Optimization of a Municipal Solid Waste incinerator

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SUMMARY: The benefits of a more stabilized combustion on a Municipal Solid Waste incinerator are many and so are the means to reach it (Fuzzy logic, MPC, cameras, etc.). Before implementing an advanced solution, the existing PID structure must be optimized. This case story on a real incinerator shows that bounds of variation have lowered with ~50% and changed the dynamic behaviour to bee more linearly.

1. INTRODUCTION

Municipal Solid Waste (MSW) incinerators are an important tool in waste management world wide. When waste is incinerated, the volume is reduced by approximately 90% and the weight by 75%-80% (Christensen, T.H., 1998). For the past 50 years, increasing focus has been placed on utilizing the hot flue gases from the incineration process in boilers. The energy extracted in the boiler is used for producing heat, process steam or power, thereby contributing to the reduction of energy based on fossil fuels and the emission of CO_2 and is also an important element in the economy of operating an MSW incinerator. This has resulted in a desire of the operators to stabilize their energy production and operate the turbine in the most efficient way to maximize production. The stabilized combustion also contributes to lowering emissions, e.g. because the flue gas cleaning system can be operated more precisely. It is therefore of great economical and environmental interest to the MSW operators to stabilize their combustion process.

One way of stabilizing the combustion process is mentioned in Zipser S. & al (2004) where an image analysis of an infrared camera is used for optimizing the combustion process. Another way is Model Predictive Control (MPC) implemented on a first principle model (Leskens M. & al (2005.). In Müller B. & al (1998) a fuzzy logic control is used together with image analysis to ensure controlled burn-out of the refuse.

Before the operators invest in advanced control strategies and/or measurements, the existing PID control strategy has to be optimized (Baxter, D. and R. El Asri, 2004) and so should the parameters in the PID-controllers. The upgrading of a control strategy is visualized in Figure 1. The PID controller is the most common controller and consists of a **P**roportional, **I**ntegral and **D**erivative action. A more detailed explanation of PID is given later in this paper, see section 2.

It is possible to make PID control strategy adapt to the actual working conditions by simple means, e.g. by the use of gain scaling the set points.

More advanced control techniques, e.g. MPC, has the strength to optimize the control signals based on two or more inputs/measurements and use system constraints at the same time. The optimization method is often based on a cost function. This cost function can be solved in different ways, for example as a quadratic problem.

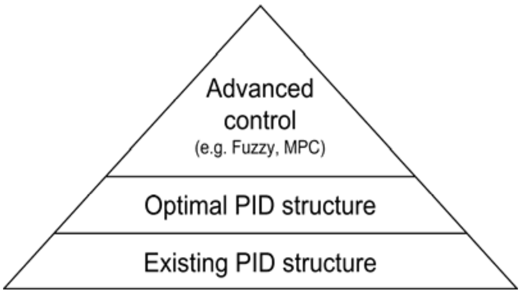

Figure 1. The control strategy upgrading triangle.

Good understanding of the process dynamics is a *must* in developing the optimal PID control strategy. A classical way of understanding the combustion process on a travelling grate stoker is the concept of the travelling pot furnace (Marskell, W.G. and J.M. Miller, 1946). By way of this assumption the reaction front of the combustion will move from the top of the waste bed and downwards. According to van Kessel & al (2004) the grate movement increases the ignition in the waste bed, starting a new state "complete combustion". This is especially seen when burning solid fuels with air preheating.

Burning Municipal Solid Waste is difficult because a capacity of unburned waste with changing calorific value lies behind the reaction front. Controlling the reaction rate and the unburned capacity of waste is the key to coping with the disturbance in the boiler production. The control strategy has to secure the right amount of waste to be in the right place at the right time. At the same time the ignition rate of the unburned waste must be controlled.

This case study is based on I/S Fasan, line 4 (Fasan 4); an MSW incinerator with a nominal steam production of 30.5 t/h (steam parameters: 405 °C and 52 Bar), see

Figure 2. The MSW incinerator is located near the town of Næstved in Denmark and is owned and operated by the local municipality. The incinerator was commissioned in 2005. The plant has three lines in operation, supplying district heating and power to the grid. The incinerator was chosen because it represents a state-of-the-art MSW incinerator with a configuration and instrumentation which is typical of its size.

Capacity - Nominal load	
Steam flow	30.5 t/h,
Steam data	405°C, 52 bar
Waste	8 t/h with 12 MJ/kg
Configuration	
Grate	One-line BS grate - air cooled
Furnace	Water cooled
Boiler	Three radiant passes and a vertical superheater section
Turbine	ABB

Figure 2. Photo and overall plant data for I/S FASAN, line 4, Næstved, Denmark.

2. PID-CONTROLLER

The PID controller is a very commonly used type of controller. It is relatively robust and the need for process information is limited. The controller consists of a proportional part (P), an integral part (I) and a differential part (D). The theoretical transfer function from the error between the process value end the set point to the control signal is in the Laplace domain (Jannerup, O. and P.S Sørensen, 2000)

$$G_c(s) = K_p\left(1 + \frac{1}{\tau_i s} + \tau_d s\right) \qquad (1)$$

PID controllers can be successfully used for control systems, which can be described by a First-Order-Plus-DeadTime (FOPDT) and can easily be tuned by one parameter λ (Dahlin, E.B., 1968). Also, more difficult dynamics, e.g. inverse response, can be controlled very well with PID-controllers. A tuning method for such systems is proposed in Chien, I-L. & al (2003).

When the control signal from the controller is proportional to the error, the controller is called a P controller.

The control signal from a PD controller is proportional to the error and the derivative.

A PI controller acts proportionally to the error and integrates the error to eliminate a stationary offset by increasing/decreasing the control signal.

When all parts (P, I, D) are used in calculating the control action, the controller is called a PID controller, but this is often used as a general term to describe the controller type.

3. EXPERIMENTAL STUDY

To examine the difference between two PID structures, a real plant (Fasan 4) has been used. Using the same plant to test the two structures eliminates the effect of different mechanical layout/design. The existing PID structure at Fasan 4 has been optimized and the new PID structure was switched on 31st March 2008. In this section the differences between the two PID structures are described, and the conditions for the used dataset are explained.

3.1 The old PID structure

A sketch of the PID structure commissioned in 2005 can be seen in Figure 3. It can be seen that the oxygen content is controlled by means of two controllers. The first oxygen controller controls the amount of secondary air and is a fast controller with a PD configuration. The second oxygen controller controls the grate speed and is slower, with a PI configuration.

Based on the actual steam production, the energy controller corrects the amount of primary air and the grate speed.

The outputs of the oxygen and energy controllers correct the set points for the underlying controllers (cascade coupled) which control the aggregates, e.g. the air dampers and the hydraulic valves. When controllers are coupled in cascade series, it is done to cope with non-linearities in e.g. flow control, where temperature or pressure may change.

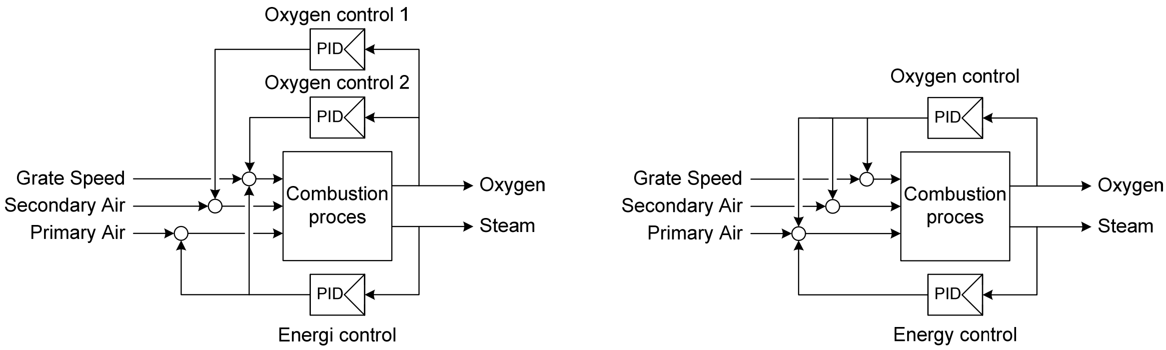

Figure 3. The left figure shows the PID structure commissioned in 2005, referred to as the old PID structure. The figure to the right shows the improved PID structure activated in the spring of 2008, referred to as the new PID structure.

When comparing the two PID structures shown in Figure 3, it can be seen that the energy control in the new PID structure does not have a influence on the grate speed. Removing the direct influence on the grate speed from the energy controller means that the task of correcting the grate speed now relies on the oxygen controller. If the steam production decreases, the energy controller will increase the amount of primary air. If there is "enough" capacity of unburnt waste available in the combustion zone, the waste will ignite and burn. In case no unburnt waste is available in the combustion zone, the oxygen will increase, resulting in an increased grate speed. The increase in grate speed will increase the supply of waste to the combustion zone. When the oxygen drops again, the grate speed will decrease. In this way the supply of waste is secured. The oxygen controller is supported by a bed layer control.

3.3 Dataset

The datasets used in this paper consist of two log types, one dataset with a time interval of 1 minute and another with a time interval of 1 hour. The 1 minute log is an average log and the 1 hour log is also an average log, but the1 hour log has also saved the maximum and minimum value recorded each hour.

The data analysis on the long term is based on the average, maximum and minimum registered in the one hour log from 1/3-2008 00:00 to 2/5-2008 23:00, after which Fasan 4 was shut down due to the annual maintenance.

The dynamic analysis is performed on the basis of two datasets of 2 days' duration based on 1 minute values (2880 samples). Only a very condensed extraction of the result of this analysis is included in this paper.

4. RESULTS AND DISCUSSION

4.1 Performance

The PID structure at Fasan 4 has been optimized and the new PID structure was switched on the 31st March 2008 11:35. An improvement is obvious after the first couple of hours, after the parameters in the new PID-controllers were manually tuned, see Figure 4.

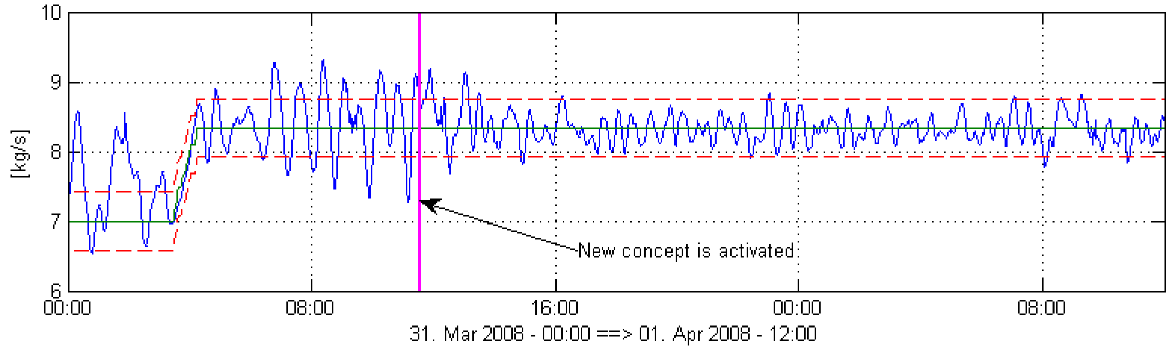

Figure 4. The new concept is activated. The green line is the set point. The blue line is the actual steam production. The red lines are +/- 5% of the set point.

Figure 4 shows the effect on the short term, i.e. the first 24 hours after implementation. The stability of the control concept must be estimated on the long term as shown in Figure 5. Figure 5 is based on the 1 hour log. The period left of the magenta line shows Fasan 4 controlled by the old PID structure and on the right side of the blue line is Fasan 4 controlled by the new PID structure. The period between the magenta and the blue line is excluded from the statistical evaluation, because the data in this period is based on the oxygen analyzer at the chimney instead of the otherwise used analyzer after the economizer, which had a malfunction during this period and was subsequently replaced (indicated by the blue line). Using the analyzer at the chimney resulted in an increased time delay of the oxygen controller by ~3½ minute, making the control slower.

Statistical information about the evaluation for the long horizon is shown in Table 1. Beside the average value, Table 1 also contains the median as a way of removing the influence of large disturbances/outliers. From Table 1 it can be seen that the maximum and minimum values of the variation have moved closer to the average steam production by ~2% equal to a relative ~50% change in the mean value. The normal set point used by the operators is 8.33 kg/s.

Making a test of the hypothesis (t-test of $H_0 : \mu_{old} = \mu_{new}$) that the means are equal between the old PID and the new PID structure for both the maximum and the minimum steam production can be rejected, as shown in Table 1. The rejection of the hypothesis validates that the upper and lower bound of the variations has changed by ~2% (relative ~50%). The rejection of the hypothesis can also be made when the variance is not assumed to be equal (Behrens-Fisher problem).

Figure 5 shows some outliers, especially downwards in the period controlled by the old PID structure. These outliers can be due to “load breakdowns” or reduction of set point. The term “load breakdown” describes a situation where the calorific value decreases in such a way that the control can not maintain the steam production. This state is often overcome by reducing the load/steam set point.

The calorific value for the period is also shown in Figure 5 where upward peaks can be seen. These peaks are due to a mix of an actual rise in the calorific value or a situation where the waste goes in the chute before the weight is logged in the system. This is a failure that may occur during periods of manual crane operation. False (or no) registration of the crane weight disturbs the mass-based calculation of the calorific value. Fasan 4 is designed to maintain full thermal load with a calorific value from 9.1 MJ/kg to 14 MJ/kg. As shown in Figure 5, the calorific value is lower than 9.1MJ/kg for a longer period on the right side of the blue line, increasing the risk of “load breakdowns”. Being able to operate at a lower calorific value than 9.1 MJ/kg without reducing the thermal load and with a reduced number of load breakdowns (see Figure 5) indicates that the new PID structure is more robust.

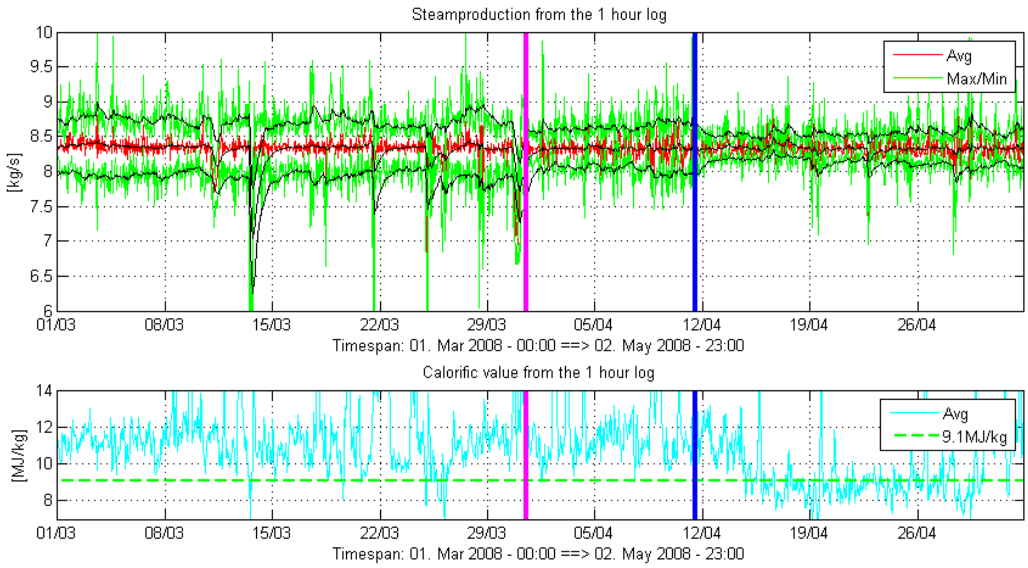

Figure 5. Data from the 1 hour log. The upper plot shows the steam production. The red line is the hourly average steam production. The green lines are the maximum or minimum for each hour. The black lines are exponential smoothened data for the last 12 hours for the maximum and minimum values. The vertical magenta line (31/3-2008 11:35) represents the end of the old PID structure. The vertical blue line (11/4-2008 12:00) represents the start of the new PID structure. The cyan coloured line in the lower plot shows the hourly average of the calculated calorific value based on a crane weight and energy balance. The green dotted line represents the lower calorific design limit for full thermal load (9.1 MJ/kg).

Table 1. Statistical evaluation of the old PID structure and the new PID structure. Mean, median, standard deviation of the steam production (Steam) and the calorific value (CV) are shown. The original dataset is plotted in Figure 5.

		Old PID	New PID	Difference	$H_0 : \mu_{old} = \mu_{new}$
$\mathrm{Steam}_{\mathrm{Max}}$	mean	8.697	8.556	-0.140 (-1.6%)	Rejected
	median	8.702	8.521	-0.181 (-2.1%)	
	std	0.3973	0.2161		
$\mathrm{Steam}_{\mathrm{Avg}}$	mean	8.293	8.331	-0.038	Rejected
	median	8.351	8.323	0.028	
	std	0.3966	0.1590		
$\mathrm{Steam}_{\mathrm{Min}}$	mean	7.869	8.120	0.251 (3.2%)	Rejected
	median	7.974	8.149	0.175 (2.2%)	
	std	0.5243	0.2002		
$\mathrm{CV}_{\mathrm{Average}}$	mean	11.967	9.807	-2.160	Rejected
	median	11.179	9.161	-2.018	
	std	4.6171	2.5944		

4.2 Evaluation of the PID structures

If the oxygen content in the flue gas and the steam production are plotted against each other, a linear relation between the two variables appears. A control reaction from the oxygen controller will therefore be correlated with the steam production, especially when they have influence on the same input to the combustion process.

In Figure 6 the impulse response function from the oxygen to the steam production is shown. The figures are based on a causal dataset. The dataset can be made into a non-causal dataset by prewhitening the impulse response function, according to Madsen, H. (2007). Prewhitening removes the cross-correlation between the input and output by a filtering procedure (Madsen, H., 2007).

Figure 6 shows that the steam production with the old PID structure tends to oscillate in contrast to the new PID structure, in which impulse response reaches a steady state after 15 minutes. The change in the dynamical behaviour is also clearly seen in Figure 4. The signature in the right sub-plot in Figure 6 looks more like a first order response; this type of responses are easier for a PID controller to control, see section 2.

Figure 6 is based on non-prewhitened data but the equivalent prewhitened data display a similar trend signature (not shown). The signature of the new PID structure decays from time=0 compared to time=5 minutes on the non-prewhitened data (see Figure 6). This indicates that a fast oxygen controller will result in a stable energy production.

Thornhill, N.K. & al (1998) shows that the impulse response can be used to evaluate a PID-controller in closed loop. The effect of the oxygen and energy controller in closed loop is estimated by the means of autocorrelation, cross-correlation and impulse response function (figures not shown in this paper). This shows that the oxygen controller has a significantly larger influence on the oxygen content with the new PID structure compared to the old PID structure. In contrast, the energy controller has significantly less influence on controlling the steam production.

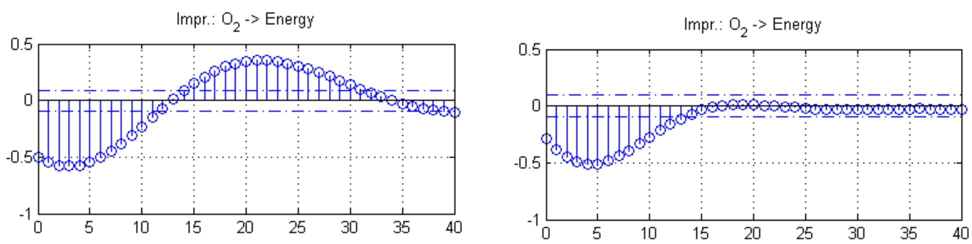

Figure 6. The left sub-plot shows the impulse response function with oxygen as input and steam production as output for the old PID structure. The right sub-plot shows the same, but for the new PID structure. The Y-axis shows the size of the coefficient (normalized data) and the X-axis the time lag in minutes.

5. CONCLUSIONS

This full-scale study, including analysis of the data with time series analysis, shows that optimizing the PID structure on an MSW incinerator can improve the stability of the steam production in more than one way. First, the new PID structure minimizes the variations in the steam production using the oxygen controller more actively. Secondly, the new PID structure is more robust and can handle low calorific value without increasing the risk of a “load breakdown”.

The analysis of the impulse response shows that dynamics have changed to a response which tends to be more like a first order response, which makes it easier for a PID controller to cope.

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