1.64 MW**

Energy for pyrolysis = **0.55 MW**

Total system energy requirement = 7.4 + 1.64 + 0.55 = 9.59 MW

System is energy deficient by = 9.59 – 6.08 = 3.51 MW