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Dynamic Limit Based Model Predictive Control of a Flash Drying Unit

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Abstract: A model predictive controller was implemented at the flash drying unit of the Anglo American Platinum Polokwane smelter. The controller uses a mix of standard model predictive control technology and dynamic control limits (based on the rate of change of the hot gas generator average bed temperature) to improve temperature stability across the unit. The improved temperature control resulted in the ability to process 1.80% more concentrate through the unit without the need to increase coal feed, as well as reducing the number of coal feeder trips that result from high temperature events.

Keywords: Flash dryer, model predictive control, PGM processing, smelting.

1. INTRODUCTION

Smelting is one of the main steps in Platinum Group Metal (PGM) processing (Ndlovu, 2014). Once ore containing PGMs have been mined and concentrated (through crushing, milling, and flotation) the concentrate is enriched through smelting, typically in an electric furnace (Sinisalo and Lundström, 2018). The concentrate moisture content is too high for the furnace which can cause equipment damage. Therefore, an important processing step is to dry the concentrate, in this case using a flash drying unit.

The control of a flash drying unit is not straightforward, owing to the comparatively slow dynamics of the hot gas generator (HGG) and the fast dynamics of the flash dryer (FD), coupled with significant interactions and nonlinearities in the process. Effective control of the flash drying unit is however imperative as undue maintenance stoppages may become a risk to overall smelting production plans.

Model predictive control (MPC) is the most successful form of advanced process control (APC) used in the process industries (Samad, 2017; Bauer and Craig, 2008). MPC is regarded as a mature technology and the one most often used for improving the economic performance of a processing plant (Craig et al., 2011). MPC also handles interactions directly through the structure of the model matrix. All these factors make MPC suited to this application.

To handle the nonlinearities present (primarily) in the response from the coal feeder to the HGG average bed temperature, the rate of change (ROC) of the HGG average bed temperature is controlled rather than controlling the temperature itself.

There are works that focus on the MPC of pneumatic dryers (see e.g. Satpati et al. (2017)), but mostly without regard for the supply of the drying gas. There are also works that focus on the MPC of fluidized bed combustors (e.g. Zlatkovikj et al. (2022)), but most focus on fluidized bed boilers (not HGGs), and mostly without regard for the end-use of the hot gas or steam. The integrated control and optimization of the flash drying unit, comprising the HGG and flash dryer, are paramount to overall unit optimality.

Olivier and French (2019) presented the control of a flash drying unit, consisting of decoupled proportional, integral, and derivative (PID) controllers. De Clerk and De Vaal (2012) presented an implementation of MPC of a flash drying unit at the Anglo American Platinum Waterval smelter, without dynamic limit changes. Apart from these, the authors are unaware of publications focused on the control of a flash drying unit, and none focused on MPC of such a unit apart from De Clerk and De Vaal (2012).

2. PROCESS DESCRIPTION

A process flow diagram of the flash drying unit is shown in Fig. 1. A coal-fed hot gas generator, making use of fluidized bed combustion technology, generates hot air at a temperature of around 700 °C (Van Manen, 2006), which is fed into the flash dryer to dry out the wet concentrate feed.

Coal is fed into the HGG via a screw feeder onto the fluidized beds. The beds also contain sand (being inert) that is suspended into a fluid state by the flow of air into the HGG. A fixed-speed fan blows the air into the HGG, the flow-rate of which is regulated by means of a damper.

When coal feed is increased, the coal needs to heat up and combust, which then increases the HGG bed temperature reading. When the coal feed is decreased, less combustion occurs and the HGG bed temperature reading decreases. The combustion kinetics are different for increases versus decreases in coal feed, which is one of the main sources of nonlinearity in the process.

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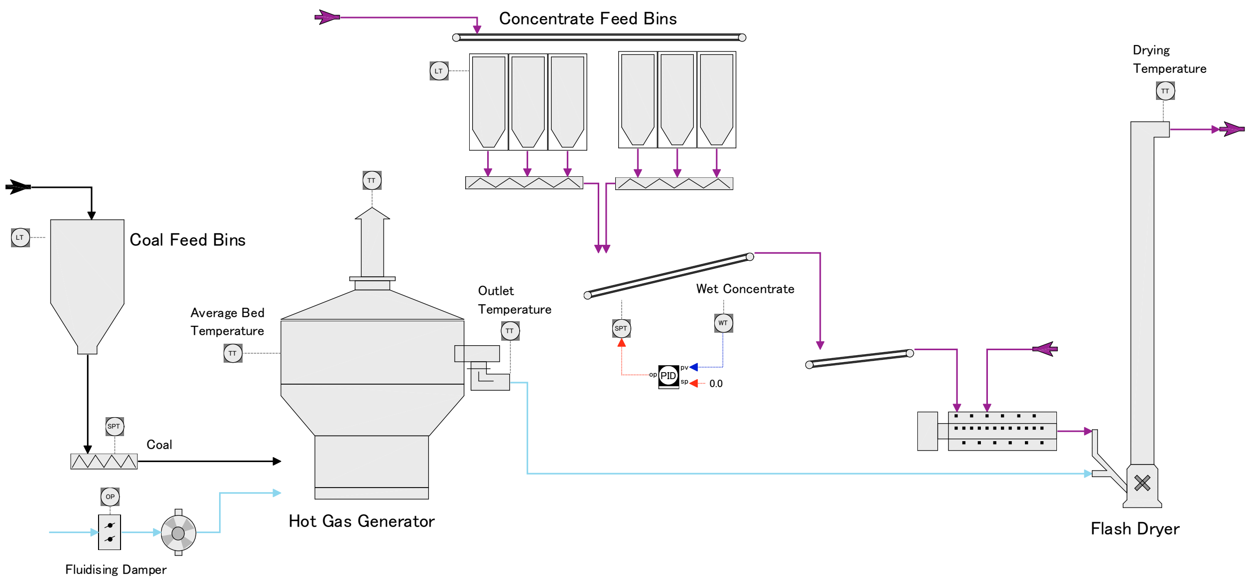

Fig. 1. Flash drying unit process flow diagram.

In the flash dryer (also called a pneumatic dryer) there is simultaneous pneumatic conveying as well as heat and mass transfer along the dryer tube (El Hallaoui et al., 2019). The FD has a relatively short residence time, which makes it suitable for processing even temperature sensitive materials.

Wet concentrate is fed into the flash dryer via a conveyor feeding into a mixer. The FD reduces the concentrate moisture content from roughly 12 – 18 % (Van Manen, 2006) to less than 0.5 % (De Clerk and De Vaal, 2012). The flash dryer product is separated through a collection of cyclones into the concentrate product (which goes into the product bin) and an air stream. The air stream is first passed through a bag-house, to remove any dust, and then ejected through a stack. The dried concentrate is then ready to be sent to the smelting furnace.

The key process variables to maintain are:

- **The FD outlet temperature**, interchangeably called the drying temperature, must be strictly maintained above a low limit as it is an inferential for the amount of concentrate drying taking place. Maintaining this temperature above its low limit indicates that the concentrate product moisture content is low enough.
- **The HGG outlet temperature.** There is no setpoint for this temperature, but it should be maximized so as to ensure that maximal energy is transferred to the flash dryer. The high limit of this variable is an equipment safety consideration soit is important to maximize without violating the high limit.
- **The HGG average bed temperature.** The fluidized bed temperatures are measured individually but there is a high correlation between their values. As such, the average bed temperature is controlled. This temperature should also be high enough to ensure adequate energy available to be transferred to the flash dryer.

  If the fluidized bed temperature goes too high, clinker formation starts inside the beds (Basu and Sarka, 1983), requiring unit stoppages for maintenance interventions. FD unit stoppages in turn necessitate periods of reduced smelter throughput.

The main handles available to regulate and optimize the process are:

- **The mass of wet concentrate** fed into the flash dryer. The main aim of the unit is to process wet concentrate, this is the main variable to be maximized, and it has the fastest response to the drying temperature. It is important to manipulate the mass of wet concentrate to compensate for fast disturbances, e.g. changes in the feed moisture, and slow disturbances, e.g. fluctuations in HGG outlet temperature caused by changing coal quality. The wet concentrate feed can be maximized to the point where the drying temperature reaches its low limit.
- **The amount of coal fed into the HGG.** Enough coal needs to be fed into the HGG to generate enough heat to be transferred to the flash dryer, effecting the concentrate drying. Energy generation and transfer however needs to be optimized such that maximal concentrate processing can occur at minimal coal feed.
- **HGG fluidizing damper.** When increasing the damper opening the air flow and therefore the amount of energy transferred to the flash dryer will increase. The response is then an increase in the HGG outlet temperature and a reduction in the HGG average bed temperature.

The optimization objectives may be stated as (in decreasing order of importance):

(1) Maximize the wet concentrate feed,
(2) Maximize the energy transferred to the FD (by maximizing the HGG outlet temperature) (Van Manen, 2006), and
(3) Minimize the coal feed.

3. CONTROLLER DESIGN

The main justification for using MPC over e.g. decoupled PIDs is the interactions in the process. The typical variable pairing in this process (as is also used by Olivier and French (2019)) is to let the HGG coal feed control the HGG average bed temperature, let the damper control the HGG outlet temperature, and let the wet concentrate feed control the drying temperature. At a specific operating point however, it may be possible to open the damper further while feeding more coal into the HGG. This has a negligible effect on the HGG average bed temperature while transferring more energy to the flash dryer, resulting in the ability to process more wet concentrate feed.

An additional controller design consideration is that when the HGG recovers from a coal feeder trip the average bed temperature is far below the low limit. However, the average bed temperature rate of range will be abnormally high. During this ramp up phase the MPC is in an inherently nonlinear abnormal condition. Given that a linear industrial MPC package is used, during this state the standard control action is to increase the coal owing to the violation of the average bed temperature low limit.

Eker and Nikolaou (2002) notes that the effective design of a linear feedback controller can be adequate for the control of a nonlinear process. As the temperature ROC is a leading indicator of the movement of the present value, the ROC model has a faster settling time than the model to the actual temperature. The time to steady-state is therefore encountered sooner implying that the steady-state prediction error can be fed into the feedback loop sooner and thereby handling nonlinearities better.

To take advantage of this approach, the ROC can be controlled directly in the MPC. Therefore, during the stable operating regime the average bed temperature ROC should be controlled on average around zero, and during the abnormal event scenario, the temperature ROC should never go into a “runaway” region. The only intermediate calculation required is to dynamically update the HGG average bed temperature ROC limits based on the temperature present value relative to its low and high limits.

3.1 Rate of change limit calculation

The ROC controlled variable (CV) low and high limits are based on the value of the HGG average bed temperature, as shown in Fig. 2.

When the HGG average bed temperature is below its low limit the ROC must be positive such that the temperature will increase and be restored to within limits. The low limit of the ROC in this scenario is a tuning parameter termed the “Driving factor” and determines how aggressively the controller will drive the temperature back above its low limit. When the HGG average bed temperature is above its high limit the ROC must be negative such that the temperature can be decreased back within limits. The ROC high limit in this scenario is set to the negative of the “Driving factor”.

There is a constant offset between the ROC low and high limits, called the “Limit gap”. This is another tuning parameter and specifies what constitutes normal movement for the ROC. Increasing this value will increase the amount of movement allowed in the ROC, i.e. increasing the amount of buffering, but if this value is too large one may find that the temperature could cycle between its limits as the controller becomes more sluggish.

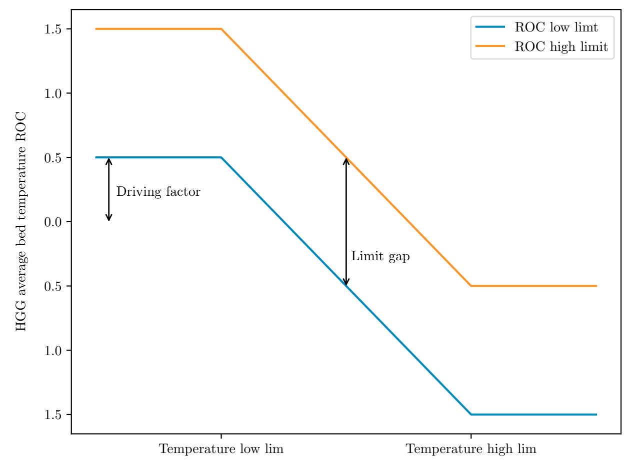

Fig. 2. Rate of change low and high limits as functions of the HGG average bed temperature.

When the HGG average bed temperature is within its limits, the ROC low and high limits change in a linear fashion between the minimum and maximum values.

The low and high ROC limits are dynamically calculated using the HGG average bed temperature low and high limits along with the present value, as well as the limit gap and driving factor using a simple lookup table transform.

3.2 Controller structure

The MPC controlled variables are the drying temperature (T_d), the HGG outlet temperature (T_o), and the HGG average bed temperature ROC (T_{ROC}). The manipulated variables are the feed of wet concentrate into the FD (f_c), the damper opening, which regulates the airflow into the HGG (f_a), and the feed of coal (fuel) to the HGG (f_f).

The output and input vectors are respectively given by

$$y = \begin{bmatrix} T_d \\ T_o \\ T_{ROC} \end{bmatrix}, \; u = \begin{bmatrix} f_c \\ f_a \\ f_f \end{bmatrix}. \tag{1}$$

The only optimization objectives are on the manipulated variables, for which the linear programming optimization weights are given by:

$$LP = \begin{bmatrix} -5 \\ 0 \\ 1 \end{bmatrix}, \tag{2}$$

indicating that the wet concentrate feed should be maximized (with the largest weight) and the coal feed should be minimized. The weights need to ensure that marginal improvements in concentrate feed need to outweigh marginal improvements in coal feed (both reported in their unscaled engineering unit values).

The conceptual step response models used in the controller design are shown in Fig. 3. The original models to the HGG average bed temperature are included for illustrative

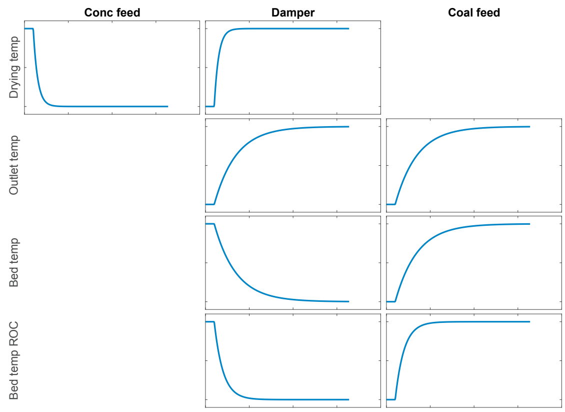

Fig. 3. Conceptual controller model matrix.

purposes even though the HGG average bed temperature is not controlled directly. The HGG average bed temperature is only controlled indirectly via its ROC.

An industrial MPC package is used to implement the controller. Qin and Badgwell (2003) provides further details about common industrial MPC packages, like objective functions used, how constraint violations are penalized, and model representations (among others).

4. PERFORMANCE ANALYSIS

The primary goal of using the ROC-based MPC is to improve the bed temperature stability and in turn improve the overall unit operation (viz. maximizing throughput while minimizing energy usage). The quantified improvements of continuous operation achieved at Polokwane Smelter are presented in Section 4.1.

Improving the bed temperature stability has the added benefit of reducing the number of coal feeder trips that originate from high bed temperature excursions during unit start-ups and large disturbances. This same MPC philosophy was implemented across all Anglo American Platinum's flash drying units, and the improvement in start-up performance achieved at Mortimer Smelter is shown in Section 4.2.

4.1 Continuous operation results

The changes to the MPC at Polokwane Smelter were commissioned at the end of November 2021. The data presented in this section are 5-minute sampled data from 08:00 on 29 November 2021 until 08:00 on 28 February 2022. This comprises roughly 3 months' worth of data, which represents a large enough sample for a sensible comparison, without being so large that equipment degradation becomes a significant consideration.

The data only include instances where the process is deemed to be running, which is defined as periods where the flash dryer feeding fan status is on, the concentrate feed and HGG outlet temperature are above minimum thresholds, the concentrate feed PID controller is tracking its setpoint, and the damper is not shut. Considering these conditions provides 6,167 data points where the APC was off (514 hours) and 9,192 data points where the APC was on (766 hours).

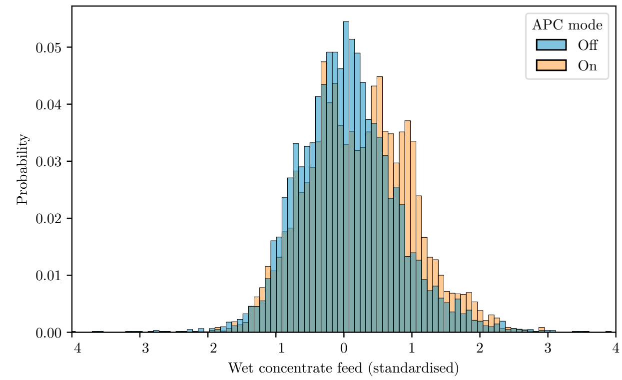

Fig. 4. Concentrate feed distributions by APC mode.

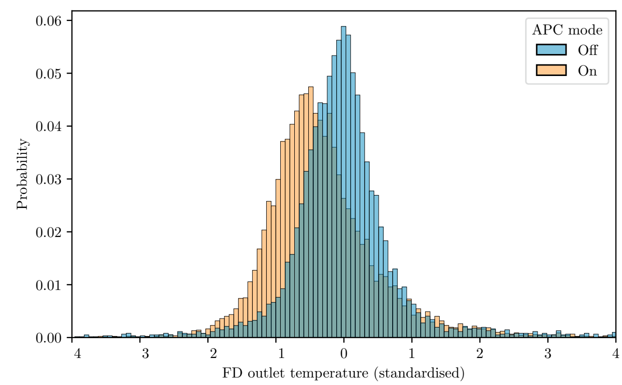

Fig. 5. Flash dryer outlet temperature distributions by APC mode.

Fig. 4 shows the wet concentrate feed distributions by APC mode for the period under evaluation, standardised such that the APC off data has a mean of 0 and standard deviation of 1. The mean wet concentrate feed is 1.80% higher with the APC on (in tons per hour).

Differing conditions like moisture content of the wet concentrate feed or coal calorific values dictate how much wet concentrate feed can be processed. The drying temperature, as shown in Fig. 5, is therefore a good inferential indication of whether the maximal amount of concentrate is being processed (when this temperature is minimized). The mean drying temperature is lower with the APC on.

The primary indication that the maximal amount of energy is transferred to the FD is the HGG outlet temperature. This temperature should be maximized, but the high limit is an equipment safety consideration. Fig. 6 shows the HGG outlet temperature distributions; that look rather similar with the APC off and on.

Through appropriate use of the fluidising damper however the APC is able to achieve this HGG outlet temperature at lower coal feed by minimizing the HGG average bed temperature. Fig. 7 shows the coal feed distributions. There is no weightometer for the coal feed, but an energy balance based calculation indicates that the specific coal consumption (the tons of coal required per ton of con-

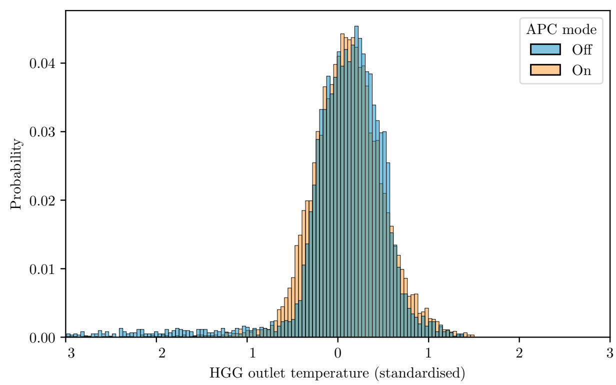

Fig. 6. HGG outlet temperature distributions by APC mode.

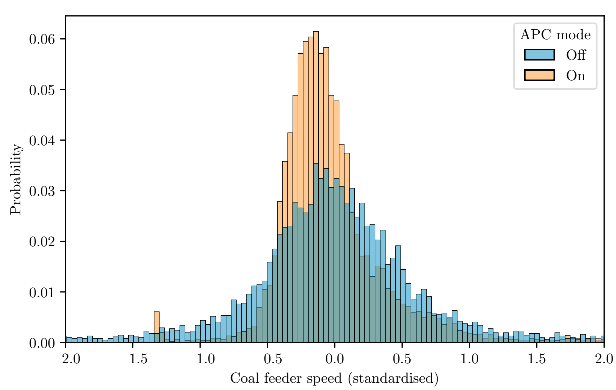

Fig. 7. Coal feeder speed distributions by APC mode.

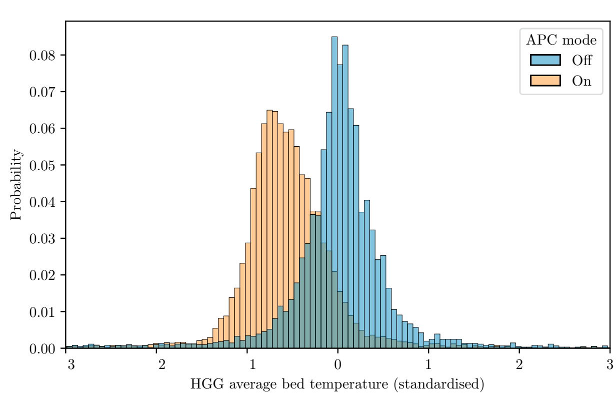

Fig. 8. HGG average bed temperature distributions by APC mode.

centrate dried) reduced by around 1 %. This calculation is however less reliable than what a direct measurement would have been.

Fig. 8 shows the HGG average bed temperature distributions and Fig. 9 shows it trended over time (showing roughly three and a half days' of data). From the time-series trend it is clear that the APC off mean is higher than the APC on mean, but also that there is less high frequency movement when the APC is on.

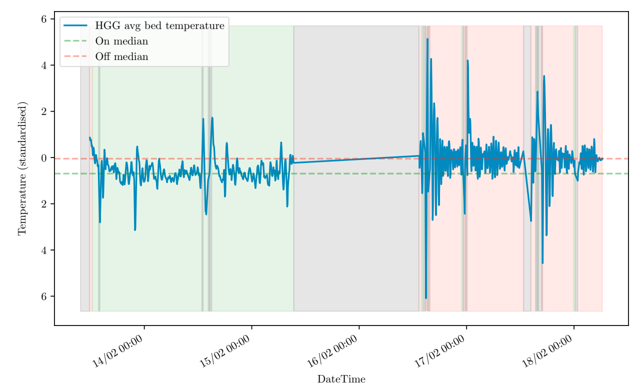

Fig. 9. HGG average bed temperature time-series plot by APC mode, showing the value in blue, the data median when the APC is on in green (dashed), and the data median when the APC is off in red (dashed). The background is also shaded in similar colors, with gray indicating that the process is off.

4.2 Start-up operation results

The MPC controller not only lead to an improvement over the decoupled PID controllers, but the ROC-based philosophy was also an improvement over pre-existing MPC controllers, especially with regards to the stability of the HGG average bed temperature after starting the coal feeder or when recovering from a significant disturbance.

The most striking improvement in coal feeder start-ups over a pre-existing MPC was seen at Mortimer Smelter. Fig. 10 shows an example of how the APC can go into a trip cycle when controlling the temperature directly. Fig. 10 shows 2 start-ups, one without the ROC control (in blue) and one with the ROC control (in orange). The start-up events are time-shifted such that the points where the process is considered to be on coincide.

The top panel in Fig. 10 shows the HGG average bed temperatures and the bottom panel shows the coal feeder speeds (all normalised). When the HGG average bed temperature exceeds its high limit the coal feeder trips and goes to 0. This causes the HGG average bed temperature to decrease significantly. When the temperature has decreased sufficiently the coal feeder starts up again. The APC ramps up the coal feed quickly to increase the temperature. Once the temperature starts to approach its high limit the APC will reduce the coal feeder speed. The temperature is however increasing so rapidly that the coal feeder speed reduction is insufficient to prevent a subsequent coal feeder trip.

When the ROC control is active the APC cuts back on the coal feeder speed much sooner. Even though the bed temperature is not yet approaching the high limit, the ROC is already violating its high limit. Cutting back so much sooner allows the bed temperature to turn before exceeding the trip limit and therefore preventing further coal feeder trips.

This "coal feeder trip cycle" is primarily owing to the nonlinear nature of the process. Tuning the controller to be sufficiently aggressive in this regime makes it too aggressive during normal operation. With the ROC control

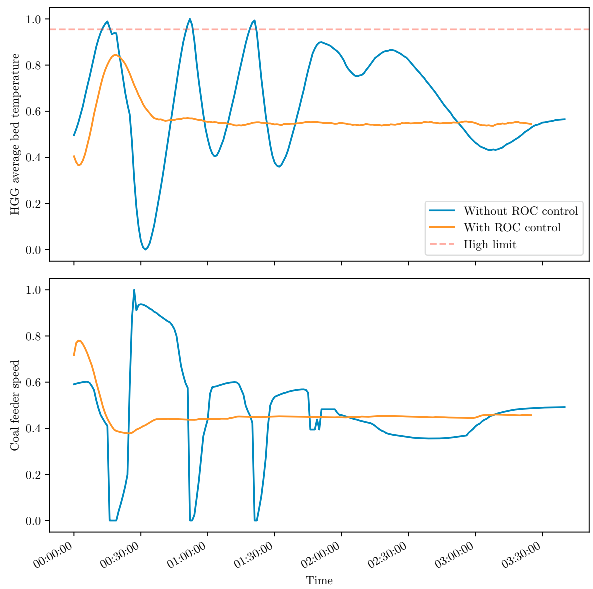

Fig. 10. HGG average bed temperature (top panel) and coal feeder speed (bottom panel) during and after a coal feeder trip cycle.

it is possible to maintain a single tuning set that is much more robust to the nonlinearities associated with operating over such a wide expanse (viz. during start-up and normal operation).

At Mortimer Smelter, 1-minute data for roughly 1 month around the commissioning period was taken to assess the start-up operation improvement. This gave 13,806 data points of operating without the ROC control and 8,922 data points with the ROC control. Without ROC control there were 155 coal feeder trips (89 minutes mean time between trips) and with ROC control there were 21 (425 minutes mean time between trips).

Considering only the first 60 minutes after starting the coal feeder (to quantify only post start-up trips) there were 102 trips in 2,763 minutes without the ROC control (27 minutes mean time between trips) and 14 trips in 1,346 minutes with the ROC control (96 minutes mean time between trips).

5. CONCLUSION

The optimization of a flash drying unit is not straightforward owing to the fast FD dynamics compared to the slow HGG dynamics, the nonlinearities present in the HGG temperature responses, and the process interactions.

This work presents an approach to optimizing the flash drying unit by controlling the HGG average bed temperature ROC as opposed to controlling the temperature directly. When controlling the temperature ROC the limits have to be determined dynamically.

This approach allowed better temperature control across the FD unit resulting in the ability to process 1.80 % more wet concentrate without the need to increase the coal feed. The improved temperature stability also reduces the number of coal feeder trips encountered after start-ups and significant disturbances.

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