THE EVALUATION OF ULTRA FINE COAL TREATMENT OPTIONS AT THE WESTERN COAL COMPLEX

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DECLARATION

I declare that this dissertation is my own unaided work. It is being submitted to the Degree of Master of Science in Engineering to the University of the Witwatersrand, Johannesburg. It has not been submitted before for any degree or examination to any other University.

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ABSTRACT

The aim of this research project was to test the response of ultra fine coal sourced at Klipspruit Colliery to froth flotation and the response of the froth flotation products to dewatering using two different types of filter presses, namely the Tecnicas Hidraulicas (TH) and the Ishigakhi presses. During test work, some difficulty was experienced with coarse material feeding the froth flotation pilot plant. This led to pilot plant modifications. Further process complexities necessitated laboratory scale flotation test work on the Klipspruit coal to be carried out.

The results for both the laboratory scale and pilot plant test work for froth flotation indicated that froth flotation as applied to the Klipspruit fines was not economically feasible because neither the required quality of the product (calorific value of 27.80 MJ/kg) nor the product yield of 50% could be achieved when subjected to a primary and secondary stage of froth flotation. The coarse material, which fed the pilot plant and the Ishigakhi filter press, gave low moisture values (12.3%) not typical of ultra fine coal moisture values. However when fed with very fine particle size distributions, prior test work with the Ishigakhi showed that moisture values below 20% could be achieved. The moisture values obtained for very fine particles using the TH filter press on product thickener underflow material sourced at Goedehoop colliery reached values below 20%. Thus both of the two dewatering options, i.e. the Ishigakhi filter press equipment or TH filter press equipment for the ultra fine coal dewatering, can be utilized. Since the filter rate is the determining factor specifying filter press size, it was determined that a larger TH filter area is required in

comparison with the Ishigakhi press. Based upon the pilot and laboratory scale test work undertaken and the assessment of the results, it appears that both dewatering options could be successfully employed on a technical basis for the dewatering of coal flotation products, tailings and the arising raw ultra fine fraction.

Froth flotation for Klipspruit ultra fine coal was deemed unfeasible for both pilot plant and laboratory scale tests conducted. For this reason a capital expenditure for the construction of a froth flotation plant at the Western Coal Complex Phola plant was not considered feasible since Klipspruit coal forms part of the feed that will feed the Phola plant.

In conclusion, following dewatering using either the TH filter press or the Ishigakhi filter press, it was established that both froth flotation concentrate and unbeneficiated ultra fines gave acceptable total moisture results (below 20%). These dewatered raw ultra fines may therefore be blended into inland product as thermal coal to be utilised by Eskom for power generation. Based upon this premise, it is estimated that profits of 76.5 million Rand could be generated by blending Klipspruit ultra fine coal into thermal coal production at the new Phola plant.

Keywords: froth flotation, dewatering, laboratory scale flotation, froth flotation pilot plant modifications, filter rate.

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1. INTRODUCTION

South African coal mines produce 252 213 kilo tones tons of coal per year which is then separated through screening and beneficiation into various products leaving a very high proportion of fines fraction which is then discarded in various ways. The fines may be divided into categories based upon size, namely, slimes, naturally-arising and generated fines due to automated mining methods. In addition the mining of poorer quality coal seams typically increases the fines percentage of the run of mine coal. Given the large quantities of fines being produced and discarded, efforts are being made to examine methods of upgrading these products, which could serve to produce industry marketable products. If these ultra fines were to be left unbeneficiated, larger discard dumps would result which is a considerable environmental concern, especially upon mine closure.

The reactive nature of discard coal and its associated mineralogy (sulphur percentages) and the high level of surface disturbance due to mining activity result in potentially significant pollution problems which may persist for many years after mining have ceased. Methods exist to rehabilitate the surface and to ameliorate pollution, but the long-term success of these interventions depends on the nature of post mining land uses (Limpitlaw, 2005). There are instances where land has been inappropriately utilized by farmers leading to significant environmental problems. This may include water pollution, spontaneous combustion and severe safety and health hazards.

The most difficult issues at present involve the handling and upgrading of coal fines and ultra fines. If fines and ultra fines are discarded due to inadequate processing technologies, excessively large discard dumps could result. Existing cleaning processes for upgrading ultra fine coal includes, froth flotation, selective flocculation, water only cyclones, dense medium cyclones, tabling and oil agglomeration (DW Horsfall, 1976). Froth flotation is the best-established method of cleaning ultra fine coal and is extensively used overseas. Anglo American has a successful froth flotation plant at Goedehoop colliery where the coal quality is similar to the coal quality investigated in this paper. It is for this reason that the preferred method of ultra fine coal cleaning is froth flotation. The coal used for this research was sourced from a resource block called the western coal complex referred to by Anglo American as Klipspruit coal. This coal was feeding Rietspruit colliery and was used for this research. Anglo American project pipeline includes Klipspruit coal feeding to Phola plant and the testwork on the ultra fines would assist in clarifying the capital expenditure required for the fines treatment plant for Phola.

The western coal complex is located south east of Mpumalanga and is less prolific than the Witbank coalfield counterpart. This coal field was investigated because the major junction of the export rail lines linking the inland coalfields to Richards Bay is located in Ermelo, in addition, the Majuba power station buys in all of its coal locally while Eskom's previously decommissioned Camden power station will need an annual 6 million tons of coal when it is re – commissioned in the near future. This location allows two markets to be evaluated namely Europe where the coal product requires a quality of 27.80 MJ/kg and the local inland market where the power station requires coal of a quality of 22.40 MJ/kg. The seams evaluated for this project was two seam and four seam coal (see figure 1).

The current research seeks to examine the froth floatability (by means of a Jameson cell) of ultra fine coals resulting from mining operations in one colliery Klipspruit Colliery in the Witbank Coalfield, and to establish the dewatering potential of the flotation products (flotation concentrates) using two different filter presses. The results will be used to model upscale versions of the processes for implementation at the Western Coal Complex joint venture between Anglo American and BHP Billiton (Phola plant) and to calculate the techno-economic impact these developments would have for the industry.

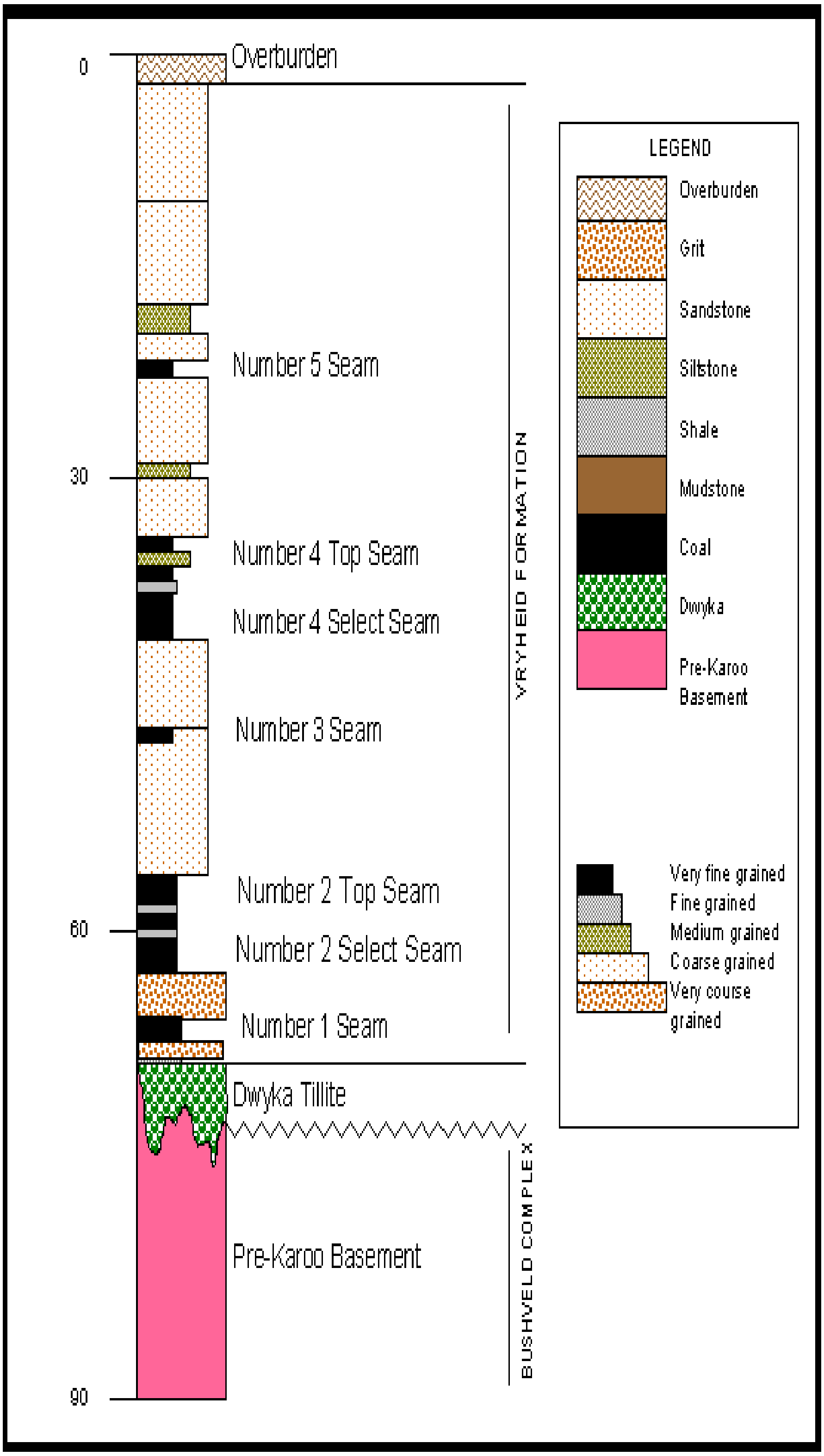

Figure 1: Western coal complex geology borehole log (data supplied by Anglo American).

1.1 PROBLEM STATEMENT

Coal in the finer particle size ranges (below 1 mm) has always suffered from poor cleaning efficiency. In addition, ultra fine coals from different mining areas respond differently to beneficiation and dewatering. Some of the major factors contributing to this situation include mechanized mining methods, suitability of froth flotation, the quality and condition of the coals in different areas (some may be oxidized or weathered if sourced from shallow opencast seams) and the resultant dewatering issues (de Korte, 1991). This problem has been exacerbated in recent years with the increased amount of high ash fines due to continuous mining machines and the mining of poorer quality coal seams. For this reason the recovery of fine coal has received increased attention over the past few years (Phillips, 1998).

Fine coal generally represents a significant portion of run-of-mine coal but unfortunately only a small percentage is recovered. In some cases, ultra fines can constitute up to 9% of Run of Mine (ROM) coal produced, which can lead to significant potential yield loss and significant slimes dam capacity requirements if left unbeneficiated.

A lower cleaning efficiency of fine coal is due in part to initial poor size separation of the fine-sized raw coal from the coarse coal streams, which results in misplaced high ash fines reporting to the coarser size streams and vice versa.

For effective froth flotation to occur, sufficient liberation of the ultra fines is necessary. A further aspect affecting froth flotation is particle surface oxidation. When the particle surface is oxidized the reagent's ability to change the particle surface charge is reduced. For this reason, there is a need to test and optimize reagents for froth flotation.

One of the most difficult and costly steps in coal preparation is the removal of moisture from the surfaces of fine coal particles. Although thermal dryers can effectively reduce moisture, these units require large capital expenditure. As a result, coal producers resort to mechanical systems for fine coal dewatering. Unfortunately, the performance of mechanical dewatering equipment diminishes sharply for particles finer than 0.1 mm. Hence the need arises to test various dewatering filter presses to compare and establish the most suitable dewatering method for these ultra fine size fractions.

Historically coal flotation has had a fairly precarious past in the Witbank area due to poor performance (reagent regimes) and the consequent difficulties in dewatering the froth flotation product. Coal dewatering is one of the most important factors for improving the product quality especially on finer coal after flotation, and as such it is critical to the success of the overall process (van der Walt, 2005). The process efficiencies and cost benefits of these methods have not proved to be economical in a number of instances**.** Grootegeluk has decommissioned their froth flotation plant for this reason.

1.2 AIMS AND OBJECTIVES

The aims and objectives of this project were:

- To establish the floatability of Klipspruit ultra fine coal.
- To optimize flotation in terms of the targeted product quality (CV of 27.80 MJ/kg) and a targeted recovery of 50%, reagent dosage rate and reagent ratio. This optimization would be highly dependant on the ability of the Klipspruit ultra fine coal to float (froth flotation) by using reagent regimes that have been proven for Witbank Coals.
- To select the optimum dewatering method for the ultra fines. For this purpose two filter presses will be investigated in terms of filter capacity required (dry kg/m^2) and moisture results obtained.
- To utilize the flotation and dewatering results for scaling up purposes for both with or without flotation option.
- To investigate the potential income of marketing the un-beneficiated Western Coal Complex ultrafine coal by utilizing local inland markets for example Eskom.
- To investigate the value chain cost benefit of undertaking ultra fines beneficiation from the Western Coal Complex and to evaluate the techno-economic impact of the potential marketability of the final products for utilizing local inland markets.

2. LITERATURE SURVEY

Coal contains a significant amount of impurities and for marketing and sales purposes coal beneficiation treatment is required to remove these impurities. A typical coal processing plant consists of crushing the as-mined coal and reducing the particle size from 300mm to less than 100mm. This typically occurs in two or three crushing phases namely, primary crushing where three hundred millimetres run of mine coal is reduced to hundred and fifty millimetres and secondary crushing where hundred and fifty millimetres is reduced to hundred millimetres. This crushed coal is then separated into 4 different size fractions by means of sizing screens (see table 1). Wet vibratory screens are typically used for separating larger coal sizes and sieve-bends are typically used to separate the coal into fine and ultra fine sizes.

The first size fraction is less than hundred millimetres and bigger than twenty millimetres. This size fraction of the run of mine, which could range from twenty to fifty percent of the crushed coal, is normally treated by means of a bath type separator like a Wemco drum. The Wemco drum separator consists of a cylindrical drum that rotates on bearings. Magnetite is used as medium to allow for density separation. The raw coal and medium enter at one end of the drum. The clean coal floats and flows over a weir at the discharge end of the drum, where it is drained, rinsed and sprayed for medium recovery. The sinks are lifted from the bottom of the drum by lifters fixed to the inside of the rotating shell. The sinks are discharged into a trough, where a current of medium conveys them to a drain and rinse screen. This is one of the most effective means of coarse coal separation in the industry and is utilised in most dense medium plants (England, 1980).

The second size fraction is less than twenty millimetres and bigger than one millimetre. This size fraction is typically treated by means of dense medium cyclones. The raw coal to be treated is suspended in a fine medium (finely milled magnetite) and this pulp is fed tangentially through a tapered inlet to a short cylindrical section. The shape and design of the inlet varies among manufacturers which has a significant effect on the volume capacity and, to a lesser extent, the efficiency of this process. The short cylindrical section also carries the central vortex finder, which prevents short-circuiting within the cyclone. Separation is made in the cone shaped part of the cyclone by the action of centrifugal force and centripetal forces. The discard portion of the raw coal leaves the cyclone at the spigot opening and the clean coal leaves at the overflow top orifice via the vortex finder. This is the most effective separator for coal and, depending on the cyclone diameter, coal up to seventy five millimetres can be treated (England, 1980).

The third size fraction is smaller than 1mm and bigger than hundred microns. This size fraction is normally treated by means of spiral concentrators. The separation process is based on gravity separation. In this case, the flow of material down the spiral results in heavy material staying on the inside turns with less dense material moving to the outside of the spiral. Separation is achieved by means of adjustable blades positioned in the flowing stream which split the flow and separate out product of the required quality. The separated fractions are collected in pipes and run into launders to collecting tanks. Spirals typically have low separation efficiency but are widely used in industry due to their low capital and operating costs.

The fourth size fraction is smaller than hundred microns (0,1mm) and is typically discarded due to poor quality and difficulty of separation. A description of a typical froth flotation plant is given in section 2.1. Although much research regarding improvements in fine coal cleaning has been performed over the last twenty years, the processing of material smaller than one mm (fines and ultra fines) remains difficult and results in very poor separating efficiencies when compared to those of coarser sizes (Horsfall, 1972). Table 1 lists the sizes of the various categories of coal for comparison. Whilst some cleaning devices have higher efficiencies than others, they often involve greater economic expense, whether real or perceived. Not only is there little consensus in the coal industry as to the overall best cleaning methods for particles finer than 1mm, there is little agreement as to what particle size ranges are best suited for each of the devices employed (Horsfall, 1972). A general guide to coal sizing is given in Table 1 (Amcoal training department, 1999).

In general, the smaller the particles of coal, the more difficult or more expensive it is to separate them from non-coal or stone particles. On the other hand, the smaller the particle, the more coal is liberated from stone. This has the effect of both increasing the yield of saleable material and diminishing the selectivity needed in the separation. The liberation effect has been demonstrated empirically by Horsfall (1976) and has been confirmed by Falcon (1978) using mineralogical examination. Liberation of ultra fine coal may be achieved by non-selective beneficiation processes. This has been proved by Franzidis and Schroeder-Buys (1985) and Panopoulos (1984; 1985).

Attempts to exploit this phenomenon within Anglo American Corporation were made in 1977, when samples of the minus hundred and fifty micron fraction and the fraction between hundred and seventy five and hundred and fifty micron were tested on the newly developed Leeds cell. The results showed clearly that lower volumes of ash, for equivalent yields, were obtained from the ultrafine coal than from the fine coal; but neither quality was sufficient to justify implementation of the procedure (Horsfall, 1986).

Therefore it follows that, for any coal-cleaning device, separation efficiency is reduced as the particle size reduces. An additional problem that is seldom properly addressed is that of high ash fine particles (slimes) reporting to a coarser size stream from which they are not properly removed or cleaned. Controlling the feed rate to the screen to below the screen capacity can reduce this problem. This problem of misplaced sizes is often the source of much of the high ash and low yields found in fine coal cleaning (de Korte, 2001).

Table 1: A general guide to coal sizing (Amcoal training department, 1999)

Grade	Size mm	Mesh	Treatment process
Coarse	+ 10	+ 3/8 inch	Dense medium bath separators
Intermediate	10 x 0.5	3/8 inch x 28 mesh	Dense medium cyclones
Fine	0.5 x 0.15	< 28 mesh	Spiral concentrators
Ultra fine	0.15 x 0	100 mesh x 0	Froth flotation

Cleaning of ultra fines is complex since the only well established global cleaning method for this purpose is froth flotation. However, froth flotation is often not selective enough on South African coal, which can result in poor qualities or cause misplacement of product in the tails.

2.1 HISTORIC DEVELOPMENT OF COAL

In terms of historical developments, fine coal processing on even a moderate scale had its beginnings as recently as 1940 and 1950 (Fourie, 1977). Some of the common methods and sizes for coal processing are illustrated in Figure 1, but the actual particle size at which one processing method gave way to another was never well defined and continues to vary from one company to another.

Coarse coal was formerly often cleaned in a jig but is now most commonly cleaned in a heavy media vessel (mainly Wemco Drums). Wemco drums can process coal particles with a top-size of 300 mm in diameter and a bottom size of 6 mm. However, it is becoming increasingly popular to simplify circuitry by cleaning the coarse size along with the intermediate size in a heavy media cyclone. The main advantages of large diameter dense medium cyclones are increased separation efficiency and sharper cut point (Swanson 2009).

The intermediate sizes were formerly cleaned in jigs or Deister tables but are now almost totally processed in Heavy Media Cyclones (HMC). Prior to 1950 fine sizes were often discarded and in later years were cleaned on either Deister tables or in water-only cyclones. Since heavy media cyclones replaced Deister tables for the intermediate sizes, it was desirable to use a separate cleaning device for the fine size only in order to reduce magnetite consumption in the HMC circuit. For this reason, as well as their low operating costs, spiral concentrators (spirals) were chosen even though their efficiency is similar to that of Deister tables on the minus 1 mm coal (Deurbrouck and Palowitch, 1963).

Historically the ultra-fine sizes were added back raw to the washed product or were discarded (Coulter, 1957). Today the ultra-fines are still discarded in many coal plants producing power plant fuel (steam coal). However, some steam coal plants and most metallurgical (coking coal) plants have incorporated froth flotation in their beneficiation process in order to achieve financial profits derived from the sales of ultrafine flotation concentrate (Coulter, 1957). Conventional cell flotation (a trough shaped tank with agitators) is the type of flotation most often employed, although column flotation (Jameson cell) has gained greater recognition in recent years.

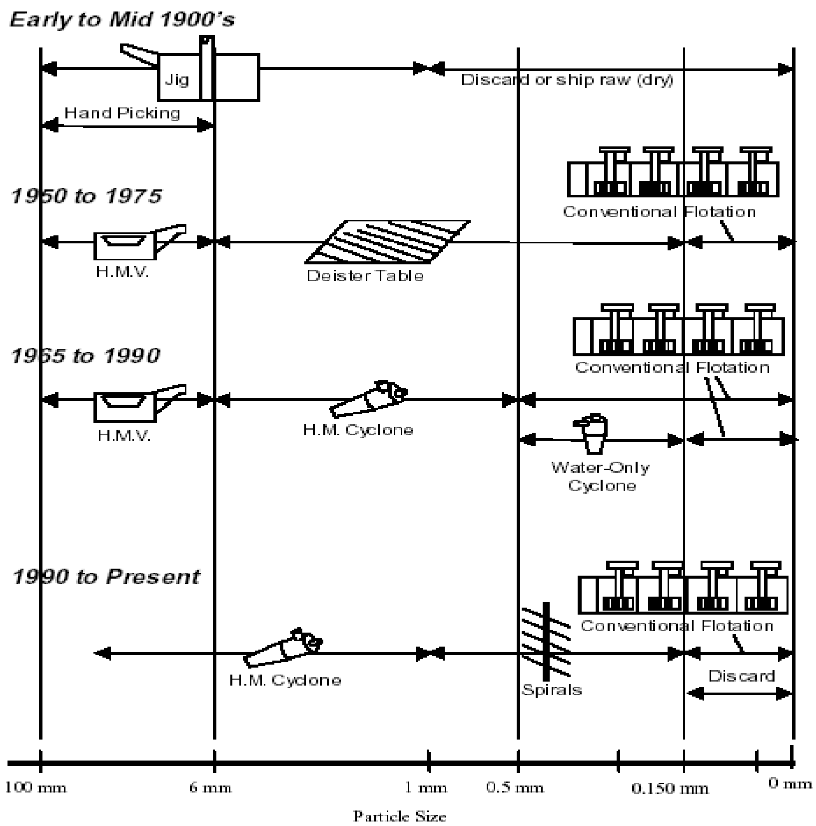

Figure 2: Common coal processing methods and sizes (Phillips, 1998).

2.2 ULTRA FINE COAL FROTH FLOTATION AND DEWATERING

Froth flotation is a process for separating ultra fine minerals from gangue by taking advantage of differences in their hydrophobicity (England, 1980). Hydrophobicity differences between valuable minerals and waste gangue are increased through the use of surfactants and wetting agents. The selective separation of the minerals makes processing complex multi-mineral ores economically feasible. The flotation process is used for the separation of a large range of sulfides, carbonates and oxides.

Froth flotation commences by comminution (that is, crushing and grinding) which is used to increase the surface area of the ore for subsequent processing. It also breaks the ore in order to separate the desired mineral and gangue in a process known as liberation. This then permits the desired mineral to be separated from the other gangue minerals. The ore is ground into a fine powder and mixed with water to form a slurry. The desired mineral is rendered hydrophobic by the addition of a collector. The particular chemical depends on which mineral is being refined. This slurry (more properly called the pulp) of hydrophobic mineral-bearing ore and hydrophilic gangue is then introduced to a water bath which is aerated, creating bubbles. The hydrophobic grains of mineral-bearing ore escape the water by attaching to the air bubbles, which rise to the surface, forming a froth. Typical collectors used in the industry include xanthates, dithiophosphates and paraffin (Phillips, 1998).

Flotation can be performed in rectangular or cylindrical mechanically agitated cells or tanks, flotation columns, Jameson cells or deinking flotation machines. Mechanical cells use a large mixer and diffuser mechanism at the bottom of the mixing tank to introduce air and provide mixing action. Flotation columns use air sparges to introduce air at the bottom of a tall column while introducing slurry above. The countercurrent motion of the slurry flowing down and the air flowing up provides mixing action. Mechanical cells generally have a higher throughput rate, but produce material that is of lower quality, while flotation columns generally have a low throughput rate but produce higher quality material.

The Jameson cell uses neither impellers nor sparges, instead combining the slurry with air in a downcomer where high shear creates the turbulent conditions required for bubble particle contacting.

A typical froth flotation circuit would consist of a conditioning tank, rougher cells and scavenger cells. Various flotation reagents are added to a mixture of ore and water (called pulp) in a conditioning tank. The flow rate and tank size are designed to give the minerals enough time to be activated. The conditioner pulp is fed to a bank of rougher cells which remove most of the desired minerals as a concentrate. The rougher pulp passes to a bank of scavenger cells where additional reagents may be added. The scavenger cell froth is usually returned to the rougher cells for additional treatment, but in some cases may be sent to special cleaner cells. The scavenger pulp is usually barren enough to be discarded as tails. More complex flotation circuits have several sets of cleaner and re-cleaner cells, and intermediate re-grinding of pulp or concentrate (see figure 3). Coal surface properties and ash percentage determine the coal floatability and particle size distribution of the ultrafine coal has a direct correlation with the coal's dewatering ability.

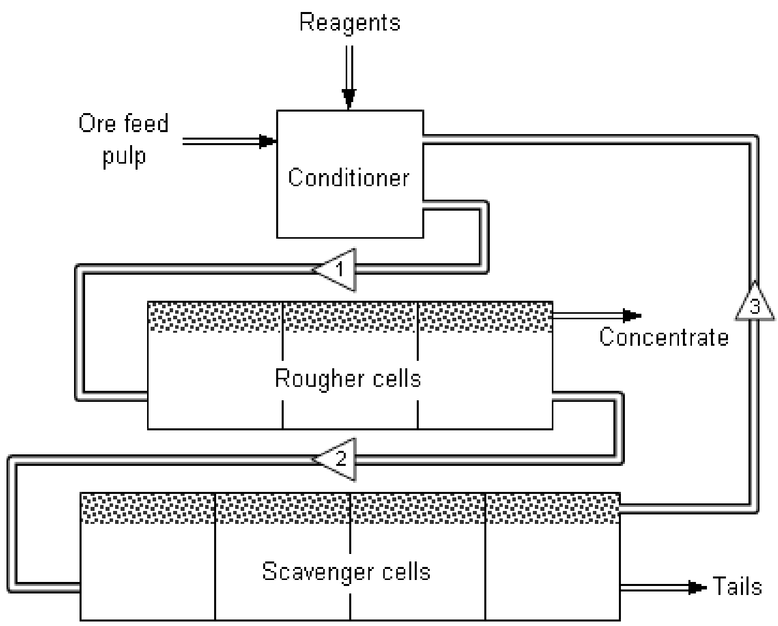

Figure 3: A typical coal flow diagram (Phillips, 1998).

Coal floatability and surface properties strongly depend on the coal rank. Very hydrophobic bituminous coals float well using only a simple frother such as MIBC or different frothers. Paraffin is used most commonly in industry due it being so readily available. In contrast, lower rank coals prove difficult to float. In this case surface oxidation is normally the major contributing reason for the low floatability achieved (Laskowski, 2003).

Ultra fine coal dewatering ability is strongly dependant on particle size distribution. A finer particle size distribution will lead to higher surface moisture obtained. This surface moisture has a direct affect on the marketability of the flotation concentrate since higher surface moistures will lead to lower calorific values. The coal prepared by a coal preparation plant is usually wet; this coal requires drying to satisfy customer requirements. The main uses to which coal is applied necessitate a fairly dry material, e.g. general industrial fuels, export coals, power station fuel and raw material for processing, such as coke oven charges. Another reason for dewatering is to recover water for re use (England, 1980).

Water is bound in different ways to the coal with which it is in contact. It is important to differentiate between two types of moisture held by coal namely inherent moisture and free moisture. Inherent moisture is held in the capillaries and cracks of the coal particles and is retained by cohesive forces such as surface tension. Free or surface moisture can be removed by draining, but inherent moisture can only be removed by heating the coal particles to a temperature higher than the boiling point of water (England, 1980).

Inherent moisture can be further defined as the water held both in the pores of the coal and on the surface of the coal. The moistures held in the pores of the coal are referred to as being ‘internally bound’. ‘Externally bound’ moisture is a thin film of water bound to the surface of the coal by means of adhesion force. This water would be present even if there were no other coal particles present (England, 1980).

Free moisture is further described as occupying the open spaces between coal particles and is not bound to the coal surface by any forces of attraction. It is bonded to the film of inherent moisture by means of cohesion force. The presence of free moisture depends on other coal particles being present, the smaller the coal particles, the greater the surface area presented. Such increased surface area can carry more free moisture, which is why fine coal has higher moisture content than large lumps of coal.

Free moisture can be removed by gravity or mechanical means while the inherent moisture can be removed by thermal drying. Free moisture is most commonly removed by mechanical means (England, 1980). China and Japan are the world's largest suppliers of dewatering equipment. They specialise in filter press manufacturing. Filtration by means of mechanical filter presses has proven very effective on ultra fine coal and is used mainly by China, the USA, and Japan.

Filtration is the process whereby solids are separated from a liquid by means of a porous medium that retains the solids but allows the liquid to pass. Suspensions of very fine material are treated in a filter. The slurry originating from a thickener cannot be readily handled. Similarly, the process of froth flotation yields frothed fines, which is usually thickened by a thickener to about thirty percent solids by mass. These suspensions can be de-watered by using filters (England, 1980).

In terms of the current research, the floatability of the fine coal from Klipspruit colliery which forms part of the Western Coal Complex will be determined first whilst the second part of the feasibility study will be to establish the most suitable dewatering method of the Klipspruit ultra fine coal. These results will be used for scaling up purposes for the new Western Coal Complex (Phola plant) joint venture between Anglo American and BHP Billiton to establish the techno-economic impact of the processes. Significant financial gain implications could arise if proven feasible. During sampling sufficient quantities of sample are required to enable representative results. As a general guide the following table indicates the quantity of sample required per sample increment.

Table 2: The quantity of sample required per sample increment (Amcoal Training Department, Coal Preparation, 1999).

Nominal top size of coal D (mm)	Mass per increment, kg, minimum	Width of opening in sampling device mm
10 to 0 mm	0.6	30
15	0.9	38
20	1.2	50
30	1.8	75
35	2.1	88
65	3.9	163
100	6.0	150
120	7.2	300
150	14.2	375
200	33.3	500

3. RESEARCH METHODOLOGY

This chapter describes the source of samples under investigation, their characterization and the equipment that was used to test them. A summary of the problems and experiences gained in the course of the work will also be presented.

Klipspruit plant received coal from Rietspruit Colliery as this coal will form part of the new Phola plant feed stream. The aim of undertaking research on this coal was to establish the recovery and dewatering potential of Klipspruit ultra fine coal (minus 0.1 millimetres).

Crushed coal was fed via a feed conveying system to a sizing screen where the coal is split into three, -75 +25 mm fed the Wemco drum plant whilst –25 + 1 mm fed the dense medium cyclone section, the –1 mm was send to classification cyclones where the – 0.1 mm portion was removed and send to the Jameson cell pilot plant unit (see figure 4). The Jameson pilot unit consisted of a rougher stage and scavenging stage. The Jameson pilot plant product concentrates were sent to an Ishigakhi filter pilot plant press for dewatering. These pilot plants were specially designed for the amount of feed based on Rietspruit plant throughput.

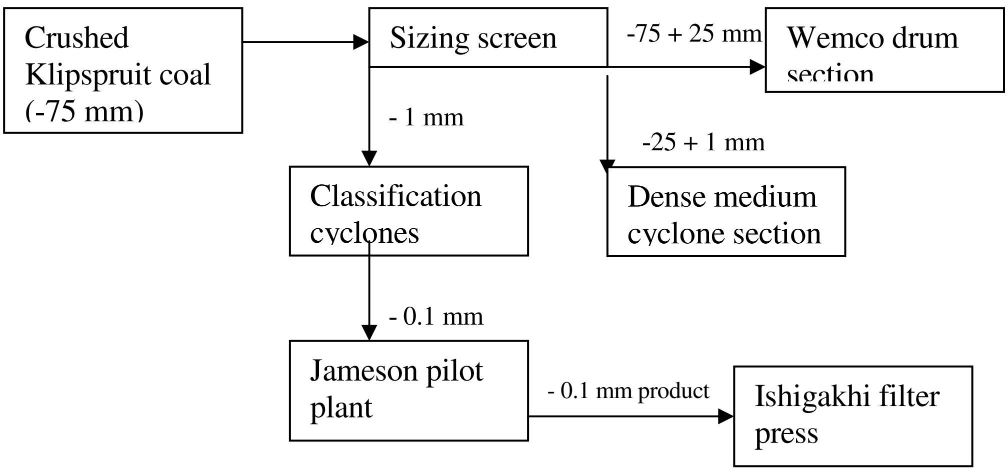

Figure 4: The Rietspruit plant flow block diagram.

Due to excessive oversize material reporting to the Jameson pilot plant, a sieve-bend screen was installed prior to the Jameson pilot plant. This screen removed material bigger than 0.25 mm due to inefficiencies in the classification cyclones.

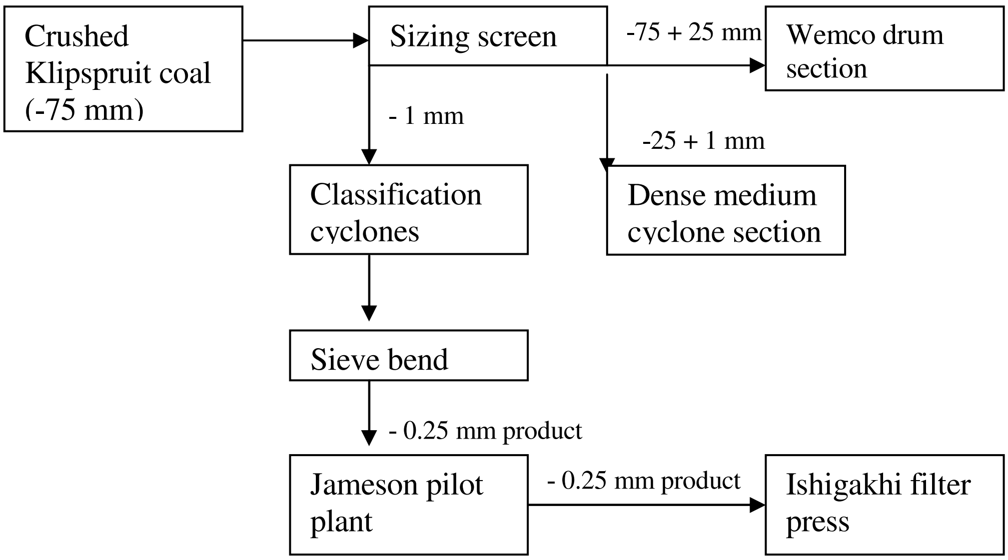

Figure 5: The modified pilot plant flow sheet.

Goedehoop plant ultra fine froth flotation product was also investigated for dewatering trails. In this case, dewatering was undertaken with a TH filter press. This was done to allow both filtration trails to run simultaneously due to project time limitations.

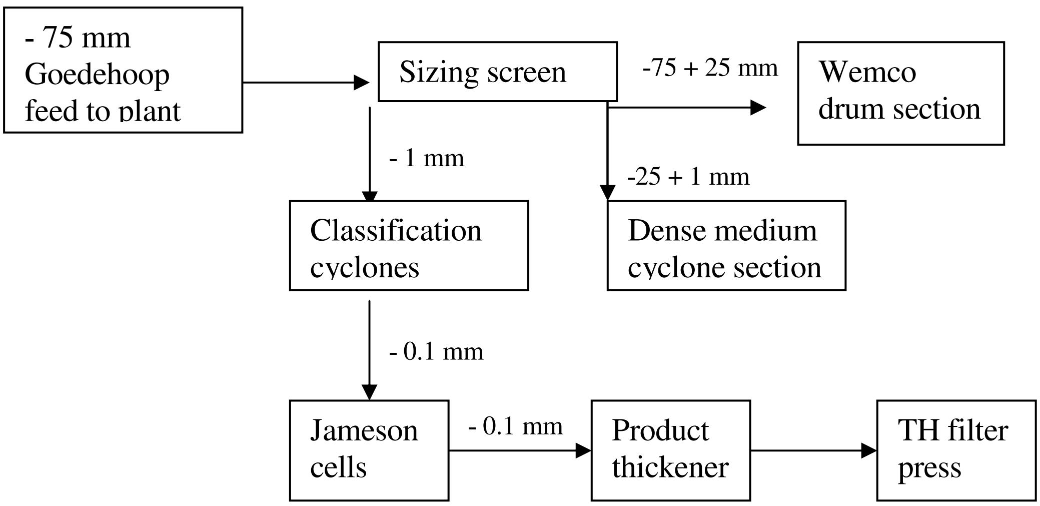

Figure 6: Goedehoop plant flow block diagram.

3.1 SAMPLING, PROCESSING AND SAMPLE CHARACTERISATION

Ultra fine coal samples from the following sampling point sources were investigated:

- Klipspruit ultra fine coal sourced from Rietspruit plant was sampled at the Jameson-cell pilot plant and analysed for froth flotation yield and dosage optimization. Dewatering was done at Rietspruit on Klipspruit ultra fines with an Ishigakhi filter press on all pilot plant concentrate samples arising from the Jameson-cell pilot plant (see figure 4).
- Klipspruit ultra fine coal sourced from Rietspruit plant was sampled at the modified pilot plant from the sieve bend underflow and analysed for froth flotation yield and dosage optimization. Dewatering was done at Rietspruit on Klipspruit ultra fines with an Ishigakhi filter press on sieve bend underflow material (see figure 5).
- Klipspruit ultra fine sieve bend underflow was sampled in buckets and was used to do test work at Goedehoop laboratory in a Denver flotation cell for froth flotation yield and dosage optimisation (see figure 5).
- Goedehoop plant flotation concentrate thickener underflow was sampled for dewatering tests with the Technicas Hydricas filter press (see figure 6).

<u>Klipspruit ultra fine froth flotation sampling program – initial Jameson cell flow sheet (figure 4)</u>

Samples of Klipspruit ultra fines were sourced from the pilot plant Jameson-cells at Rietspruit plant. Samples of Jameson-cell feed, concentrate and tailings were taken. The test work was conducted at various reagent ratios and dosage rates. In total, 37 pilot plant tests was performed where the ratio of paraffin to frother was varied in ratios of 50:50, 75:25, 82:18 and 90:10. Dosage rates of 1000 g/ton, 2000 g/ton, 3000 g/ton and 4000 g/ton were investigated at the various reagent ratios. The aim of this test work was to achieve a product with a Calorific Value (CV) of 27.80 MJ/kg and a combined concentrate yield of 50%.

Klipspruit ultra fine sampling program for dewatering – initial jameson cell flow sheet (figure 5)

The Rietspruit plant Ishigakhi filter press was used to dewater the Klipspruit pilot plant sieve bend underflow material. A fixed membrane pressure of 12 bar, air blow time of two minutes and feed rate of 7.8 m^3/hr was used for all dewatering trials. Three different airflow rates were investigated namely, 30, 80 and 90 M^3/hr. Dewatering trials on both primary and combined concentrate were conducted.

Klipspruit ultra fine sampling program – modified pilot plant flow sheet (figure 5)

The pilot Rietspruit plant was modified by the installation of a sieve bend screen prior to the Jameson-cells. Four samples were taken from the Jameson-cell feed, product and concentrate. Ekofol and paraffin was used with a reagent ratio of seventy five percent collector (paraffin) and twenty five percent frother at various dosage rates (1000, 2000, 3000 and 4000 g/ton). Due to process complications at Rietspruit plant the test work program was aborted.

Laboratory scale flotation test work

Klipspruit ultra fine material taken from the modified flotation pilot plant (figure 5) at Rietspruit colliery sieve bend underflow underwent laboratory scale flotation. A matrix indicating the test work that was conducted on Klipspruit ultrafine coal is given in table 2. The reagents used, the various reagent dosage rates and collector to frother reagent volume ratios are listed in Table 3 below.

Table 3: Flotation matrix for Klipspruit coal laboratory scale test work.

Klipspruit Coal flotation matrix.		
Reagent	Dosage kg/ton	Reagent ratio – Collector : frother
Ekofol (frother) & paraffin (collector)	1 2 4	60:40; 75:25; 80:20; 95:5 60:40; 75:25; 80:20; 95:5 60:40; 75:25; 80:20; 95:5
Centifroth (frother) & paraffin (collector)	1	60:40; 75:25; 80:20; 95:5
Centifroth & no collector	1 2 4	60:40; 75:25; 80:20; 95:5 60:40; 75:25; 80:20; 95:5 60:40; 75:25; 80:20; 95:5
Betafroth (frother) & Betachem (collector)	1 2	95:5 95:5

Goedehoop flotation concentrate dewatering test work

The samples taken at Goedehoop flotation plant product thickener underflow were used for dewatering tests with the Technicians Hydricas filter press. Surface moisture analysis was analysed on filtration cakes. (See figure 6)

All flotation reagents were pre mixed by the suppliers as the plant design had only one reagent tank. For this reason, the normal procedure of adding collector to the feed and then allowing time for feed conditioning before frother is added could not be achieved.

3.2 ANALYSIS

The following sample analyses were conducted on flotation feed, flotation concentrate and flotation tailings for both flotation pilot plant test work and laboratory scale flotation:

Ash content (SABS Method 926)

About one gram of coal was weighed into a dish and placed in a cold furnace ventilated by air; the temperature was slowly raised to 800 degrees Celsius and maintained at this temperature until constant weight was obtained. The dish was left in open air until it is completely cooled down. The ash content is the mass of the ash expressed as a percentage of the amount of coal taken.

Calorific value (SABS method 929)

The heating value of coal was determined by burning one gram of coal in a calorimetric bomb filled with oxygen under pressure. The bomb was placed in the calorimeter with a fixed amount of water and the increase in temperature was measured. The calorific value was obtained from the corrected rise in temperature, the effective heating capacity of the calorimeter and the weight of coal used. The calorific values were measured in mega joules per kilogram (MJ/kg). The calorific value determined by this procedure was the gross calorific value and included the heat generated when water vapor was formed and condensed as a product of combustion.

The following sample analysis was conducted on flotation dewatered material for both filter presses under investigation:

Moisture content (SABS methods 924, 925)

Approximately one gram of coal was weighed into a dish and heated in an oven at 105 degrees Celsius until the moisture was driven off. The dish was taken out, covered with its lid, cooled in desiccators and weighed to determine the loss in mass. The loss in weight, expressed as a percentage of the coal used, is the moisture content.

3.3 TEST EQUIPMENT

The materials required to enable this research project are listed below:

- Anglo Coal in-house Jameson-cell pilot plant primary and secondary cell
- Classification cyclone overflow from Rietspruit plant for supply of froth flotation feed to the pilot plant Jameson-cell (for both the initial pilot plant and modified pilot plant)
- A feed tank with a pump and pipelines connected to the primary jameson cell
- A water supply source is necessary to maintain flotation feed tank level (water source must be consistent in quality throughout the test series)
- A laboratory filter press to dewater samples from flotation feed primary concentrate and secondary concentrate as well as tailings samples
- A laboratory surface moisture furnace set at 40 degrees Celsius
- Flotation reagents make up tank
- A laboratory Denver cell with a 5 litre container
- A Raymond sample pulverizer
- Laboratory cal 2 k calorific value analysis equipment
- Oxygen cylinder to feed the CV analysis equipment
- A laboratory scale for weighing ash and CV samples
- A pilot plant high rate thickener
- A feed pipe feeding a pilot plant filter press
- Product thickener underflow from a reference colliery.
- Ishigakhi pilot plant filter press
- Tecnicas Haudraulicas (TH) pilot plant filter press

A brief description of the dewatering filter press equipment is given below:

The TH filter press, as supplied by Prepquip, had the following major advantages: less moving components compared to the Ishigakhi filter press, is fully automated, has easy operating philosophy and requires less maintenance. It dewaters the fines through constant air flow across a series of membranes of which the coal pulp is encapsulated. Once the volume of air passes over the pulp the membranes, the membranes are retracted through an electric hydraulic principle and the dry pulp is deposited on a slow feeding conveyor belt. [NB: This filter press had been tested at Landau colliery and proved to give surface moisture values of twenty to thirty percent, hence its application in the research programme].

The Ishigakhi filter press was supplied by Seprotech. The operating principle includes generating higher surface pressures on the membranes through an electric mechanical duel chain mechanism. This filter press has more moving components and thus requires more intensive maintenance and labour strategies to achieve a financial model similar to the TH filter press. This filter press had been tested at Goedehoop colliery and proved to give surface moistures between 17% and 28%.

3.4 THE PILOT PLANT FLOTATION CIRCUITS AND PROBLEMS EXPERIENCED

3.4.1 Initial flotation circuit (illustrated figure 4)

Flotation feed material was received from Rietspruit plant. This material was fed to the primary flotation cell where the fast floating material reported to the froth product. The remaining proportion of the coal which is usually less hydrophobic sank to the primary cell underflow (tails) due to gravity forces.

The primary tailings fraction was fed to the secondary flotation cell where a second stage of separation occurred. The second stage is generally referred to as a scavenger stage since remaining product is scavenged from the primary cell discard. A schematic flow sheet depicting the Jameson cell flow sheet of the Rietspruit plant is shown in figure 7.

PRIMARY FLOTATION CELL SECONDARY FLOTATION CELL

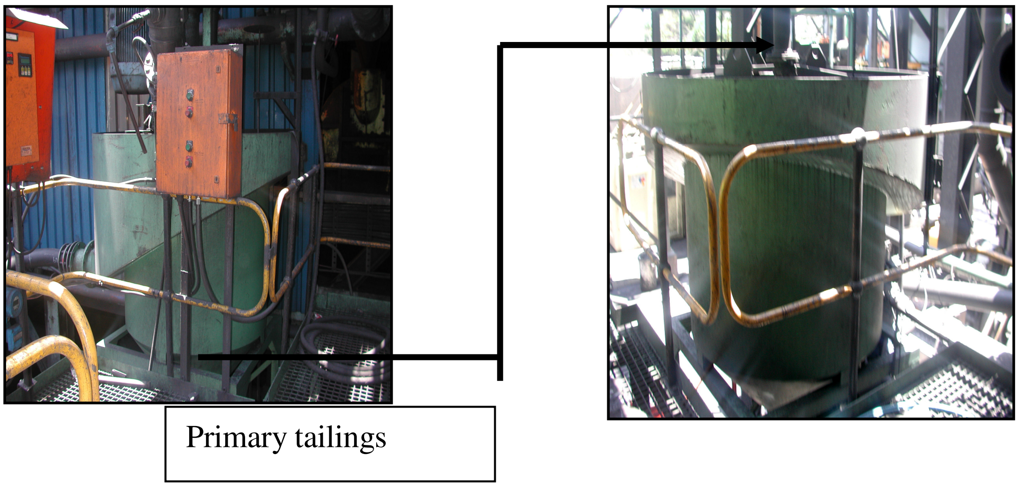

Figure 7: Schematic illustration of a Jameson cell flow sheet.

3.4.2 Problems with the initial flotation circuit

One of the main problems experienced in the initial flotation circuit was the occurrence of coarse material in the Jameson cell feed. The Jameson cell feed size distribution was examined and it was found that Rietspruit had an abnormally coarse particle size distribution compared to other reference particle size distributions, Goedehoop colliery ultra fines feed size distribution was used for a reference particle size distribution. Comparative size distributions are illustrated in Figure 8.

Coarse material was difficult to float due to increased gravitational forces on such particles. During flotation in the primary cell, coarser material of good quality reported to the tailings fraction whilst flotation in the second cell was hardly possible due to a heavily loaded and stable froth. The results obtained for some of the secondary cell yields were negative which supports the previous statement (appendix A1). It can therefore be concluded that the Jameson cell is fit for coarse floatation but the limiting particle size is approximately one millimetre in diameter. Other problems included low proportions of solids from the feed and erratic feed conditions. Erratic feed, in turn, caused varying froth height, mass recovery and product quality.

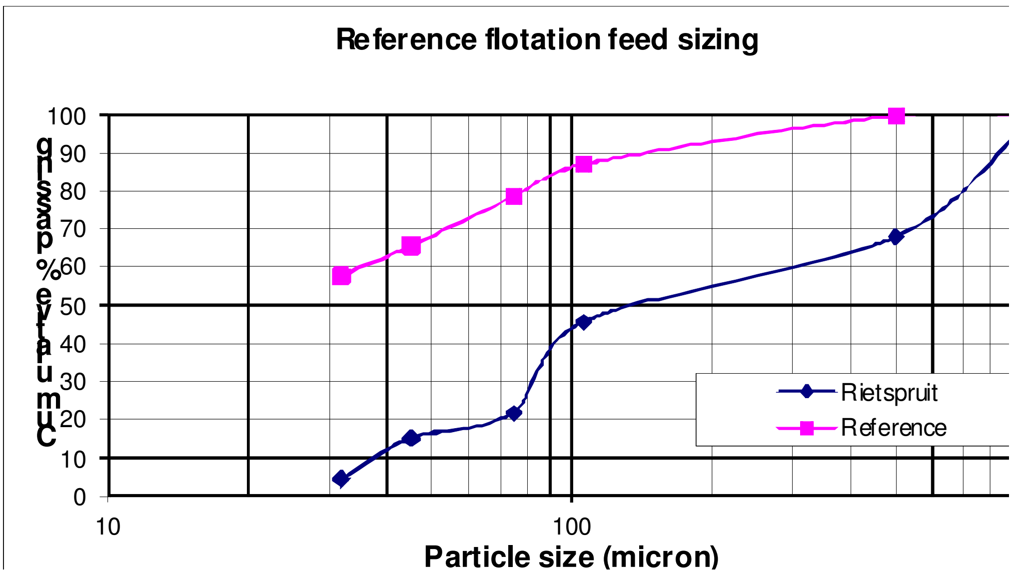

Figure 8: Klipspruit ultra fine particle size distribution prior to sieve bend installation.

3.4.3 Modified flotation circuit (figure 5)

The feed to the sieve bend was obtained from the classification cyclone overflow fraction of Rietspruit fines section. The undersize from the sieve bend was transferred to the Jameson cell via a 4/3 Warman-slurry pump. Air and reagents were added to the cell from a common collector and frother tank. The floated coal reports to the froth product launder. The primary tailings (slower floating coal and shale) which are less hydrophobic, were transferred to the second cell for further processing (figure 9).

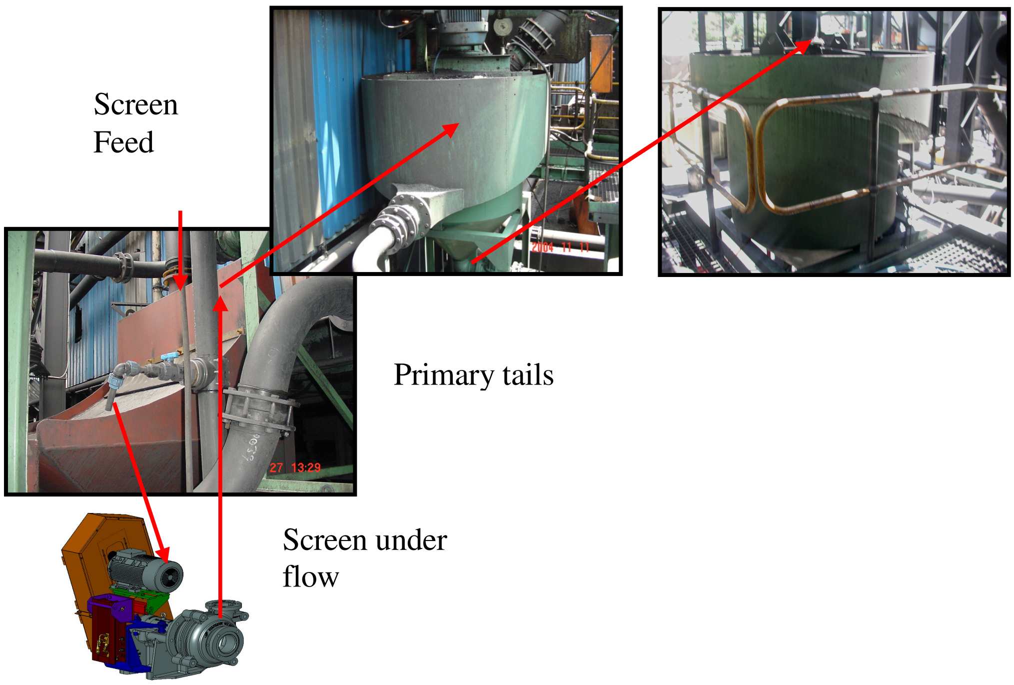

Figure 9: The modified pilot flotation plant

3.4.4 Difficulties experienced with the revised circuit (figure 5)

Due to further process difficulties experienced at the Rietspruit plant, the pilot plant facility could not be utilized fully and as a result only four test runs were conducted. The results show that the yields were less than 50% and that the product had an average ash value of 13.63% and an average CV value of 27.52 MJ/kg. Therefore a decision was made to perform batch flotation tests in a laboratory Denver flotation cell at Goedehoop Colliery using samples taken at the sieve bend underflow (figure

5). The pilot plant sizing results, however, showed that the sieve bend functioned well and that oversized material was removed efficiently. The oversize (+0.25 mm) accounted for only 5.56% of the flotation feed (see Figure 10). Therefore misplacement of oversize in the undersize was mainly due to the presence of elongated particles.

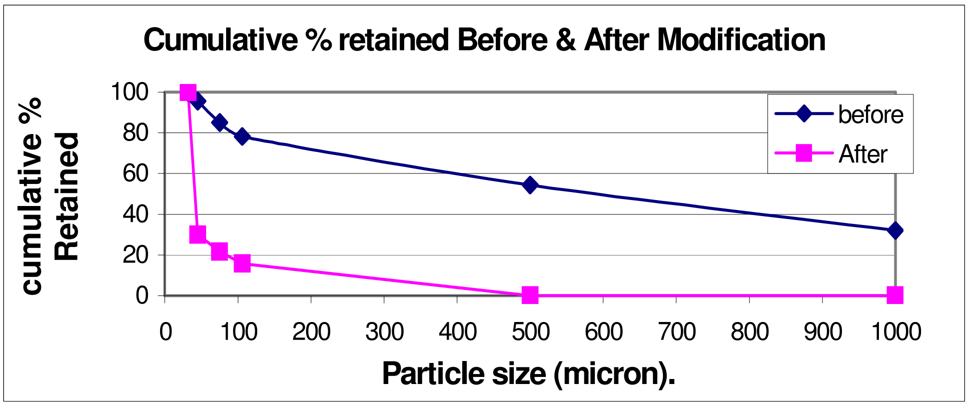

Figure 10: The particle size distribution of the Jameson cell feed before and after modification

A shift in the quality of the Klipspruit material was noticed after the circuit modifications. A quality decrease of 3.72 MJ/kg was observed (Table 4). This shift in quality was due to deportment of lower quality coal into the smaller size fractions again emphasizing the criticality of ensuring the proper segregation of particles prior to further processing.

Table 4: Quality decrease in the Klipspruit ultra fine coal pre and post modifications.

	Ash %	**CV (MJ/Kg)**
Before	28.07	21.90
After	39.10	18.18

3.5 DATA GATHERING

Sampling can be defined as the extraction of a small sample of coal from a larger bulk of coal such that it is representative of the larger bulk all the sources of samples are mentioned in section 3.1.

3.7. DATA ANALYSIS

Data was gathered and each feed, product (primary or secondary) and tailings streams from the jameson cell and laboratory Denver cell ash results was used to determine the percentage yield on the primary rougher stage and the secondary scavenging stage. The two yields were combined to determine the combined product yield as well as combustible recovery. The following formulae were applied to obtain combined product yield and combustible recovery.

$$\text{Primary stage yield} = \frac{\text{Tailings ash \% - Feed ash \%}}{\text{Tailings ash - primary concentrate ash \%}}$$

$$\text{Secondary stage yield} = \frac{\text{Final tailings ash \% - primary tailings ash \%}}{\text{Final tailings ash \% - secondary concentrate ash \%}}$$

$$\text{PCR} = \text{Primary yield \%}\left(100 - \frac{\text{Primary conc ash \%}}{\text{100 - feed ash \%}}\right)$$

$$\text{SCR} = \text{Secondary yield \%}\left(100 - \frac{\text{Primary conc ash \%}}{\text{100 - feed ash \%}}\right)$$

$$\text{TCR} = \text{PCR}(100 - \text{PCR})\left(\frac{SCR}{100}\right)$$

* Where PCR = primary combustible recovery

SCR = Secondary combustible recovery

TCR = Total combustible recovery

Conc = concentrate

The test work was conducted at various reagent ratios and dosage rates. In total, 37 pilot plant tests was performed where the ratio of paraffin versus collector was varied in ratios of 50:50, 75:25, 82:18 and 90:10. Dosage rates of 1000 g/ton, 2000 g/ton, 3000 g/ton and 4000 g/ton were investigated at the various reagent ratios. The aim of this test work was to achieve a product CV of 27.80 MJ/kg and combined concentrate yield of fifty percent.

Dewatering was conducted on the combined concentrate of the flotation pilot plant at a fixed membrane pressure of 12 bar and various feed pressures with an Ishigakhi filter press. Dewatering was also conducted on reference material with a Tecnicas Haudrilicas (TH) filter press at various air blow times and feed pressures. The surface moisture results were compared. The proportion of solids in the filter press feed and particle size distribution of the filter press feed influenced the results obtained. The filtration rates of the filter presses were also compared to enable scaling up calculations.

4. RESULTS AND DISCUSSION

The Western Coal Complex ultra fine coal sourced from Rietspruit plant was evaluated this coal will form part of the feed to one of Anglo American's new operations, Phola plant. Two processing routes were evaluated namely sending raw as arising ultra fine coal to a froth flotation plant followed by dewatering of product ultra fines this coal will be at a quality of 27.80 MJ/kg and will form part of the export market. The second processing route is to send raw as arising ultra fines at a product quality of 22.50 MJ/kg directly to dewatering and to blend it into a local Eskom product market.

The coals ability to be recovered by means of froth flotation was evaluated samples (feed, concentrates and tails) were taken from a Jameson pilot plant operation receiving feed (ultra fines) directly from the Rietspruit plant. This plant was later modified by means of installing an oversize protection screen prior to the Jameson pilot plant and samples was taken from the feed (sieve bend underflow), concentrates and tails. The product concentrates from the Jameson pilot plant were dewatered by means of an Ishigakhi filter press. Dewatering was done on product concentrates prior to the installation of the sieve bend and after the installation of the sieve bend. Due to difficult operational conditions at Rietspruit plant sampling from the jameson pilot plant was seized and samples from the underflow of the sieve bend (flotation feed) was taken and send to a laboratory to evaluate ultra fine coal yield and quality by means of using a Denver laboratory flotation cell.

Dewatering results of Goedehoop ultra fine coal was used as a reference since any filter press ability to dewater is not dependant on the coal quality, but only on the particle size distribution. This was done because of the difficult operational conditions at Rietspruit plant. Dewatering was done with the Ishigakhi filter press at Goedehoop plant on froth flotation concentrates. Dewatering was also done with a Technicas Hydrilicas (TH) filter press on Goedehoop froth flotation concentrates to enable a comparison between the two filter presses in terms of dewatering ability. A financial evaluation for the feasible processing route is described in chapter 4.4.

4.1 FLOTATION RESULTS

4.1.1 PILOT PLANT FLOTATION RESULTS PRIOR TO MODIFICATIONS

In total 37 pilot plant tests were performed where the volume ratio of paraffin (collector) and frother were varied in ratios of 50:50, 75:25, 82:18 and 90:10. Dosage rates of 1000 g/ton, 2000 g/ton, 3000 g/ton and 4000 g/ton were investigated at all the various ratios. The highest total product yield was achieved at a dosage rate of 4000 g/ton and a reagent ratio of 82/18 (figure 11).

In general as the recovery increases the selectivity decreases. It is for this reason that the lowest product ash values are obtained at a dosage rate of 1000 g/ton for all ratios tested. The product ash values also increase as the dosage rate increases due to decreased selectivity (Figure 12). The combustible recoveries also increase as the dosage rate increases due to the higher yields achieved at higher dosage rates. Therefore some compromise between yield and product quality must be struck. The most optimal result was achieved at a dosage rate of 1000 g/ton with a reagent ratio of collector to frother of 75:25. At this reagent ratio and dosage rate the product ash percentage was 12.11%, the CV was 28.12 MJ/kg and the total product yield was 34.54%, which is below the targeted yield recovery of 50%.

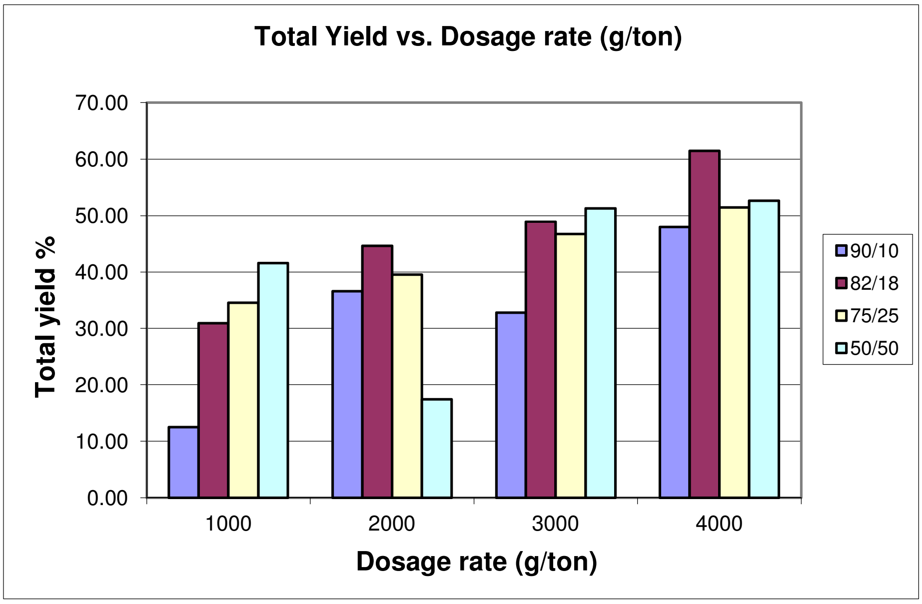

Figure 11: The relationship between yield and dosage rate at various reagent ratios.

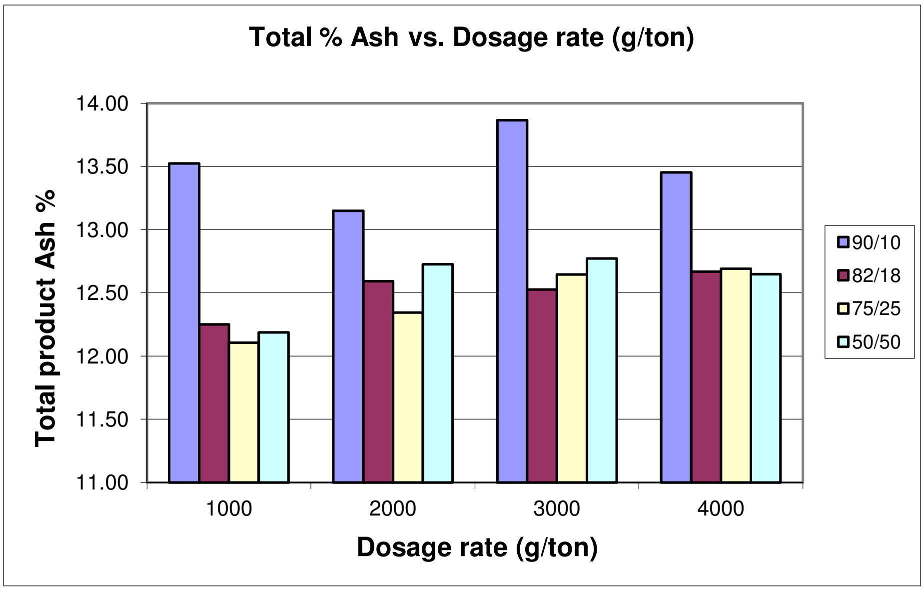

Figure 12: Product quality vs. dosage rate.

4.1.2 LABORATORY SCALE FLOTATION RESULTS

Klipspruit coal that will feed the new Phola plant was investigated to establish its floatability. The aim was to create a product, which had a calorific value of 27.80

MJ/kg (13.5 to 14 % ash) at a minimum recovery of 50%, this would allow for an export market to be utilised. A Denver flotation cell was utilized for the laboratory scale flotation tests. Klipspruit coal was sourced from the Rietspruit plant's modified pilot plant circuit. A slurry sample from the sieve bend underflow was utilized for laboratory scale tests. The table below indicates the reagents and reagent volume ratios tested.

Table 5: Flotation matrix for Klipspruit coal.

Klipspruit Coal flotation matrix.		
Reagent (collector & frother)	Dosage kg/ton	Ratio of Collector : Frother
Paraffin & Ekofol	1,2 & 4	60:40; 75:25; 80:20; 95:5
Paraffin & Centifroth	1	60:40; 75:25; 80:20; 95:5
No collector & Centifroth	1,2 & 4	60:40; 75:25; 80:20; 95:5
Betafroth & betachem	1 &2	95:5

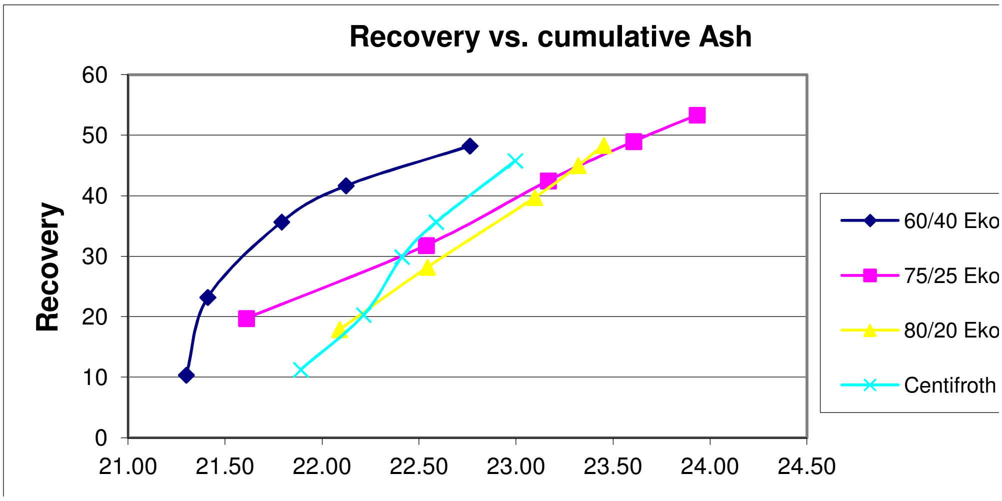

Figure 13: Grade recovery curves for Klipspruit ultra fine coal (ash %).

The results indicated that the mass recoveries (product yield) are all well below 50% (figure 13). The required product quality is 27.80 MJ/kg (13.5 to 14% ash) could not be achieved at any of the tested dosage rates for all the tested reagents (table 5). Therefore it can be concluded that flotation of Klipspruit ultra fine coal was not selective enough to allow for the required product quality (Figure 13). The required quality could not be achieved at 1kg/ton, 2 kg/ton or 4 kg/ton. For this reason a cleaner stage was applied to the primary stage product concentrate. However, due to the complexity involved in a cleaning stage as well as the lower economic viability of successive cleaner stages, this approach was not considered to be acceptable as it is not general practice in the coal industry to apply cleaner stages for thermal coals. In addition, the cleaner stage did not result in the required product CV (Table 6).

Table 6: Results for the cleaner stage on Klipspruit Ultra fine coal.

75/25 Paraffin & Ekofol 2 kg/ton in 1st then 1kg/ton in second stage	Ash %	CV MJ/kg
Feed	23.87	23.53
Concentrate	15.50	26.79
Tailings	27.60	22.08
Overall Yield %	30.83	
Overall combustible recovery %	34.22	

Thus it can be concluded that flotation on Klipspruit material was not selective enough to allow for the required product CV and product yield. It is important to note that this coal can be rendered as part of the export product coal if lower densities for separation are used at the Wemco drum section and dense medium cyclone section. This is not advisable since the yield loss on the coarse coal fractions outweighs the gain of fine coal product. This is due to the low recovery on the fine coal. Other economic factors like the reagent price also affect the economic viability of a ultra fine coal treatment plant this in combination with low yields as a result of lower rank coal directly influences the profit margin. Consequently this will not be catered for in the plant design for the Western Coal Complex (WCC). A summary of all of the results is given in appendix A3.2 up to A3.4.

4.2 DEWATERING RESULTS OF THE ISHIGAKHI FILTER PRESS

4.2.1 INITIAL ISHIGAKHI TEST WORK AT RIETSPRUIT COLLIERY

Dewatering was done at Rietspruit plant on Klipspruit ultra fines with an Ishigakhi filter press on all pilot plant concentrate samples. A fixed membrane pressure of 12 bar, air blow time of two minutes and feed rate of 7.8 m^3/hr was used for all dewatering trials. Three different airflow rates were investigated, namely, 30, 80 and 90 M^3/hr. Dewatering trials on both primary and combined concentrate were conducted.

Due to coarse material feeding the Ishigaki filter press at the Rietspruit plant**,** low moisture values were obtained. However, the moisture results were not representative of the true ultra fine material that will be present in the WCC preparation plant fines circuit. This was due to worn classification cyclone spigots causing oversize material to report to the ultrafine coal fractions. The particle size distribution of the combined concentrate in Figure 14 illustrates this problem clearly, where only 64.64 % of the material is smaller than 106 microns. It was found that the lowest moisture results were achieved at airflow of 80m^3/hr (average moisture 12.31%). Higher cycle times resulted in lower moisture values. The membrane squeeze was kept constant at 12 bar pressure. It must however be mentioned that the maximum achievable air blow pressure was 6 bar due to air supply and air flow problems. Appendix B1 summarises the results obtained.

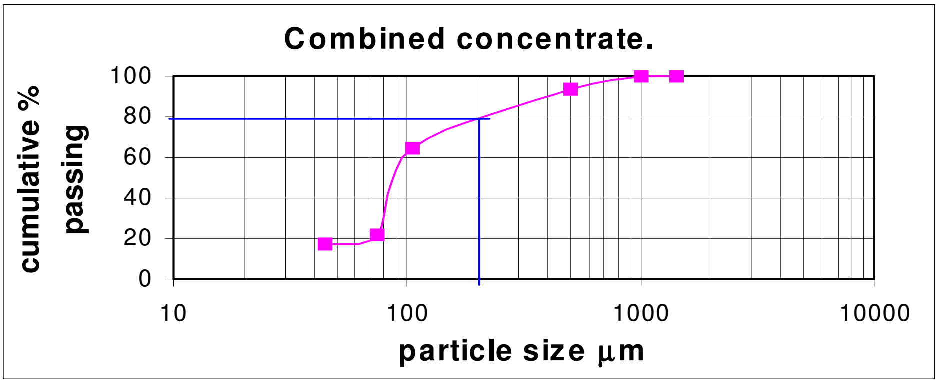

Figure 14: Size distribution of the combined concentrate sampled at the Rietspruit

4.2.2 DEWATERING TRAILS AT RIETSPRUIT AFTER PILOT PLANT MODIFICATIONS

Due to low feed solids and erratic feed conditions, the pilot plant was modified to allow for more representative moisture results. The flotation feed from the sieve bend underflow was pumped to a sump. From the sump the material was transferred to a pilot scale high rate thickener. The thickener underflow fed the Ishigaki filter press. A test matrix for the thickener underflow is given in Appendix B2.

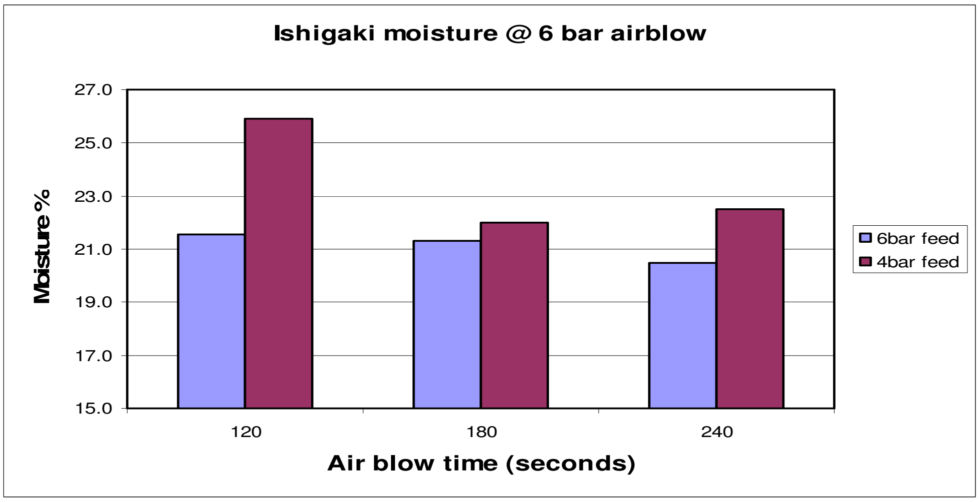

Figure 15: The moisture results of the Ishigakhi filter press at 4 and 6 bar feed pressure.

It was observed that the moisture decreased as the air blow time increased at 6 bar feed pressure. The lowest moisture values were achieved at an air blow time of 240 seconds, an air blow pressure of 6 bar and a feed pressure of 6 bar. The results for 4 bar feed pressure gave higher moisture values (Figure 15). It was observed that a finer particle size distribution (psd) gave higher moisture values. The moisture values are given in Table 7 and in the psd graphs in Figure 16. Without filtration this material would typically be pumped to a slimes dam to settle at a 70 percent moisture value.

Table 7: Test conditions and moisture results.

Run NO.	Feed Pressure (Bar)	Feed SG	Cloth state -	Cake Blow (Bar)	Cake Blow (Secs)	Surface moisture %
30	4	1.178	Clean	6	120	26.1
32	4	1.178	Used 3X	6	180	22.0
33	6	1.176	Clean	6	120	20.1

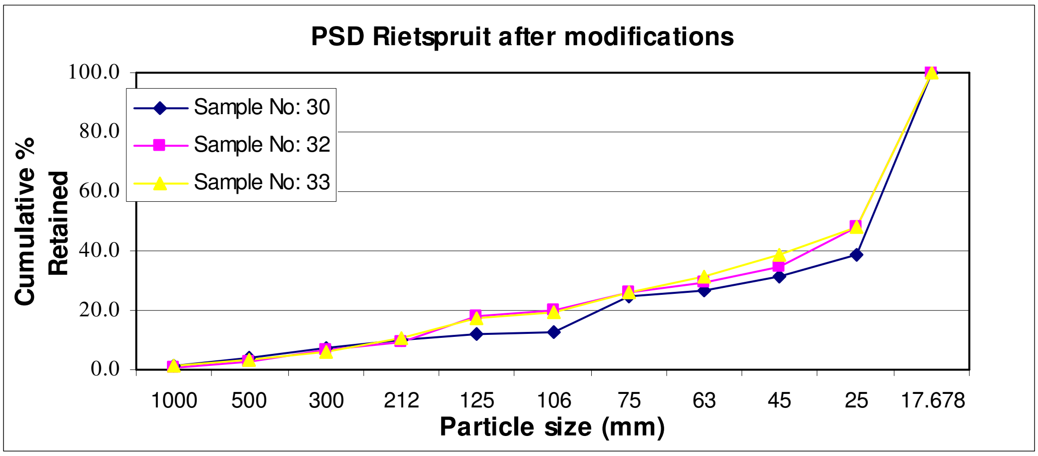

Figure 16: Particle size distribution of the Rietspruit samples after plant modifications.

4.2.3 DEWATERING TRIALS ON GOEDEHOOP ULTRA FINES WITH THE ISHIGAKHI FILTER PRESS

Dewatering trials with the Ishigakhi filter press were performed at Goedehoop Colliery. Since Goedehoop ultra fine material serves as a reference, the data was utilized to enable comparison and effective assessment of the two available dewatering options. During the test work the following parameters were utilized:

- Time allowed for filter plate membrane squeeze,
- Air blow pressure

- Cake blow time allowed per filter press cycle (Table 8).

Table 8: Test parameters for dewatering trials on reference material.

Description	Value
Squeeze time (minutes)	**1 and 1.5**
Air blow pressure (bar)	**4 and 8**
Cake blow time (minutes)	**1, 2,3 and 4**

It was observed that higher air pressure and longer cycle times gave lower moisture values. A further important parameter was found to be the membrane squeeze pressure**.** Generally a high squeeze pressure leads to a high moisture reduction, so a squeeze pressure of 14-bar was utilised. After the squeeze cycle, the membrane pressure reduces and the remaining pressure keeps the filter cake in position for the air blow cycle. This pressure drop is a function of the initial squeeze pressure, the cake permeability and cake volume. A summary of the results is given in Appendix B3.

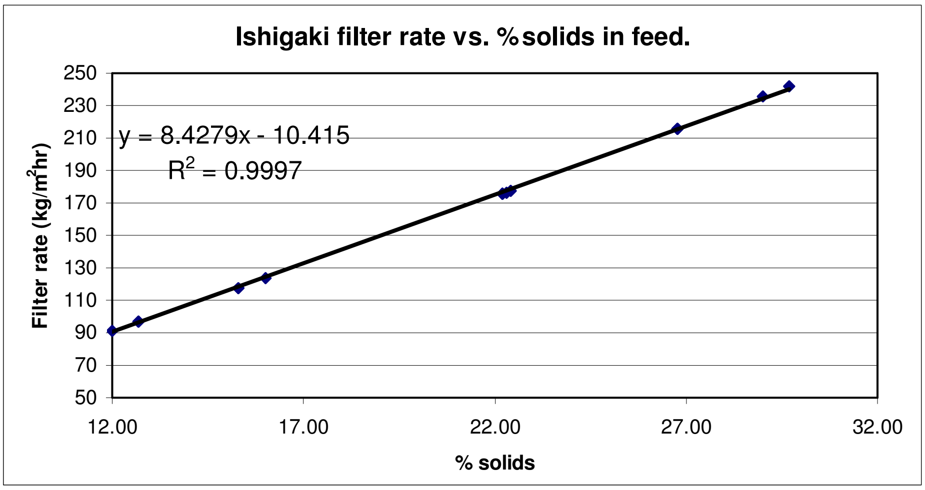

Figure 17: Filter rate versus percentage solids in the filter press feed.

The filter rate was found to be a function of the percentage solids in the feed. The filter rate increased linearly with the percentage solids in the feed (Figure 17). It was also observed that the machine settings of the Ishigakhi filter press play a significant role in the dewatering process. A summary of the different machine settings is given in Table 9. From the data summarized in Appendix B1, B2 and B3, it may be concluded that higher membrane squeeze pressure, air blow pressure and feed pressure result in lower moisture values. A photograph illustrating the Ishigakhi filter press is shown in figure 18.

Table 9: Various filter press settings on reference material and Rietspruit plant after plant modifications.

Description of parameter	**Goedehoop settings used**	**Rietspruit plant after modifications settings**
Membrane squeeze pressure (bar)	14	12
Air blow pressure (bar)	4 & 8	6
Feed pressure (bar)	6	4 & 6

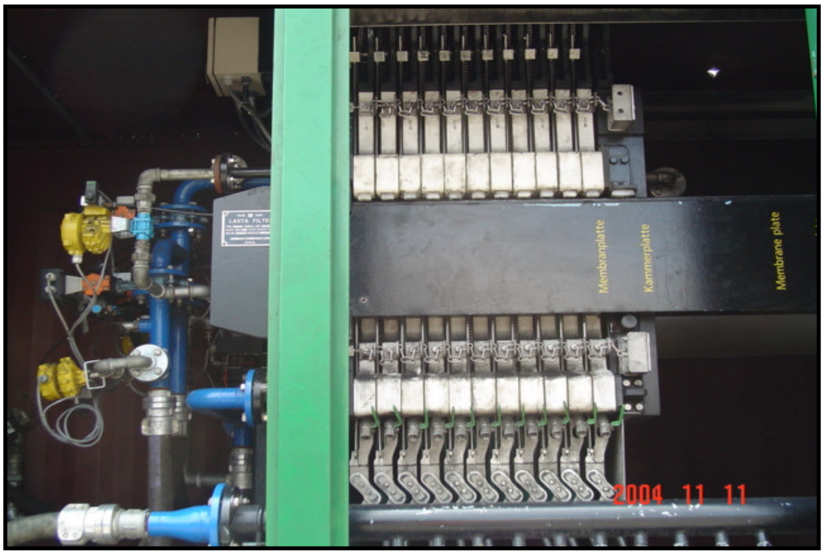

Figure 18: The Ishigakhi filter press (Seprotech).

4.3. DEWATERING RESULTS OF THE TECHNICAS HIDRAULICAS FILTER PRESS

4.3.1 TECHNICAS HIDRAULICAS (TH) TEST WORK ON GOEDEHOOP ULTRA FINE COAL.

Goedehoop plant flotation concentrate thickener underflow was sampled for dewatering tests with the Technicas Hidraulicas filter press. A photograph in Figure 19 indicates the pilot plant Technicas Hidraulicas.

Figure 19: The Technicas Hidraulicas (TH) pilot plant filter press.

A series of trials were carried out on Goedehoop ultra fine coal concentrate with the Technicas Hidraulicas (TH) and it was observed that longer air blow time led to lower surface moisture values. However, although surface moisture was significantly reduced with longer air blow time, this resulted in lower filtration rates (Figure 20). It is postulated that this could be due to variations in particle size distribution and mineral content of the flotation product possibly influenced by the variation in the amount of collector used.

The most optimal results were achieved at 7 bar feed and 4 bar air blow pressure with average surface moisture of 17 %. Higher air blow pressures resulted in cake cracking (at 8 bar air blow). Since air follows the route of the least resistance, the air will travel through the crack instead of through the filter cake. In these instances, the filter cake moisture ranged between 20 and 22% (van der Walt and Terblanche 2005). It is clear that both dewatering filter presses investigated are capable of reducing the surface moisture to equal to or less than 20% which is required for both the export and inland market.

The filtration rate (dry kg/m^{2}hr) was investigated and it was observed that a higher percentage solids in the feed resulted in a higher filtration rate at shorter air blow times (Figure 21). An increased amount of fines in the feed resulted in lower filter rates and higher surface moistures (e.g. Test 19 which had 50% of feed material below 10 microns indicated by figure 20). The normalized data for filtration rate is indicated in Figure 22. The filtration rate of the TH is lower than the filtration rate of the Ishigakhi filter press (figure 17) and this rate would directly impact on the capital requirement in terms or required dewatering surface area. Lower filter rates requires larger surface areas to achieve a specified throughput compared to higher filtration rates that would require a smaller surface area for the same specified throughput.

A finer particle size distribution leads to higher surface moistures and lower filtration rates (blue and red graphs in figure 20). The red data points represent 50 % of feed material below 10 microns and the blue data points represent 30 % of feed material below 10 microns. Thus it can be concluded that the size fraction below 10 microns plays a significant role in the filter rate and surface moisture.

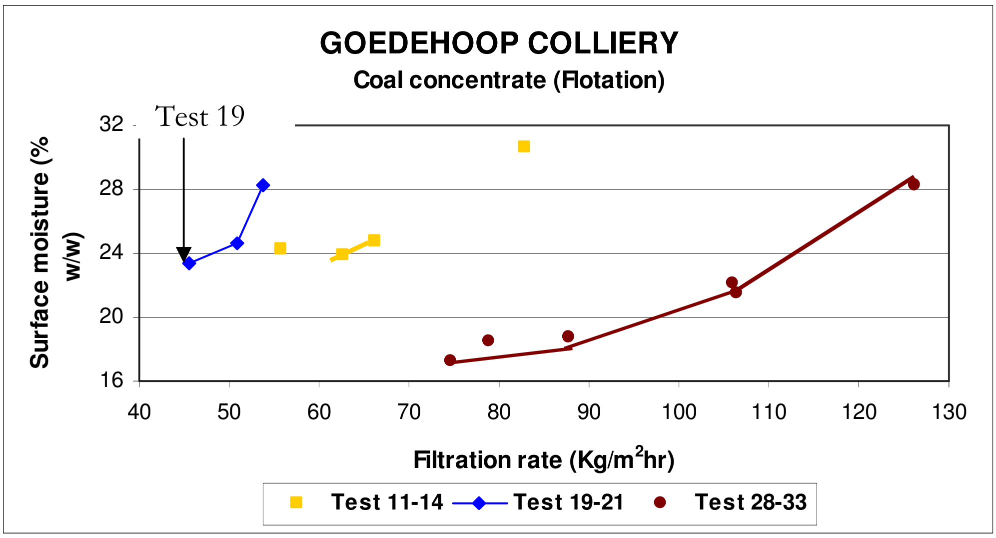

Figure 20: Surface moisture versus filtration rate.

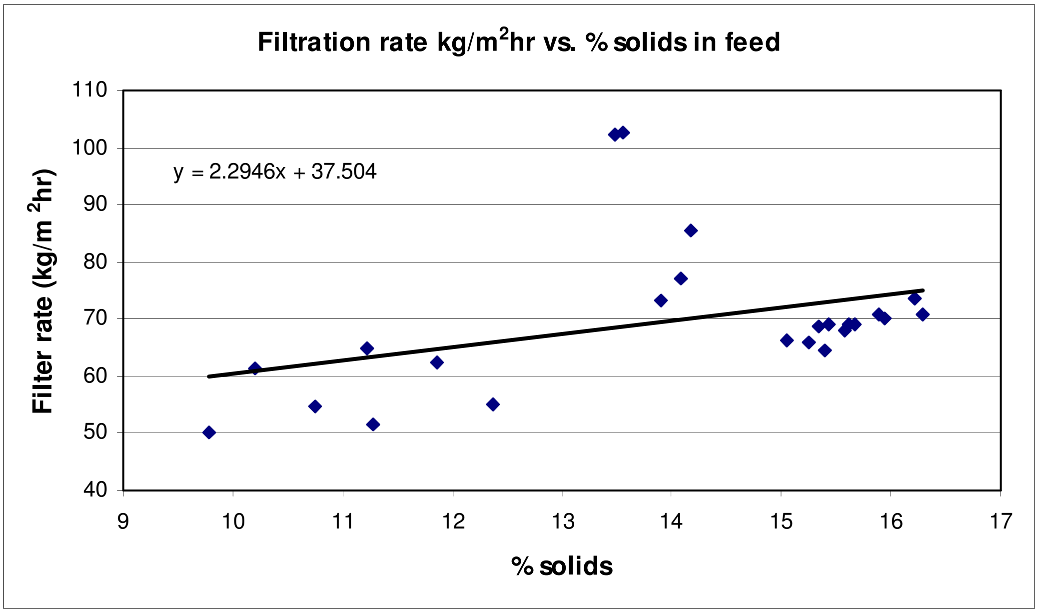

Figure 21: TH filter rate versus percentage solids in the feed (flotation concentrate).

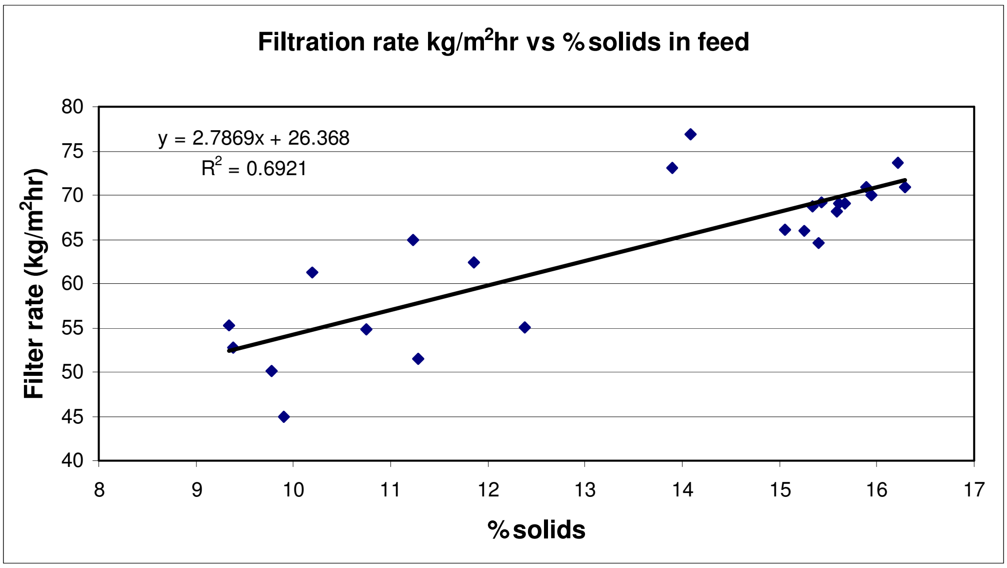

Figure 22: TH normalized data for filter rate versus percentage solids.

4.4 FINANCIAL EVALUATION

From the results listed in table 6, it is clear that Klipspruit ultra fines are not recoverable by means of froth flotation because of the poor quality of the raw (unwashed) ultra fines. This is a direct result of the low product recovery. A capital investment of four hundred million rand is required for both flotation and dewatering, with a running cost of three hundred thousand rand per month results in a payback period exceeding 5 years. Therefore only dewatering on raw as arising fines will be considered and this will then be selectively blended into the Eskom market. This option will lead to a payback period of 1 year and 2 months, which is a more viable option in terms of the reduced amount of capital required for only installing a filter press plant of 80 million Rand compared to 400 million Rand for a Jameson froth flotation plant with a filter plant.

However dewatering of ultra fines with both dewatering filter presses was feasible, as shown in table 7, 8 and table 9. These results indicate that a new combined washed plant would be able to sell 'unwashed ultra fines' as 'dewatered arising ultra fines' locally to Eskom to increase the tonnages of products being used for power generation. This process could generate 76.5 million Rand per annum (Table 10) and, as such, it would have significant impact on the revenue gained of coal products sold to the local Eskom market. A blending model satisfying Eskom to product specifications is shown below. This model also satisfies the particle size requirement for Eskom.

Table 10: Blending percentages for reaching Eskom quality

	Wemco	Cyclone	Spirals	Ultra fines	Total
Blending %	45.00	41.00	9.00	4.62	99.62
Product Quality (Mj/kg)	23.60	24.22	22.00	20.00	23.45
Surface Moisture %	6.00	6.20	15.00	20.00	7.50

The Phola plant which is situated 60 kilometers from Klipspruit and Goedehoop Mines, in the region of Witbank could utilize filtration to dewater raw as arising fines and blend them into the Eskom product. This plant receives a quarter of Klipspruit coal as run-of-mine product. This is then crushed and stockpiled as part of the feed to the plant stream, some of which go for export and some to the Power generating facility, Eskom.

This dissertation research shows that the Phola plant would be able to sell untreated, dewatered, ultra fines as part of the Eskom product. The product would be produced by means of blending the larger sized middling products with the now de-watered ultra fine coal. This would serve to increase the marketable tonnages from Phola Plant to Eskom, and provide a means whereby an otherwise useless sized coal fraction would be of commercial value to industry.

The potential income calculation is summarised in Table 11. A net profit of 76.5 million rand per annum is achieved. All ROM fine coal is sold as product. This financial study is based on the assumption of two hundred Rand per ton of Eskom quality saleable coal. To date, the results of this research has led to the installation of three dewatering units in three other collieries

Table 11: Marketing of as arising ultra fine coal to the inland market (Eskom).

Description	**Total**
Tons produced per annum	450000
Yield	100%
Capital cost	80000000
10 % Interest on capital	8000000
Ops expenditure per annum	13500000
R Income per ton	200
Income per annum	90000000
Payback period years	1.13

5. CONCLUSIONS

This research program was undertaken to establish the most effective means of beneficiating ultra fine coal from one opencast colliery, Klipspruit Colliery, Witbank. In the absence of finding a best process for beneficiation, the research has established that de-watering using two types of presses can be achieved to varying extents to produce marketable products. Specifically, the results have lead to the following conclusions:

1. The **desired quality and yield** that was required from Klipspruit ultra fine coal after beneficiation using froth flotation (i.e. yield of 50% and product quality of 27.80 MJ/kg) could not be achieved (see section 4.1).

2. For this reason froth flotation on Klipspruit ultra fine coal **did not prove economically feasible as it could not be optimised.** The highest quality achieved in froth flotation yielded a calorific value of only 26.79 MJ/kg (table 6).

3. In terms of **de-watering**, the most optimal moisture content on the reference ultra fine slurry material (Goedehoop coal) was obtained with the **Ishigaki** filter press. This was achieved at a membrane pressure of 14 bar and an air blow pressure of 8 bar. This moisture result was 15.9 % from an original as-received moisture content of 70 %. The moisture values decreased as the total cycle time increased. Higher-pressure settings on the Ishigaki press lead to even lower moisture values.

4. Dewatering results conducted on the **Tecnicas Hidraulicas filter press** at Goedehoop using ultra fine slurry material gave moisture values ranging between 17.3% and 30.6% depending on the particle size distribution and machine settings. The most optimal surface moisture results for the Tecnicas Hidraulicas press were achieved at a 4 bar air blow for 360 seconds.

5. Surface moisture and filtration rates were found to be dependent on the particle size distribution. **Finer particle size distributions** lead to higher surface moistures and lower filtration rates (kg/m^2hr).

6. The **filtration rate for the Ishigaki press was greater** than that of the Tecnicas Hidraulicas press thus implying a significant difference in filter area requirement for the Tecnicas Hidraulicas compared to the Ishigaki press. This implies that there would be greater capital expenditure required to achieve the same extraction rate using this process relative to the Tecnicas Hidraulicas filter press.

7. In terms of **the up-scaling of the de-watering processes**, the Ishigakhi filter press is recommended to be the most economical due to a higher dewatering capacity for a smaller filter area. On the basis of this research three up-scaled de-watering units have been installed successfully on other collieries.

8. The investigation into the **potential income of marketing the beneficiated Western Coal Complex ultrafine coal** indicated that the capital investment to undertake de-watering for export coals would not be feasible. The payback period for export equipment exceeded 5 years. On the other hand, calculations based upon processes set for the inland market indicated a payback period of 1 year and 2 months for both processes. The **latter option is there the more financially viable one**.

9. From the **techno-economic evaluation study,** it was found the Phola plant would be able to sell dewatered untreated ultra fines economically as part of the Eskom product, as produced by means of blending methods using either one of the de-watering process options.

10. The **final choice between the two de-watering processes** would depend upon the amount of capital that the company principles would wish to spend as both are viable. The TH is the more expensive of the two by approximately 6 million Rand for the same rate of throughput. **Both options however, would achieve**

the desired moisture contents in the final product. This step has the capacity to generate 76.5 million Rand income per annum for the Phola Plant in terms of additional saleable product.

11. This **step in the value chain** would lead to the need for lower mined tonnages of coal required for power generation via Eskom thereby increasing coal-fired energy sustainability, and it would result in smaller discard rehabilitation facilities leading to reduced acid mine drainage.

6. RECOMMENDATIONS

This research has led to recommendations regarding the suitability of froth flotation on ultra fine Klipspruit coal. The determination of moisture values on ultra fine coal post dewatering and the overall economics of dewatering and pre-cleaning of ultra fines lead to the following recommendations:

1. Although froth flotation is not economically viable for the Klipspruit opencast ultra fine coal, it may yet be feasible for future underground sources. This should be investigated.

2. Where moisture values play a significant role in the economics of export coal, it is recommended that the Ishigaki filter press should be used in preference to the Tecnicas Hidraulicas press. The efficiency of the Ishigaki press is dependant upon the specific settings of the operating parameters. This would require onsite trials at each potential location.

3. For other non-export-related coal applications, the overall economics should be considered prior to making a commercial decision on which filter press would be used for the dewatering of ultra fine coal.

4. It is recommended that coal mineralogy (QEMSEM) and particle size be investigated prior to future test work in order to establish liberation potential of the mineral matter and particle size distribution. De-sliming with a classification cyclone may then permit further effective upgrading of Klipspruit ultra fines coal.

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APPENDIX A1: INITIAL FLOTATION TESTWORK

Test number	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**
Feed Ash %	31.90	32.20	31.20	32.00	54.00	57.90	50.90	53.70	50.70	49.70
Primary concentrate Ash %	12.30	12.40	12.70	12.80	17.60	16.10	13.50	16.90	15.00	15.00
Primary tails ash %	43.10	48.80	40.40	46.20	53.10	62.20	66.90	63.30	62.60	64.30
Secondary concentrate Ash %	11.40	12.20	12.90	12.00	14.20	16.70	15.50	15.50	17.20	14.30
Final tails ash %	45.90	36.30	50.60	53.50	67.20	62.00	63.10	65.00	70.70	62.10
Overall product ash %	12.19	12.72	12.77	12.65	13.85	16.07	13.04	16.74	15.69	15.09
Primary Yield %	36.36	45.60	33.21	42.51	-2.54	9.33	29.96	20.69	25.00	29.61
Secondary Yield %	8.12	-51.87	27.06	17.59	26.60	-0.44	-7.98	3.43	15.14	-4.60
Overall Yield %	41.53	17.39	51.28	52.63	24.74	8.93	24.37	23.41	36.36	26.38
Primary Combustible Recovery %	46.83	58.92	42.14	54.52	-4.54	18.59	52.79	37.13	43.10	50.04
Secondary Combustible Recovery %	12.64	-88.94	39.54	28.77	48.67	-0.97	-20.38	7.91	33.52	- 11.05
Overall combustible recovery %	53.55	22.39	65.02	67.61	46.34	17.80	43.16	42.11	62.17	44.53
Test number	**11**	**12**	**13**	**14**	**15**	**16**	**17**	**18**	**19**	**20**
Feed Ash %	50.10	44.60	33.90	26.70	27.50	26.90	27.00	25.00	25.90	23.40
Primary concentrate Ash %	15.60	16.80	12.90	12.30	12.50	12.80	12.60	12.80	12.50	13.00
Primary tails ash %	57.40	56.10	48.30	32.90	33.70	33.60	32.00	30.00	33.70	30.00
Secondary concentrate Ash %	14.80	15.20	11.90	10.80	11.90	12.30	10.50	12.20	12.60	12.10
Final tails ash %	60.10	51.00	56.40	34.40	37.40	39.40	33.60	35.00	38.70	40.50
Overall product ash %	15.42	17.64	12.69	12.11	12.34	12.64	12.25	12.59	12.52	12.67
Primary Yield %	17.46	29.26	40.68	30.10	29.25	32.21	25.77	29.07	36.79	38.82
Secondary Yield %	5.96	-14.25	18.20	6.36	14.51	21.40	6.93	21.93	19.16	36.97
Overall Yield %	22.38	19.18	51.48	34.54	39.51	46.72	30.91	44.62	48.90	61.44
Primary Combustible Recovery %	29.54	43.95	53.60	36.01	35.30	38.42	30.86	33.80	43.45	44.09
Secondary Combustible Recovery %	11.92	-27.52	31.02	8.45	19.28	28.27	9.12	27.51	25.25	46.43
Overall combustible recovery %	37.94	28.52	67.99	41.42	47.77	55.83	37.16	52.01	57.73	70.05
Test number	**21**	**22**	**23**	**24**	**25**	**26**	**27**	**28**	**29**	**30**
Feed Ash %	29.60	24.60	32.50	27.90	27.60	27.70	28.30	28.50	28.10	25.50
Primary concentrate Ash %	12.90	13.40	13.50	13.80	14.10	13.90	16.00	14.10	14.00	16.10

Primary tails ash %	33.40	29.60	39.10	33.80	38.10	33.60	34.80	33.10	30.60	31.60
Secondary concentrate Ash %	11.60	11.80	15.20	12.90	13.60	11.90	11.70	11.50	11.60	12.00
Final tails ash %	31.90	31.20	41.60	41.20	44.10	35.60	32.80	31.10	36.30	33.80
Overall product ash %	13.53	13.15	13.86	13.45	14.00	13.57	16.94	15.32	12.64	15.55
Primary Yield %	18.54	30.86	25.78	29.50	43.75	29.95	34.57	24.21	15.06	39.35
Secondary Yield %	-7.39	8.25	9.47	26.15	19.67	8.44	-9.48	- 10.20	23.08	10.09
Overall Yield %	12.52	36.57	32.81	47.93	54.82	35.86	28.37	16.48	34.66	45.47
Primary Combustible Recovery %	22.93	35.45	33.04	35.27	51.91	35.67	40.51	29.09	18.01	44.32
Secondary Combustible Recovery %	-9.81	10.33	13.19	34.40	27.46	11.20	-12.84	- 13.50	29.39	12.98
Overall combustible recovery %	15.38	42.12	41.87	57.54	65.11	42.87	32.87	19.51	42.11	51.55

APPENDIX A1: INITIAL FLOTATION TESTWORK

Test number	31	32	33	34	35	36	37
Feed Ash %	29.60	27.60	26.90	24.80	24.90	30.90	25.50
Primary concentrate Ash %	15.00	12.90	14.10	12.70	12.60	12.40	13.10
Primary tails ash %	32.90	32.80	32.90	27.80	31.50	35.90	33.40
Secondary concentrate Ash %	32.40	11.50	13.00	10.80	12.60	12.10	10.70
Final tails ash %	33.60	31.50	32.50	30.60	31.20	41.50	30.60
Overall product ash %	27.54	13.22	14.15	12.01	12.60	12.28	13.78
Primary Yield %	18.44	26.13	31.91	19.87	34.92	21.28	38.92
Secondary Yield %	58.33	-6.50	-2.05	14.14	-1.61	19.05	- 14.07
Overall Yield %	66.01	21.33	30.52	31.20	33.87	36.27	30.32
Primary Combustible Recovery %	22.26	31.44	37.50	23.06	40.64	26.97	45.39
Secondary Combustible Recovery %	58.77	-8.56	-2.66	17.47	-2.06	26.12	- 18.87
Overall combustible recovery %	67.95	25.57	35.84	36.51	39.42	46.05	35.09

APPENDIX A1: INITIAL FLOTATION TESTWORK CONDITIONS

Test No.	**1**	**2**	**3**	**4**	**5 (5#)**	**6 (5#)**	**7 (5#)**	**8 (5#)**
Date	4-Oct-04	5-Oct-04	6-Oct-04	7-Oct-04	29Oct04	29-Oct-04	29Oct04	29Oct04
Avg. Feed rate m^3/hr	75.00	75.00	75.00	75.00	75.00	75.00	75.00	75.00
froth depth primary/sec. mm	200/150	200/150	200/150	200/150	200/150	200/150	200/150	200/150
Airflow m^3/hr	14.76	14.76	14.76	14.76	14.76	14.76	20.00	14.76
Primary Cell feed sg	1.03	1.03	1.02	1.02	1.02	1.02	1.02	1.02
Reagent Ratio IP/440	50/50	50/50	50/50	50/50	50/50	50/50	50/50	50/50
Dosage rate g/ton feed	1000.00	2000.00	3000.00	4000.00	3000.00	2000.00	1000.00	4000.00
System Residence time (min)	4.10	4.10	4.10	4.10	4.10	4.10	4.10	4.10
Test No.	**9 (5#)**	**10 (5#)**	**11 (5#)**	**12 (5#)**	**13**	**14**	**15**	**16**
Date	29Oct04	29Oct04	29Oct04	29Oct04	4Nov04	4Nov04	4Nov04	4Nov04
Avg. Feed rate m^3/hr	75.00	75.00	75.00	75.00	75.00	75.00	75.00	75.00
Froth depth primary/sec. mm	200/150	200/150	200/150	200/150	200/150	200/150	200/150	200/150
Airflow m^3/hr	14.76	14.76	14.76	14.76	14.76	14.76	14.76	14.76
Primary Cell feed sg	1.02	1.02	1.02	1.02	1.025	1.03	1.03	1.03
Reagent Ratio IP/440	75/25	75/25	75/25	75/25	75/25	75/25	75/25	75/25
Dosage rate g/ton feed	4000.00	3000.00	2000.00	1000.00	4000.00	1000.00	2000.00	3000.00

System Residence time (min)	4.10	4.10	4.10	4.10	4.10	4.10	4.10	4.10
Test No.	**17**	**18**	**19**	**20**	**21**	**22**	**23**	**24**
Date	4Nov04	4Nov04	4Nov04	4Nov04	11Nov04	11Nov04	11Nov04	11Nov04
Avg. Feed rate m^3/hr	75.00	75.00	75.00	75.00	75.00	75.00	75.00	75.00
Froth depth primary/sec. mm	200/150	200/150	200/150	200/150	200/150	200/150	200/150	200/150
Airflow m^3/hr	14.76	14.76	14.76	14.76	14.76	14.76	14.76	14.76
Primary Cell feed sg	1.03	1.03	1.03	1.03	1.03	1.03	1.03	1.03
Reagent Ratio IP/440	82/18	82/18	82/18	82/18	90/10	90/10	90/10	90/10
Dosage rate g/ton feed	1000.00	2000.00	3000.00	4000.00	1000.00	2000.00	3000.00	4000.00
System Residence time (min)	4.10	4.10	4.10	4.10	4.10	4.10	4.10	4.10

APPENDIX A1: INITIAL FLOTATION TESTWORK CONDITIONS CONTINUED

Test No.	25	26	27	28	29	30	31	32
Date	11Nov04	11Nov04	12Nov04	12-Nov-04	12-Nov-04	12-Nov-04	12-Nov-04	12-Nov-04
Avg. Feed rate m^3/hr	85.00	65.00	80.00	75.00	75.00	75.00	75.00	75.00
Froth depth primary/sec. mm	200/150	200/150	200/150	350/150	150/150	100/150	200/250	200/350
Airflow m^3/hr	14.76	14.76	14.76	14.76	14.76	14.76	14.76	14.76
Primary Cell feed sg	1.03	1.03	1.02	1.02	1.02	1.02	1.02	1.02
Reagent Ratio IP/440	82/18	82/18	82/18	90/10	90/10	90/10	90/10	90/10
Dosage rate g/ton feed	3000.00	3000.00	3000.00	3000.00	3000.00	3000.00	3000.00	3000.00
System Residence time (min)	3.62	4.74	3.85	4.10	4.10	4.10	4.10	4.10

Test No.	33	34	35	36	37
Date	12Nov04	16Nov04	16Nov04	16Nov04	16Nov04
Avg. Feed rate m^3/hr	75.00	75.00	75.00	75.00	75.00
Froth depth primary/sec. mm	200/100	200/150	200/150	200/150	200/150
Airflow m^3/hr	14.76	14,76	19.68	24.60	9.84
Primary Cell feed sg	1.02	1.02	1.02	1.02	1.02
Reagent Ratio IP/440	90/10	75/25	75/25	75/25	75/25
Dosage rate g/ton feed	3000.00	2000.00	2000.00	2000.00	2000.00
System Residence time (min)	4.10	4.10	4.08	4.10	4.10

APPENDIX A2 : FLOTATION RESULTS AFTER PLANT MODIFICATIONS

Test number	AT1	AT2	AT3	AT4
Feed Ash %	35.00	35.70	35.00	45.90
Primary concentrate Ash %	13.30	13.50	13.30	16.60
Primary tails ash %	45.70	44.40	42.20	53.50
Secondary concentrate Ash %	14.20	13.70	14.40	13.70
Final tails ash %	51.30	50.30	52.40	55.20
Overall product ash %	13.51	13.56	13.79	16.20
Primary Yield %	33.02	28.16	24.91	20.60
Secondary Yield %	15.09	16.12	26.84	4.10
Overall Yield %	43.13	39.74	45.07	23.85
Primary Combustible Recovery %	44.05	37.88	33.23	31.75
Secondary Combustible Recovery %	23.85	25.02	39.75	7.60
Overall Combustible recovery %	57.39	53.42	59.77	36.94

APPENDIX A 3.2: LABORATORY SCALE FLOTATION TESTS AT A DOSAGE RATE OF 1 KG/TON

Reagent Ratio	75/25 Paraffin & Ekofol			
Time	% Solids	Cumulative Recovery	Cumulative Ash %	Cumulative CV (Mj/kg)
10	25.52	14.79	18.18	25.75
20	95.08	22.99	18.75	25.53
40	44.66	30.37	19.06	25.41
60	19.45	37.79	21.79	24.34
100	21.73	45.33	25.10	23.05
Tails	6.49	100.00	35.46	19.02

A detailed laboratory froth flotation procedure is given below this procedure was used for all laboratory froth flotation test conditions:

1. The weight of filtration paper and sample containers were determined. Each filtration paper and sample container were labeled prior to use.
2. The feed slurry sampled was diluted from 32.1% solids to 7% solids by means of the addition of dilution water.
3. The diluted feed slurry was preconditioned for 50 seconds prior to reagent addition.
4. The liquid density of reagent consisting of a pre-determined volumetric ratio of collector and frother was determined by means of determining the weight of 100 milliliters of reagent.
5. The required amount of reagent was added to the preconditioned slurry at various dosage rates (1kg/ton, 2kg/ton or 4kg/ton).
6. One minute was allowed for froth formation per test prior to sampling of froth thereafter froth was collected every ten seconds in each sample container.

7. The tails was collected in a separate sample container
8. The mass of all slurry samples collected during the test was measured and recorded.
9. All slurry samples were filter pressed and dried at fifty degrees Celsius until all surface moisture was driven off. Filter papers prepared in step 1 was utilized during this process.
10. The mass of the dry solids was used to determine cumulative percentage of mass recovery.
11. The ash percentage and calorific value for the dry samples in step ten was measured. This data was used to determine the cumulative ash% and CV value.

Recovery grade curves were drawn from the data gathered in step 11.

The reason the above laboratory scale froth flotation methodology was included is because froth flotation response time and froth formation is unique for each coal source. The Anglo American thermal coal procedure had to be adapted to standardize results in order to maintain reproducibility. This procedure allowed for reproducible and comparative lab results.

APPENDIX A 3.3: LABORATORY SCALE FLOTATION TESTS AT A DOSAGE RATE OF 2 KG/TON

40% Ekofol & 60% Paraffin				
Time	% Solids	Cumulative Recovery	Cumulative Ash %	Cumulative CV
10	16.38	10.34	21.3	24.67
20	15.26	23.19	21.41	24.54
40	12.83	35.66	21.79	24.45
60	9.33	41.64	22.12	24.33
100	6.60	48.23	22.76	24.08
Tails	4.28	100.00	40.64	15.426
25 % Ekofol & 75 % paraffin				
Time	% Solids	Cumulative Recovery	Cumulative Ash	Cumulative CV
10	17.28	19.71	21.61	25.57
20	14.29	31.78	22.54	25.01
40	11.80	42.47	23.17	24.50
60	8.27	48.95	23.61	24.25
100	5.60	53.32	23.93	24.10
Tails	3.84	100	40.01	18.21
20 % Ekofol & 80 % Paraffin				
Time	% Solids	Cumulative Recovery	Cumulative Ash	Cumulative CV
10	15.81	17.88	22.09	24.82
20	13.67	28.17	22.54	24.65
40	11.50	39.68	23.10	24.44
60	8.85	44.99	23.32	24.34
100	7.41	48.33	23.45	24.29
Tails	4.16	100.00	39.77	17.67
Pure frother Centifroth				
Time	% Solids	Cumulative Recovery	Cumulative Ash %	Cumulative CV
10	14.01	11.26	21.89	24.17
20	12.57	20.28	22.21	23.97
40	10.68	29.85	22.41	24.01
60	10.14	35.65	22.59	24.07
100	8.87	45.76	23.00	23.86
Tails	4.84	100.00	39.65	17.92

Appendix A3.4: Laboratory scale flotation tests at a dosage rate of 4 kg/ton

40 % Ekofol & 60 % paraffin				
Time	% Solids	Cumulative Recovery	Cumulative Ash %	Cumulative CV
10	20.09	13.28	21.60	25.12
20	18.84	27.21	22.78	24.64
40	18.11	44.72	24.59	23.92
60	15.19	50.70	25.08	23.74
100	12.79	59.52	25.80	23.46
Tails	4.60	100.00	36.36	19.32
25 % Ekofol & 75 % Paraffin				
Time	% Solids	Cumulative Recovery	Cumulative Ash %	Cumulative CV
20	19.16	17.34	22.15	24.2
40	17.09	33.28	23.52	23.82
60	14.01	45.93	24.40	23.45
100	12.52	51.28	24.63	23.34
Tails	4.44	100.00	40.01	17.01
20 % Ekofol & 80 % Paraffin				
Time	% S	Cumulative Recovery	Cum Ash	Cum CV
10	16.29	17.19	22.10	24.51
20	4.78	22.50	22.74	24.36
40	10.59	33.98	23.50	24.14
60	12.15	40.37	24.13	23.92
100	5.71	45.71	24.54	23.77
Tails	3.99	100.00	43.79	16.17
5 % Ekofol & 95 % Paraffin				
Time	% S	Cumulative Recovery	Cum Ash	Cum CV
20	16.6	22.71	21.53	24.81
40	15.11	30.40	21.37	24.52
60	18.95	35.08	21.14	23.30
100	12.68	40.43	21.25	23.42
Tails	4.06	100.00	38.69	17.65
100 % frother (Centifroth)				
Time	% S	Cumulative Recovery	Cum Ash	Cum CV
20	11.02	12.79	25.48	22.83
40	10.13	24.77	25.52	22.77
60	10.06	32.70	25.55	22.69
100	9.95	42.89	25.64	22.59
Tails	5.67	100.00	39.07	17.10

APPENDIX B1: INITIAL ISHIGAKI DEWATERING TESTWORK AT RIETSPRUIT PLANT

Name	Description	Fill time minutes	Total cycle time minutes	Cloth used	Airflow m3/hr	Feed RD	% Solids in feed	Total Moisture%	CV (MJ/kg)	Ash %
Test1	Combined concentrate	3.41	10.13	0	30	1.09	23.26939	14.50	28.24	12.90
Test2	Combined concentrate	4.14	11.13	0	30	1.08	20.87542	14.70	28.77	12.00
Test3	Combined concentrate	6.21	13.10	0	30	1.06	15.95197	13.30	28.83	12.00
Test4	Combined concentrate	6.3	13.74	0	30	1.05	13.41991	20.80	29.08	11.60
Test5	Combined concentrate	6.11	13.29	0	90	1.05	13.41991	13.25	28.44	12.40
Test6	Combined concentrate	6.41	14.02	1	90	1.05	13.41991	16.55	28.65	12.50
Test7	Combined concentrate	7.11	14.32	2	90	1.05	13.41991	17.85	28.84	12.00
Test8	Combined concentrate	7.13	14.36	0	90	1.05	13.41991	16.05	28.75	11.90
Test9	Combined concentrate	6.37	13.59	1	90	1.06	15.95197	12.75	28.76	12.20
Test10	Combined concentrate	6.45	13.52	0	90	1.05	13.41991	13.40	27.59	12.60
Test11	Combined concentrate	3.16	10.12	1	80	1.07	18.4367	16.30	28.67	12.80
Test12	Combined concentrate	3.18	10.11	2	80	1.07	18.4367	12.20	27.78	12.60
Test16	Primary Concentrate	7.33	14.14	2	80	1.04	10.83916	12.80	26.85	15.40
Test17	Primary Concentrate	6.59	14.42	0	80	1.04	10.83916	13.90	27.19	15.10
Test18	Primary Concentrate	8.09	16.09	1	80	1.04	10.83916	13.50	26.54	15.40
Test19	Primary Concentrate	8.14	16.09	0	80	1.04	10.83916	10.90	26.93	15.20

Test20	Primary Concentrate	8.24	16.57	0	80	1.02	5.525847	13.00	28.6	13.20
Test21	Primary Concentrate	7.18	15.51	1	80	1.03	8.208297	9.70	27.99	13.90
Test22	Primary Concentrate	7.2	15.53	3	80	1.03	8.208297	12.50	27.22	14.30

APPENDIX B2 : MOISTURE RESULTS OF ISHIGAKI FILTER PRESS AFTER PILOT PLANT MODIFICATIONS AT RIETSPRIUT

Run No.	Cycle time (Min/sec)	Feed pressure (Bar)	Feed SG	Feed time (Min/Sec)	Cloth Used	Cake blow (Sec)	Surface moisture %
30	9.39	4	1.178	2.35	0 times	120	26.10
31	9.31	4	1.178	2.31	2 times	120	25.70
32	11.42	4	1.178	3.4	3 times	180	22.00
33	10.35	6	1.176	3.25	0 times	120	20.10
34	11.29	6	1.176	4.25	2 times	120	23.00
35	12.17	6	1.176	4.14	3 times	180	21.30
36	12.38	6	1.176	4.12	0 times	240	20.50
37	12.25	6	1.176	3.23	2 times	240	20.60
38	12.53	4	1.176	3.48	3 times	240	22.50

* Note above for flotation concentrate at a cake blow pressure of 6 bars.

APPENDIX B3: MOISTURE RESULTS OF ISHIGAKI PRESS ON GOEDEHOOP ULTRA FINES

NO	SQUEEZE 14BAR	Cake Blow (min)	Cake Blow (bar)	6 bar feed min:sec	Total Cycle Tme (min)	Thickener u/f	Flotation Conc	Mixed	filtratrate quality %	Cloth Wash	Cake Moisture	Cake Density	CV	Ash	%SOLIDS Feed	Particle size	Seprotech Moisture
A	B	C	D	E	F	G	H	I	J	K	L	M	N	O	P	Q	O
1	1min	3	4	4:07	12:00	Y	N	N	0.15	Clean	24	1.252	27.4	15.7	29.94	N	23.7
2	1min	3	4	4:20	12:00	Y	N	N	0.15	No	25.8	1.275				N	25.8
3	1min	3	4	4:45	12:32	Y	N	N	0.15	No	20	1.276	27.29	15.9	26.77	Y	21.3
4	1min	3	4	5:01	13:43	Y	N	N	0.15	No	23.3	1.276				Y	24.89
5	1min	2	4	4:20	10:42	Y	N	N	0.15	Clean	23	1.29	27.69	15.1	29	Y	23.71
6	1min	2	4	4:58	11:45	Y	N	N	0.15	No	23.7	1.29				Y	24.58
7	1min	2	4	4:14	12:12	Y	N	N	0.15	No	No	1.34	27.66	15.3	30	N	21.83
8	1min	2	4	4:27	11:38	Y	N	N	0.15	No	No					N	
9	1min	1	4	4:14	9:40	Y	N	N	0.15	Clean	30	1.31	27.53	15.6		Y	27.6
10	1min	1	4	4:41	9:56	Y	N	N	0.15	No	24.8	1.31				N	26.4
11	1min	1	4	4:01	9:45	Y	N	N	0.15	No	22	1.33	26.47	18.1		N	26.65
12	1min	1	4	4:12	10:01	Y	N	N	0.15	No	22.7	1.35				Y	27.03
13	1.5min	4	4	4:12	12:47	Y	N	N	0.15	Clean	18.1	1.25	26.41	18.3		Y	19.57
14	1.5min	4	4	aborted			N	N	0.15								
15	1.5min	4	4	4:27	13:31	Y	N	N	0.15	No	24	1.32	26.55	18		Y	20.3
16	1.5min	5	4	4:45	14:29	Y	N	N	0.15	No	21.5	1.3				N	22.1
17	1.5min	4	4	9:50	19:40	N	Y	N	Clean	Clean	21.5		30.3	8.9	11.27	Y	
18	1.5min	4	4	8:40	19:10	N	Y	N	Clean	No	21.1		30.26	8.8	10.89	Y	
19	1.5min	4	4	6:26	15:18	N	Y	N	Clean	Clean	20.5	1.28	28.48	12.2	16	Y	
20	1.5min	4	4	6:40	15:35	N	Y	N	Clean	No	17.6	1.2	25.91	18.2	12	Y	
21	1.5 min	4	8	2:00	10:09	N	Y	N	Clean	No	17.2	1.28	28.89	12.7	22.2	Y	17
22	1.5 min	3	8	2:10	9:31	N	Y	N	Clean	No	15.03	1.21	28.89	12.7	22.2	N	15.03
23	1.5 min	2	8	2:16	8:20	N	Y	N	Clean	No	15.93	1.24	28.89	12.7	22.2	N	15.93
24	1.5 min	1	8	2:20	7:16	N	Y	N	Clean	No	15.89	1.22	28.89	12.7	22.2	N	15.89
25	1.5 min	1	8	2:40	7:34	N	Y	N	Clean	No	18.21	1.23	29.18	13	22.3	N	18.21
26	1.5 min	1	8	3:34	8:38	N	Y	N	Clean	Clean	11.2		28.83	13	15.3	Y	OFF SITE
27	1.5 min	2	8	3:53	10:08	N	Y	N	Clean	No	10.8		28.66	13.2	15.3	N	OFF SITE
28	1.5 min	3	8	4:15	11:27	N	Y	N	Clean	No	10		29.19	11.3	15.3	N	OFF SITE
29	1.5 min	1	8	4:32	9:27	N	Y	N	Clean	Clean	13.5		29.86	11.6	12.7	Y	OFF SITE
30	1.5 min	2	8	5:18	11:12	N	Y	N	Clean	No	16		29.22	11.9	12.7	N	OFF SITE

APPENDIX B4: MOISTURE RESULTS OF TH PRESS ON GOEDEHOOP ULTRAFINES

Test Nr	Feed SG	Filter time (sec)	Air blow (sec)	Press & dry	Airblow (Bar)	Cycle Time (min)	Furnace Moisture %	Filter rate (Kg/m2/hr)	Flotation Reagents
8	1.035	480	0	0	5	10.4	31.3	66.53	C
9	1.035	480	360	0	5	16.4	23.9	47.20	C
10	1.035	420	360	0	5	15.4	24	67.11	C
11	1.035	540	240	0	5	15.4	23.9	61.3	C
12	1.037	600	360	0	5	18.4	24.2	54.82	C
13	1.039	600	0	0	5	12.4	30.6	80.61	C
14	1.039	600	0	240	5	15.9	24.8	64.94	C
15	1.032	540	0	360	5	16.9	27.3	55.24	C
16	1.039	540	0	480	5	18.9	25.5	51.47	C
17	1.041	600	240	0	5	15.9	25.9	62.46	C
18	1.043	600	360	0	5	17.9	24.9	55.11	C
19	1.032	840	0	240	5	19.9	28.3	52.79	C
20	1.034	840	0	360	5	21.9	24.6	50.12	C
21	1.034	840	0	480	5	23.9	23.4	44.96	C
22	1.055	360	360	0	8	14.4	20	69.03	C+P
23	1.057	360	0	360	8	13.9	16.9	73.73	C+P
24	1.056	360	360	0	8	14.8	17.5	70.89	C+P
25	1.057	360	360	0	4	14.8	18.4	70.95	C+P
26	1.054	240	360	0	8	12.8	16.8	64.54	C+P
27	1.053	240	0	360	8	11.8	17.7	66.12	C+P
28	1.047	360	0	120	8	9.8	21.5	102.8	C+P
29	1.047	360	60	0	8	10.1	22.1	102.41	C+P
30	1.049	360	360	0	4	15.1	17.3	73.08	C+P
31	1.050	360	240	0	4	13.1	18.7	85.52	C+P
32	1.049	360	300	0	4	14.1	18.5	76.93	C+P
33	1.043	360	0	0	0	7.9	28.3	120.91	C+P
34	1.056	360	360	0	4	15.1	21.5	69.97	C+P
35	1.054	360	360	0	4	15.1	23.8	65.93	C+P
36	1.055	360	360	0	4	15.1	22.9	68.14	C+P
37	1.055	360	360	0	4	15.1	21.9	69.12	C+P
38	1.054	360	360	0	4	15.1	23	69.2	C+P
39	1.054	360	360	0	4	15.1	22.3	68.72	C+P

Note: C = Centifroth and P = Paraffin