Integrated Real-time Optimization of Boiler and Post-combustion System in Coal-based Power Plants via Extremum Seeking

Eugenio Schuster, Carlos Romero, Zheng Yao, Fengqi Si

Abstract— AES Cayuga Unit 1 is a $160MW$ unit, equipped with a low nitrogen oxide (NO_x) firing system and an anhydrous ammonia (NH_3), $TiO_2/V_2O_5/WO_3$ selective catalytic reduction (SCR) system for NO_x emission control. An ammonium bisulfate (ABS) probe was retrofit to the SCR to monitor ABS formation in real-time with the ultimate goal of minimizing air preheater (APH) plugging (ABS concentration) by regulating the APH bypass damper. Recent work on static optimization of coal-based power plants has played a crucial role in improving overall efficiency. However, static optimization falls short in dealing with real-time scenario changes (i.e., cycling unit load, coal quality, firing system maintenance conditions, subsystem failures, plant aging, etc.). Extremum seeking (ES) is proposed in this work to optimally tune boiler operation in order to minimize NO_x production in real-time. The effectiveness of the ES adaptive controller in keeping the system at an optimal operation point in presence of input disturbances and system changes is demonstrated through simulations based on identified models of the boiler, SCR and APH systems.

I. INTRODUCTION

During the last few years, the ever-increasing demand for cost-efficient power generation and stronger environmental regulation has motivated implementation of process optimization strategies in coal-fired power generation. Given that coal is an important element in the energy source portfolio in the U.S.A., process optimization for stack emission reductions and efficiency improvements in coal-fired boilers plays an important role in minimizing operational and maintenance (O&M) costs, and maximizing performance and unit availability. One area that has received significant attention is tuning and optimization of the combustion process for nitrogen oxides (NO_x) emission and unit thermal performance improvement.

Selective Catalytic Reduction (SCR) is a key component in utility companys plans for NO_x emission control. Over $100GW$ of coal-fired generation in the U.S. is expected to have SCR capabilities [1]. SCR systems rely on the chemical reduction of NO_x with ammonia (NH_3) over the surface of a catalyst. A theoretical one-to-one NH_3/NO_x molar ratio would result in conversion of these reactants to environmentally benign molecular nitrogen and water vapor. However, this ideal condition is not always met, resulting in over-conditioning of NH_3 (NH_3 slip), with associated operating
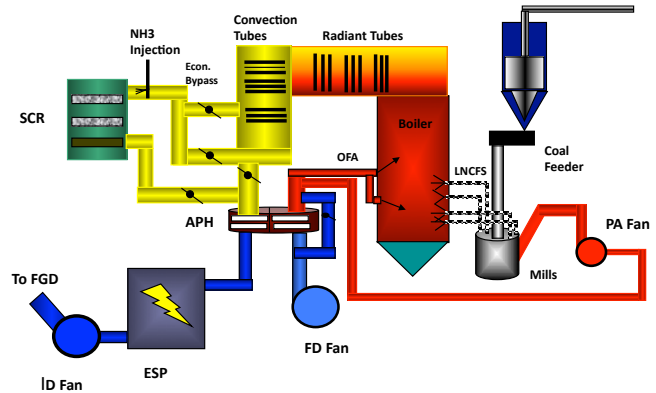
Fig. 1. Conceptual diagram of AES Cayuga Unit 1.

cost penalties. On the one hand, the price of NH_3 has more than doubled in recent years. On the other hand, high NH_3 slip in coal-fired boilers has an adverse impact on cold-end equipment located downstream of the SCR reactor. The concerns include ammonium bisulfate (ABS) deposition, and plugging and corrosion potential of air preheaters (APHs). Controlling and mitigating APH fouling is imperative in coal-fired boilers, since it precludes continued operation of the unit, requiring forced shutdowns for APH cleaning, with the associated lost in unit availability and financial penalty.

The SCR post-combustion NO_x control technology is usually retrofit on boilers equipped with low-NO_x firing systems and on high-dust, high-temperature configurations, with the SCR system located in front of the APH and the dust collection equipment. Many of these SCR systems are implemented and used for compliance with stringent NO_x regulations (year-round stack NO_x levels below the 0.15 lb/MBtu mark), hence, are designed and operated to achieve high NO_x reduction efficiencies of over 90 percent, while minimizing NH_3 slip to below $2ppm$. The challenge of the SCR technology is to achieve cost-effective high levels of NO_x emission performance, while constraining its detrimental impact at the boiler back-end. SCR process variables, such as flue gas temperature and residence time, and NO_x inlet concentration and the level of reagent conditioning all affect SCR performance, and are conditioned by the operating conditions of the boiler. An integrated approach to the optimization of the combustion and post-combustion systems offers an alternative to meet this challenge. Such approach should consider, in a coordinated fashion, the optimal operation of the boiler firing system, SCR reactor, APH and net unit thermal performance.

There is a large list of reported experiences where combustion optimization has proved to be an effective method
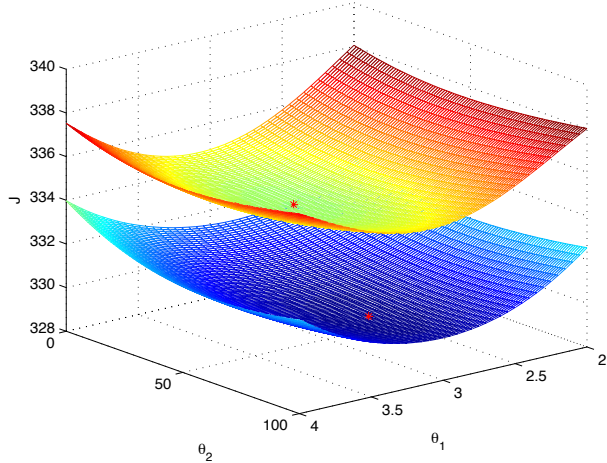
Fig. 2. Cost function J as a function of O_2 (θ_1) and SOFA (θ_2) for two levels of coal flow (Case 1: coal flow = $15ton/hr$ (upper), Case 2: coal flow = $5ton/hr$ (lower)).

Fig. 3. System Identification.

to reduce NOx emissions, while mitigating its impact on the net unit heat rate [2]. With the growing population of SCR systems already retrofitted and projected to be retrofitted in the US, an extension of such optimization techniques to the combined SCR/APH systems is worth exploring. Recent work on static optimization of coal-based power plants has played a crucial role in improving overall efficiency [3]. However, static optimization falls short in dealing with real-time scenario changes (i.e., cycling unit load, coal quality, firing system maintenance conditions, subsystem failures, plant aging, etc.). Taking into account all possible scenarios during an offline static optimization is simply unfeasible. When the plant conditions change in real-time and depart from the conditions considered during the offline static optimization, the optimal set-points are not longer optimal and need to be recomputed online. A real-time adaptive controller has the potential of overcoming this limitation.

This papers reports the results of a feasibility study for a combined boiler/SCR/APH real-time optimization performed at AES Cayuga Unit 1, for cost-effective NO_x emission control. AES Cayuga Unit 1 is a $160MW$ unit, equipped with a low-NO_x firing system and an anhydrous ammonia, $TiO_2/V_2O_5/WO_3$ SCR system for NO_x emission control (see Fig. 1). An ABS probe was retrofit to the SCR to monitor ABS formation in real-time. A recently proposed control system upgrade includes a control strategy provision for the APH bypass damper. Such control strategy regulates the ABS deposition location by manipulating the average cold-end APH temperature with the ultimate goal of minimizing APH plugging (ABS concentration). A real-time, adaptive, extremum-seeking (ES) controller [4] has been proposed in this feasibility study to optimally reduce the combined cost of unit NO_x compliance in real-time. An integrated system composed by boiler, SCR and APH is considered in this study with the ultimate goal of synthesizing a controller that regulates a set of boiler inputs (e.g., O_2, separated overfire (SOFA) registers, burner tilt, SOFA tilt, etc.), the NH_3 flow to the SCR system and the APH bypass damper opening in

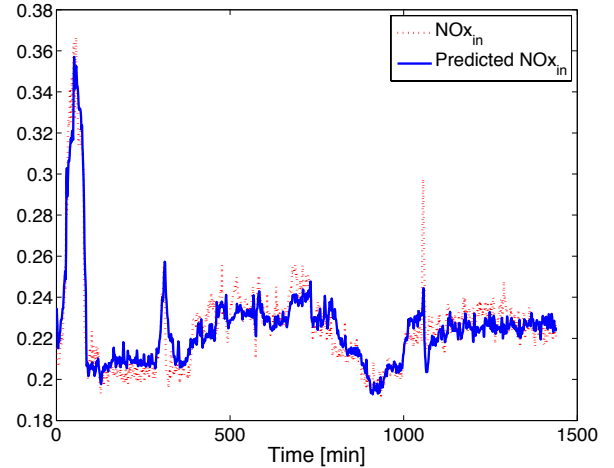order to optimally control in real time and in a coordinated fashion both continuous emission monitoring (CEM) NOx and ABS deposition location within the APH, avoiding NH3 slip and minimizing APH heat rate (HR) penalty.

The paper is organized as follows. Section II motivates the need for real-time dynamic optimization, an overall-economics-control paradigm, and subsystem integration in coal-based power plants. Section III describes the model used to simulate boiler, selective catalytic reactor and air preheater. The fundamentals of the extremum seeking adaptive control method is introduced in Section IV . The control approach together with simulation results are presented in Section V . Conclusions are stated in Section VI.

II. Motivation

A. The Need for Real-time Dynamic Optimization

The surfaces in Fig. 2 represent the cost function

$$J = K_{NO_{x_{in}}} NO_{x_{in}}^2 + K_{HR_B} HR_B^2, \qquad (1)$$

which depends on two parameters: excess O_2 (θ_1) and SOFA register opening (θ_2) ($NO_{x_{in}}$ and HR_B denote the NO_x level at the SCR inlet (Boiler outlet) and the boiler heat rate penalty respectively). The surfaces in the figure have been plotted using operational data obtained through parametric field tests at Cayuga Unit 1 [3]. The upper surface corresponds to a coal flow through the top mill of $15ton/hr$. In this case, the optimal O_2 value is 3.1 and the optimal SOFA value is 42%, while the optimal value of the cost function is $J_{min} = 334$. The lower surface was obtained for a coal flow of $5ton/hr$. In this case, the optimal O_2 value is 3.0 and the optimal SOFA value is 68%, while the optimal value of the cost function is $J_{min} = 329.72$. As it can be noted from the figure, when the plant conditions (coal flow) change, the optimal operation point needs to be recomputed. Coal-based power plants require the development of a real-time adaptive optimal controller that allows for real-time optimization, reacting to any plant change and keeping the system at an optimal operation point.
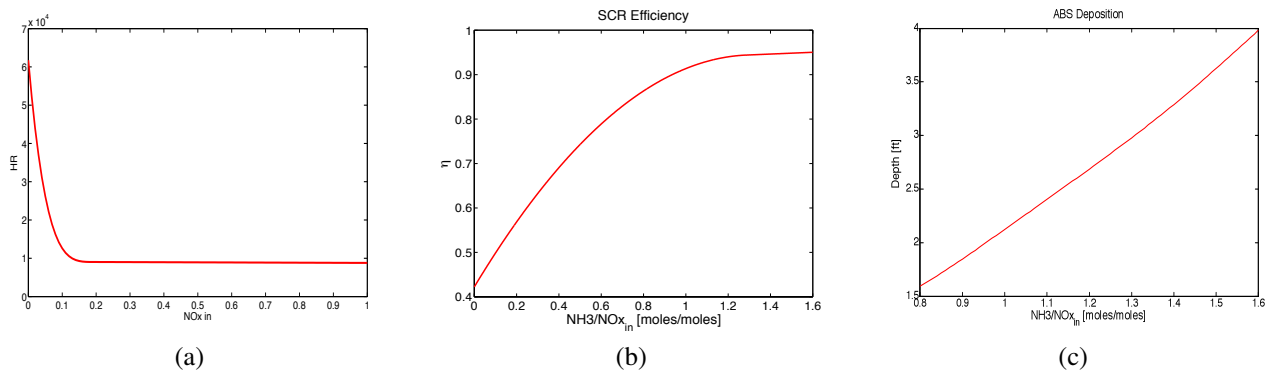
(a)   (b)   (c)

Fig. 4. $NO_{x_{in}}$ vs. HR_B correlation (left); SCR efficiency vs. $NH_3/NO_{x_{in}}$ (middle); ABS deposition depth vs. $NH_3/NO_{x_{in}}$ (right).

B. The need for an Overall-Economics-Control Paradigm

The growing performance requirements in present coal-based power plants demand a transition from the current set-point-control paradigm to an overall-economics-control paradigm, where the goal will be not to regulate plant subsystems around predefined set-points, but to optimize an economic function for the plant. By assigning a dollar value to the CEM NO_x emission, NH_3 cost, NH_3 slip impact, and net unit heat rate penalty, adaptive optimal controllers can be used to minimize in real-time a cost function weighting competing variables by relaxing set-points associated with the many individual control loops currently present in coal-fired power plants. An additional benefit is that the weight factors associated with each one of these variables could be changed over time according to seasonal economical changes.

C. The Need for Subsystem Integration

In order to conciliate competing objectives and achieve in real time an optimal operating point based on the overall economics of the plant, subsystem integration is a must. For instance, ABS deposition depth is currently controlled in coal-fired units equipped with SCRs by regulating the APH bypass damper opening. However, ABS deposition depth is also affected by other parameters, such as the $NH_3/NO_{x_{in}}$ ratio. The variation of this ratio can be seen as a perturbation for the ABS deposition depth control loop in present control architectures. However, taking into account this interdependence may lead to obtaining a combined SCR/APH optimal operation point. For instance, for a given $NO_{x_{in}}$, a decrease in NH_3 injection would result in a decrease in $NH_3/NO_{x_{in}}$, and in turns, in the ABS formation temperature. Therefore, a decrease in NH_3 injection could prevent opening of the APH bypass damper and its associated heat rate penalty. However, a decrease in NH_3 injection could result in an increase in $NO_{x_{CEM}}$. There is then a tradeoff between the NO_x emission level at the stack and the heat rate penalty due to the opening of the APH bypass damper. If a small increase of $NO_{x_{CEM}}$ is tolerated, it can help control the ABS deposition depth by reducing the NH_3 flow rate, in lieu of opening the bypass damper.

III. MODELING

A. Boiler Dynamic Model

In order to illustrate the potential of adaptive ES control for minimization of both the level of NO_x at the SCR inlet and the corresponding net unit heat rate penalty, it is necessary to simulate the dynamics of the boiler. Previous efforts towards both static and dynamic modeling for boiler optimization include first-principles [5], [6], data-driven [7], [8] and hybrid [9], [10] approaches. In this work we have chosen to identify a simplified dynamic model for the boiler from data obtained at Cayuga Unit 1. In contraposition to static models that are obtained from steady-state data, transient data is needed to identify dynamic models. Based on sampled measurements of the boiler variables, an ARMAX model is proposed:

$$A(q)y[k] = \sum_i B_i(q)u_i[k - nk_i] + C(q)e[k]$$

where y denotes the measured output (level of NO_x at the SCR inlet, $NO_{x_{in}}$), u_i, for $i = 1, \ldots, 5$, denote the measured inputs (excess O2, SOFA registers, burner tilt, SOFA tilt, top mill coal flow), and e denotes non-measurable noise. The variable q denotes the time shift operator, i.e., $qy[k] = y[k + 1]$, $q^{-1}y[k] = y[k - 1]$, where k denotes the sampling time. The polynomial $A(q)$ has order na, the polynomials $B_i(q)$ has order nb_i for $i = 1, \ldots, 5$, and $C(q)$ has order nc. The variable nk_i denotes the delay order for each of the inputs i. The coefficients of these polynomials are obtained by solving the following minimization problem:

$$\min_{A, B_i, C} \left(y^*[k] - y[k] \right)^2 \qquad (2)$$

where $y[k]$ denotes the output predicted by the model above and $y^[k]$ represents the real output obtained from direct measurements. Fig. 3 shows the difference between predicted and measured system output after solving the above minimization problem.

From parametric tests it was possible to infer the correlation between $NO_{x_{in}}$ and HR_B shown in Fig. 4 (a).
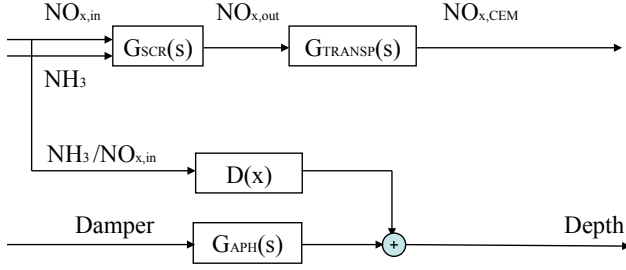
Fig. 5. SCR/APH model structure.

Fig. 6. Extremum seeking control scheme.

B. SCR/APH Dynamic Model

The SCR/APH model structure is shown in Fig. 5. The SCR system is modeled as

$$NO_{x_{out}} = e^{-T_{SCR}} G_{SCR}(NO_{x_{in}}, NH_3)$$
$$= e^{-T_{SCR}} \frac{1}{\tau_{SCR}s + 1} \eta(NH_3/NO_{x_{in}})NO_{x_{in}}, \quad (3)$$

where the efficiency $\eta(NH_3/NO_{x_{in}})$, plotted in Fig. 4 (b), was modeled using operational data obtained through parametric field tests at Cayuga Unit 1. The channel connecting the SCR and the stack was modeled as a transport line, where

$$NO_{x_{CEM}} = e^{-T_{TRANSP}} G_{TRANSP}(s)NO_{x_{out}}$$
$$= e^{-T_{TRANSP}} \frac{1}{\tau_{TRANSP}s + 1} NO_{x_{out}}. \quad (4)$$

The time delays T_{SCR} and T_{TRANSP}, as well as the time constant τ_{SCR} and τ_{TRANSP} were identified from experimental data obtained at Cayuga Unit 1.

The ABS deposition depth is determined by both the formation temperature and the APH temperature. The response of the deposition depth to the bypass damper is modeled as a simple first-order transfer function G_{APS}. The response of the deposition depth to the $NH_3/NO_{x_{in}}$ ratio is modeled by a nonlinear function D shown in Fig. 4 (c) obtained by parametric field tests at Cayuga Unit 1.

IV. EXTREMUM SEEKING CONTROL

Extremum seeking control, a popular tool in control applications in the 1940-50's, has seen a resurgence in popularity as a real time optimization tool in different fields of engineering [4]. Extremum seeking is applicable in situations where there is a nonlinearity in the control problem, and the nonlinearity has a local minimum or a maximum. The parameter space can be multidimensional.

The continuous-time implementation of ES is depicted in Fig. 6, where s denotes the Laplace transform variable. Without loss of generality, the static nonlinear block $J(\theta)$ is assumed to have a minimum J^* at $\theta = \theta^*$. The purpose of the ES optimization algorithm is to use the gradient information of the static map $J(\theta)$ to drive θ to θ^* so that the cost functional $J(\theta)$ is driven to its minimum J^*. The variable $\hat{\theta}$ in Fig. 6 denotes the estimate of the unknown optimal parameter θ^* provided by the ES algorithm. The probing signal $a\sin(\omega t)$, with $a > 0$, added to the estimate $\hat{\theta}$ and fed into the plant helps to get a measure of the gradient information of the map $J(\theta)$. The high-pass filter preserves
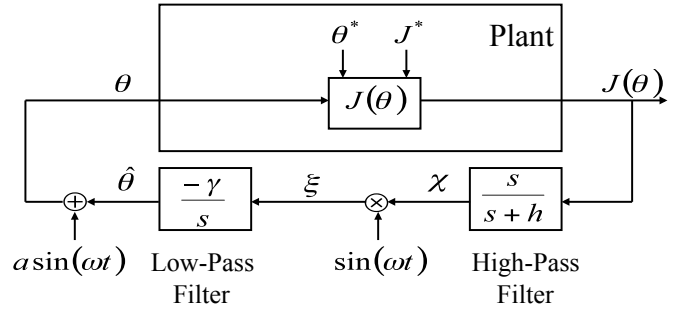
only the perturbation in the cost functional, J, caused by the perturbation in the θ parameter introduced by the probing signal. The demodulator picks the component of the filtered perturbed cost functional χ with the same frequency, ω, as the probing signal. The resulting signal, ξ, which can be seen as proportional to the gradient of the map, $J(\theta)$, is used by the pure-integrator low-pass filter to update the θ parameter in order to drive the cost functional J closer to its minimum. The ES algorithm in Fig. 6 can be written equivalently as

$$\dot{\chi} = -h\chi + \dot{J} \quad (5)$$
$$\xi = \chi \sin(\omega t) \quad (6)$$
$$\dot{\hat{\theta}} = -\gamma \xi \quad (7)$$
$$\theta = \hat{\theta} + a\sin(\omega t) \quad (8)$$

A proper choice of ω allows for extremum-seeking and plant time-scale separation, which is critical for the scheme.

V. CONTROL APPROACH

The control scheme in Fig. 7 summarizes the non-model-based optimal adaptive control strategy for both boiler and SCR/APH systems. The proposed control scheme is based on extremum seeking and a dual loop control approach. The first non-model-based adaptive controller is proposed to regulate the boiler inputs (e.g., excess O_2, SOFA registers, burner tilt, SOFA tilt) to minimize both NO_x emission at the SCR inlet and the boiler heat rate penalty. The second non-model-based adaptive controller is proposed to regulate NH_3 flow to the SCR system and the APH bypass damper opening in order to optimally control in real-time and in a coordinated fashion both the CEM NO_x and ABS deposition within the APH, avoiding NH_3 slip and minimizing the APH heat rate penalty. The effectiveness of the extremum seeking adaptive controllers, in keeping the system at an optimal operation point in the presence of input disturbances and system changes is demonstrated through dynamic simulations.

A. Boiler Real-time Optimization

By defining the cost function

$$J_B = K_{NO_{x_{in}}} NO_{x_{in}}^2 + K_{HR_B} HR_B^2, \quad (9)$$

Fig. 2 shows that there exists an optimal operating point for which a minimum is achieved. By carefully selecting the coefficients $K_{NO_{x_{in}}}$ and K_{HR_B}, different weights can be assigned to the competing objectives $NO_{x_{in}}$ and HR_B.
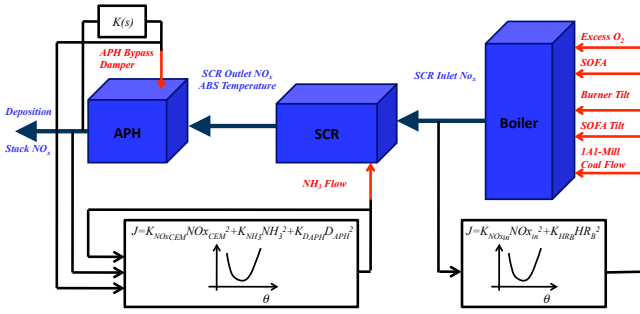
Fig. 7. Control configuration.

An example to illustrate the performance of the non-model-based adaptive ES controller is shown in Fig. 8, where the time evolutions of $NO_{x_{in}}$ and heat rate penalty predicted by the identified dynamic model are used to compute the cost function J_B. In this case, the coal flow is considered an "uncontrolled" input and a step perturbation of magnitude $10 ton/hr$ is introduced at $t = 2,500 min$. Although ES can handle multi-input systems, only the variable O_2 is controlled in this example in order to simplify the exposition. Out of the four possible "controlled" inputs (excess O_2, SOFA registers, burner tilt, SOFA tilt) available in our simplified boiler model, the variable O_2 is chosen after carrying out a sensitivity study based on parametric data. A real-time optimization is carried out based on extremum seeking to modify the value of excess O_2 in order to drive the value of the cost function J_B to its minimum. The level of SOFA registers, and burner and SOFA tilts are kept constant during the simulation. Before the coal flow step, the ES introduces a modification to the O2 level (Fig. 8 (a)) in order to drive the system to the point where the cost function is minimized (Fig. 8 (b)). After the coal flow step, the ES scheme reacts by modifying the O_2 level (Fig. 8 (a)) in order to recover to a cost function minimum. It should be noticed that not only the minimizing values of O_2 are different, but also the cost function minimum values also differ. Fig. 8 (c) and Fig. 8 (d) show how the NO_x and heat rate penalty values are changed by the ES controller. The potential of ES resides on its capability of computing the optimal value in real-time without the aid of a model.

B. SCR/APH Real-time Optimization

We propose a combined ES/PID control architecture. While the NH_3 flow is directly controlled by an ES controller, the APH bypass damper opening (D_{APH}) is controlled by a PID controller driven by the ABS deposition depth error. By defining the cost function as

$$J_{SCR} = K_{NO_{x_{CEM}}} NO_{x_{CEM}}^2 + K_{NH_3} NH_3^2 + K_{D_{APH}} D_{APH}^2, \quad (10)$$

the extremum-seeking controller regulates the NH_3 flow in order to minimize $NO_{x_{CEM}}$, NH_3 flow itself, and D_{APH}. The measurable APH bypass damper opening D_{APH} is a direct indication of the APH HR penalty. The introduction of the bypass damper term in the cost function is a step forward in the efforts to coordinate the SCR and APH control loops. The user-defined weigh factors $K_{NO_{x_{CEM}}}$,

K_{NH_3} and $K_{D_{APH}}$ regulate the tradeoff between $NO_{x_{CEM}}$ minimization and NH_3 and D_{APH} minimizations. Fig. 9 shows the simulation results. The level of NO_x at the SCR inlet is varied as shown in Fig. 9 (a). It is possible to note that while the ABS deposition depth (Fig. 9 (f)) is regulated by the PID at a fixed set-point ($2.5 ft$), the level of CEM NO_x (Fig. 9 (e)) has the freedom to fluctuate (set-point relaxation) in order to minimize the cost function J_{SCR} shown in Fig. 9 (b). The minimizing variable is the NH_3 flow rate shown in Fig. 9 (d). Note that the bypass damper opening D_{APH} shown in Fig. 9 (c) is kept at a relatively low value.

VI. CONCLUSIONS

A real-time, non-model-based, adaptive control scheme has been proposed in a feasibility study carried out at AES Cayuga Unit 1 to optimally reduce the combined cost of unit NO_x compliance in real-time. The proposed control scheme is based on an extremum-seeking dual-loop control approach. The first non-model-based adaptive controller is proposed to regulate the boiler inputs (e.g., excess O_2, SOFA registers, burner tilt, SOFA tilt) to minimize both NO_x level at the SCR inlet and boiler heat rate penalty. The second non-model-based adaptive controller is proposed to regulate NH_3 flow to the SCR system and the APH bypass damper opening in order to optimally control in real-time and in a coordinated fashion both the CEM NO_x and ABS deposition within the APH, avoiding NH_3 slip and minimizing the APH heat rate penalty.

A coordination of both the SCR and APH control systems has been proposed in order to enhance the overall performance of the system. This coordination is achieved through the relaxation of some of the control set-points ($NO_{x_{CEM}}$ in this case). The simple idea behind this approach is that by relaxing some of the control set-points we may be able to gain more than what we lose. This coordination approach leads to the definition of tradeoffs, which are optimally approached using extremum-seeking. In this work, an extremum-seeking controller has been proposed to regulate the NH_3 flow to the SCR system while a PID controller has been proposed to regulate the APH bypass damper opening. Note that an alternative approach would be to control both the SCR and the APH systems by extremum seeking. Along this line, the extremum seeking real-time optimizers could also control the set-points for both $NO_{x_{CEM}}$ and ABS deposition depth instead of directly controlling the NH_3 flow and the APH bypass damper opening. Conventional PID loops would be indeed in charge of controlling these variables based on the set-points received by the extremum seeking controllers.

In order to illustrate in simulations the effectiveness of the proposed control scheme in optimally solving performance tradeoffs in real time, simple dynamic models for the boiler, SCR system and APH system were identified using experimental data from Cayuga Unit 1. Although more complex models could have been derived, simplicity was preferred in this feasibility study since the models were developed with the unique goal of carrying out simulations and were not used for control design (controller not based on model).
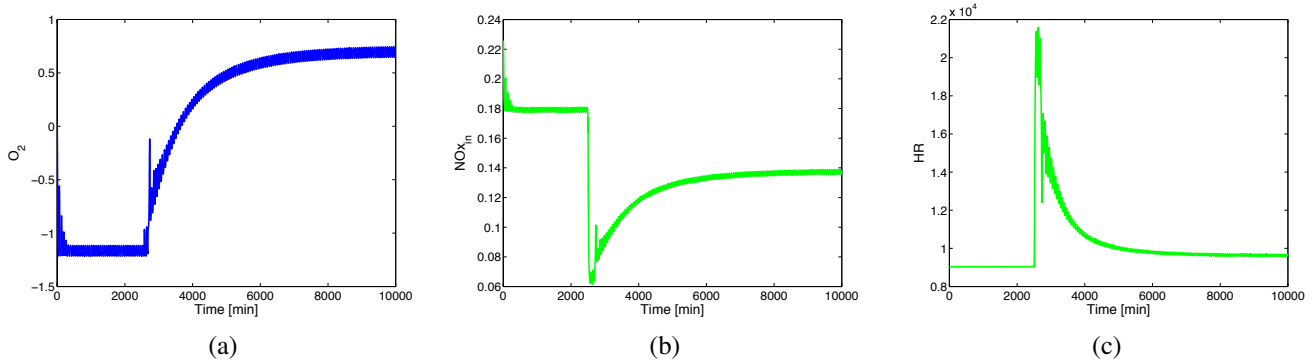
Fig. 8. Boiler operation optimization via extremum seeking. Note that the excess O_2 plot denotes corrections of the nominal value around which the linearized boiler model has been identified.
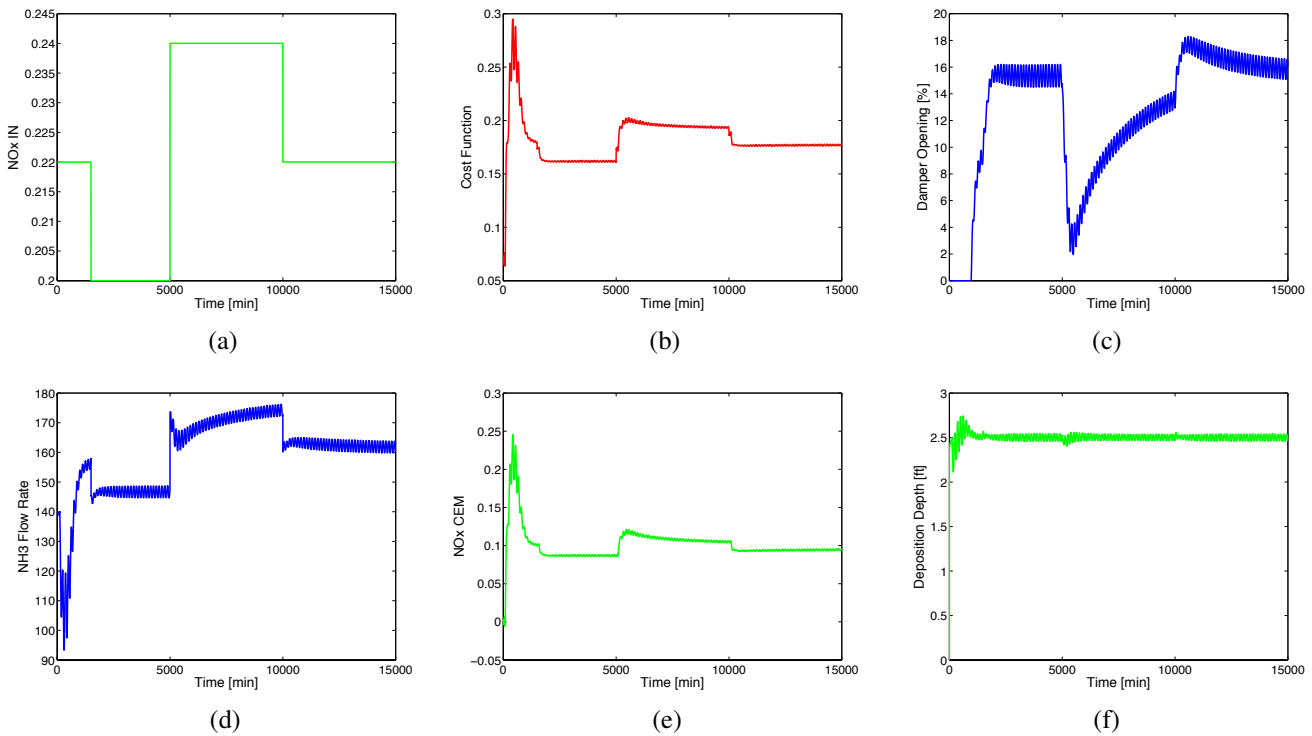

Fig. 9. SCR/APH operation optimization via extremum seeking.

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