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Flash Dryer Unit Optimization through Advanced Process Control

by

Niel de Clerk

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Niel de Clerk

96012031

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Synopsis

In line with Anglo American's asset optimization and environmental policies, the coal burning flash dryer operations at its smelters have been identified for potential optimization by means of advanced process control. For this project, the process and related literature were studied in detail and a revised control philosophy, which includes modifications to the existing basic control structure as well as a hybrid rule and model-predictive advanced control layer, was developed, installed and tested on one of these flash drying operations.

Since commissioning of the APC, the flash dryer's average throughput has increased by more than 6 %, despite higher feed moistures. Furthermore, even though coal consumption has increased slightly, the operation efficiency has improved by almost 5 %. This was made possible by improving the stability of the drying column outlet temperature by approximately 40 %, which in turn enabled the selection of a more optimal setpoint. Recent data has shown that APC utilization now exceeds 95 %. This is indicative of a successful controller installation with good site acceptance.

Keywords: flash dryer, platinum processing, model predictive control

Acknowledgements

I would like to thank Anglo American for sponsoring this project and for allowing the study to be conducted at its operations.

This report is dedicated to my faithful, supportive and patient wife, Anneke, and my two children, Natanya and Johnathan. I love you.

All honour be to God Almighty who has created this marvelous universe with all of its fascinating sciences that we live in.

Contents

Nomenclature

a:	Step coefficient (in the DMC algorithm)
ANSI:	American National Standards Institute
APC:	Advanced Process Control(ler)
APET:	Anglo Platinum Expert Toolkit
BIC:	Bushveld Ingeous Complex
BMR:	Base Metals Refinery
CRO:	Control Room Operator
CV:	Controlled Variable
DCS:	Distributed Control System
DMC:	Dynamic Matrix Control
DV:	Disturbance Variable
FBC:	Fluidized-bed combustion
g/t:	Gram per ton
h:	Impulse coefficient (in the DMC algorithm)
HAZOP:	Hazard and Operability Study
HGG:	Hot Gas Generator
HIRA:	Hazard identification and Risk Assessment.
ID:	Induced Draft
iDX:	Industrial Data Exchange
ISA:	The instrumentation, Systems and Automation Society
KPI:	Key Performance Indicator
ktpa:	Kilo ton per annum
LP:	Low Pressure

MJ/kg:	Mega Joule per kilogram
mm:	Millimeter
MPC:	Model Predictive Control
MV:	Manipulated Variable
MW:	Mega Watt
OP:	Output (e.g. damper opening)
OPC:	OLE (Object Linking and Embedding) for Process Control
PGM:	Platinum Group Metal
PID:	Proportional, Integral and Derivative (controller)
PLC:	Programmable Logic Controller
PMR:	Precious Metals Refinery
POC:	Products of Combustion
RC:	Ratio Controller
SCADA:	Supervisory Control and Data Acquisition
SPD:	Speed (transmitter)
STC:	Self Tuning Control
tph:	Ton per hour
TT:	Temperature transmitter
U:	Control Horizon (in the DMC algorithm)
UG2:	Upper Group 2
V:	Prediction Horizon (in the DMC algorithm)
VSD:	Variable Speed Drive
WT:	Weightometer

1 Introduction

Platinum processing can be divided into concentrating, smelting and refining operations. A concentrator receives thousands of tons of ore, containing less than 6 g/t platinum group metals (PGMs), per day. Through multiple steps of liberation (crushing and milling) and beneficiation (flotation), the concentrators produce a concentrate that contains approximately 100 g/t PGMs. This concentrate is smelted in electrical arc furnaces to produce a matte, containing approximately 1 000 g/t PGMs, which can be fed to the refineries where each base and precious metal is isolated and purified to an acceptable level.

Depending on the mass pull (the percentage of ore treated at the concentrator that is passed on to the smelter as concentrate) and grade of the concentrate, the production pressure on a smelter varies. It is imperative for the safe and stable operation of a furnace that it receives a consistent uninterrupted feed supply. Given the temperatures at which furnaces operate, and the consequential risk for hydrogen explosions, all measures should be taken to avoid moisture from entering the furnace, either from a leak in the cooling water system, or from moist feed.

For this reason, prior to entering the furnace, flash dryers are used to dry all concentrate to bone-dry levels before filling the furnace feed silo.

At Anglo Platinum there are several independent flash dryers that are responsible for keeping the furnace feed silos adequately filled with bone dry concentrate. These flash dryers require relatively frequent maintenance and hence it is important to maximize their throughput in order to allow for sufficient time to do maintenance without jeopardizing the furnace feed supply. The hot gas that is used for the transport and drying of the concentrate in the flash dryers, is produced by coal burning hot gas generators which should be operated as efficiently as possible to minimize the specific cost of the coal as well as the environmental impact associated with coal combustion.

Before optimization can commence, it is important to stabilize the process (Rademan, 2008). This is done by evaluating and improving the existing instrumentation and control system and then by supplementing it with an advanced layer where deemed necessary. Once stabilized, the process can be optimized by adjusting the operating regions and by assigning an appropriate cost function to the advanced controller.

A list of agreed key performance indicators (KPIs) is monitored in order to determine the plant performance and whether or not the new control scheme is adding value.

Initially this strategy will be tested on one of the three flash dryers at the chosen smelter. If successful, it will be rolled out to the other two flash dryers at the chosen smelter and then to the flash dryers at the other two smelters.

This dissertation will summarize the theory surrounding flash dryers and their operation, it will then describe the specific process to be optimized, followed by a detailed account of the various project stages that led to the ultimate successful deployment of the new advanced process control (APC) strategy. Finally the results will be presented and discussed prior to concluding with some recommendations.

2 Theory

In order to optimize a process, it is important to understand its main purpose and how that is achieved. The theory in this chapter starts with the industry, Platinum processing, where the process operates (2.1). Furthermore, the process of solids drying is investigated (2.2) and supplemented by a discussion on the relevant energy source, namely coal combustion (2.3). Next the environmental impact of the process is briefly discussed (2.4) and finally process control (2.5) and more specifically model predictive control (2.6) is considered as appropriate optimization tools.

2.1 Platinum Processing

Figure 1 illustrates a high level flow diagram of a typical platinum processing process. The various steps are briefly discussed in this section. (Johnson Matthey, *sa*; Anglo Platinum, *sa*; Cawthorn, 1999)

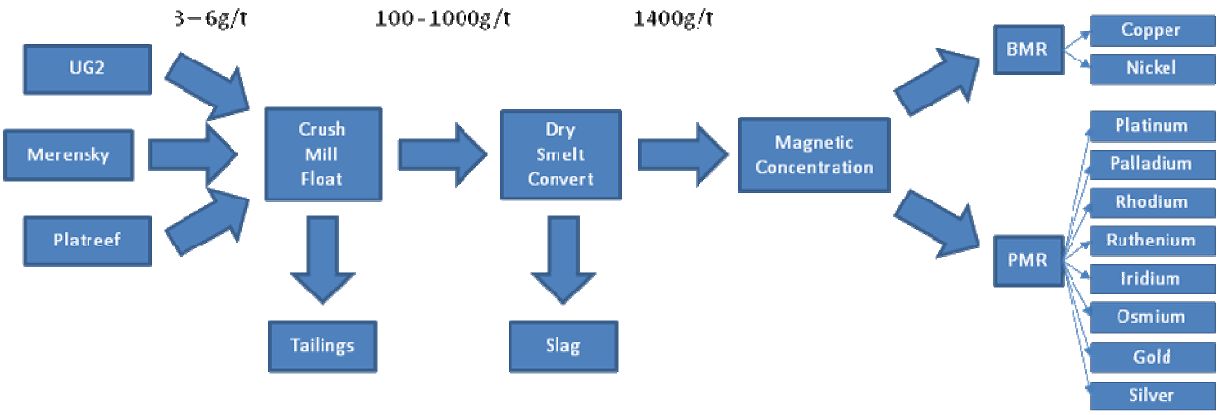

Figure 1: Platinum processing overview

2.1.1 Mine

The largest PGM deposits in the world are found in South Africa in the Bushveld Igneous Complex (BIC). Amongst its various distinct layers, only the Merensky, Upper Group 2 (UG2) and Platreef layers contain concentrations of PGMs that are economically viable to mine. The Merensky and UG2 reefs are mainly extracted from underground while Platreef is mined using open-pit methods. BIC ore typically contains between 3 to 6 g/t PGMs, hence it is necessary to process between 10 and 25 tons of ore to obtain a single ounce of platinum.

2.1.2 Concentrate

At a concentrator, the ore is crushed and milled to reduce the rock sizes and thereby liberate the minerals that contain the PGMs. After mixing with water and reagents, and floating with air, a concentrate is produced that now contains between 100 and 1000 g/t PGMs.

2.1.3 Smelt

At a smelter, the concentrate is dried (the focus of this project) and smelted in an electric furnace that operates at temperatures that may exceed 1 500 ^{o}C. During this process, the molten liquid separates into a PGM rich matte and an unwanted slag layer. By removing iron and sulphur from the matte, a converted matte that contains in excess of 1400 g/t PGMs is produced.

2.1.4 Refine

Next, the converted matte is split into a non-magnetic and a magnetic portion. The non-magnetic portion is processed at a Base Metals Refinery (BMR) to recover the contained Nickel and Copper, while the magnetic portion is separated and purified at a Precious Metals Refinery (PMR) to produce individual PGM products (ingots, grain or sponge) with purities in excess of 99.95 %.

2.2 Solids Drying

Solid materials are dried when entrained volatile liquid, most notably water, is removed by means of evaporation. Drying equipment types are classified based on the following design and operating metrics (Sinnott, 1998):

- Batch (Stationary) or continuous (Drum, rotary, conveyor or suspended particle)
- Feed Physical state: Liquid, slurry, or wet solid
- Method of solid transport: belt, rotary, or fluidized
- Method of heat transfer: conduction, convection or radiation

The concentrate flash dryers are therefore classified as follows:

- Continuous suspended particle
- Wet Solid Feed
- Fluidized transport (Short retention time)
- Convection heating
- High capacity

Other typical products that are dried in flash dryers include ores, coal, clays and chemicals. These can be in the form of slurries, pastes, sludges, filter cakes, powders or granules. The feed should be relatively consistent in particle size and not too big. Flash dryers (also referred to as pneumatic dryers) dry the solids by dispersing it into an upward flowing stream of hot gas. This leads to extremely short contact times (seconds) that are only effective with very small particle sizes. By increasing hold-up times in the cyclones, the residence time can be altered. Flash dryers generally have low thermal efficiency. This type of dryer has a relatively small physical footprint, but, because of the large fans, they tend to consume more power than some of the other drying methods.

The flash drying definition offered by the supplier of the flash dryer under consideration for this project supports the above (Drytech, 2001): *"Flash drying is the process of using a heated gas to pneumatically convey the feed up a flash tube and into a primary air separation device, most commonly a cyclone. Air is induced or forced into the feed area through a hot gas generator where it entrains the feed and flashes off the moisture as it conveys. The product is fed into the throat of the feed area in a controlled fashion. The particles attain a velocity in the order of 80 % of the conveying velocity. Secondary dust collection such as multiclones, bag houses, scrubbers and electrostatic precipitators are required for fine products".*

Flash dryers are fed and discharged automatically and continuously by means of vibrators, screw-feeders, rotary airlocks, double flap valves and gravity chutes.

After backmixing, the feed is contacted with the air stream via a venturi or disintegrator. The gas, which can be heated by steam, electricity, coal, liquid fuel or gas, is mostly contacted directly with the feed in a concurrent fashion.

Table 1 lists the most important strengths and weaknesses associated with flash drying.

Table 1: Strengths and weaknesses of flash drying technology

Strenghts	Weaknesses
Very intimate contact with gas stream	Loss of power will cause product to fall into dryer base
Excellent transfer of energy	Attrition and impact may cause size reduction
Ideal for heat sensitive, explosive or reactive products	High energy costs for fans and dust collection
Small real estate requirements	May not be suitable for products with a high degree of bound moisture.
Flash tube is flexible, can be routed within plant constraints	May be susceptible to high wear if improperly accounted for
Low maintenance - few moving parts	

2.3 Atmospheric Fluidized Coal Combustion

2.3.1 Background

The global oil crises of the 1970s, combined with environmental pressures in the 1980s resulted in the development of various coal based technologies, including fluidized coal combustion (Stratos Tavoulareas, 1991). The flash dryer under consideration makes use of hot gas that is generated in a hot gas generator by means of fluidized coal combustion. Such hot gas generators are ideal for direct drying applications like mills, drum and flash dryers, fluidized bed furnaces/driers and spherodizers (Loesche, *sa*).

Combustion is the process of burning the carbon, hydrogen and sulfur in a fuel in the presence of air to form carbon dioxide, water vapor and sulfur dioxide. Even though the minimum air requirement for complete combustion can be estimated, it is common practice to supply excess air. The amount of excess air that was supplied can be calculated if the oxygen and nitrogen compositions in the gaseous products of combustion (POC) are measured. (Perry & Green, 1984))

$$A_x = \frac{[O_2]_{POC} - 0.5([C]_{POC} + [H_2]_{POC})}{0.266[N_2]_{POC} - [O_2]_{POC} + 0.5([C]_{POC} + [H_2]_{POC})} \tag{1}$$

Where

A_x is the amount of excess air as a fraction

$[O_2]_{POC}$ is the concentration of oxygen in the POC

$[C]_{POC}$ is the concentration of carbon in the POC

$[H_2]_{POC}$ is the concentration of hydrogen in the POC

$[N_2]_{POC}$ is the concentration of nitrogen in the POC

2.3.2 Combustion Efficiency

Furthermore, combustion efficiency (excluding carbon losses) can be calculated if the carbon monoxide and carbon dioxide composition of the POC are known (Patumsawad, *sa*).

$$\epsilon_{CE} = \frac{[CO_2]_{POC}}{[CO_2]_{POC}+[CO]_{POC}} \tag{2}$$

Where

ϵ_{CE} is the combustion efficiency

$[CO_2]_{POC}$ is the concentration of carbon dioxide in the POC

$[CO]_{POC}$ is the concentration of carbon monoxide in the POC

This relationship assumes that CO and CO_2 are the only carbon products formed during combustion. Experimental work has shown that this efficiency can be expected to be in the order of 99.8 %.

If the flue gas composition, fractional excess air, and the ultimate analysis of the fuel are known (unlikely in a production environment), combustion efficiency can be calculated more accurately as

$$\eta_{CE} = \left(\frac{B}{C}\right) \tag{3}$$

Where

B is the mass fraction of burnt coal in the fuel

C is the mass fraction of total carbon in the fuel

Experimental work (Patumsawad, *sa*) has shown that this efficiency can be expected to be in the order of 91 to 95 %. It was also found that this efficiency can be increased by increasing the amount of excess air (Figure 2). However excessive air flow velocity can ultimately lead to a reduction in efficiency as a result of higher carbon losses (Figure 3).

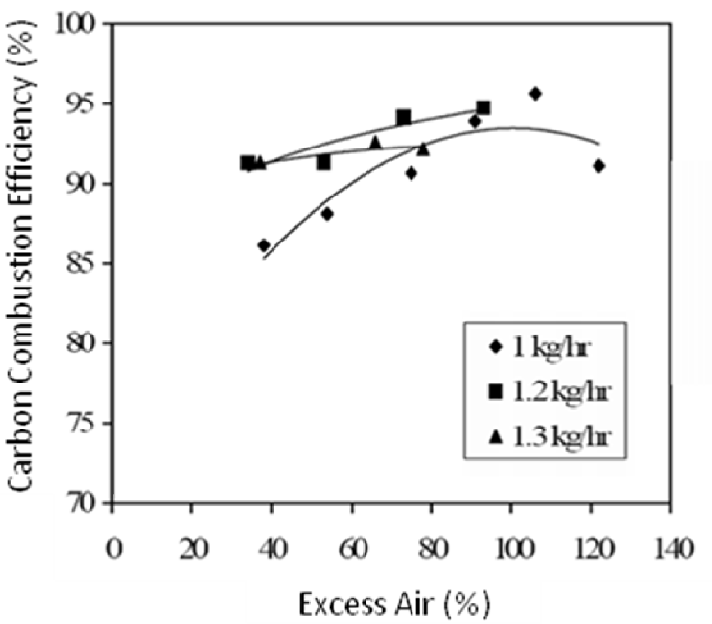

Figure 2: Combustion efficiency versus excess air at different feed rates. (Patumsawad, *sa*)

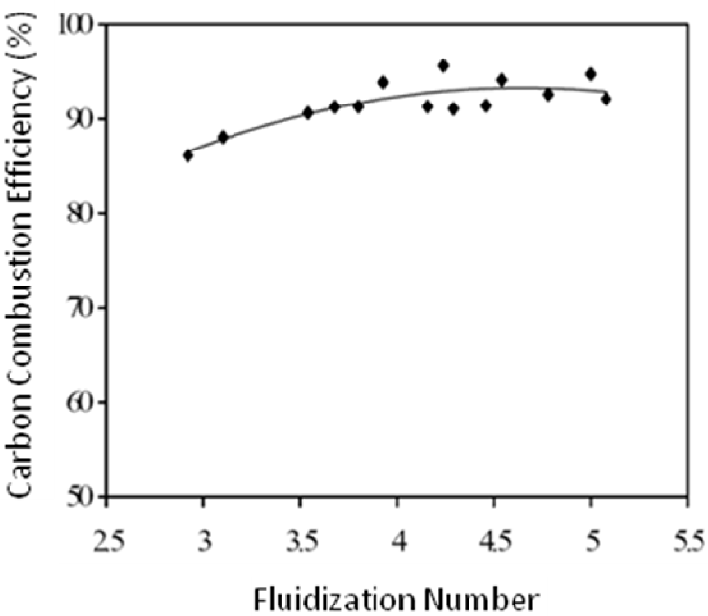

Figure 3: Combustion efficiency as a function of fluidization number (the ratio of fluidizing velocity with minimum fluidization velocity) (Patumsawad, *sa*)

2.3.3 Coal Composition

The composition of a typical coal sample is shown in Table 2.

Table 2: Proximate and ultimate analyses of coal (Patumsawad, *sa*)

Particle size range	1.4-4.7 mm
Proximate analysis, wt% (dry basis)	
Volatile	38.15 %
Fixed carbon	58.87 %
Ash	2.98 %
Moisture content, wt% (as received)	5.9 %
Ultimate analysis, wt% (dry basis)	
Carbon	80.13 %
Hydrogen	5.31 %
Oxygen	9.88 %
Nitrogen	0.96 %
Sulphur	0.74 %
HHV (dry basis)	33 MJ/kg

This compares with the upper end of the coal consumed by the studied process as this coal is mined in South Africa where carbon content of bituminous coal varies between 45 to 85 % and calorific value between 23 to 33 MJ/kg. (Mintek, 2007)

2.3.4 Comparison and Advantages

In pulverized coal firing, coal is crushed into very fine powder, sprayed into the boiler and ignited to form a long, lazy flame (Zactruba, 2009). In other systems, coal simply rests on grates. Fluidized-bed combustion (FBC) is the process of combusting a solid fuel (like coal with particle sizes of approximately 3 to 6 mm) in a bed of inert solid particles (e.g. sand with average size of 850 µm) that are held in suspension (fluidized) by the injection of air from the bottom of the bed at a typical velocity of 1 to 2 m/s. The result is a turbulent mix of gas and solids that provides a tumbling action, similar to that of a bubbling fluid, which benefits the chemical reactions and heat transfer (to heat up the fuel, drive off the moisture, preheat the combustion air and heat the flue gas). Typically the fuel supply rate is controlled with a screw-feeder.

FBC offers the following advantages over the other conventional combustors (FBC, *sa;* Zactruba, 2009)):

- Due to fuel-bed additives (e.g. limestone or dolomite) that capture most of the sulfur in the fuel, SO_2 emissions are minimized.
- A fluidized bed burns cleaner as it burns cooler. Typical bed temperatures are below 950 ^{o}C (compared to other methods that operate at almost 1650 ^{o}C). This ensures:
    - Minimal thermal NOx emissions (as these only form at approximately 1350 ^{o}C)
    - Reduction in ash agglomeration
    - Reduced fouling and corrosion of downstream components as a result of less volatization of the sodium and potassium contained in the coal.
    - Lower dependency on high quality coal. In fact many FBC systems operate on biofuels (like wood, ground-up railroad sleepers or even soggy coffee grounds).

- Typically, there isn't a need for external chemical pollutant emission controls such as scrubbers.
- High heat transfer and volumetric energy release rates.
- The larger fuel particles require less time, energy and facilities to prepare.
- Most ash collects at the bottom of the furnace and therefore reduces the load on downstream precipitators or filters.
- Compact furnace with a relatively simple design that supports comparably uniform temperatures.

2.3.5 Efficiency Factors

The profitability of a coal combusting process is dependent on its overall plant efficiency which in strongly reliant on a high carbon-combustion efficiency. (Perry & Green, 1984) Typically, carbon is primarily lost through the carry-over of unburned carbon particles while the losses due to incomplete combustion (to carbon monoxide) or unburned carbon removed via ash collection is much less significant. Bed temperature, excess air and fluidizing velocity are the main operating variables that affect carbon-combustion efficiency. Increasing bed-temperature and excess air will improve efficiency. This benefit needs to be balanced with unwanted effects of increased SO_2 capture, NO_x formation and ash-fusion. Too much excess air will also cool the bed and the combustion products (Fox, 1957). The carry-over of fine unburned carbon particles can be minimized by minimizing the air velocity, however care should be taken to still maintain minimum velocity required for fluidization. Larger beds will have lower velocities for the same air flow. It has also been shown that the fuel feed rate can affect carbon efficiency in that typically higher feed rates result in higher bed temperatures, but this also increases coal loading which normally leads to higher carbon loss associated with elutriation.

Experimental work (Patumsawad, *sa*) has shown that there is not neccesarily a significant temperature profile across the bed, or even directly above the bed due to volatile burning or freeboard combustion as shown in Figure 4. Unfortunately, due to a lack of instrumentation,

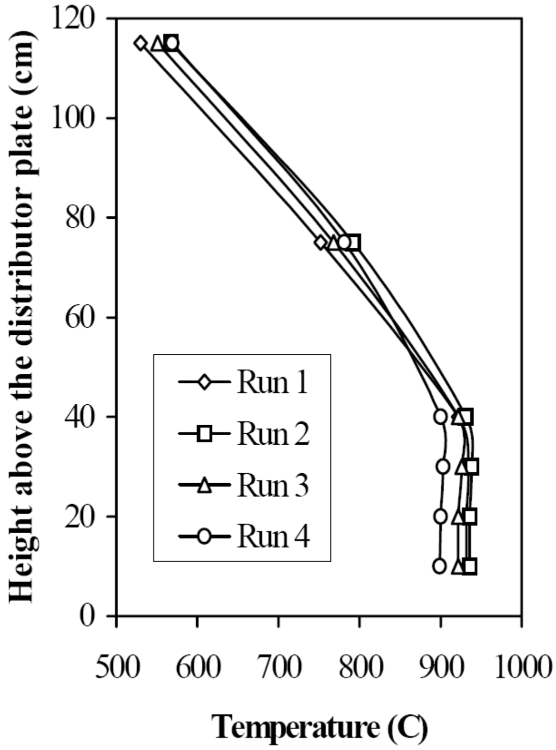

Figure 4: Temperature profile above distributor plate of laboratory fluidized bed reactor.

2.4 Environmental

Recent and ongoing energy shortages in South Africa emphasized the urgency with which energy usage should be optimized. (Anglo Platinum, 2009) Hence Anglo American has fast tracked several energy efficiency projects based on the understanding that energy security is critical to ensure continued production at its various operations. This is also in line with the commitment of the company to minimize climate change, as improved efficiency results in reduced greenhouse gas emissions. Anglo aims to reduce its specific energy consumption by 15 % by 2014. Figure 5 shows Anglo Platinum's energy footprint (2008) and also highlights the need for an optimization strategy at the flash dryers.

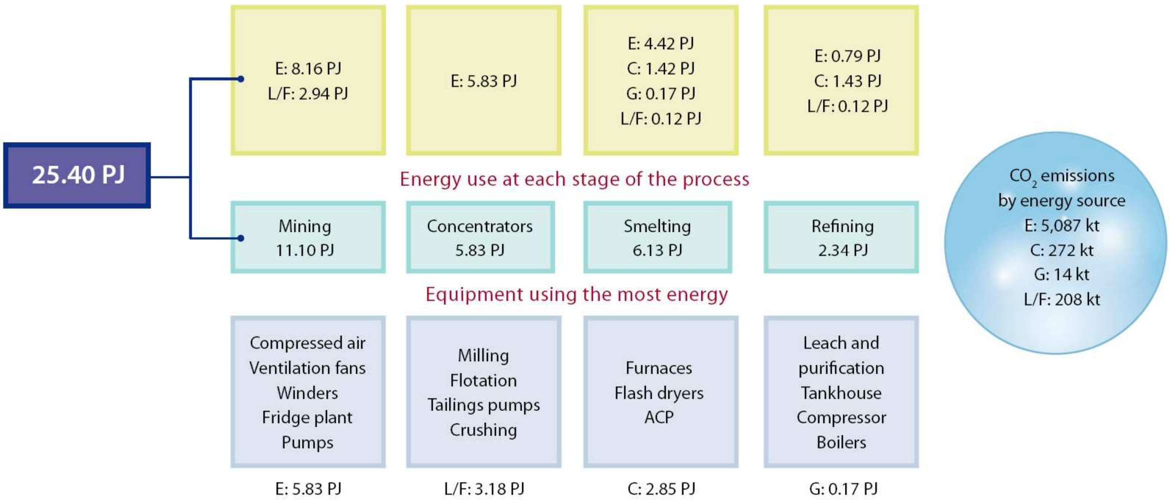

Figure 5: Anglo Platinum Energy Footprint (E: Electricity, L/F: Liquid Fuel, G: Gas)

2.5 Process Control

The engineering discipline of process control deals with the mechanisms and algorithms for controlling the outputs of a specific process. There are various incentives (2.4.1) and classes (2.4.2) of process control that will be discussed here.

2.5.1 Incentives for process control

In general, chemical and minerals processing plants exist so that certain raw materials can be transformed into desired products through the application of available sources of energy (Stephanopoulos (1984). In the case of the concentrate flash dryer, wet concentrate (the raw-material) is transformed to dry-concentrate (the products) through the combustion of coal (the energy source).

Process control is applied to help achieve this general goal while ensuring the following requirements are met:

- Safety: No process can be operated sustainably unless it is operated safely. Therefore it is critical to maintain acceptable operating conditions that will prevent damage to people, production, property and/or the environment (KBC, 2010). For example, given the combustion process that is present in the flash drying operation, process control can be used to avoid excessively high temperatures that can result in hot gas generator (HGG) bed sintering or thermal damage to the ducting and/or bag-house filters.
- Production specifications: If the product from a process is to be useful, it is important that certain quality specifications are met. For example, the dried concentrate from the flash dryer can only be sent to the furnace feed silo as

long as the drying column outlet temperature exceeds 100 ^{o}C. Process control can therefore be applied to ensure that the concentrate is sufficiently dried without over-drying and also to interrupt the concentrate feed supply as soon as the drying specification can no longer be achieved.

- Environmental regulations: Everyone has a legal and ethical responsibility to protect the environment. Process control can assist with the efficient operation of the HGG and air cleaning processes to minimize dust and carbon emissions.
- Operational constraints: Design and physical laws impose certain constraints on processes that need to be satisfied in order to maintain efficient operation. For example, in the flash dryer, a minimum bed temperature, fluidizing and induced air flow and HGG outlet temperature must be maintained in order to ensure efficient combustion, fluidization, and drying respectively. Process control can assist in monitoring and maintaining these required conditions.
- Economics: Market conditions dictate the feasibility of the operation of a plant. Profitability is driven by product demand and price as well as the cost of raw-materials, energy, capital and labour. Process control helps to reduce cost by, amongst others, improving specific energy requirements, throughput, maintenance cycles and equipment life (through reducing stresses as a result of improved stability).

2.5.2 Classes of process control

In general, process control action can be classified as follows:

- Disturbance rejection: As various measured and unmeasured disturbances act on a process, the control system needs to compensate by adjusting the available manipulated variables in order to minimize the influence on the controlled variables. For example, if the moisture content in the wet concentrate increases, the controller needs to reduce the concentrate feedrate or increase the HGG outlet temperature in order to rebalance the energy supply and demand relationship.
- Stabilize an unstable process: As time progresses, a stable (self-regulating) process that has been disturbed by a bounded input will return to specific value and stay there, e.g. it will produce a bounded output. A process that does not exhibit this behaviour is said to be unstable (Figure 6). Control systems are required to move processes to their desired setpoints (setpoint tracking) and to stabilize unstable processes.

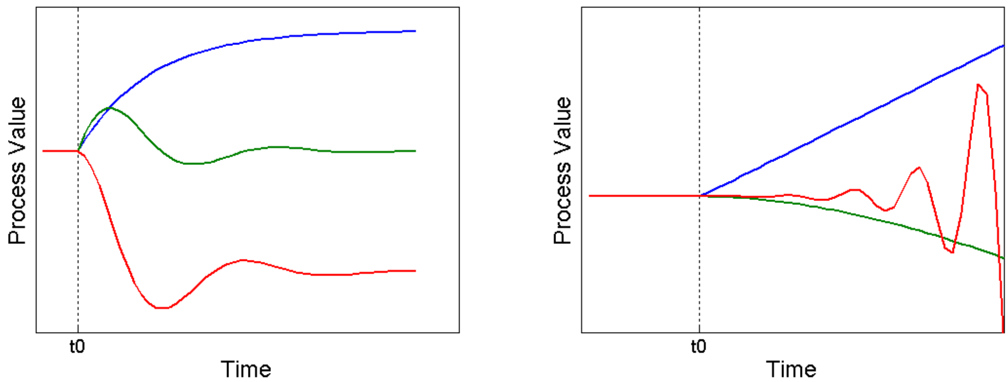

Figure 6: Examples of Stable process responses (left) vs Unstable process responses (right)

- Optimize: It is typical for operators to be more conservative when manually operating a process. They will tend to maintain a relatively large safety margin between a process variable and its limits mainly because a human operator normally needs more time to respond to disturbances that might cause the process to violate its limits. Also, it is not humanly possible to effectively manipulate several process inputs to control several process outputs, especially in the presence of process interactions. Hence it is rather common for operators to focus just on one process output while allowing another to drift. Process control systems can however overcome these limitations since they are faster to respond to disturbances and, based on internal models, can effectively deal with multivariate integrated processes.

- Interlocks: Having the necessary permissive checks built into the process control system is vital for the safe operation and ordered startups and shutdowns of a process.

From this analysis, and based on existing industry examples (Rademan, 2008) it is clear that a flash drying operation can benefit from APC.

2.6 Model Predictive Control

Model Predictive Control (MPC), or more specifically Dynamic Matrix Control (DMC) is the chosen control algorithm for this project. This decision was based on several examples of similar applications that are documented in literature (2.6.1) as well as on the interactive nature of the process dynamics of the flash dryer operation that includes significant dead times.

2.6.1 Examples of MPC applications.

Various control algorithms exist for controlling chemical processes like flash drying. Of these the most common is arguably Proportional-Integral-Derivative (PID) and ratio controllers that are normally executed by a programmable logic controller (PLC) or distributed control system (DCS). Almost all other types of controllers are referred to under the term APC. These are often executed externally to the PLC/DCS on a networked server and can also be one of various different types like supervisory, fuzzy, inferential, feed-forward, multi-variable, model-predictive (MPC) and so forth. Hybrid solutions that combine more than one algorithm also exist. MPC, including its derivative DMC is especially widely used (more than 10 000 current applications) in the petrochemical industry (Tay, 2007). It is however also starting to penetrate the minerals processing industry as is shown in the work by Le Page, Tade & Stone (1998) where PID, DMC and STC (self-tuning-control) were compared in the regulation of a calciner furnace temperature. Here it was shown that DMC delivered the best performance, closely followed by the results of optimized PID controller. The STC performed poorly. This work also illustrates the benefits that can be achieved by simply tuning existing PID controllers. In addition, Altafini & Furini (1997) discusses how model-based control was applied to an industrial starch flash dryer in pursuit of improved regulation robustness. This process also exhibited some undesired behaviour that had to be remedied. Soft sensors were developed in order to approximate the output starch moisture since no online measurements were available. Tay (2007) reports on an operation that deployed MPC to de-bottleneck a dryer process. The immediate value that was generated justified a plant-wide rollout. Cristeaa, Baldeaa & Agachia (2000) tested the applicability of MPC on an industrial batch dryer. Through simulation it was shown than MPC holds clear potential to improve the operation of such a process. Results from an initiative to develop a MPC controller for an infrared dryer also indicated that, through its proper handling of process interactions, such a controller produces good results for setpoint tracking of the moisture and temperature of the exit material, while effectively rejecting measured and unmeasured disturbances (Abukhalifeh, Dhib, & Fayed, 2003)

2.6.2 DMC Algorithm

The MPC algorithm predicts future process behaviour with reference to a mathematical process model. From here it can calculate the optimum set of allowable control moves that will minimize the controller error. The model takes the interaction among the various process variables into account and thereby helps to improve plant stability and to achieve production targets that were set to meet business expectations (Tay, 2007). The basic DMC algorithm is widely documented in literature like Seborg, Edgar & Mellichamp (1989) and Marlin (1995) and is neatly summarized by Hurowitz (1998), who also indicated that the most popular form of MPC is the DMC algorithm, which was developed by Cutler and Ramaker in the late 1970's. Note that for this project, the DMC controller was deployed using the AspenTech DMCplus commercial software. Even though the algorithm in this software package is more advanced than the algorithm explained here, the core principles remain the same.

The DMC algorithm uses linear step-response models, as shown in Figure 7 and obtained either through physical step tests or through theoretical approximations, to predict the future behaviour of each controlled variable (CV) based on the past moves in each manipulated variable (MV).

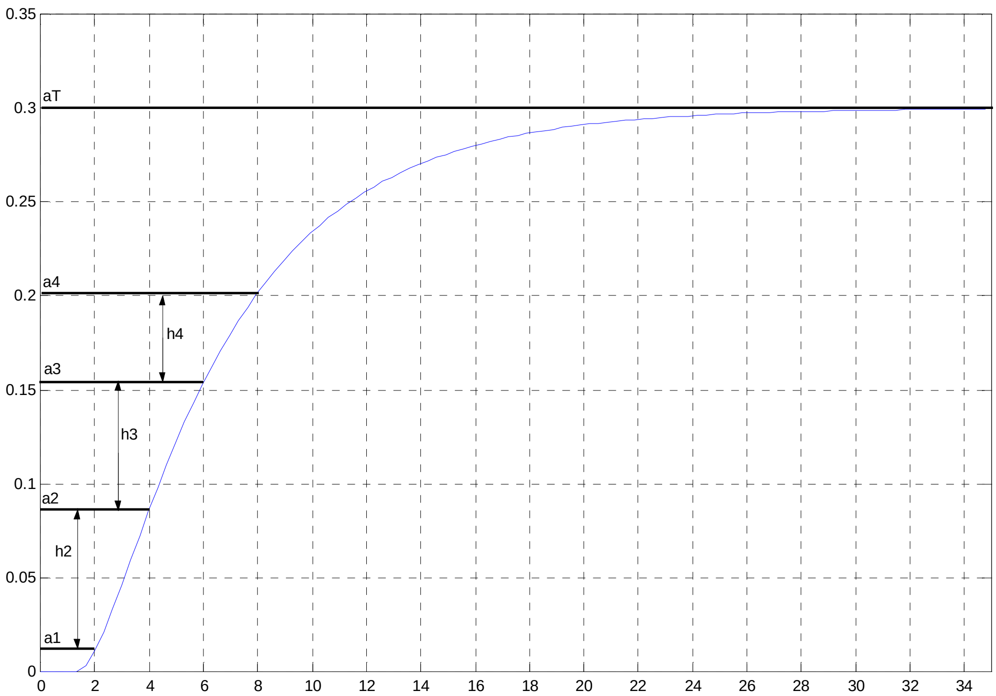

Figure 7: An example process response in reaction to a unit step.

Such models are expressed either as a series of step (a) or impulse (h) coefficients. These coefficients are combined with the list of recent MV changes to predict the expected future CV trajectory according to the principle of linear superposition which states that the total output is the sum of the effects of each individual input. Knowing the future trajectory of the CV and considering targets, limits, priorities, and allowable move sizes, a quadratic programming optimization algorithm can be used to determine the optimal future moves for each MV in order to minimize the overall controller objective function which typically has the form shown in Equation 4.

$$J(n_c, n_p) = \sum_{i=1}^{n_p} \sum_{j=1}^{n_y} \delta_j(k)\left[\hat{y}_j(k+i|k) - w_j(k+i)\right]^2 + \sum_{i=1}^{n_c} \sum_{j=1}^{n_u} \lambda_j(k)\left[\Delta u_j(k+i-1)\right]^2 \quad (4)$$

Where

$n_c = Control\ horizon$

$n_p = Prediction\ horizon$

$n_y = Number\ of\ CVs$

$n_u = Number\ of\ MVs$

$k = Time\ of\ calculation, typically\ current\ time$

$\delta_j(k) = Penalty\ on\ j^{th}\ CV\ error\ at\ time\ k$

$\hat{y}_j(k+i|k) = Predicted\ value\ for\ j^{th} CV\ at\ time\ k+i\ as\ calculated\ at\ time\ k$

$w_j(k+i) = Reference\ value\ for\ j^{th}\ CV\ at\ time\ k+i$

$\lambda_j(k) = Penalty\ on\ \mathrm{j}^{\mathrm{th}}\ MV\ movement\ at\ time\ k$

$\Delta \mathrm{u}_j(k+i) = Change\ in\ \mathrm{j}^{\mathrm{th}}\ MV\ at\ time\ k+i$

Hence at each sampling interval, the DMC algorithm takes the following action:

- Measures the current CV values.
- Assumes that no future MV changes will take place, i.e. as if open loop.
- Uses the convolution model matrix to predict V (prediction horizon) future values for each CV based on past moves in the MVs.
- Assumes that the various limits, targets etc will remain constant for the following V time steps.
- Calculates the objective function for the following V time steps.
- Minimizes the objective function by determining the optimal allowable changes to be made in the MVs during the following U (control-horizon) time steps.
- Implements the first calculated changes in each MV.
- Records these changes as the most recent past MV moves.

3 Process Description

As shown in Table 3 Anglo Platinum has three smelter complexes each with one or more flash dryers. This project considers one of the 6 tph (tons of moisture removed per hour) flash dryers at the Waterval Smelter. The process flow (3.1), original control strategy (3.2) and interlocks (3.3) are discussed in this chapter.

Table 3: Anglo Platinum Smelter Capacities (De Villiers, 2010)

	Waterval	**Mortimer**	**Polokwane**
Capacity	**650 ktpa**	**180 ktpa**	**650 ktpa**
Flash Dryers	**2x 6 tph** **1x 12 tph**	**1x 6 tph**	**2x 12 tph**
Furnaces	**2x 34 MW**	**1x 18 MW**	**1x 68 MW**

3.1 Process Flow

The concentrate from various concentrators is delivered to the smelters and stored on concentrate stockpiles. Before this concentrate can be fed into the smelting furnaces, it needs to be thoroughly dried to prevent any moisture from entering the furnaces. This drying is performed by a number of flash dryers that operates in parallel. Figure 8 shows the process flow of one of these.

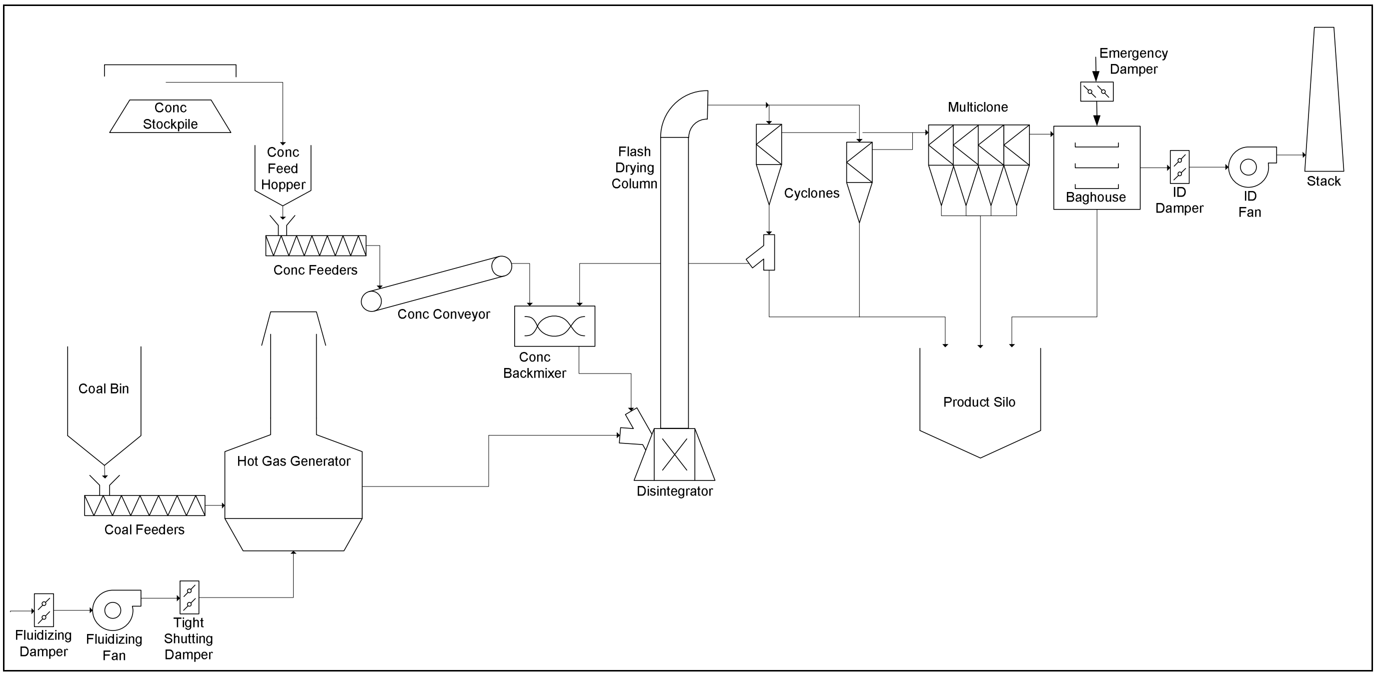

Figure 8: Flash drying circuit process flow diagram

Using overhead cranes, concentrate is fed into the concentrate feed hoppers from the concentrate stockpile. From here it is transported with screw feeders onto the concentrate conveyor that delivers it to the back-mixer where it is mixed with a portion of the previously dried concentrate. This mixture is fed into the flash dryer through the disintegrator for drying with hot gas. The hot gas for the flash dryer is generated in the HGG. It operates by combusting coal on a fluidized silica bed. The heated fluidizing air that passes through this bed combines with air that is sucked in through the HGG chimney prior to being fed to the flash dryer. Air flow through the HGG and flash dryer is achieved by a fixed speed exhaust, induced draft (ID), fan at the outlet side of the flash dryer that sucks the air through the system and a fluidizing fan that forces air through the HGG bed.

After drying, the concentrate is separated from the exhaust gas by means of two cyclones, a multi-clone and a bag house as shown.

3.2 Original Control Strategy

Prior to the start of this project, the PLC based control strategy, illustrated in Figure 9, was in use. This system remains in place in case APC control is not available, due to a communication failure for example.

- A PID controller that manipulates the speed of the concentrate feeders with a negative gain to maintain the flash dryer outlet temperature at a setpoint of 115 ^{o}C. This setpoint was chosen based on the limited stability that can be achieved without APC control that requires a relatively large safety margin to be maintained to avoid low temperature feed stops.
- A ratio controller that minimizes dead time by adjusting the concentrate conveyor speed in accordance with the speed of the concentrate feeders.
- A PID controller that manipulates the fluidizing damper opening with a positive gain to also maintain the flash dryer outlet temperature, but at a setpoint of 125 ^{o}C. This setpoint was chosen based on the maximum temperature that can be tolerated at the bag house.
- A PID controller that manipulates the speed of the coal feeders with a positive gain to maintain the HGG average bed temperature at a setpoint of 900 ^{o}C.
- The ID fan damper opening is adjusted manually, but normally set to 45 %.

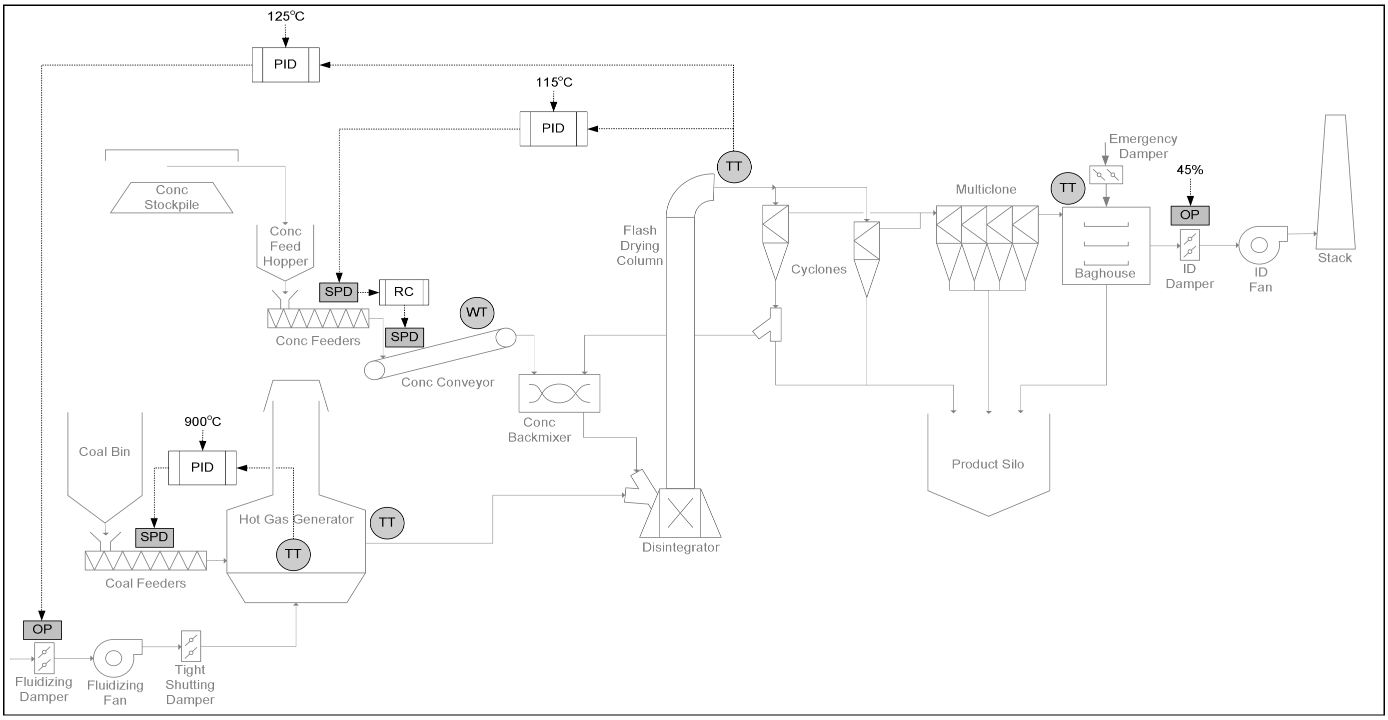

Figure 9: Original control strategy.

This strategy is sub optimal since it struggles to effectively deal with the process dead times or the interaction between the process variables. It also omits critical instrumentation (like the concentrate feedrate). These limitations support the need for an improved control strategy.

3.3 Process Interlocks

Familiarization with the process interlocks of a plant provides a good understanding of the various limits and risks involved in the process. Any optimization strategy must be subject to these interlocks and will therefore not jeopardize the safe operation of the plant. The following process interlocks exist in the PLC for equipment protection purposes (Le Roux, 2010).

- Bag house emergency damper. This damper is opened when the bag house temperature exceeds 135 °C. At the same time the flash dryer will be tripped. The interlock resets once the bag house temperature has cooled down to below 130 °C. This will prevent damage to the filter bags which have a limited tolerance for high temperatures.
- The concentrate feeders, conveyor and back-mixer are stopped when the flash dryer outlet temperature drops to below 100 °C and resets once the temperature has risen above 110 °C. This will prevent moist concentrate from entering the product silo.
- If the disintegrator current exceeds 150 A for more than three seconds, the concentrate feed will be stopped until it has dropped to below 100 A for more than five seconds. This condition is typically an indication of an overload situation in the disintegrator and therefore this action must be taken to prevent damage to the disintegrator motor. Typically it is not required to interrupt the hot gas supply as well since the overloaded disintegrator already restricts the air flow.
- If any metal is detected on the concentrate conveyor, the concentrate feed will be stopped until the metal is removed and the metal detector has been reset by the field operator. This will prevent damage to the downstream equipment like the back-mixer and disintegrator.
- The tight shutting damper is closed in case the flash dryer outlet temperature exceeds 135 °C or if the hot gas generator outlet temperature exceeds 840 °C.

It resets at 130 ^{o}C and 785 ^{o}C respectively. This condition is named the alternative caretaker mode and serves to protect the bag-house and the HGG outlet ducting. By closing the tight-shutting damper, the entire air supply to the flash-dryer and the bag-house is sucked in through the HGG stack. This air is supplied at ambient temperature and therefore rapidly cools down the entire process downstream of the HGG.

- When the HGG average bed temperature exceeds 980 ^{o}C the coal feed will be stopped (reset at 975 ^{o}C) but the fluidizing air will continue normally to allow rapid bed cooling to take place. If any single bed temperature exceeds 1080 ^{o}C the HGG will trip, causing the flash dryer to trip as well. This is to prevent bed clinkering which occurs at high bed temperatures.
- Coal feed to the HGG will be stopped whenever the fluidizing air supply is interrupted (including alternative caretaker mode). This is to prevent accumulation of coal in the HGG due to insufficient oxygen for complete combustion as this may result in a rapid unmanageable increase in the bed temperature once oxygen supply is restored, which could then lead to bed clinkering.
- The flash dryer will trip if any of the bag house drop-out section levels goes to the high-high state to prevent damage to the bag house. If the actual pressure in the bag house is too low (vacuum) the flash dryer will trip as well to prevent imploding of the bag house structure. The differential pressure across the bag

house will trip the flash dryer if it goes too high, indicating a possible choke or bags that are not clean enough and restricting the air flow through the plant.

4 Project

The project was executed according to the sequence illustrated in this chapter. A proper understanding of the process was developed (4.1). This was then used to formalize, in conjunction with site management, the KPIs of the process to which any optimization strategy has to align (4.2). Next the current control strategy was investigated and improved where possible (4.3). To fully optimize the process, it was decided to deploy an APC strategy (4.4), and for this some additional infrastructure had to be installed (4.5). Using the new infrastructure and considering the defined KPIs, the controller design could be commenced (4.6). To support the chosen control algorithm, it was necessary to develop mathematical models of the process (4.7). No changes to the process are allowed without the necessary risk assessment (4.8). Finally the new controller was deployed (4.9) and refined (4.10). After successful implementation, the site operators and engineers were trained so that they could take ownership of the new system (4.11). In conclusion, the results obtained by the new controller justified the roll out to other similar processes (4.12).

4.1 Process familiarization

By means of a literature study, site visits, interviews, and some data analysis, a thorough understanding of the process was obtained as illustrated in the previous sections of this document. Any issues or limitations that might hamper or prevent the successful implementation of such optimization strategies also had to be identified and corrected or accommodated. These issues, as described in sections

4.1.1 to 4.1.8, were noted from literature, while interacting with the process and from interviews (including the project hazard and operability study (HAZOP)) held with site personnel and support contractors.

4.1.1 HGG

In the HGG, coal is combusted in the presence of air on a fluidized silica bed. Initially the coal is ignited through the use of low pressure (LP) gas, after which the combustion is sustained through a continuous supply of coal and air. A fluidizing fan supplies air to the bottom of the HGG. This air has three main purposes, it is used to fluidize the bed, supply oxygen to the combusting coal and become the heated gas from the HGG outlet. The balance of the ID fan's air requirement is sucked in through the HGG chimney.

Properly controlling the fluidizing air supply is important. Insufficient air supply will cause the fluidized bed to collapse and can result in incomplete combustion. By supplying too much fluidizing air, combustion efficiency is depleted and some of the heated gas, or even some of the sand, will escape through the HGG chimney – this leads to energy losses and environmental risks. Note however that there are currently no measurements (flow rate/direction, temperature or pressure) in the HGG chimney, and therefore the only indication that the HGG is optimally operated is if one can just notice, with the human eye, a very slight plume puffing out of the HGG's chimney – this indicates that the fluidizing air is perfectly balancing the ID fan's air requirement. Unfortunately there are no air flow measurements in this process and that the only indication of air flow is therefore the respective openings of the fluidizing and ID fan dampers. The ability to measure the volume of fluidizing air that is introduced to and present in the system will greatly improve the controllability of the process.

One of the biggest risks when operating the HGG is sintering of the silica bed. This occurs at around 1 200 ^{o}C. At the same time it is important to note that a minimum temperature of 600 ^{o}C is required to maintain the coal combustion. Hence a target average bed temperature of 900 ^{o}C has been chosen. This temperature is measured by four thermocouples equally distributed around the perimeter of the fluidized bed. The main means of controlling the bed temperature is through adjusting the coal supply. However, it is occasionally required to completely stop the coal supply to bring runaway bed temperatures back under control. The need for these interventions should be minimized as the coal feeders are more prone to choking when they are stopped (Rademan, 2008). It should be noted that the coal supply varies quite substantially with regard to quality, density, and size distribution. Stabilizing these factors, for example by improving the coal screening, will definitely improve the process stability. A sudden supply of fine coal into the HGG will rapidly ignite and cause significant spikes in the bed temperature – an event that is difficult to control due to the inherent change in the process dynamics (reduced dead time, increased rise time) – hence increasing the risk of sintering the HGG bed. Such particles can also easily be swept along with the hot gas, through the flash dryer, and end up burning the bags in the bag-house. It is believed that coal particles that are approximately 10 to 20 mm in size are optimal. Similar to the air supply, the actual quantity of coal supplied is not measured and therefore the only indication of coal feed rate is the coal feeder speed setpoints. This obviously introduces significant inaccuracy that will limit the benefits that can be achieved by a control system.

Historically, a local PID controller controlled the bed temperatures. Because this controller had only one output to the coal variable speed drives (VSDs), the bed temperatures were balanced by means of an adjustable mechanical gearbox. Nowadays the local controller no longer exists, and the temperatures are controlled by means of a PID loop in the PLC. This means that the PLC can now do feed balancing by adjusting the outputs to the coal feeders individually. The mechanical gearboxes are therefore no longer required. These gearboxes present unnecessary maintenance overhead and changes to their gearing configurations during maintenance impacts the control system negatively as this changes the process gains.

4.1.2 Concentrate Feed

The wet concentrate is collected from the concentrate stock yard using two operator driven overhead cranes and then dropped into one of three flash dryer feed hoppers. There are variable screw feeders underneath these hoppers that extract the concentrate onto the variable speed concentrate feed conveyor belt. Largely due to the human element that is introduced by the overhead cranes, the hoppers tend to run empty (especially during shift change over or when all three flash dryers are running). This results in large drying column outlet temperature excursions that are rather difficult to contain.

4.1.3 Back-Mixer

The back-mixer runs at a fixed speed and is responsible for mixing all the wet concentrate with a portion of the previously dried concentrate that is circulated back from the first cyclone. This serves a few purposes:

- Improved transferability of the concentrate which in turn prevents blockages
- Forms an air plug that prevents cold air from being sucked in from the concentrate feed rather than hot air from the HGG.

Note that since the circulated concentrate is already dry, it does not significantly affect the thermal efficiency of the flash dryer. Hence, no real benefit is expected from manipulating the size of the circulation load.

4.1.4 Flash Drying Column

The flash dryer column purely provides a space for the heat transfer and drying to occur. The temperature differential across this column is indicative of its drying capacity. It can be maximized by reducing the column outlet temperature and/or by increasing the HGG outlet temperature. Notice should be taken of the air temperature profile across the entire system. Significant (>50 ^{0}C) drift in the HGG outlet temperature and/or large temperature drops between the flash dryer and the bag house is normally indicative of leakages where cold air, instead of hot air from the HGG, is sucked in. It is also important to note that a minimum HGG outlet temperature is required to ensure sufficient energy to dry the minimum amount of concentrate feed that can be supplied by the concentrate feeders.

4.1.5 Cyclones

Two cyclones are used in parallel to separate the bulk of the concentrate from the hot gas directly after the drying column. The underflow from each cyclone (the concentrate) is collected in a double-flap valve before transfer to the product bin – a portion of the first cyclone's underflow is returned to the back-mixer. These double-flap valves have been found to choke up at times. There is also a piping bend that feeds the cyclones that tends to choke up. Furthermore, it has been found that the pressure transmitters on these cyclones are not always operational and accurate. Finally, it has also been noted that, to some extent due to the relatively new IsaMill circuits at the concentrators, the particle size distribution of the feed has changed somewhat. This resulted in an uneven split of material to the cyclones and also to higher than expected concentrate in the cyclone overflows. Some flow dynamic analysis is underway to investigate and correct this problem. The cyclones' overflows pass through the multi-clone before it reaches the bag house.

4.1.6 Bag House

Despite being relatively far downstream, the bag house forms a critical part of the overall process and can easily become a bottleneck that prevents further capacity increases. If the bag house is not properly designed or operated, it can lead to PGM losses via the stack, decreased production and feed shortages to the furnace. The correct type of filter bags should be utilized and all leaks should be eliminated (Gore, 2010).

The bag house consists of nine compartments that operate in parallel. Each has its own isolation valve, collection of bags and pulsing solenoids and valves. The bag house is controlled by a sequence that continuously loops through each compartment, isolates it by closing the isolation valve, pulses the bags in order to shed any collected concentrate dust and then reopens the isolation valve before continuing to the next compartment. The efficiency of this dedusting sequence is often negatively impacted when the isolation valves don't close or re-open properly (there is no feedback), or when the pulsing solenoids and valves are not fully operational. Inefficient operation leads to build up in the bag house which ultimately results in plant downtime. The concentrate dust that is collected from the bags is accumulated at the bottom of the bag house from where it is dropped into pneumatic transfer pods for delivery to the product bin. Here too, proper isolation sequences and operation are critical to avoid concentrate accumulation and ultimately further plant downtime.

It is imperative that the bag house doors are always properly closed to ensure a well sealed system with no air leaks that jeopardize the air suction of hot air all the way from the HGG through to the bag house.

It should be noted that the bags in the bag house carry an extremely high replacement cost in addition to the costs involved with plant unavailability while replacing them. Hence it is important to protect these bags as far as possible. As mentioned, this is achieved by an emergency damper that opens up as soon as the feed air temperature exceeds 130 ^{o}C, thereby ensuring ample supply of cold air to prevent the bags from burning. Opening this damper does however sometimes cause an "air-lock" on the hot air received from the flash dryer and this can take a long time to correct – resulting in excessive production loss. One option would be to consider some bypass valve that could be used to temporarily break this "air-lock" and thereby rapidly reduce the air supply temperature to the bag house.

Finally care should be taken not to operate the bag house at temperatures (also considering cold spots) beneath the dew point of the vapours in the air. Condensation of these vapours could form water and even sulphuric acid which can cause serious corrosion damage in the longer term. The exact dew point has not yet been determined, however given the negative pressure environment, it is expected to be below 100 ^{o}C.

4.1.7 ID Fan and Damper

This is a fixed speed fan and the damper position is rarely adjusted. It should be noted that this setting affects the entire air-flow through the system and hence all the dynamics of the system. It should only be adjusted after careful consideration as it could negatively impact the control system tuning.

4.1.8 Product Silo

It is imperative that the flash dryers keep the product silo relatively full at all times to ensure uninterrupted operation of the furnaces. Concentrate supply rates have been much higher (at lower grades) in the past than they have been recently - mainly due to optimization initiatives at the concentrators. This opens up the opportunity for possibly running fewer flash dryers at the same time which will enable better maintenance scheduling. It also provides the option of adding a silo management controller that will manipulate the flash dryer throughput according to the level in the product silo. This can lead to reduced start and stop cycles as well as reduced HGG caretaker periods. Hence this can enable more efficient operation, reduced LP gas costs, lower environmental impact, less thermal stresses on the equipment and less risk of HGG bed sintering as a result of temperature spikes that sometimes follow HGG caretaker mode.

4.2 Key Performance Indicators and Controller Objectives

Based on the gained process knowledge and understanding, and in collaboration with site management, the following KPIs were defined. These will guide the performance assessments as well. The first three are process objectives while the next two are safety objectives and the final one is a control objective.

4.2.1 Stabilize Drying Column Outlet Temperature (Process Objective)

Once a stable drying column outlet temperature has been achieved, the setpoint of this temperature (and hence the low-temperature feed stop safety margin) can be reduced, effectively enabling higher throughput due to the increased temperature differential and reduced coal demand while still avoiding concentrate feed stops. Unnecessary feed stops result in wasted energy (as the HGG is operating without drying) and potential mechanical damage to the feed belts (Rademan, 2008). To measure the stability of this temperature, the following metrics can be calculated:

- Average accumulated flash dryer outlet temperature control error.
- The concentrate feed stop count due to the flash dryer outlet temperature dropping below 100 ^{o}C.

4.2.2 Maximize Concentrate Throughput (Process Objective)

Consistently maximizing the concentrate throughput through the flash dryers will provide optimal utilization of these assets and should provide more opportunity to run with only some of the flash dryers at any given point in time, resulting in reduced operating cost and improved maintenance scheduling. It should be noted that the concentrate moisture varies over time. This moisture is only measured daily, but as it will affect the work load on the dryer, it should also be taken into account. To measure the concentrate throughput and moisture removed, the following metrics can be calculated:

- Total concentrate dried: Accumulated concentrate feedrate.
- Average concentrate feedrate: Total concentrate dried divided by plant run time.
- Total moisture removed: Accumulated concentrate feedrate multiplied by the corresponding daily moisture sample. The single moisture measurement per day provides limited accuracy, but it does ensure a reasonable estimate.
- Average moisture removal rate: Total moisture removed divided by plant run time.

4.2.3 Minimize Specific Coal Consumption (Process Objective)

Reducing the amount of coal required to dry a ton of concentrate has cost saving and reduced environmental impact implications. Note that the actual coal feedrate is not measured; hence it will be approximated by using screw feeder speed. To measure this objective, the following metric will be calculated:

- Total coal consumed: Accumulated coal feeder speed multiplied by feedrate correcting factor
- Average coal consumption rate: Total coal consumed divided by plant run time.
- Overall specific coal consumption: Total coal consumed divided by total moisture removed.
- Average specific coal consumption: Overall specific coal consumption divided by plant run time.

4.2.4 Minimize HGG Bed Temperature High Limit Violation (Safety Objective)

Avoiding large periods of extreme high HGG Bed temperature will reduce the risk of bed sintering. This can be calculated as follows:

- Average accumulated HGG average bed temperature control error.
- Totalize the time that the average HGG Bed temperature exceeded 950 ^{o}C.

4.2.5 Maintain minimum Air to Coal ratio (Safety Objective)

Ensuring sufficient air to coal will minimize the risk of incomplete combustion that can potentially lead to CO buildup and/or coal accumulation on the bed. Note that since neither actual air nor coal feed rates are measured, this can only be approximated by calculating the air damper position relative to the coal feeders' speed. Historical data has shown that a ratio of 40 % is more than adequate. This can be calculated as follows:

- Totalize the time that the time that the air to coal ratio is less than 40 %.

4.2.6 Maximize APC utilization (Control Objective)

Once it has been proven that the plant performance is improved by APC, it is important to sustain and maximize its utilization. A decrease in APC utilization is indicative of a reduction in confidence of the site in its ability. This may be because of infrastructure issues, model drift or other tuning issues and needs to be promptly corrected to maintain the relationship between the production and control engineers as well as the return on capital investment of installing the APC in the first place. This will be calculated as follows:

- APC Utilization: Time that plant was running under APC control divided by the time that the plant was running.

4.3 Base layer control investigation

As part of the base layer control analysis, it was found that even though the concentrate mass flow is measured, it was not used as part of the control strategy. Instead, the flash dryer outlet temperature was controlled by directly manipulating the concentrate feeders. This introduced unwanted and unnecessary non-linearity and uncertainty into the system. Given the variability of the concentrate moisture and levels in the feed hoppers, it is inaccurate to assume that a unit change in feeder speed will always result in the same unit change of concentrate mass flow and hence the same unit change of flash dryer outlet temperature. To address this problem, a new concentrate mass flow PID controller was installed and the existing flash dryer outlet temperature PID controller was modified to now supply the setpoint to this new mass flow controller (Figure 10)

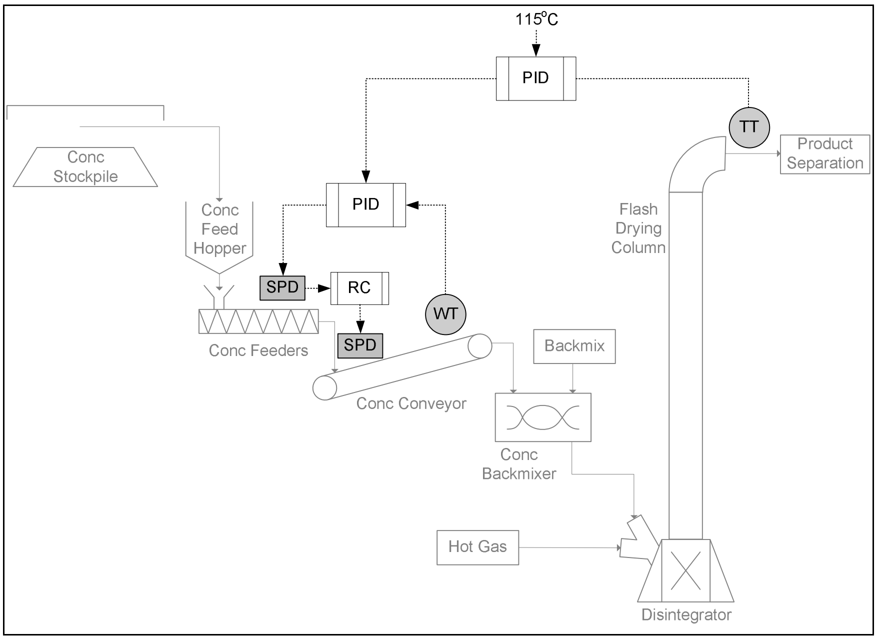

Figure 10: New concentrate mass flow PID controller

Advantages of this new PID controller include:

- Better flash dryer outlet temperature control due to the reduction in non-linearity of the process through the inclusion of additional process knowledge.

- Better feed disturbance detection and rejection due to the more direct measurement of the concentrate feed rate, especially when the feed hopper runs empty.
- Better use of the entire range of the concentrate feeders. Prior to installing the new PID controller, these feeders had a relatively low high limit to avoid chokes in the downstream ducting, however that limit could now be relaxed in favour of the more appropriate high limit on the concentrate mass-flow.

4.4 Motivation for APC

Following a thorough investigation of the process, it was evident that the existing base layer control strategy is inadequate to extract the full potential of the process. Despite initial improvements, the process still exhibited significant instability (as can be seen in the time series charts shown in Appendix A) that lead to frequent process interruptions and hence reduced throughput. This can be explained by the interaction that exist between the process variables, e.g. coal feed and fluidizing damper adjustments not only affects HGG bed temperature but also HGG outlet temperature and hence drying column outlet temperature, as well as the dead times that are present as a result of, for example, a long feed conveyor belt. The resulting instabilities necessitated selection of conservative operating ranges and hence the process could not be operated at maximum efficiency. Consequently, it was decided to introduce an APC strategy that would primarily stabilize the process despite the presence of interaction and delays and furthermore improve throughput and efficiency.

4.5 Infrastructure

Anglo Platinum has a well developed APC platform that ensures that all control projects are executed in line with the necessary standards to guarantee reliable and sustainable controllers. This controller for this project was therefore also designed and deployed using these tools.

4.5.1 PLC Interface

The PLC, with its supervisory, control and data acquisition (SCADA) interface to the control room operator (CRO), will always remain primarily responsible for the safe operation of the process. All advanced and supervisory control must therefore communicate with the plant via the PLC and through a dedicated interface that can be enabled or disabled by a single master switch. This ensures that all the plant safety interlocks remain intact and that the CRO or other defined process interlocks can switch off the APC whenever deemed necessary. The standard APC block that is deployed in Anglo Platinum's PLCs offers the following functionality (Lombard, de Villiers & Humphries, 2010):

- Heartbeat functionality that will switch off the APC controller whenever communication with external APC software is lost.
- Additional interlocks that need to be true in order for APC to be active. These can be determined internally or externally to the PLC.
- On-Off switch. The APC will be enabled as long as this switch is on and all interlocks, including the heartbeat, are healthy.

- Support for up to 4 MVs and 4 CVs. These are linked to the appropriate instrument and PID controller blocks. Furthermore, an operator or engineer with the necessary authentication can specify limits and targets for each of these.
- SCADA faceplate that provide the operator with relevant APC specific information like current detected process state and messages listing the control actions that are taken along with reasons where available.

4.5.2 APC Server and Software

Anglo Platinum utilizes the Anglo Platinum Expert Toolkit (APET) to host all of its APC controllers. APET is a Gensym G2 based application that enables the rapid capturing of an asset model consisting of plant areas, process-cells and units according to the ANSI/ISA S88/95 standards. Each unit in turn contains various pieces of equipment with their associated instruments and controllers. APET facilitates the implementation of various types of APC controllers. Depending on the chosen control algorithm, the controller will execute either internally to APET or externally on a 3rd party software platform. Either way, all control actions are always governed by the process states and timers defined in APET. APET also enables powerful hybrid solutions that combine different control algorithms based on different process states. All communication between APET and the PLC and the 3rd party software is OPC based and facilitated by the iDX middleware product (de Clerk, de Villiers & Humphries, 2010).

4.5.3 MPC Server and Software

For this project, it was decided to implement an Aspen DMC+ model predictive controller. This software resides on a separate server and communicates with APET using Aspen's CimIO protocol.

4.6 Controller Design

4.6.1 Controlled and Manipulated Variables

From the defined KPIs, and provided the existing instrumentation it is evident that the controller should and can only have the following controlled variables (CVs):

- HGG Bed Temperature
- HGG Outlet Temperature
- Flash Dryer Outlet Temperature

Furthermore, it can only have the following manipulated variables (MVs):

- Coal Feed - through adjusting the coal feeders' speed.
- Fluidizing Air Supply – through adjusting the fluidizing damper position.
- Concentrate Feed – through adjusting the concentrate mass flow setpoint.
- Induced draft – through adjusting the ID fan damper position.

Figure 11: Process Inputs and OutputsFigure 11 illustrates a conceptual input-output diagram of the flash drying process.

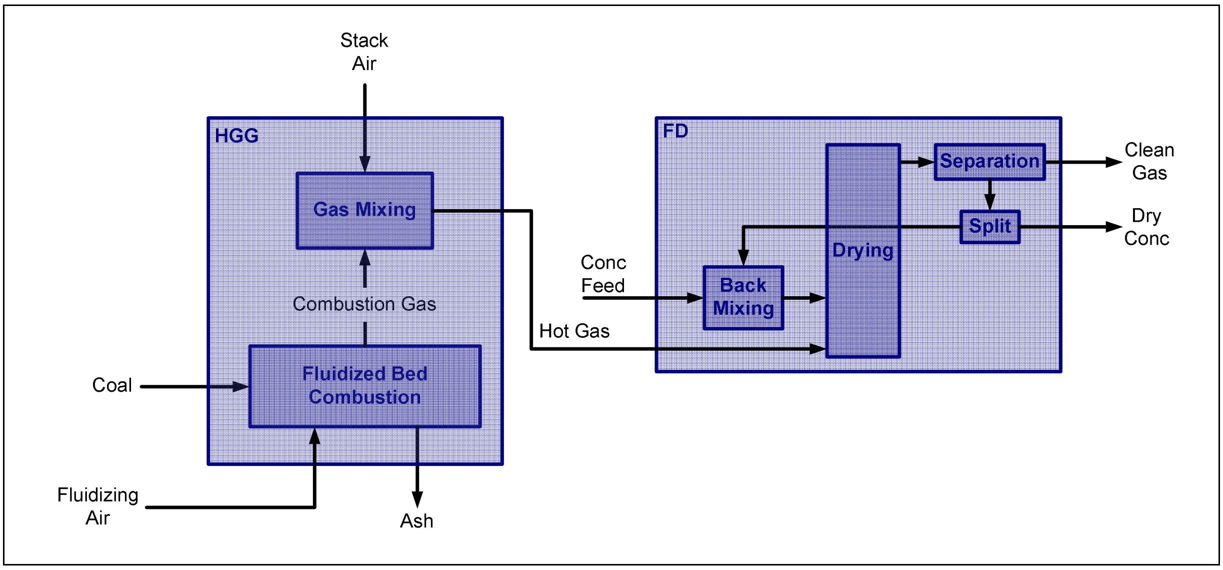

Figure 11: Process Inputs and Outputs

4.6.2 HGG Stabilization Reasoning

Stabilizing the HGG operation is important for the following reasons

- A consistent stable hot gas supply to the flash dryer, with only sufficient energy to meet the moisture drive-off demands, is required in order to allow downstream stability.
- In order to sustain the coal combustion in the HGG a minimum bed temperature of 600 ^{o}C has to be maintained. In case the temperature drops below this, the bed will be "lost" and will have to be reignited using LP gas – this is an expensive and complex procedure.
- Bed sintering is one of the biggest risks surrounding the HGG operation and has to be avoided as the HGG will be out of operation for several days if this occurs. Bed sintering takes place at temperatures exceeding 1200 ^{o}C.

- The ducting that channels the hot gas from the HGG to the flash dryer is not equipped with lining that can resist excessive temperatures. Hence, in order to avoid equipment damage, it is important to avoid HGG outlet temperatures that exceed 780 ^{o}C.
- Unstable temperatures expose the equipment to unnecessary thermal fatigue (Fox, 1957) which will increase maintenance costs in the long run.

4.6.3 HGG Disturbances

Disturbances that affect the HGG operation include the following:

- Coal quality (composition, calorific value, moisture content) and size distribution. These are unmeasured disturbances that significantly affect the amount of energy that is added to the system for each turn of the coal feeders. Better quality coal provides more energy and fewer impurities. Size distribution and moisture content impacts on the transportability of the coal. Coal fines ignite much faster than the bigger chunks and also bodes the risk of being swept away through the flash dryer and into the product separation stage where they can damage the bags in the bag house.
- Ambient temperature. Seasonal changes, humidity and the time of day influence the amount of heat that is absorbed by preheating the combustion air.
- Air flow balance in the HGG. The HGG always operates under negative pressure. Hence all air, demanded by the ID fan suction, which cannot be supplied through the HGG bed by the fluidizing fan, is sucked in via the HGG

stack. Since none of these air flows are measured, it is not possible to determine the ratio of stack air versus fluidizing air.

4.6.4 Drying Column Stabilization Reasoning

Stabilizing the drying column operation is important for the following reasons:

- The primary objective of the flash dryers is to supply bone-dry concentrate to the smelting furnaces. Ensuring that the column outlet temperature consistently exceeds 100 ^{o}C will achieve this objective.
- However, operating at column outlet temperatures that significantly exceeds 100 ^{o}C constitutes wasted energy (as long as it does not lead to downstream condensation) and should therefore be avoided.

4.6.5 Drying Column Disturbances

Disturbances that affect the drying column operation include the following:

- Concentrate moisture (Rademan, 2008).
- Concentrate feed hopper level. If the feed hoppers run empty, there is no concentrate to absorb the energy in the hot gas and hence the column temperature will increase rapidly.
- HGG outlet temperature. This disturbance will be minimized by proper control of the HGG.

4.7 Model Identification

This section discusses the process dynamics identification process that was followed in order to obtain the mathematical input output model of the process that forms the basis of the dynamic matrix controller.

4.7.1 Fundamental gain matrix

Prior to step testing, it is important to list what types of responses to expect given an initial fundamental understanding of the process. Table 4 lists the expected direction of movement of the CVs in response to positive step changes in the respective MVs.

Table 4: Fundamental Gain Matrix

CVs / MVs	HGG Bed Temperature	HGG Outlet Temperature	Drying Column Outlet Temperature
Coal	⇧	⇧	⇧
FD Damper	⇩ ?	⇧ ?	⇧ ?
Conc	⊘	⊘	⇩
ID Damper	⊘	⇩	⇩

- From a simple energy balance, it is clear that increasing coal feed will increase the HGG bed temperature as there is more carbon available for combustion. If the HGG bed temperature increases, the HGG outlet temperature will increase as the flue gas temperature is higher. Finally, provided all else remains constant, the drying column outlet temperature should increase as well.

- The effect of the fluidizing damper is less predictable, especially given the lack of instrumentation to assist in accurately monitoring key parameters like air flow rates and excess oxygen. Assuming excess oxygen conditions, the following hypothesis holds: Increasing fluidizing air will further increase the oversupply of oxygen and therefore not promote further combustion, but rather cool down the HGG bed. Also take note that the thermocouples that are used to measure bed temperature are located in between the air supply and the primary combustion zone and will therefore also be cooled directly by the increased air. With increased fluidizing air, there will be more flue gas (offset by less air sucked in via the HGG chimney) but at a slightly colder temperature. Hence, the HGG outlet temperature is expected to initially rise slightly but may then return to its previous value (as the decreased temperature offsets the increased flow rate). Any change in the HGG outlet temperature should reflect in the column outlet temperature.
- Feeding more concentrate into the system will have no effect on the HGG as it is upstream to the concentrate feed point, however the increased drying demand will cause a definite reduction in the drying column outlet temperature.
- Increasing the ID fan damper should not have any effect on the HGG Bed temperature, it will however result in more cold air being introduced via the HGG chimney (for mixing with the hot flue gas) and hence the HGG and drying column outlet temperatures should drop. It should be noted that by adjusting

the ID fan damper, the air flow throughout the system changes thereby altering the residence time, process dynamics and cyclone efficiencies and should therefore be done with due caution. For this reason, the ID fan was excluded for purposes of this project. This MV can be reconsidered once additional instrumentation, that will allow for proper balancing of the air flows in the system, has been installed.

4.7.2 Step testing

A Hazard Identification and Risk Analysis (HIRA) procedure was conducted to explain the step testing process to the control room operators, to identify the acceptable ranges within which the MVs may be stepped, to agree upon tolerable movements in the CVs and also to re-iterate any associated risks.

During the step testing process, all three temperature PID controllers were switched to manual and hence the coal feeder speed, fluidizing damper opening, concentrate feed and ID fan damper opening had to be adjusted manually. The process now had to be stabilized as far as possible before stepping the MVs one by one while keeping the other MVs constant. Throughout this process it was important to closely monitor all the CVs in order to observe their respective responses, but also to be prepared to take any corrective action should a CV start to move beyond its acceptable range.

Step testing the flash drying circuit proved rather difficult given the high frequency of significant disturbances (most notably the feed hopper that runs empty from time to time, but also other unmeasured disturbances in the coal supply and concentrate moisture) that result in sudden considerable ramps in the various CVs (often exceeding 20 ^{o}C per minute). Given the dead times in the order of 2 to 3 minutes, it becomes essentially impossible to wait long enough in-between steps to allow for full time to steady state. This indicated that most process responses will have to be modeled as pseudoramps. This is because pseudoramp control variables allow the DMC controller to control relatively long settling-time processes (where there is a large disparity in CV time constants and where very long open-loop response times are present) without considering steady state control issues. The feedback correction of such variables is handled by combining a contribution from the bias prediction error term with a contribution from the rate-of-change of the bias prediction error term (Hurowitz, 1998)

Step testing was conducted over several days; first manually to obtain initial models, and then by using the DMCplus Smart Step technology. Figure 12 and Figure 13 show the time series data that was recorded during one of the step testing sessions.

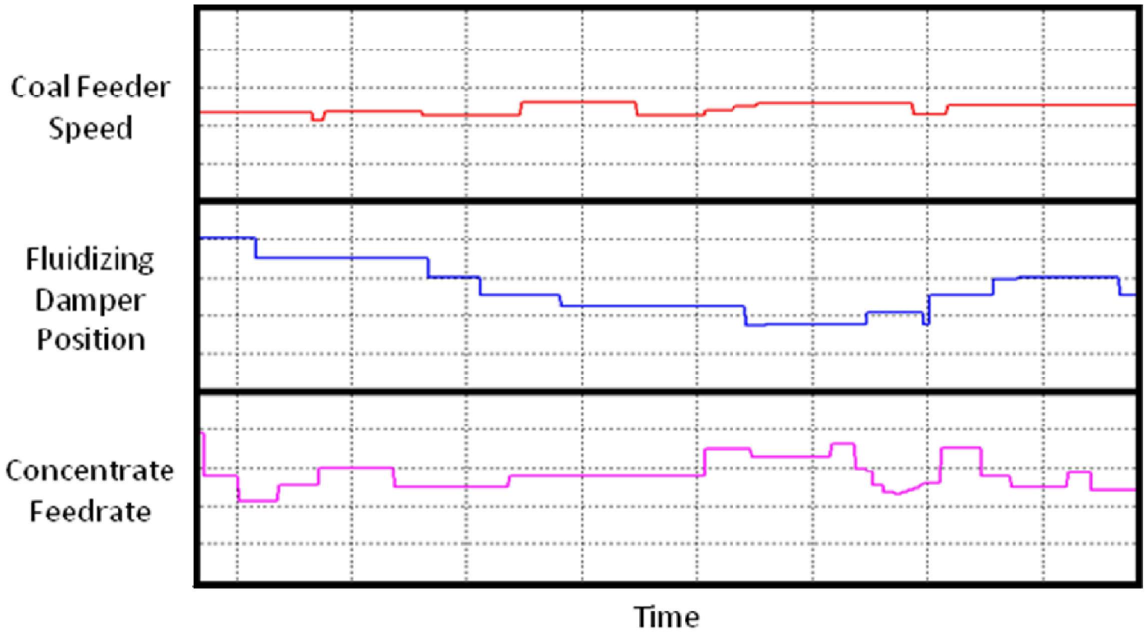

Figure 12: Example of manually introduced steps in Coal Feeder Speed, Fluidizing Damper Position and Concentrate feedrate respectively.

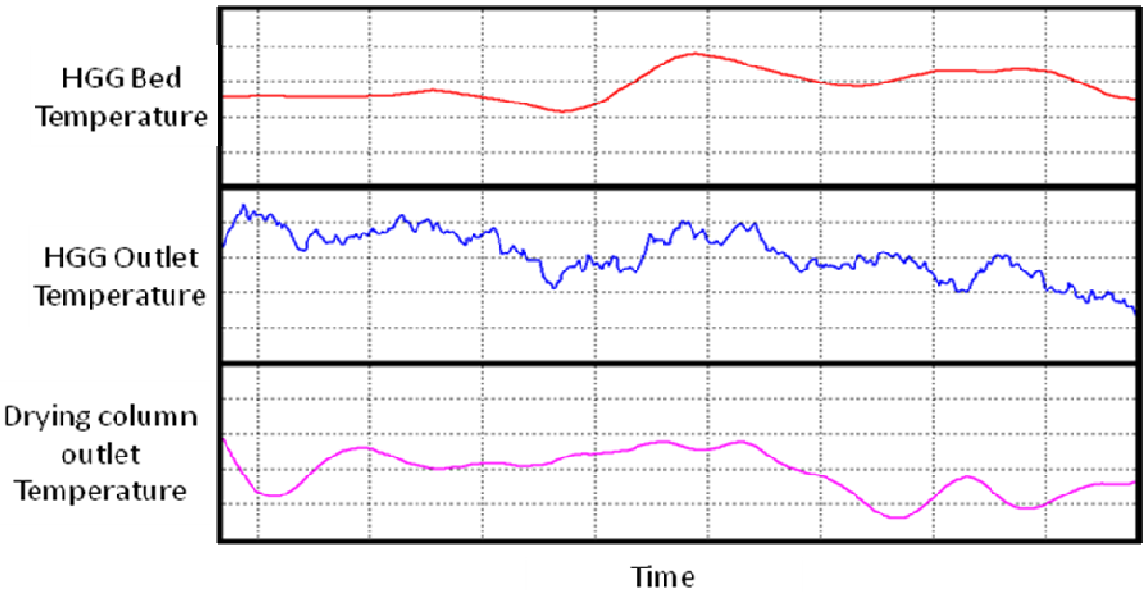

Figure 13: Example of the process responses of the HGG Bed, HGG outlet and drying column outlet temperatures respectively in reaction to manual steps shown in Figure 12

4.7.3 Final Model Matrix

Using DMCplus' model identification software to analyze the recorded step test data, the final process model matrix could be determined. This matrix is shown in Figure 14.

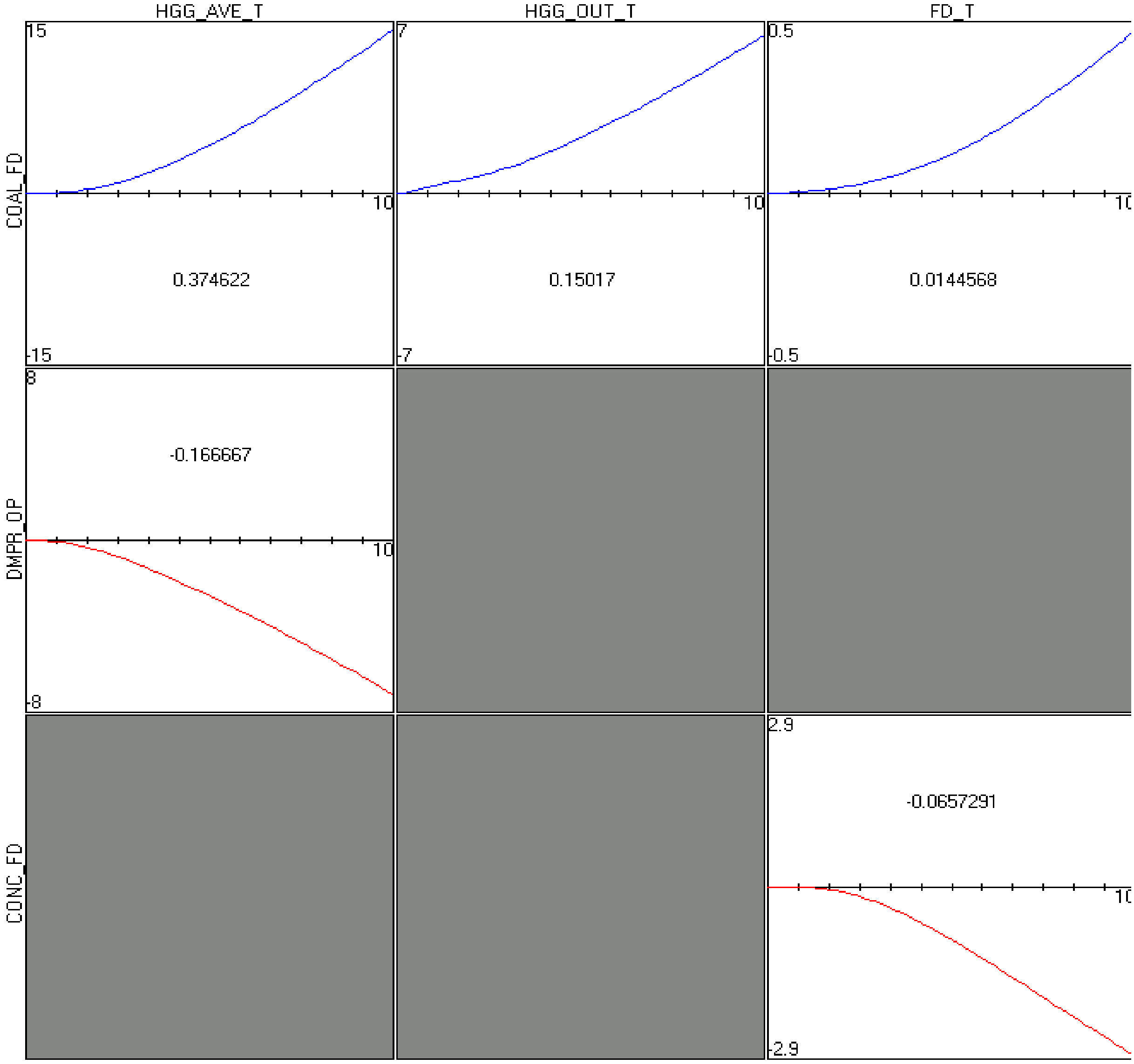

Figure 14: Final process model matrix as deployed using DMCplus

It should be noted that the gains identified corresponded well with the expected gains listed in Table 4. As explained in the previous section, the process disturbances, combined with strict operating limits, do not allow adequate time for any of the process variables to reach steady state, hence they all had to be modeled as pseudoramps. Furthermore no significant models could be found between the fluidizing damper position and the HGG outlet or drying column outlet temperatures. This too corresponds with the uncertainty listed in Table 4. Finally, as mentioned, the ID fan was excluded from the controller and therefore kept constant throughout the step testing.

4.8 HAZOP

Before commencing with the commissioning stage of the project, a detailed HAZOP had to be completed where all the relevant engineering disciplines were represented. At the start of the meeting, the controller objectives and operation were briefly explained and discussed after which all potential risks were accessed (van Rooyen, 2010). Given that existing PLC interlocks address the risks identified by previous HAZOP meetings, and that all these interlocks remain in place and active while the plant is under APC control, it was agreed that no additional interlocks (except the ones that will prevent APC operation) would be required. Furthermore, the existing basic control layer (PID controllers) will always remain as a fallback strategy when APC control is not available or if the CRO decides to switch of the APC control for whatever reason. The APC control strategy does not cater for startup conditions and will therefore only become available after the startup sequence has completed, the HGG has reached operating temperature and all the other equipment is also running. During the HAZOP, all initial acceptable CV and MV limits were also discussed and confirmed. It was reiterated that a proper air to coal ratio should be maintained in order to avoid bed sintering, a risk that is minimized by existing PLC interlocks and proper selection of air, coal and air-to-coal ratio limits. The potential risk of cold spots in the bag-house as a result of lower column outlet temperatures was raised. These cold spots could bring about sulfuric acid condensation which can cause long term corrosion issues. This is however an operational issue and will be treated as such. If this is found to be a problem, the controller limits will simply be adjusted accordingly.

4.9 Commissioning

Most infrastructure issues were already sorted out during the step testing phase, so prior to commissioning only a few final integration tests had to be performed to ensure reliable data transfer from the PLC to APET to DMCplus and back. Next the controller model was placed online so that the controller predictions and suggested control actions could be evaluated. After confirming the open loop performance, the controller was placed online and closely monitored by the control engineers and site personnel. A gradual approach was followed by first controlling just the HGG and later adding the drying column as well. Tuning the controller is an iterative process of adjusting various parameters in the DMC algorithm in order to achieve the desired controller performance in terms of setpoint tracking and disturbance rejection. These parameters include:

- General
    - Filtering
    - Limits and targets
    - Models (mainly delays and gains)
- MVs
    - Move suppression
    - Maximum moves
    - Steady state costs
- CVs
    - Limit rankings
    - Steady state equal concern errors
    - Ramp rates, horizons and rotation-factors.

4.10 Process States and intervention control

As mentioned before, hybrid control solutions can be utilized to leverage the best characteristics of different control algorithms. In this project the need for certain intervention control measures were soon identified. The need for these measures are detected in APET by means of process state expressions and are then either implemented by adjusting settings in DMCplus or by overriding the DMCplus control action altogether. Note that all process state calculations are governed by a certain persistence requirement to avoid reacting on insignificantly short spikes in the data. Table 5 lists the details of the process states that have been defined, how each is calculated and what intervention measure is taken.

Table 5: Process States, their calculation and resulting intervention control

Process State	Calculation	Intervention
Extreme Low Column Outlet Temperature	Column Outlet Temperature Below target and predicted to reach 100 °C within less than 2.5 minutes	Cut concentrate feedrate to minimum
Extreme High HGG Bed Temperature	Average HGG Bed Temperature exceeds 925 °C and predicted to reach 930 °C in less than 2.5 minutes or Individual HGG Bed Temperature exceeds 1 000 °C	Cut coal feeder speed to minimum
Concentrate Feed Disturbance	Concentrate feed conveyor is running but its mass flow PID is not positive setpoint tracking (PV < (SP – margin))	Set windup flag in DMCplus to prevent further concentrate feedrate increases
Concentrate Feed Stopped	Concentrate feed conveyor not running	Set windup flag in DMCplus to prevent concentrate feedrate changes
Coal Feed Stopped	Coal feeders not running, tight shutting damper closed or fluidizing fan stopped	Set windup flag in DMCplus to prevent coal feeder speed changes

In addition to these process states, the following adaptive measures are also taken in order to ensure the necessary controller aggressiveness to deal with different operating regions.

- When the column outlet temperature moves below its target, the DMCplus ramp rate for this CV is set to a more aggressive value and reset when the temperature moves back above its target.
- While the HGG average bed temperature is within the band of its target plus 7 ^{o}C and minus 20 ^{o}C, the move suppression factor in DMCplus on the coal feed is set to a tighter value and reset when this CV moves outside this range.

4.11 Training and handover

Effective change management is vital to the success of the new control system (Rademan, 2008). To ensure full acceptance of the new APC controller on site, the staff at various levels of the operation were involved throughout the project. Regular progress update sessions were held during which a deeper understanding of the new controller was developed and valuable feedback could be solicited. A detailed project report and operator manual were compiled. These documents captured the process operation, the details of the new controller and how to interact with it. With the help of the site metallurgists, in reference to the operator manual, the CROs of all the shifts were trained on the new additions to the SCADA that they now have to interact with. They were taught what the controller does, what is required for it to run, and how it can be switched off if required.

4.12 Roll out

After the successful commissioning of the new APC strategy on the first flash dryer at Waterval Smelter, and the realization of some initial benefits, it was logical to roll out the controller to the other flash dryers as well. At the time of writing this report, all three flash dryers at Waterval Smelter have been placed under APC control, while the APC projects for the flash dryers at the other smelters were in the planning phase.

5 Results and Discussion

In order to evaluate the performance of the newly installed control strategy, process data was collected throughout the project, starting just prior to step testing throughout commissioning and including a few months post hand-over. The time series data has been split into monthly chunks and is shown in Appendix A (Figure 22 to 31). Each month's data has also been summarized into a table (Table 7 to 16) listing the most important performance indicators for that month including a comparison between APC and PID control.

Table 6: Summarized Results

StartDate:	2010/02/01 00:00						
EndDate:	2010/11/18 23:59						
Duration:	291 days						
	Total		**PID**		**APC**		**Change**
PlantRunTime [days] With Utilization [%]	154	52.9	67	43.4	87	56.6	
Wet Concentrate Dried [Normalized] With hourly average [Normalized]	195 410	52.9	81 963	51.1	113 447	54.3	6.3 %
Moisture Removed [Normalized] With hourly average [Normalized]	160 991	43.6	66 707	41.6	94 284	45.1	8.6 %
Coal Used [Normalized] With hourly average [Normalized]	210 200	56.9	89 583	55.8	120 616	57.7	3.4 %
Specific Coal [Coal/Moisture]	1.31		1.34		1.28		-4.7 %
Accumulated HGG Temp Err [°C] With daily average [°C per day]	30 306	196.8	11 779	176.1	18 526	212.8	20.8 %
Accumulated FD Temp Err [°C] With daily average [°C per day]	6 505	42.3	3 635	54.4	2 870	33.0	-39.4 %

Table 6 shows the result of combining all the monthly results tables. It covers 291 days during which the flash dryer was operated just more than half of the time. During this time, the plant was under PID control for 43.4 % and under APC control for 56.6 %. It can be seen that, while under APC control, the process performs significantly better. This is highlighted by an 8.6 % improvement in the average moisture that was removed per hour. Furthermore, the 4.7 % improvement in specific coal consumption (units of coal used to remove one unit of moisture) offsets the slight increase in overall coal consumption.

To analyze the temperature stability, the average accumulated control error per day was calculated. Here it can be seen that the HGG Bed temperature became slightly less stable (20.8 %), while the drying column outlet temperature is now significantly more stable (39.4 %). Note that this is a direct consequence of the controller design in which the HGG Bed temperature is allowed to drift slightly in order to achieve more stable downstream conditions.

The rest of this section will show how the most important indicators have changed during the course of the project.

The blue bars in Figure 15 indicate the plant utilization per month as well as the overall average for the duration of the project. The utilization of the plant is driven strongly by furnace demand, the performance of the other flash dryers at the same smelter and maintenance schedules.

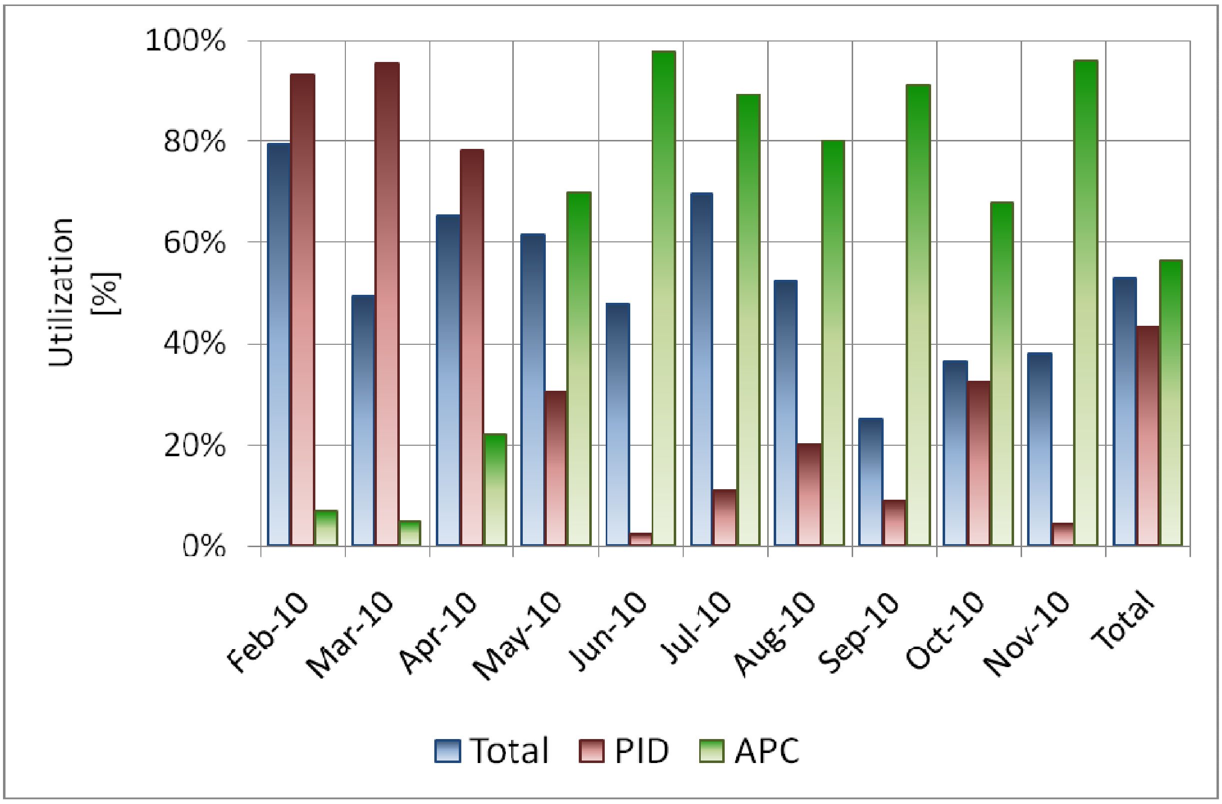

Figure 15: Plant and Controller Utilization

Note that since APC has been installed at all of these flash dryers, the plants have produced record tons and as a consequence can be operated less frequently. The red bars indicate the percentage of time that the plant was running under PID control. For the balance of the time the plant was running under APC control as indicated by the green bars. Note that since commissioning the APC in June, the APC has been consistently used for at least 80 % of the time. The exception was a short period early in October during which a communications fault prevented APC from switching on. Note that this was quickly rectified and countered with an APC utilization in excess of 95 % in November.

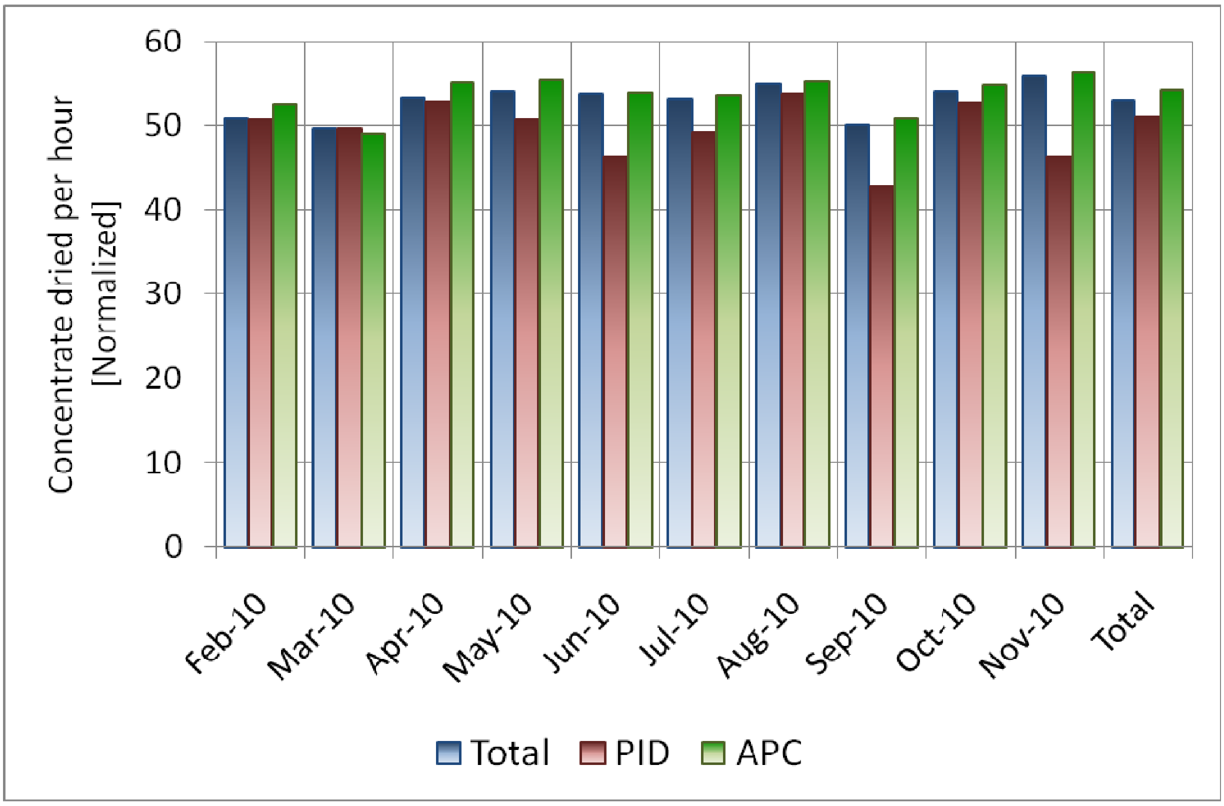

Figure 16: Average wet concentrate dried per hour

Figure 16 shows the average amount of wet concentrate that was processed per hour for each month. Note how the APC (green bars) manages to consistently achieve a higher concentrate feed rate than the PID control scheme (red bars).

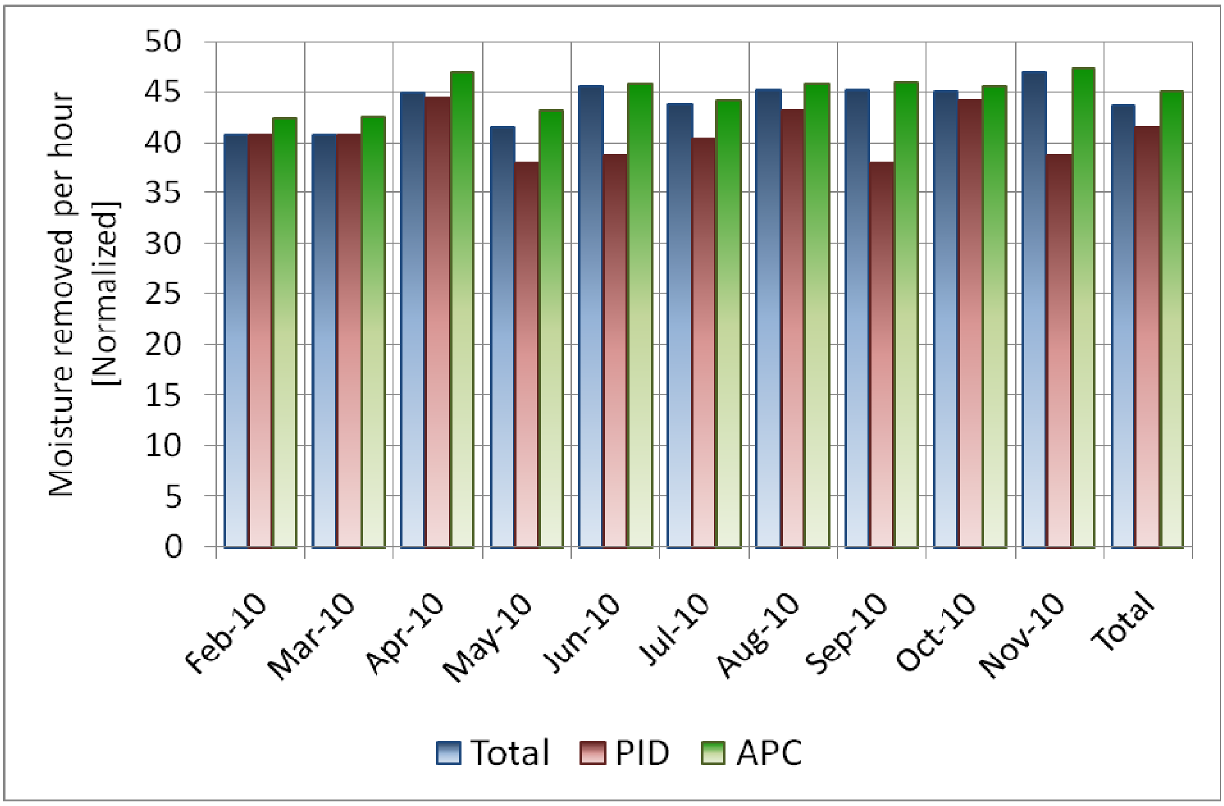

Figure 17: Average moisture removed per hour

The data for Figure 17 was generated by combining the concentrate feed rates of Figure 16 with the daily recorded moisture readings. Here it can be seen that APC not only managed to dry more concentrate, it also managed to do this in the presence of higher feed moisture. As shown in Table 6 APC did not only treat 6.3 % more concentrate, but also removed 8.6 % more moisture.

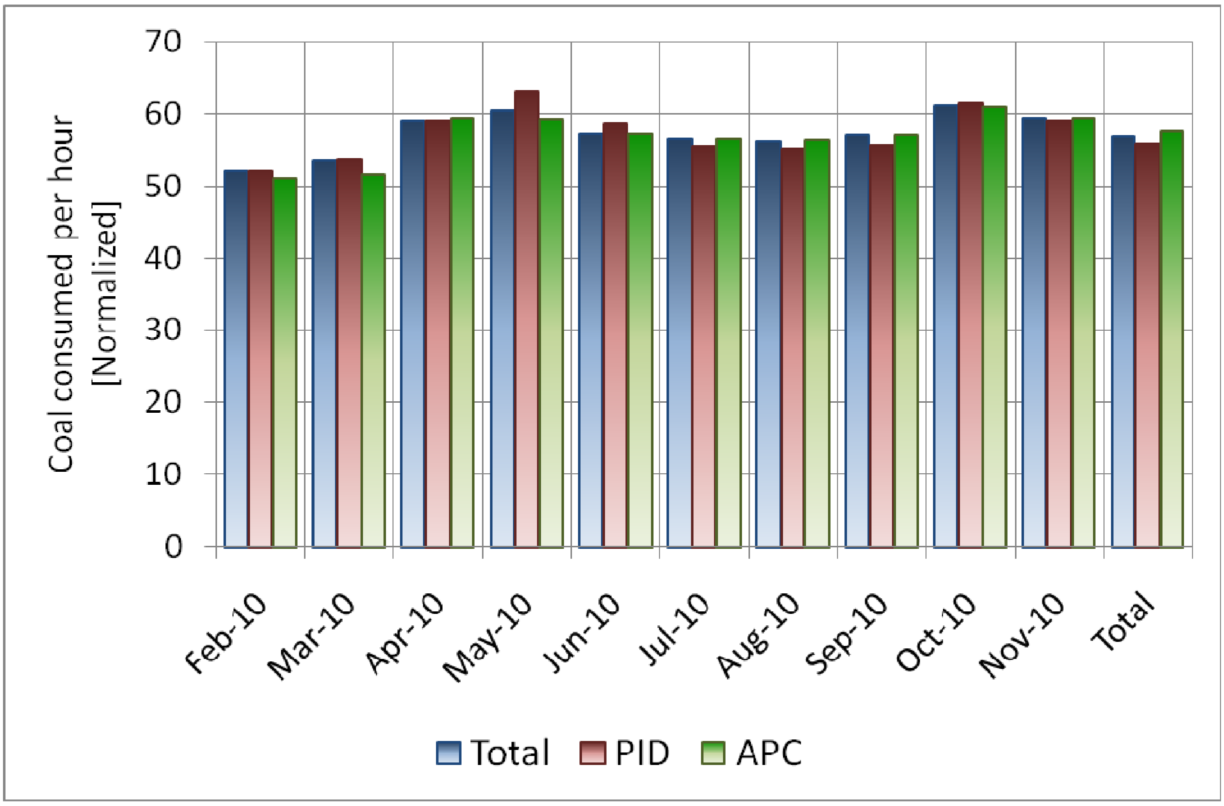

Figure 18: Coal consumed.

The exact coal consumption cannot currently be measured therefore it can only be estimated from the coal feeder speeds. I.e. it is assumed that one feeder revolution consistently supplies a fixed amount of coal. As shown with the concentrate mass flow controller, this is not a very good assumption, however it is the best possible estimate and it is expected to be more accurate over a larger time window. Figure 18 shows the data based on this assumption. Note that coal consumption is normally slightly higher when running the APC controller, however as mentioned before, this is strongly offset by the significantly improved concentrate throughput.

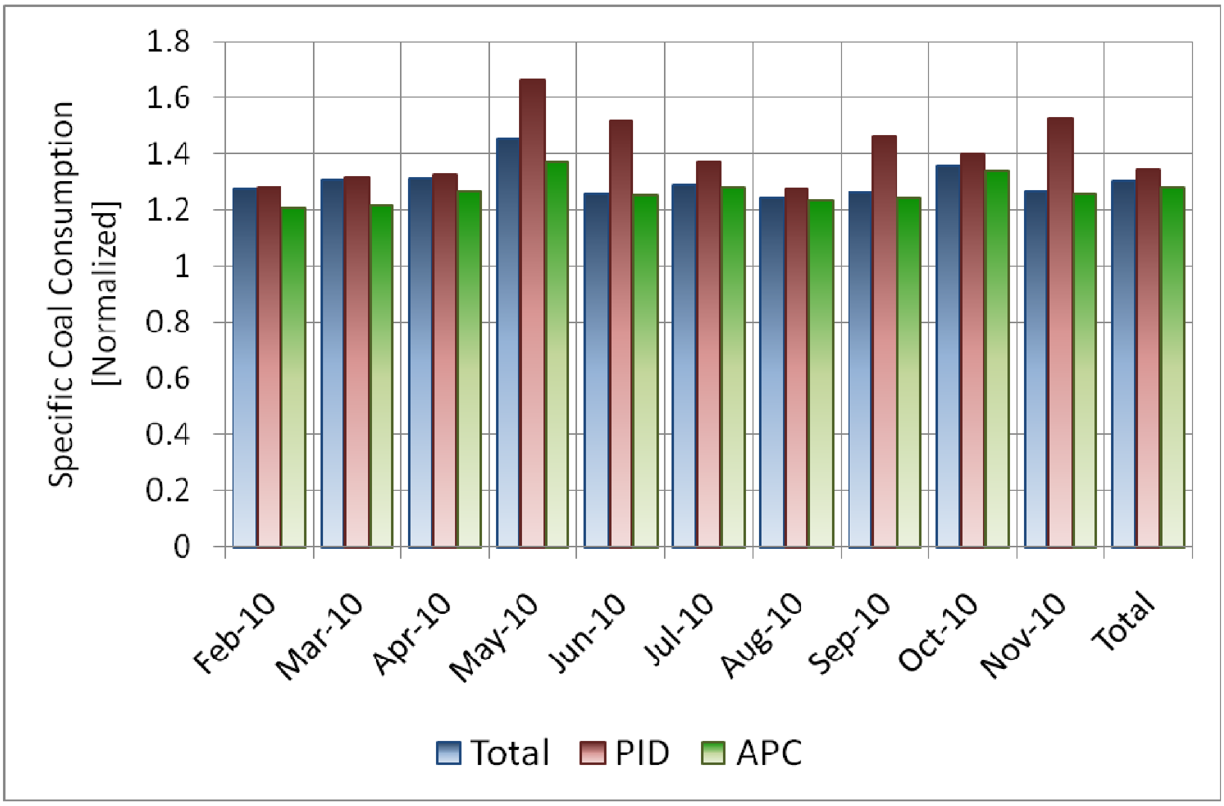

Figure 19: Specific Coal Consumption (Coal used to remove unit moisture)

Figure 19 shows the results of combining Figure 18's data with that of Figure 17. From this the specific coal consumption (how much coal is required to remove one unit of moisture) was calculated. Note that, based on this metric, APC also constantly out performs PID control as the aim is to minimize specific coal consumption.

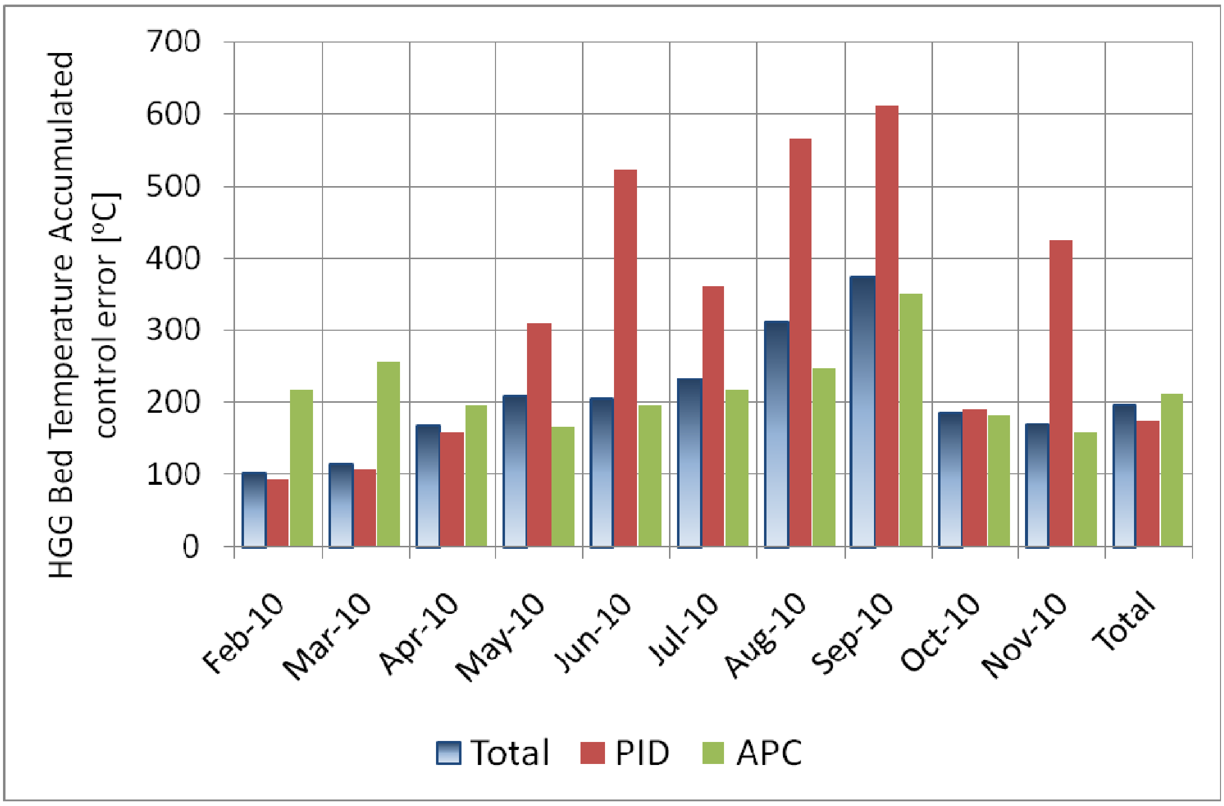

Figure 20: HGG average bed temperature accumulated control error per day

In order to evaluate temperature stability, the absolute of the temperature setpoint offset was integrated over time and divided by total run time in order to derive an accumulated control error per day. Figure 20 shows this metric over time for the HGG average bed temperature. Note that at times PID provides slightly tighter HGG bed temperature control, but this is in line with the controller design where the HGG bed temperature is allowed to slightly drift in order to stabilize downstream conditions. Also the HGG bed temperature stability is showing a positive trend even for APC control.

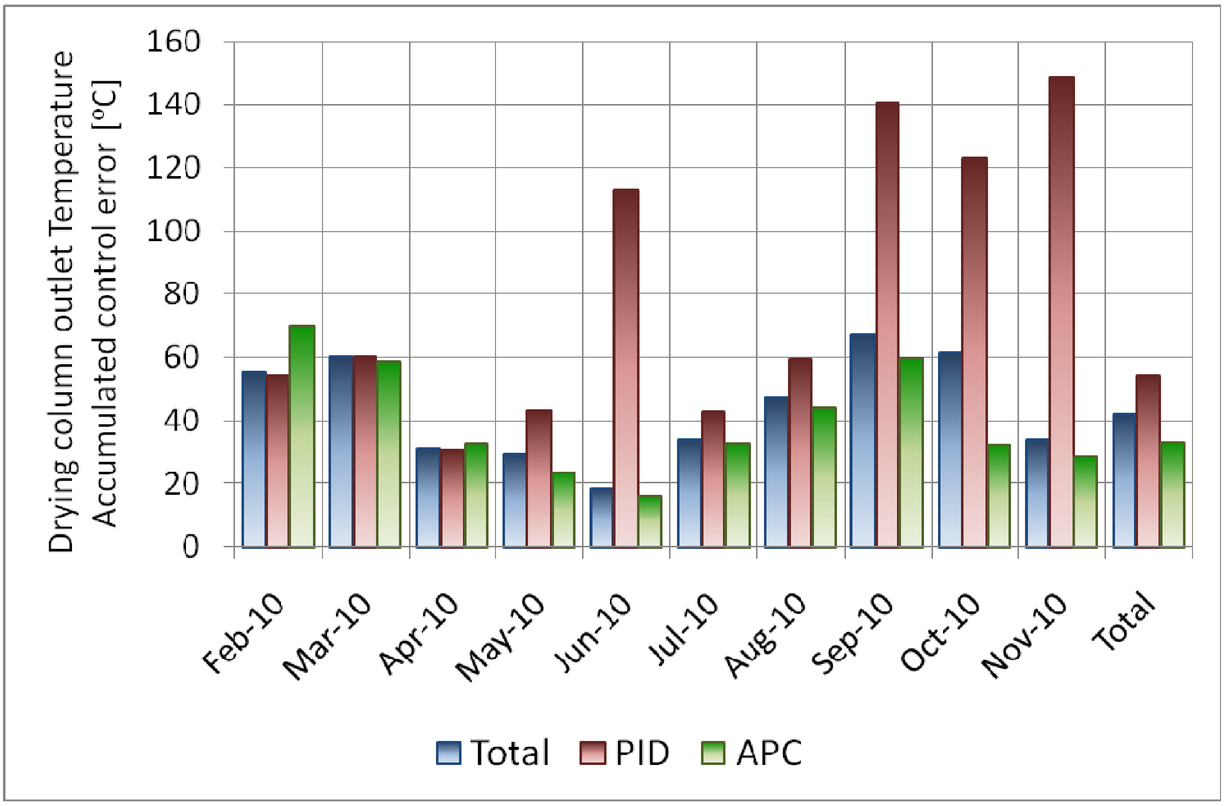

Figure 21: Flash Drying Column Outlet temperature accumulated control error per day

Figure 21 shows the same temperature stability metric, but for the flash drying column outlet temperature. Note that this is a key controlled variable of the APC solution and is hence significantly more stable than the PID control strategy. This is significant since it enabled the selection of a more optimum setpoint which in turn partly facilitated the increased production figures.

6 Conclusions

Literature has shown that flash drying operations can benefit from an advanced control strategy. This was proven in this project where a hybrid control strategy, which combines rule based and DMC control logic, was developed and installed at a PGM concentrate flash dryer. Since commissioning the APC, the flash dryer's average throughput has increased by more 6 % despite higher feed moistures. Furthermore, even though coal consumption has increased slightly, the operation efficiency has improved by almost 5 %. This was mainly made possible by improving the stability of the drying column outlet temperature by approximately 40 % which in turn enabled the selection of a more optimal setpoint. Recent data has shown APC utilization now exceeds 95 %. This is indicative of a successful controller installation with good site acceptance.

To ensure good results, a systematic approach was followed throughout the project. A proper understanding of the process, its behaviour and issues was developed from literature, interviews and data analysis. Limitations in the base layer control philosophy were indentified and addressed. For example, the need for a new concentrate feedrate controller was identified and fulfilled. Despite some improvements, it was evident that, due to process interactions and dead times, the existing base layer control strategy was inadequate to extract the full potential from the process. Consequently it was decided to develop and deploy an APC solution, like a MPC, that can address these limitations. Step tests helped to confirm the expected process dynamics that formed the basis of the DMC control matrix. Key controller objectives were formulated in collaboration with the site metallurgists to guide the tuning process and ongoing performance analysis. The project was well documented, and the operators on site were trained in order to ensure sustainable benefits. The entire project was executed with due cognisance of the relevant safety risks.

7 Recommendations

Several benefits have been derived by deploying the new APC strategy at the flash dryer. These benefits can potentially be increased even further by installing some additional instrumentation (7.1). Certain steps must also be taken to ensure that the achieved improvements are sustained (7.2). Finally it is recommended that the same strategy be deployed to similar processes to further leverage the knowledge gained during this project (7.3).

7.1 Additional instrumentation

It became apparent early in the project that fluidized combustion is a relatively complex process with various non-linearities and uncertainties. Furthermore, due to limited instrumentation, it is difficult to fully describe the entire system and therefore various assumptions have to be made and control regions has to be selected conservatively. Therefore, even though the new APC controller is already adding significant value, it is believed that even more value can be derived if the understanding of the process is further improved and if this understanding can be supported by additional instrumentation. Hence it is recommended that consideration is given to installing the following instrumentation:

1. Thermocouple at the top of the HGG Chimney. This instrument will assist in determining the extent to which the fluidizing air supply is balancing the ID fan's air requirement. As long as this temperature is close to ambient temperature, it indicates that air is sucked in through the chimney. If it

rises significantly above ambient temperature, it indicates that hot air is lost through the chimney.

2. Fluidizing air flow measurement. This instrument will assist with proper control of the fluidizing air supply to the HGG. It will remove the uncertainty surrounding actual damper position and it will enable a cascaded PID controller arrangement similar to that which was introduced on the concentrate feed.

3. Coal feed rate measurement. This instrument will assist with proper accounting of coal consumption. Direct measurement of the coal feedrate might not be possible due to space restrictions etc. However there are other options like load cells (on the supply side). Note that there are load cells on the coal feed bins, but since these are filled and discharging at the same time, it's not possible to infer the discharge rate since the feeding rate is also unknown. Depending on where and how the coal feed is measured, this instrument might also assist with proper control of the coal supply to the HGG. It can potentially remove the uncertainly surrounding the feeder speeds and to some extent the size distribution of the coal. It too might enable a cascaded PID controller arrangement. Replacement of the coal feeder gearboxes with fixed ratio alternatives should also be considered in order to reduce maintenance issues and the resulting process gain shifts.

4. Thermocouple at the bag house outlet. This instrument will assist with monitoring the temperature profile across the system. It can also indicate potential corrosion risk due to too low temperatures.

5. Air flow measurements at the FD inlet and/or outlet and the bag house exit can assist with detecting air leaks and with balancing the overall air flow through the system.

6. A CO Analyser at the output of the HGG can assist in determining the extent of the combustion and also with preventing CO build ups.

7.2 Controller monitoring and maintenance

Because a DMC controller is based on a plant model, and because plant models drift over time, it's important to develop an efficient controller monitoring strategy that will flag any need for retuning. This might require occasional step tests, but will ensure that the achieved benefits can be sustained. Note that the additional instrumentation listed above will improve the process understanding and should also be included in the process models.

Given the reduced drying column outlet temperature setpoint, it is recommended that a study be conducted in order accurately determine the dew point of the vapours passing through the bag house at normal operating conditions. The flash dryer outlet temperature setpoint/target band should then be adjusted accordingly in order to avoid an increase in equipment corrosion.

7.3 Additional deployments

This new controller strategy has already been implemented on all the other flash dryers at the same smelter. It is planned that the same be done for all the flash dryers at Anglo Platinum's other smelters and that other similar processes should also be considered for optimization by means of APC.

8 References

1. Abukhalifeh, H, Dhib, R and Fayed, M (2003) “Model Predictive Control of an infrared-dryer”, *International Conference on Physics and Control*, 1, 340-344.
2. Altafini, C and Furini, M (1997) “Robust Control of a Flash Dryer Plant”, *Proceedings of the 1997 IEEE International Conference on Control Applications,* 785-787.
3. Anglo Platinum (2009) “Environment – Energy and Water”, http://www.angloplatinum.com [2010, June 28].
4. Anglo Platinum (*sa*) “Our Business”, http://www.angloplatinum.com [2010, June 28].
5. Cawthorn, RG (1999) “The platinum and palladium resources of the Bushveld Complex” *South African Journal of Science*, 95, 481-489
6. Cristeaa, VM, Baldeaa, M and Agachia ŠP (2000) “Model Predictive Control of an industrial dryer” *Computer Aided Chemical Engineering*, 8, 217-276.
7. De Clerk, DJ, de Villiers, PGR and Humphries, G (2010) “A Sustainable Approach to Process Optimization through Integrated Advanced Control Software Standards”, paper presented at *IFAC MMM Conference*, Cape Town, South Africa.
8. De Villiers, B (2010) “Investors Presentation: Polokwane Metallurgical Complex”, http://www.angloamerican.com [2010, June 28].
9. Drytech (2001) “Flash Dryers”, http://www.drytecheng.com [2010, June 21].

10. Fox, HM (1957) "Fuel Air Ratio Regulation for Combustion Systems", *US Patent 2,812,637*, assigned to Phillips Petroleum Company
11. FBC (*sa*) "Fluidized Bed Combustion – for Clean Power Generation", http://www.fluidizedbedcombustion.com [2010, June 21].
12. Gore, WL (2010) "Platinum Flash Dryer Case Study", http://www.gore.com [2010, June 28].
13. Hurowitz, SE (1998) "Superfractionator Process Control", PhD Thesis, Texas Tech University, USA.
14. Johnson Matthey (*sa*) "Platinum production in South Africa", http://www.platinum.matthey.com [2010, June 21].
15. KBC (2010) "KBC Level 3 Training", Personal Communication, KBC Offices
16. Le Page, GP, Tadé, MO and Stone, RJ (1998) "Comparative evaluation of advanced process control techniques for alumina flash calciners" *Journal of Process Control*, 8, 287-298.
17. Le Roux, H and De Villiers, PGR (2010) "Flash Dryer History and Operation", Personal Communication, Corporate Office, Anglo American Platinum
18. Loesche (*sa*) "Hot Gas Generators", http://www.loesche.com [2010, June 21].
19. Lombard, N, de Villiers, PGR and Humphries, G (2010) "A lean Approach to managing integrated basic control software standards", paper presented at *IFAC MMM Conference*, Cape Town, South Africa.
20. Marlin, TE (1995) "Process Control", McGraw Hill.

21. Mintek (2007) "Assessment of Kwazulu-Natal province's coal mining and coal resources", Report, Trade & Investment, Kwazulu-Natal, South Africa.

22. Patumsawad, S (*sa*) "Coal Combustion Studies in a Fluidised Bed", Article, Dept Mech Eng, King Mongkut' s Institute of Technology, North Bangkok.

23. Perry, RH and Green, D (1984) *Perry's Chemical Engineers' Handbook 6th Edition*, McGraw-Hill.

24. Rademan, J (2008) "Lonmin improves smelter efficiency" *MMS Mag*, July, 24 – 25.

25. Seborg, DE, Edgar, TF and Mellichamp, DA (1989) "Process Dynamics and Control", John Wiley & Sons, USA.

26. Sinnott, RK (1998) *Coulson & Richardson's Chemical Engineering Volume 6*, Butterworth Heinemann.

27. Stratos Tavoulareas, E (1991) "Fluidized-Bed Combustion Technology" *Annual Review of Energy and the Environment*, 16, 25-57.

28. Stephanopoulos, G (1984) *Chemical Process Control. An Introduction to Theory and Practice*, Prentice Hall.

29. Tay, M (2007) "Case Study: Reduce Energy Costs, Increase Production with Model Predictive Control" *Ethanol Producer Magazine*, September.

30. Van Rooyen, Q (2010) "Flash Dryer APC HAZOP Meeting Minutes", Personal Communication, Waterval Smelter, Anglo American Platinum.

31. Zactruba, J (2009) "Differences of a Circulating Fluidized Bed Boiler and a Pulverised Coal Boiler", http://www.brighthub.com [2010, June 21].

Appendix A: Monthly Time Series Data and Summarized Tables

This Appendix contains charts of all the data, split into monthly chunks, that was collected throughout the project. Each month's data is also summarized into a table that list the key performance measures and APC vs PID comparisons.

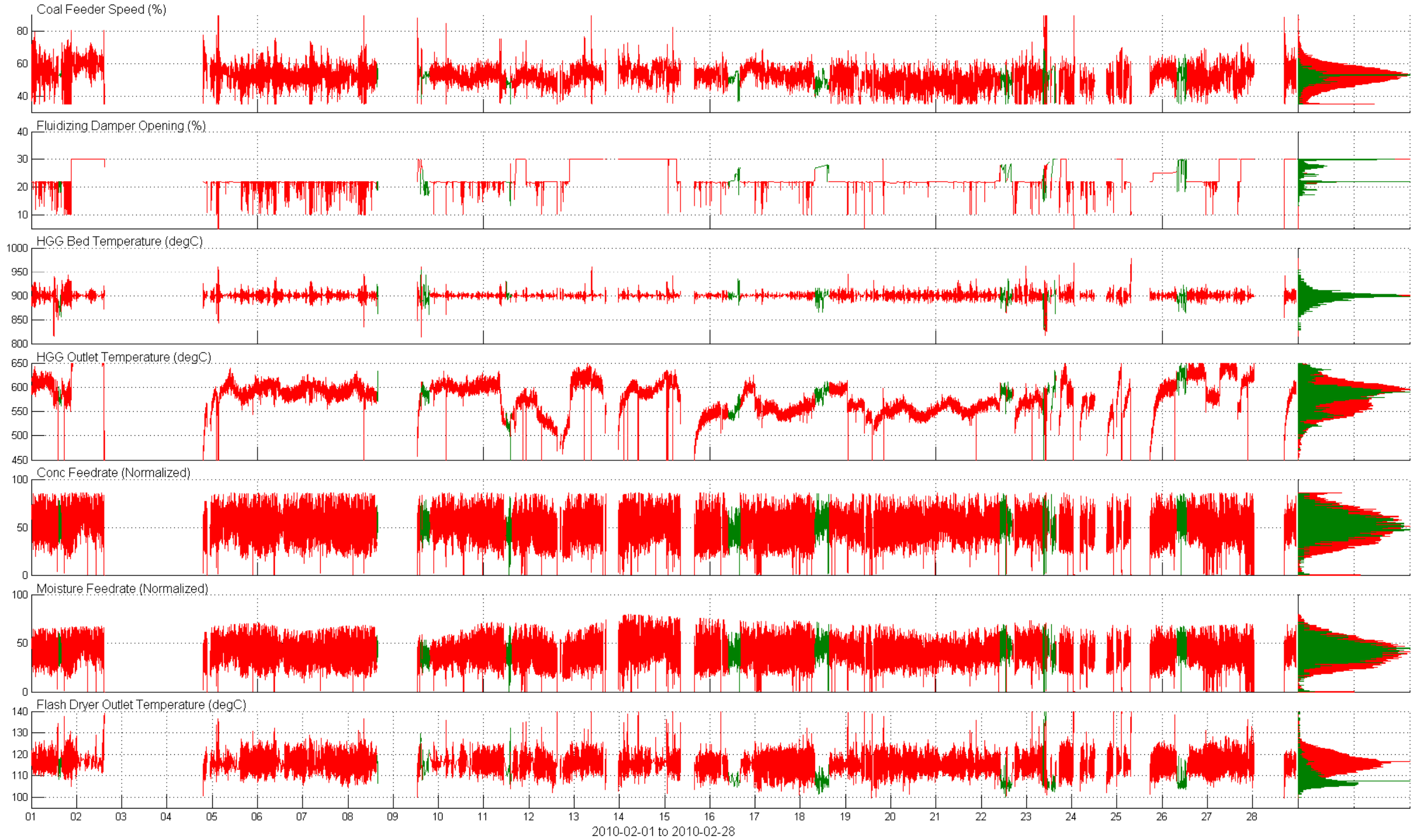

Figure 22: Time Series Data Comparing APC ON vs OFF for February 2010

Table 7: Summarized results for February 2010

StartDate	2010/02/01 00:00						
EndDate	2010/02/28 23:59						
Duration	28 days						
	Total		**PID**		**APC**		**Change**
PlantRunTime [days] With Utilization [%]	22	79.4	21	93.3	1	6.7	
Wet Concentrate Dried [Normalized] With hourly average [Normalized]	27 171	50.9	25 281	50.8	1 889	52.5	3.4 %
Moisture Removed [Normalized] With hourly average [Normalized]	21 799	40.8	20 276	40.7	1 524	42.3	4.0 %
Coal Used [Normalized] With hourly average [Normalized]	27 800	52.1	25 963	52.2	1 837	51.1	-2.1 %
Specific Coal [Coal/Moisture]	1.28		1.28		1.21		-5.9 %
Accumulated HGG Temp Err [°C] With daily average [°C per day]	2 244	100.9	1 918	92.5	326	217.3	135.0 %
Accumulated FD Temp Err [°C] With daily average [°C per day]	1 231	55.4	1 126	54.3	105	70.0	28.9 %

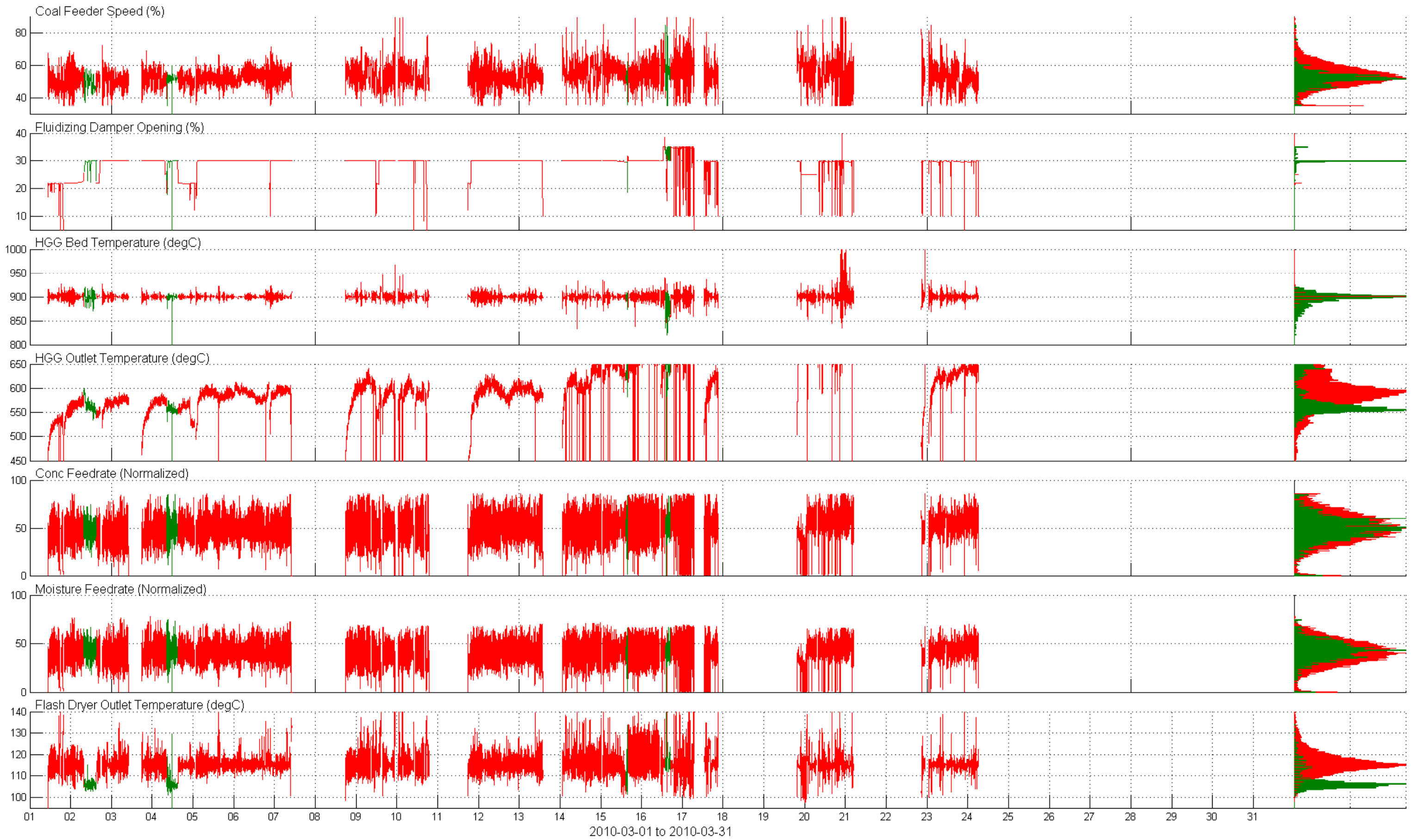

Figure 23: Time Series Data Comparing APC ON vs OFF for March 2010

Table 8: Summarized results for March 2010

StartDate	2010/03/01 00:00						
EndDate	2010/03/31 23:59						
Duration	31 days						
	Total		**PID**		**APC**		**Change**
PlantRunTime [days] With Utilization [%]	15	49.2	15	95.4	1	4.6	
Wet Concentrate Dried [Normalized] With hourly average [Normalized]	18 188	49.7	17 355	49.7	833	49.1	-1.2 %
Moisture Removed [Normalized] With hourly average [Normalized]	14 933	40.8	14 214	40.7	719	42.5	4.2 %
Coal Used [Normalized] With hourly average [Normalized]	19 563	53.5	18 690	53.6	873	51.5	-3.8 %
Specific Coal [Coal/Moisture]	1.31		1.31		1.21		-7.7 %
Accumulated HGG Temp Err [°C] With daily average [°C per day]	1 756	115.2	1 574	108.2	182	257.7	138.1 %
Accumulated FD Temp Err [°C] With daily average [°C per day]	919	60.3	877	60.4	41	58.7	-2.7 %

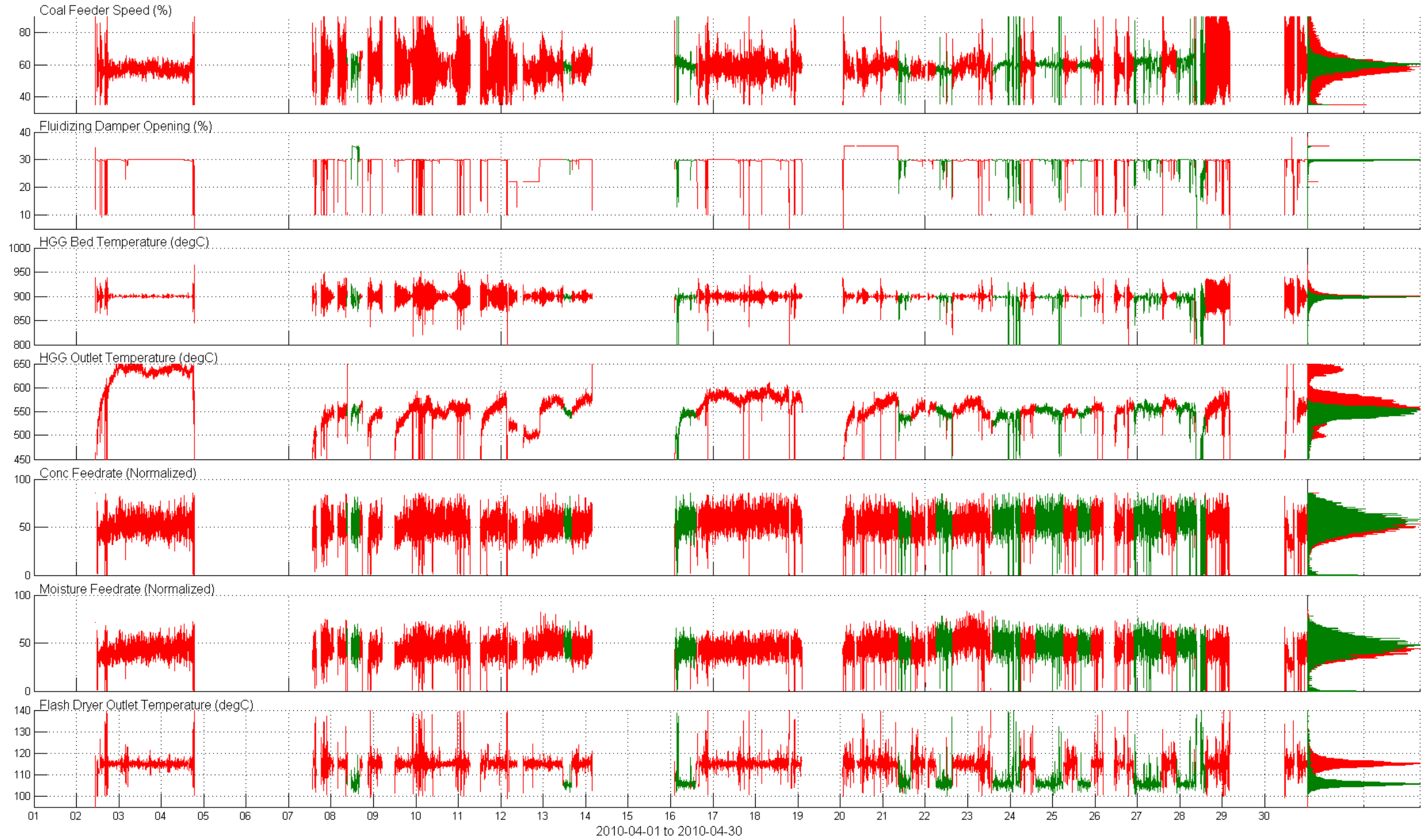

Figure 24: Time Series Data Comparing APC ON vs OFF for April 2010

Table 9: Summarized results for April 2010

StartDate	2010/04/01 00:00						
EndDate	2010/04/30 23:59						
Duration	30 days						
	Total		**PID**		**APC**		**Change**
PlantRunTime [days] With Utilization [%]	20	65.4	15	78.1	4	21.9	
Wet Concentrate Dried [Normalized] With hourly average [Normalized]	25 085	53.3	19 398	52.7	5 687	55.1	4.5 %
Moisture Removed [Normalized] With hourly average [Normalized]	21 160	44.9	16 320	44.4	4 840	46.9	5.8 %
Coal Used [Normalized] With hourly average [Normalized]	27 788	59.0	21 670	58.9	6 119	59.3	0.7 %
Specific Coal [Coal/Moisture]	1.31		1.33		1.26		-4.8 %
Accumulated HGG Temp Err [°C] With daily average [°C per day]	3 279	167.1	2 436	158.9	844	196.3	23.5 %
Accumulated FD Temp Err [°C] With daily average [°C per day]	609	31.0	469	30.6	139	32.4	6.0 %

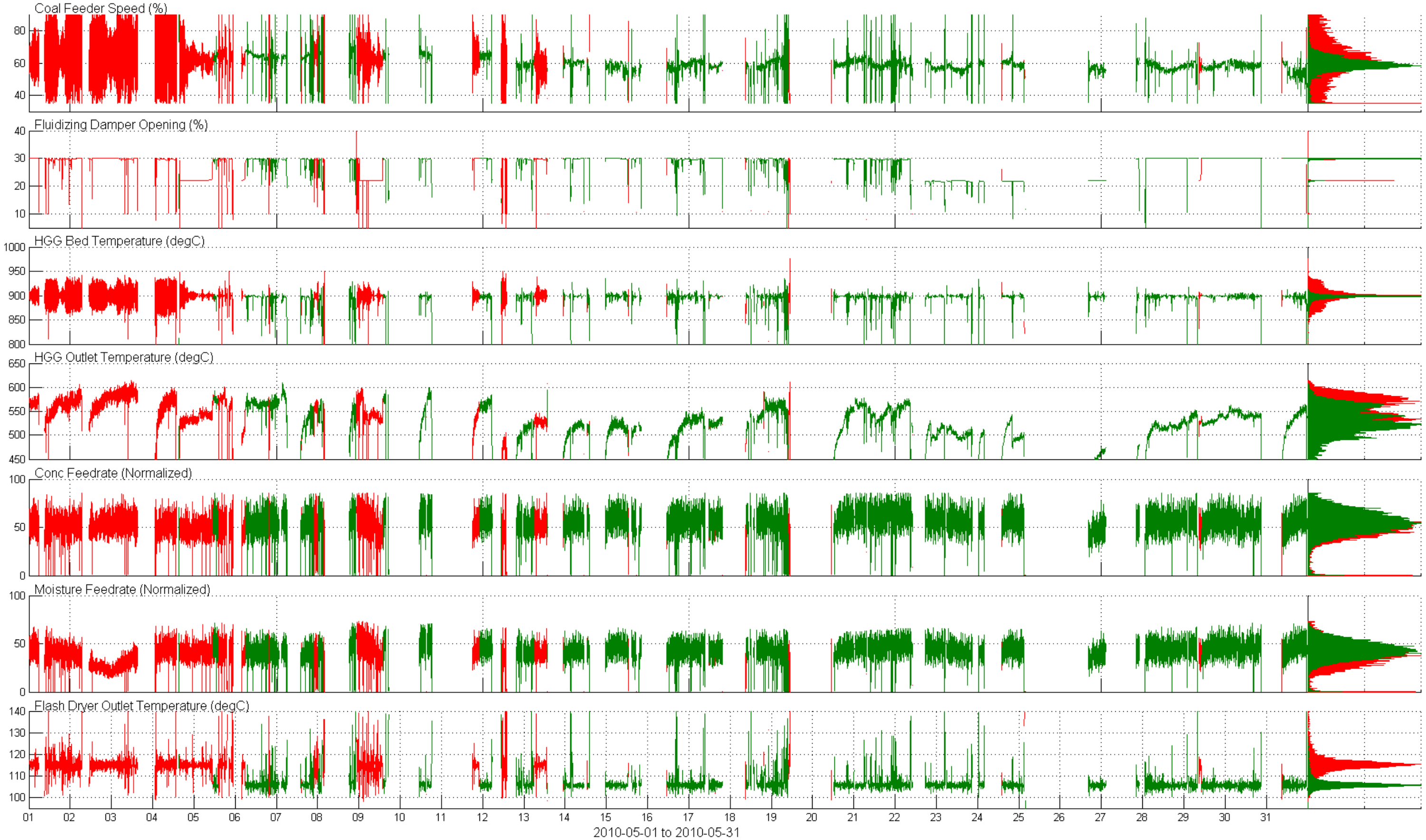

Figure 25: Time Series Data Comparing APC ON vs OFF for May 2010

Table 10: Summarized results for May 2010

StartDate	2010/05/01 00:00						
EndDate	2010/05/31 23:59						
Duration	31 days						
	Total		**PID**		**APC**		**Change**
PlantRunTime [days] With Utilization [%]	19	61.5	6	30.2	13	69.8	
Wet Concentrate Dried [Normalized] With hourly average [Normalized]	24 732	54.1	7 015	50.8	17 717	55.5	9.3 %
Moisture Removed [Normalized] With hourly average [Normalized]	19 016	41.6	5 239	37.9	13 777	43.2	13.8 %
Coal Used [Normalized] With hourly average [Normalized]	27 632	60.4	8 725	63.1	18 906	59.2	-6.2 %
Specific Coal [Coal/Moisture]	1.45		1.67		1.37		-17.6 %
Accumulated HGG Temp Err [°C] With daily average [°C per day]	3 977	208.7	1 784	309.7	2 194	164.9	-46.8 %
Accumulated FD Temp Err [°C] With daily average [°C per day]	561	29.5	248	43.1	313	23.5	-45.4 %

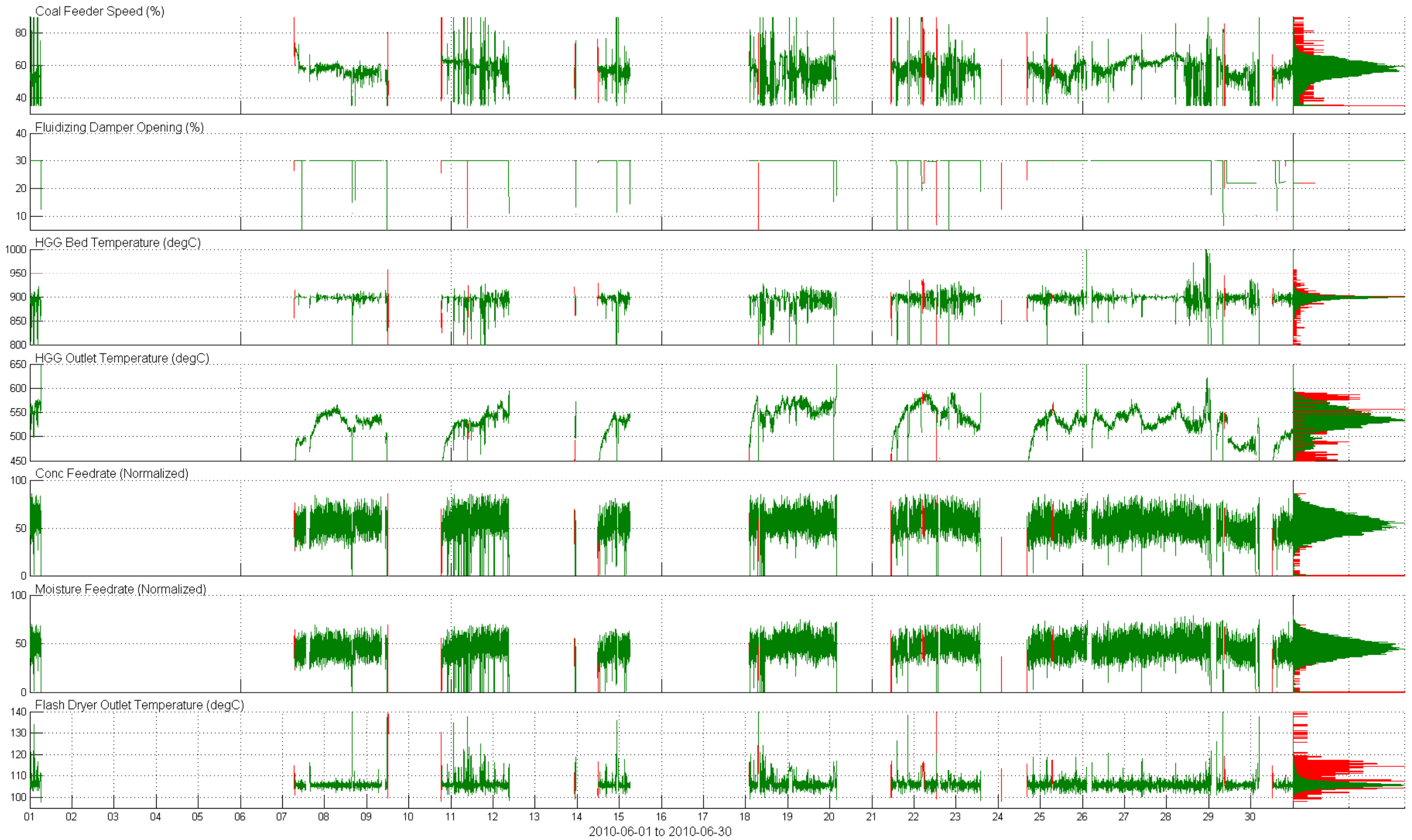

Figure 26: Time Series Data Comparing APC ON vs OFF for June 2010

Table 11: Summarized results for June 2010

StartDate	2010/06/01 00:00						
EndDate	2010/06/30 23:59						
Duration	30 days						
	Total		**PID**		**APC**		**Change**
PlantRunTime [days] With Utilization [%]	14	47.9	0	2.3	14	97.7	
Wet Concentrate Dried [Normalized] With hourly average [Normalized]	18 568	53.9	373	46.3	18 194	54.0	16.8 %
Moisture Removed [Normalized] With hourly average [Normalized]	15 736	45.6	312	38.6	15 424	45.8	18.5 %
Coal Used [Normalized] With hourly average [Normalized]	19 732	57.2	472	58.6	19 259	57.2	-2.3 %
Specific Coal [Coal/Moisture]	1.25		1.52		1.25		-17.6 %
Accumulated HGG Temp Err [°C] With daily average [°C per day]	2 938	204.5	176	522.4	2 762	196.9	-62.3 %
Accumulated FD Temp Err [°C] With daily average [°C per day]	263	18.3	38	113.2	225	16.0	-85.8 %

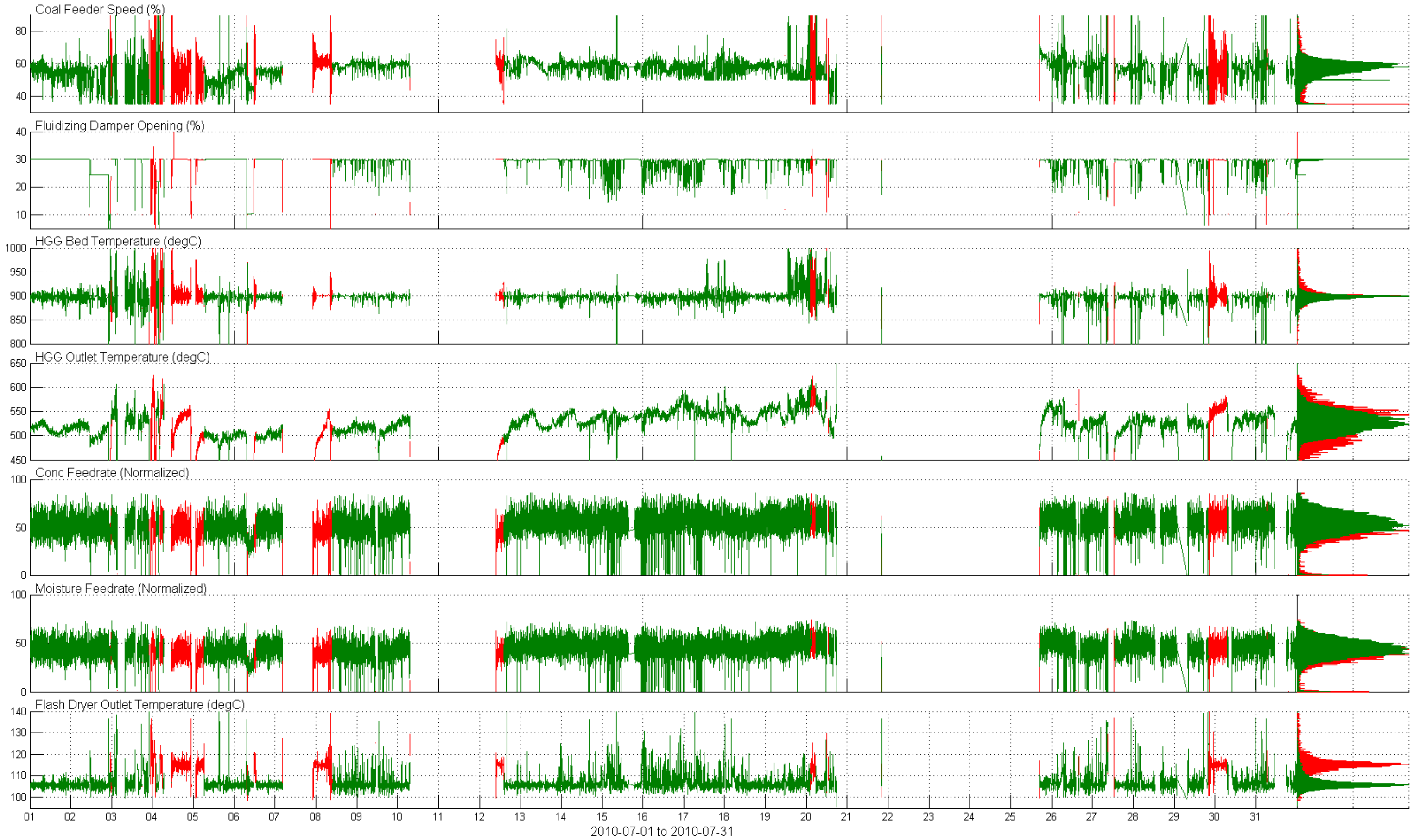

Figure 27: Time Series Data Comparing APC ON vs OFF for July 2010

Table 12: Summarized results for July 2010

StartDate	2010/07/01 00:00						
EndDate	2010/07/31 23:59						
Duration	31 days						
	Total		**PID**		**APC**		**Change**
PlantRunTime [days] With Utilization [%]	22	69.5	2	10.9	19	89.1	
Wet Concentrate Dried [Normalized] With hourly average [Normalized]	27 465	53.1	2 766	49.2	24 700	53.6	8.9 %
Moisture Removed [Normalized] With hourly average [Normalized]	22 625	43.8	2 273	40.5	20 352	44.2	9.2 %
Coal Used [Normalized] With hourly average [Normalized]	29 220	56.5	3 114	55.4	26 106	56.7	2.2 %
Specific Coal [Coal/Moisture]	1.29		1.37		1.28		-6.4 %
Accumulated HGG Temp Err [°C] With daily average [°C per day]	5 003	232.3	842	359.6	4 161	216.7	-39.7 %
Accumulated FD Temp Err [°C] With daily average [°C per day]	729	33.9	100	42.8	629	32.8	-23.4 %

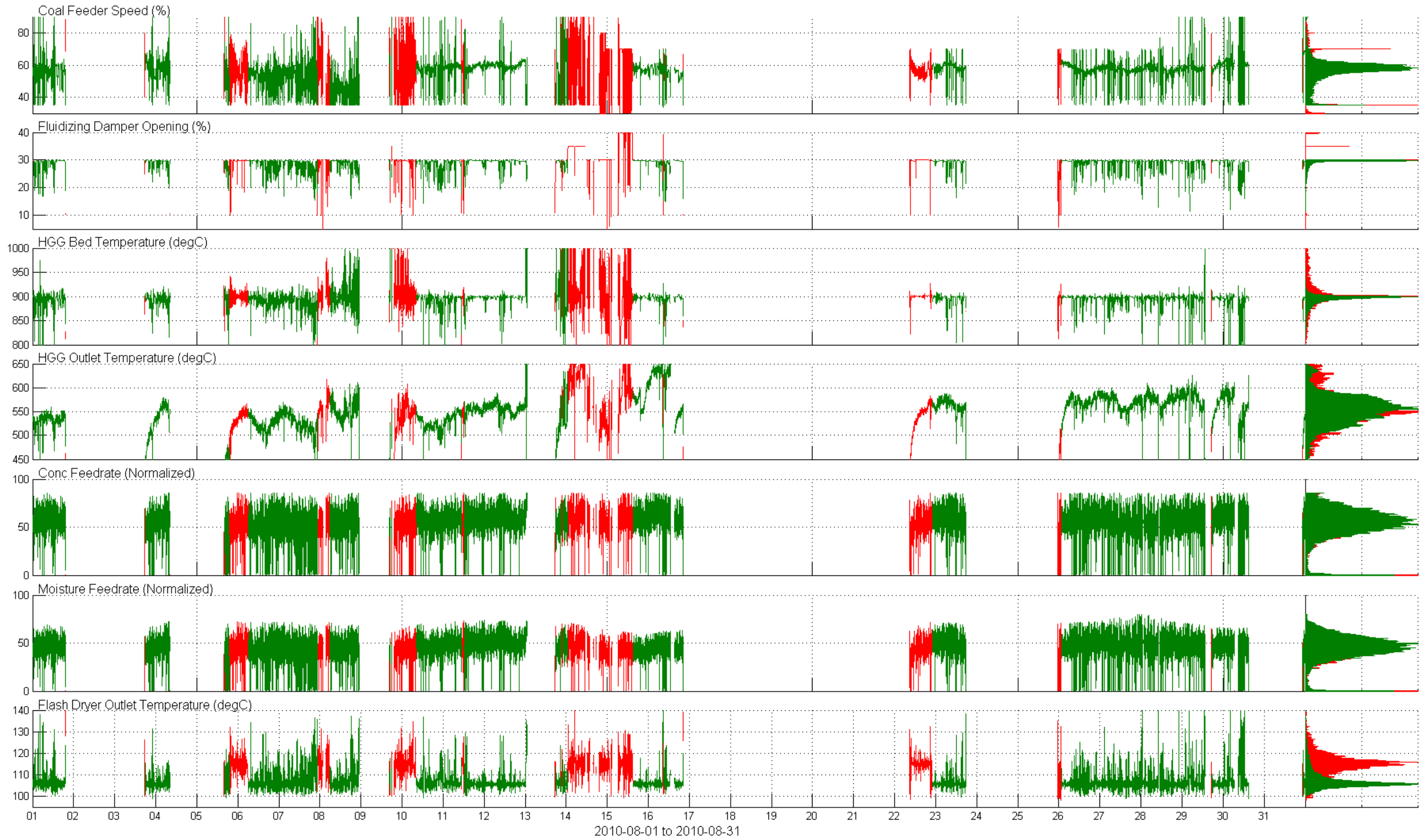

Figure 28: Time Series Data Comparing APC ON vs OFF for August 2010

Table 13: Summarized results for August 2010

StartDate	2010/08/01 00:00						
EndDate	2010/08/31 23:59						
Duration	31 days						
	Total		**PID**		**APC**		**Change**
PlantRunTime [days] With Utilization [%]	16	52.3	3	19.9	13	80.1	
Wet Concentrate Dried [Normalized] With hourly average [Normalized]	21 410	55.0	4 180	53.9	17 230	55.3	2.7 %
Moisture Removed [Normalized] With hourly average [Normalized]	17 612	45.3	3 347	43.1	14 265	45.8	6.2 %
Coal Used [Normalized] With hourly average [Normalized]	21 851	56.1	4 274	55.1	17 577	56.4	2.4 %
Specific Coal [Coal/Moisture]	1.24		1.28		1.23		-3.5 %
Accumulated HGG Temp Err [°C] With daily average [°C per day]	5 044	311.1	1 827	565.0	3 217	247.8	-56.1 %
Accumulated FD Temp Err [°C] With daily average [°C per day]	766	47.2	192	59.4	574	44.2	-25.6 %

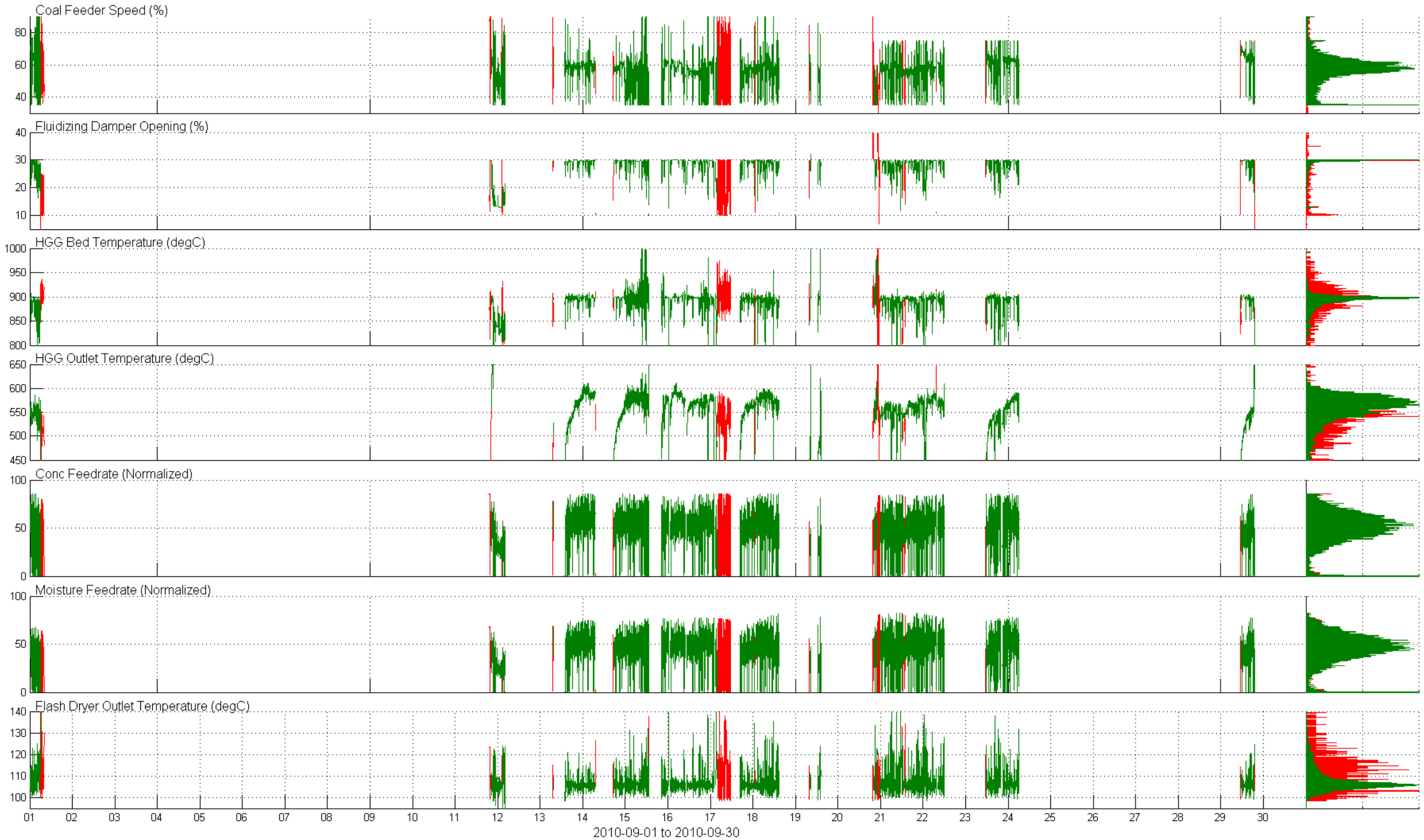

Figure 29: Time Series Data Comparing APC ON vs OFF for September 2010

Table 14: Summarized results for September 2010

StartDate	2010/09/01 00:00						
EndDate	2010/09/30 23:59						
Duration	30 days						
	Total		**PID**		**APC**		**Change**
PlantRunTime [days] With Utilization [%]	7	25.0	1	8.9	7	91.2	
Wet Concentrate Dried [Normalized] With hourly average [Normalized]	9 025	50.2	682	42.9	8 343	50.9	18.7 %
Moisture Removed [Normalized] With hourly average [Normalized]	8 145	45.3	604	37.9	7 542	46.0	21.3 %
Coal Used [Normalized] With hourly average [Normalized]	10 259	57.0	884	55.5	9 375	57.2	2.9 %
Specific Coal [Coal/Moisture]	1.26		1.47		1.24		-15.2 %
Accumulated HGG Temp Err [°C] With daily average [°C per day]	2 800	373.6	404	609.5	2 396	350.6	-42.5 %
Accumulated FD Temp Err [°C] With daily average [°C per day]	501	66.8	93	140.4	408	59.7	-57.5 %

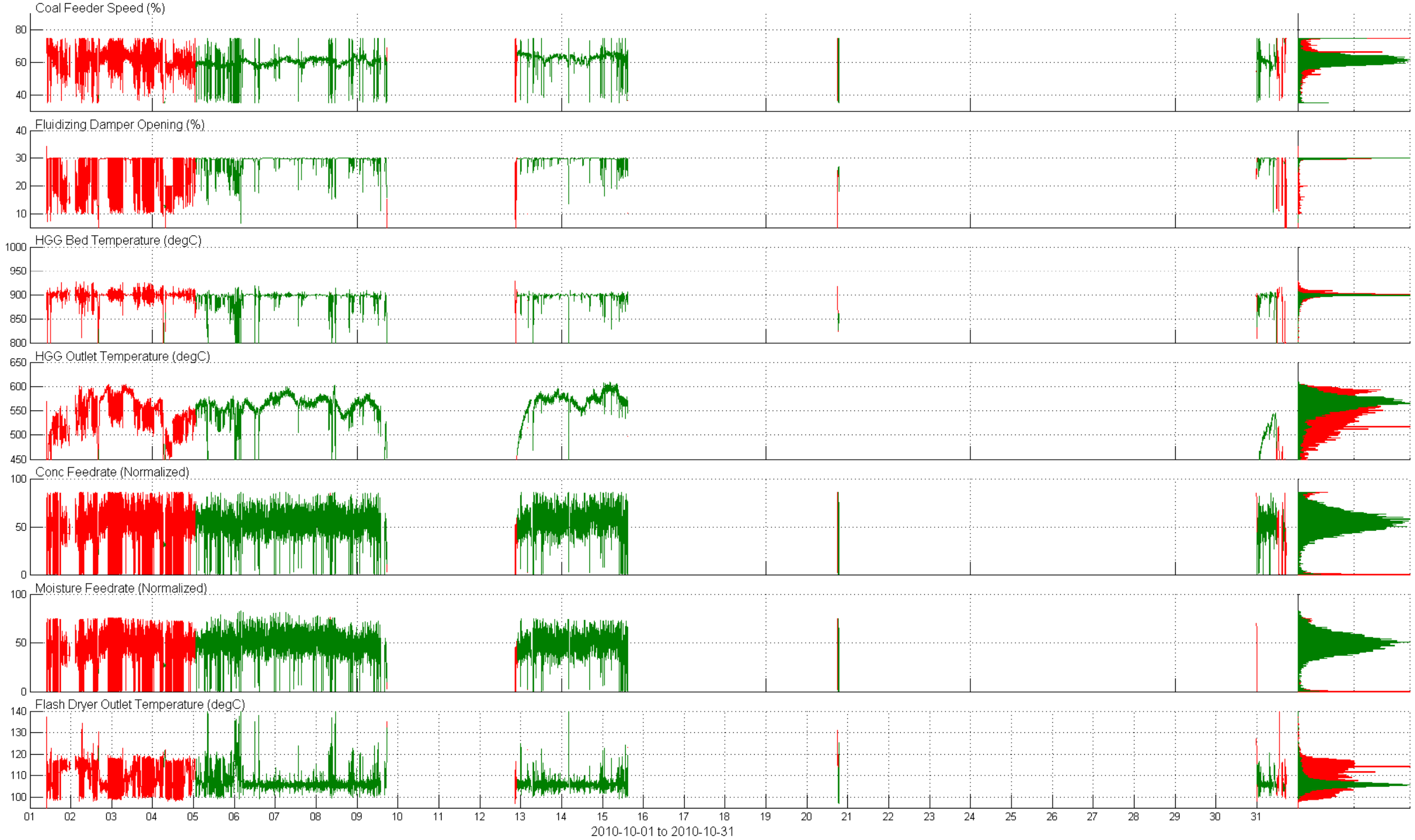

Figure 30: Time Series Data Comparing APC ON vs OFF for October 2010

Table 15: Summarized results for October 2010

StartDate	2010/10/01 00:00						
EndDate	2010/10/31 23:59						
Duration	31 days						
	Total		**PID**		**APC**		**Change**
PlantRunTime [days] With Utilization [%]	11	36.6	4	32.2	8	67.8	
Wet Concentrate Dried [Normalized] With hourly average [Normalized]	14 609	53.7	4 597	52.5	10 011	54.3	3.4 %
Moisture Removed [Normalized] With hourly average [Normalized]	12 272	45.1	3 860	44.1	8 412	45.6	3.5 %
Coal Used [Normalized] With hourly average [Normalized]	16 639	61.2	5 389	61.5	11 250	61.0	-0.9 %
Specific Coal [Coal/Moisture]	1.36		1.40		1.34		-4.2 %
Accumulated HGG Temp Err [°C] With daily average [°C per day]	2 104	185.6	699	191.5	1 405	182.8	-4.6 %
Accumulated FD Temp Err [°C] With daily average [°C per day]	695	61.4	448	122.8	247	32.2	-73.8 %

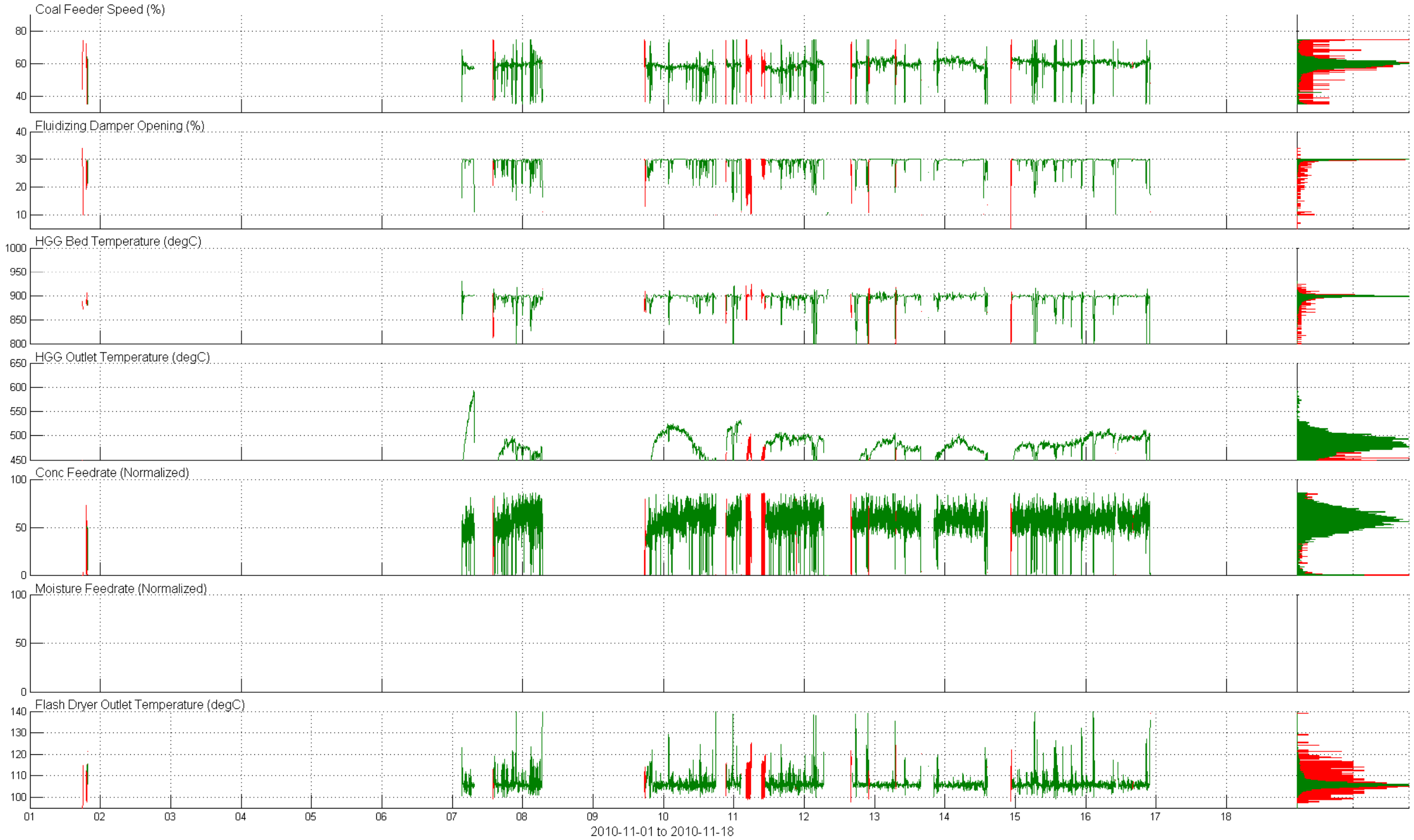

Figure 31: Time Series Data Comparing APC ON vs OFF for November 2010

Table 16: Summarized results for November 2010

StartDate	2010/11/01 00:00						
EndDate	2010/11/18 23:59						
Duration	18 days						
	Total		**PID**		**APC**		**Change**
PlantRunTime [days] With Utilization [%]	7	37.9	0	4.2	7	95.8	
Wet Concentrate Dried [Normalized] With hourly average [Normalized]	9 157	55.9	315	46.2	8 842	56.3	21.9 %
Moisture Removed [Normalized] With hourly average [Normalized]	7 693	47.0	263	38.6	7 429	47.3	22.5 %
Coal Used [Normalized] With hourly average [Normalized]	9 716	59.3	402	58.9	9 315	59.3	0.7 %
Specific Coal [Coal/Moisture]	1.26		1.53		1.25		-17.8 %
Accumulated HGG Temp Err [°C] With daily average [°C per day]	1 160	170.0	120	423.6	1 040	159.0	-62.5 %
Accumulated FD Temp Err [°C] With daily average [°C per day]	230	33.8	42	148.7	188	28.8	-80.6 %