1 Introduction

Power generation is responsible for a significant part of the total emissions of solid, liquid and gaseous pollutants in Europe. Due to a predicted higher long-term availability of solid fuels, in particular coal, compared to oil and natural gas, solid fuels will play an important part in future energy supply. Further, the demand for clean air and stringent environmental regulations are forcing consideration of alternative technologies, with the advantages of high energy conversion efficiency and reduced pollutant emissions. However, power generation by conventional coal firing has a much greater environmental impact compared to natural gas based systems. Therefore, combustion technologies have been developed which aim at an environmentally advantageous use of coal in power generation plants [5].

As a result of this, Integrated Gasification Combined Cycle (IGCC) power plants, combining gasification with a gas and steam cycle, are being developed around the world. The gasification plant considered in this study is based on the spouted fluidised bed gasification concept and can be considered as a reactor where coal is gasified with air and steam. Pulverised coal and limestone, which is required to capture sulphur, is conveyed by pressurised air and steam and spouted into the gasifier. The air and steam fluidise the solids in the gasifier and react with the carbon and volatiles from the coal, producing a low calorific value fuel gas. The remaining char (ash from coal, limestone and unreacted carbon) is removed as bed material from the base of the gasifier or elutriated as fines with the product gas. The fuel gas is then cleaned and fed to a gas turbine.

The coal gasifier is essentially a chemical reactor where coal reacts with air and steam. The products of the gasification process are a low calorific value fuel gas, which can be burnt in a suitably adapted gas turbine, and char. Limestone is also added to the vessel to capture the majority of sulphur present in the coal.

The gasifier is difficult to control since it is multi-variable and highly nonlinear, with significant cross-coupling between the input and output variables. In existing gasifiers of this type, the automatic control goes no further than closing local PI or PID flow loops around the feed system actuators. For example, on the experimental gasifier, the control loop is closed by a skilled human operator, who used his judgement to modulate the mass flow ratios and hence provide the set points for the input actuator (PID) controllers. Whilst this approach provides basic control, it is clearly not viable as a commercial proposition and there exists the opportunity to achieve significantly improved operation via the use of modern advanced control techniques.

Previous $H_{\infty}$ designs applied to the linear gasifier model are described in [7] and [6]. Both of these designs use $H_{\infty}$ mixed-sensitivity optimisation, whereas, in the present study, the $H_{\infty}$ loopshaping design method is employed, in combination with a $H_{\infty}$ optimised anti-windup compensator.

2 Nonlinear gasifier model

A detailed nonlinear gasifier model was developed by the ALSTOM Energy Technology Centre which includes all the significant plant dynamics: e.g., drying of coal and limestone, pyrolysis and volatilisation of coal, the gasification process and elutriation of fines. The model was validated using measured time histories from an experimental test facility and was demonstrated to predict the main trends in fuel gas quality [8]. Control specifications, along with the notation used, may be found in [9].

3 Controller design method

The controller design procedure used in this study is based on $H_{\infty}$ robust stabilisation combined with classical loop-shaping, as first proposed by McFarlane and Glover [4]. The resulting $H_{\infty}$ Loop-Shaping design method is essentially a two stage process. First, the open loop plant is augmented by (generally diagonal) weighting matrices to give a desired shape to the singular values of the open-loop frequency response. Then, the resulting shaped plant is robustly stabilised with respect to coprime factor uncertainty using $H_{\infty}$ optimisation. A particular implementation structure, [3], for $H_{\infty}$ Loop-Shaping controllers is shown in Figure 1. With reference to this figure, the weighting matrix $W_{1}(s)$ is chosen to add integral action and ensure reasonable roll-off rates...
for the open-loop singular values around the desired crossover frequencies. The constant weighting matrix $k$ is then used to adjust control actuation requirements to respect the various actuator rate and magnitude limits.

The controller design was carried out using the linearised model of the plant at 100% operating load. For the controller discussed in this paper, the open-loop plant was initially augmented with diagonal input and output scaling matrices:

$$
s_{\text{input}} = \text{diag}(5, 10, 3, 1.75)$$

$$s_{\text{output}} = \text{diag}(\frac{1}{4500}, 1, \frac{1}{600}, 5);$$

in order to reflect variations in control power between different actuators as well as the use of different units to measure different controlled outputs. The weighting matrices for the $H_\infty$ Loop-Shaping controller were chosen as:

$$k = \text{diag}(0.25, 0.25, 0.25), W_1 = \frac{s + 0.05}{s} \times I_{4 \times 4}$$

These weights were straightforward to choose, making the design process quick and intuitive. The approach taken in this study was to first select the input and output scaling matrices depending on the actuator position range and the relevant output magnitude, then $W_1$ (generally starting with $W_1 = \frac{s + 1}{s} \times I_{4 \times 4}$, and finally $k$ (generally starting with $k = \text{diag}(1, 1, 1, 1)$).

The second stage of the $H_\infty$ Loop-Shaping design method involves the use of $H_\infty$ optimisation to robustly stabilise the shaped plant against a particular type of uncertainty description, based on stable perturbations to each of the factors in a coprime factorisation of the plant. For a plant $G(s)$ with normalised left coprime factorisation, $G = M^{-1}N$, a perturbed plant model can be written as

$$G_p = (M + \Delta_M)^{-1}(N + \Delta_N)$$

where $\Delta_M$ and $\Delta_N$ are stable unknown transfer functions which represent the uncertainty in the nominal plant model $G$. This arrangement removes the usual restriction included in other uncertainty models for the nominal and perturbed plants to have the same number of unstable poles. The objective of robust stabilisation is then to compute a feedback controller which stabilises the family of plants $G_p$. A bound for the ‘size’ of the uncertainty in $G_p$ can be defined as

$$\| [\Delta_N \Delta_M] \|_\infty \leq \epsilon$$

where $\epsilon$ is then the stability margin. The optimal robust stabilisation controller $K_{\infty}(s)$ is thus the controller which maximises this stability margin for a given shaped plant. This controller can be computed explicitly by solving two Riccati equations, thus avoiding the iterative procedures required in general $H_\infty$ optimisation. It has also been shown theoretically that for $\epsilon > 0.25$, (a) the original loop-shapes are largely preserved thus retaining the desired nominal performance, [4], and (b) the closed-loop system will have good robust performance properties [3]. For our design, the controller $K_{\infty}(s)$ gave a value of $\epsilon$ equal to 0.264, thus guaranteeing closed-loop stability for coprime factor uncertainty of at least 26%. Finally, the constant pre-filter $K_{\infty}(0)W_2$ is formed to ensure zero steady state tracking error, assuming integral action in $W_1$.

The order of the controller is equal to the order of the plant plus the order of the weights i.e. 21 states. This was reduced using standard model order reduction techniques, in this case balanced residualisation methods [1], to 10 with no loss of performance.

The resulting controller was found to be sufficiently robust across the plant operating regime to not require any controller scheduling or switching scheme.

4 $H_\infty$ optimised anti-windup

From an analysis of initial nonlinear simulation results with the above controller it was concluded that some form of anti-windup compensation was required for the present application, due to the frequent actuator saturation (both position and rate) encountered during the specified disturbance tests. The method used to design the anti-windup compensator was again based on $H_\infty$ optimisation, [2], and is an extension of the model-based anti-windup method, where the direct model is synthesised via an $H_\infty$ optimisation.

The key idea in this approach is to compensate the plant output $y$ before it is used in the controller. The closed-loop is formed by the original controller and the direct model as shown in Figure 2, where $e$ is the feedback error, $u$ is the output from the controller, and $\hat{u}$ is the actual input to the plant.

![Figure 2: Direct model and controller closed-loop](image-url)
make the signal:

\[ \ddot{u} - u = S(s)\ddot{u} - S(s)K(s)e \]

small (where \( S = (I + KG_m)^{-1} \)) while also minimising the error \( e \). The optimisation problem to be solved is thus:

\[
\min_{\text{Stab } G_m} \left\| \begin{bmatrix} W_{aw1}S & -W_{aw1}SK \\ W_{aw2}G_m & -W_{aw2}G_mSK \end{bmatrix} \right\|_\infty
\]

where \( W_{aw1} \) and \( W_{aw2} \) are frequency dependant weights, chosen to emphasise the importance of the compensated signals over different frequency ranges. The ‘standard plant’ representation is

\[
\min_{\text{Stab } G_m} \| \mathcal{F}(P, G_m) \|
\]

where

\[
\begin{align*}
0 & \quad \quad 0 \\
I & \quad -K
\end{align*}
\]

\[
P(s) = \begin{bmatrix} W_{aw1} & -W_{aw1}K \\ 0 & 0 \end{bmatrix} - \begin{bmatrix} -W_{aw1}K \\ W_{aw2} \end{bmatrix} - \begin{bmatrix} I & -K \\ -K & -K \end{bmatrix}
\]

In this setup, the exogenous inputs ‘\( w \)’ are \( e \) and \( \ddot{u} \), the error signals to be made small (‘\( z \)’) are \( W_1(\ddot{u} - u) \) and \( W_2G_m(\ddot{u} - u) \), \( K(s) \) is the loop-shaping controller, and the anti-windup compensator, \( G_m(s) \) is driven by \( \ddot{u} - u \). \( W_1 \) is usually chosen to be a low pass filter, and \( W_2 \) a high-pass filter. This optimisation problem was easily solved using Matlab.

Such an approach does not guarantee that the synthesised anti-windup compensator \( G_m \) is stable, which is essential in this situation. In practice, however, no difficulty has been found in selecting weights that result in a stable \( G_m \).

Weights were selected as

\[
W_{aw1} = \frac{1}{s + 6} \times I_{4 \times 4}, \quad W_{aw2} = 6 \times I_{4 \times 4}
\]

These weights again were straightforward to choose, with just a few iterations on the initial choice for both, (starting with \( W_{aw1} = \frac{1}{s+7} \times I_{4 \times 4} \) and \( W_{aw2} = 1 \times I_{4 \times 4} \)), making the design a relatively simple task.

The price to be paid for the inclusion of the anti-windup compensator in the design is the increase in the order of the system. However, it was seen to be easy to reduce the 29th order anti-windup compensator to 8 states with no loss of performance, again using a balanced residualisation. This makes the overall order of the controller and anti-windup compensator 18 states.

5 Nonlinear simulation results

For all of the figures included here, the response from the \( \mathcal{H}_\infty \) design is plotted using a solid line, with the response from the PI design plotted using a - - . line. Maximum and minimum limits for both controlled outputs and actuator position limits are plotted with a - - line. Figures 4 to 5 show the nonlinear output and actuator responses for a change in coal quality of +14%, with the gasifier at 100%. This is the maximum positive coal quality variation that can be tolerated for the 300 second simulation. However, it can be seen that the CVGAS is still increasing at the end of the simulation. If the simulation time is increased to 25000 seconds, 7% is the maximum positive coal variation that can be tolerated. It is clear that this limitation results from a simple lack of control power, i.e. the lower limit of WCHR being reached. A similar result is observed for the PI design.

![Figure 3: Plant representation](image)

![Figure 4: Controlled outputs, coal quality disturbance of +14%, gasifier operating at 100%](image)

![Figure 5: Actuators, coal quality disturbance of +14%, gasifier operating at 100%](image)
of 0.04 Hz, with the gasifier acting at 50%. Although CVGAS briefly exceeds its limits, the steady-state values of all of the controlled outputs are acceptable. It can be seen that both WCHR and WCOL are operating on their rate-limits. It is clear that, although the PI controller regulates CVGAS more closely, the $H_\infty$ controller regulates both PGAS and TGAS more closely.

![Figure 6: Controlled outputs, sine pressure disturbance, gasifier operating at 50%](image)

Figure 6: Controlled outputs, sine pressure disturbance, gasifier operating at 50%.

![Figure 7: Actuators, sine pressure disturbance, gasifier operating at 50%](image)

Figure 7: Actuators, sine pressure disturbance, gasifier operating at 50%.

Figures 8 to 9 show the responses to a step pressure disturbance of -0.2 bar, with the gasifier operating at 0%. Although CVGAS exceeds its limit briefly at the time of the disturbance, this is quickly corrected, and the steady-state responses show all of the controlled outputs maintained within their limits.

Figures 10 to 12 show the responses to a ramped load change, starting from 50%, and increasing to 100%. From 10 it can be seen that CVGAS briefly exceeds its limit, but recovers, as does the temperature. Figure 11 shows that the coal input is operating on its maximum limit. Figure 12 shows good tracking of the reference demand.

![Figure 8: Controlled outputs, step pressure disturbance, gasifier operating at 0%](image)

Figure 8: Controlled outputs, step pressure disturbance, gasifier operating at 0%.

![Figure 9: Actuators, step pressure disturbance, gasifier operating at 0%](image)

Figure 9: Actuators, step pressure disturbance, gasifier operating at 0%.

![Figure 10: Controlled outputs, ramp test](image)

Figure 10: Controlled outputs, ramp test.

Table 1 compares the integrated absolute error (IAE) for each of the controlled outputs for the simulations at the 100%, 50% and 0% operating points, for both step and sine pressure disturbances of the
Overall, CVGAS exceeded the limits during the step disturbance tests only transiently (for $\approx 4$ to 5 seconds) when the step disturbance is applied. It then recovers to a good steady-state response. In general, it appears that WCHR rate-limiting is the cause of CVGAS exceeding its limits.

The maximum and minimum coal quality disturbance (up to a specified maximum change of $\pm 18\%$) that produce results that are within acceptable limits with simulations run for 25000 sec are as follows: It can be seen that, overall, similar results are produced for the two approaches.

### 6 Conclusions

A combined $H_\infty$ loop-shaping controller and $H_\infty$ optimised anti-windup compensator were designed for the ALSTOM nonlinear gasifier benchmark problem. The power of the $H_\infty$ optimisation framework to produce robust controllers with a minimum of design effort was clearly demonstrated by the performance of the controller, which compared well with a classical design tuned by an experienced operator on the actual plant.

### References


