

## ***Executive Summary***

### **The Role of CFD in Process Development, Design and Optimization**

*Written by Scott Miller & Claire Schmit, AECOM Process Technologies*

Most coal-fired electric generating units (EGU) today are operating in unfamiliar territory. With low-cost natural gas and increased generation from renewable energy, utilities are constantly challenged to respond to the market through leaner, more flexible operation while simultaneously maintaining regulatory compliance. This article summarizes how Computational Fluid Dynamics (CFD) has become an indispensable and cost effective engineering tool for optimizing the performance and economics of coal plants to better respond to environmental and market pressures.

***Full Story....***

### **The Real-Time Measurement of Sulfur Bearing Vaporous Compounds (“SO<sub>3</sub>”, sulfuric acid, ammonium bisulfate) and Its Application for Use for Process and Environmental Control**

*Written by Daniel T. Menniti; Breen Energy Solutions*

In the hydrated lime injection process, the measurement of “SO<sub>3</sub>” is a key parameter to allow the plant operations to control the Hydrate injection rate. In addition to the measurement of “SO<sub>3</sub>” it is important to know the impact of the concentrations of any vaporous compound within the plant ducts or equipment. One can utilize this in addition to the “SO<sub>3</sub>” value to understand the transition of species of the vaporous compounds and employ this knowledge to establish true Minimum Operating Temperature (MOT) of an SCR system, mitigate air heater fouling and control mercury and/or visible emissions (blue plume). This article outlines the application of the SO<sub>3</sub> measurement technology application in various utilities.

***Full Story....***

### **Maintaining and Operating an Electrostatic Precipitator in a world of cycling coal assets**

*Written by Jesse D. Price, Southern Environmental Inc.*

Traditionally, coal-fired electric generating units (EGUs) and their associated air pollution controls have been designed to be base-loaded, operating at steady-state full-load conditions for a period of time lasting up to a year or more. In today’s generation mix, however, these coal-fired EGUs are often used as peaking units, cycling on and off at a much greater frequency than designed. This places stresses and strains on the entire boiler plant, including the air pollution control equipment. This is especially true of the particulate emissions control devices, namely the dry electrostatic precipitator (ESP) or the fabric filter (FF), which are the focus of this article.

***Full Story....***

### **EPA Re-Thinks OTM-37 for Condensable Particulate**

*Written by Scott Evans, Clean Air Engineering*

R&D testing has already been done with OTM-37 method. Time after time in paired tests against Method 202, OTM-37 has shown lower results, sometimes significantly lower. So why aren’t we all using OTM-37 at this point? This article will discuss exactly what both methods are and what the future holds regarding these test methods.

***Full Story....***

**[Continued on page 2](#)**

# WPCA Worldwide Pollution Control Association www.wpca.info NEWS

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## Evaluation of SCR Operation for a Gas Conversion Project

*Written by Rod Beittel, Babcock Power*

Many owners of coal-fired boilers are opting to convert them to co-fire gas, or to fire gas exclusively. Fuel conversion projects require detailed consideration of the fuel influence on steam generator performance, as well as on “back end” equipment including the ID fans, air heaters, and emission control equipment.

*Full Story....*

## The Role of CFD in Process Development, Design and Optimization

Written by Scott Miller & Claire Schmit, AECOM Process Technologies

### ABSTRACT

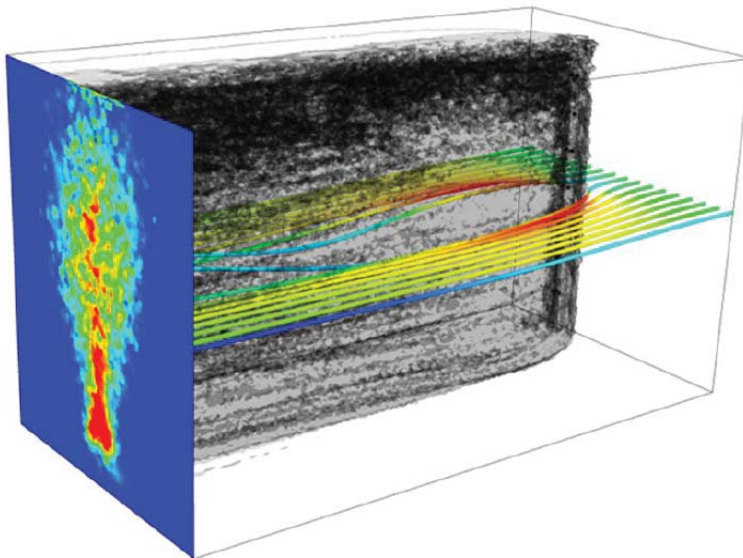
Most coal-fired electric generating units (EGU) today are operating in unfamiliar territory. With low-cost natural gas and increased generation from renewable energy, utilities are constantly challenged to respond to the market through leaner, more flexible operation while simultaneously maintaining regulatory compliance. This article summarizes how Computational Fluid Dynamics (CFD) has become an indispensable and cost effective engineering tool for optimizing the performance and economics of coal plants to better respond to environmental and market pressures.

### INTRODUCTION

CFD is a valuable engineering tool used to develop, evaluate, optimize and design fluid flow processes by

simulating the complex behavior and interaction of liquids and gases. Once predominantly used in the design of large, intricate processes, CFD is now helping utilities and independent power producers consider various operating conditions and design refinements that allow them to quickly respond to market demands.

For 25 years, AECOM Process Technologies has used CFD to develop, refine and implement pollution control processes for coal-fired power plants. Initially focused on air pollution control, CFD has been and remains vital to the development, design, troubleshooting and upgrade of processes that control acid gases, heavy metals (such as mercury), particulate matter and nitrogen oxide emissions. Today, we also use CFD to optimize the performance and economics of these same processes, helping plants remain competitive and agile in their response to current market demands.



*Figure 1: Development of a dispersion lance for improved performance of dry sorbent injection technologies (particle trajectories and concentration slice plane shown).*

### DEVELOPMENT

CFD is used to advance the development of technology that addresses both established and emerging market needs. It has been instrumental in progressing research and development (R&D) programs at a pace not thought possible by eliminating a significant amount of trial-and-error experimentation; hence, streamlining the transition from conceptual design to commercial implementation. For example, CFD was used to develop technologies to control emissions from fossil fuel combustion gas, including innovative NO<sub>x</sub> control and flue gas desulfurization (FGD) systems, technologies for sulfur trioxide control, and numerous sorbent- and catalyst-based mercury control technologies. For all of these technologies, CFD analysis was performed to determine optimal equipment size, configura-

tion, and operating conditions to achieve the desired system performance.

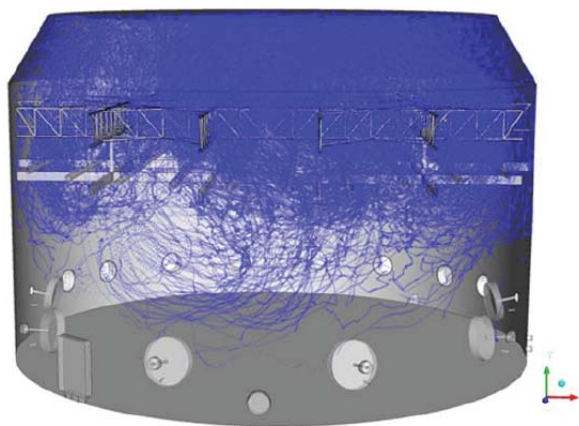
More specifically, CFD used as an engineering tool can help ensure system designs provide necessary contact between reactants and/or catalysts while addressing any associated pressure and temperature requirements. The analysis minimizes the potential for design flaws that negatively impact performance or contribute to the misinterpretation of results. Efficiently advancing design concepts while minimizing the potential for rework is especially important given budget and schedule constraints of most R&D projects.

## DESIGN

For pollution control processes and equipment beyond the development stage, CFD can be used in the design phase to maximize performance and reliability. It has been used to design new and/or upgrade modifications to:

- Particulate control devices;
- Multi-phase absorbers for the control of acid gas emissions;
- Fixed-structure reactors for the oxidation or adsorption of nitrogen oxides and heavy metals;
- Sorbent injection processes (including the injection lances) for the in-line removal of nitrogen oxides, heavy metals and acid gases;
- Oxidation air systems for in-situ forced oxidation of calcium sulfite to gypsum; and
- Associated flow straightening and mixing devices.

For preliminary design, CFD is indispensable in refining aspects of a process or piece of equipment to achieve an acceptable balance of performance and cost. By engaging with



*Figure 2: Design of an in-situ oxidation air system for the production of gypsum in a wet FGD process (air bubble trajectories shown).*

other engineering disciplines, fabricators, and construction personnel early in the process, CFD engineers address factors such as constructability and maintainability to deliver designs that conform to client and project objectives.

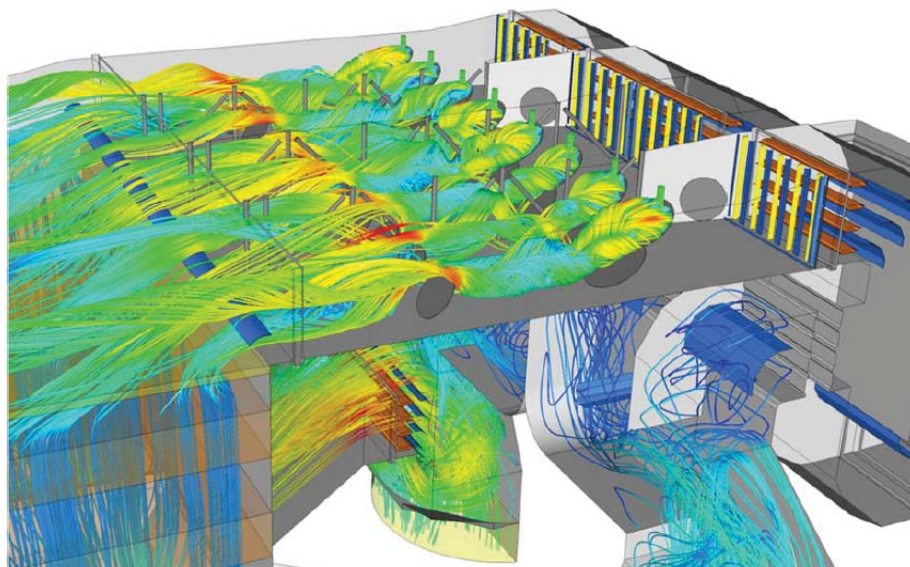
For detailed design, engineers and designers use CFD results to aid in the development of the engineering deliverables required by fabricators and construction contractors for final implementation. For example, CFD derived-data (e.g. temperatures, operating pressures, loads, etc.) are often used in the specification of materials and design of support structures. In addition, the final design's impact on performance can also be assessed, and if necessary, addressed through additional CFD runs and design iterations.

## EVALUATION

When a process fails to meet performance requirements or contributes to operations and maintenance (O&M) challenges that add cost or decrease reliability, CFD proves invaluable in root cause diagnoses. CFD can efficiently and effectively evaluate process performance over a range of operating conditions and configurations without many of the constraints associated with costly and time consuming physical testing and sampling programs. In particular, CFD can assess conditions that cannot easily be evaluated otherwise, such as expected future operating conditions after a system upgrade or retrofit.

Because coal-fired power plants have operated for years as base-load units, their design and that of their pollution control processes was often limited to a subset of operating conditions. However, coal plant operations have transitioned to conform to the current power market. Many facilities are swinging loads to respond to demand and operating at reduced generating outputs that were once rare. As a result, inadequacies in the design of pollution control processes have caused reductions in both system efficiency and reliability. Symptoms of the problem can often be identified through analytics or visual observations, but the cause cannot. CFD offers an effective way to assess operational impacts by evaluating process or equipment performance at varying conditions.

There are numerous examples of CFD use in performance assessments, including evaluating performance when operating at reduced capacity, after a change to the number of in-service parallel systems, or after changes to the influent conditions from a turbine upgrade, fuel switch or other modification.



*Figure 3: Use of CFD to assess pre-SCR NH3 injection and distribution considering elimination of SCR bypass (NH3 streamlines shown).*

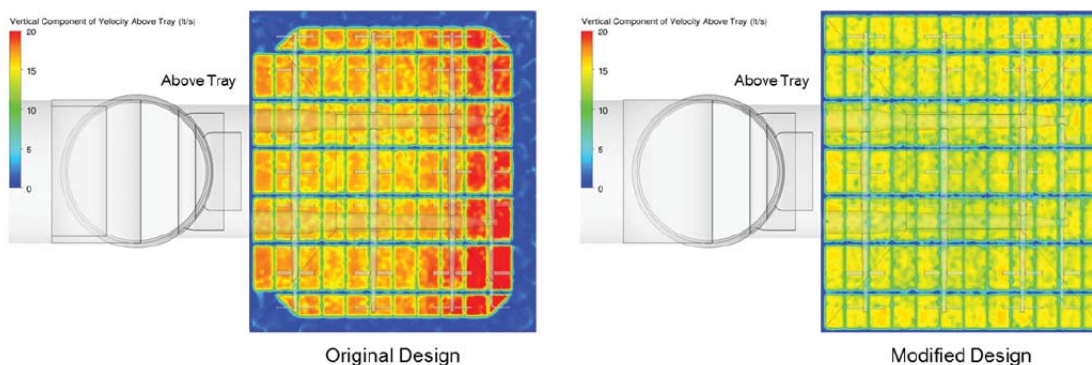
**OPTIMIZATION**

Evaluating the operation of a process or piece of equipment is the first step in optimizing performance; however, the true value often comes from determining changes to design or operation that would minimize (if not negate) the detrimental impact of identified issues. In other words, the merit of a CFD analysis is the ability to find a solution to resolve an identified issue. CFD can systematically evaluate numerous iterations of potential process modifications to arrive at the most flexible and economically justifiable solution.

In the power sector, CFD has been an integral component in the upgrade and retrofit of air pollution control technologies. It has helped ensure continued compliance with increasingly stringent regulatory limits by contributing to the improved performance of existing processes while minimizing the associated capital investment. For example, CFD was used to

optimize numerous FGD processes to achieve greater levels of sulfur dioxide removal, improved reliability, and improved economics. This was done through:

- Modification or implementation of flow straighteners to reduce recirculation zones, improve gas velocity and temperature profiles, and minimize pressure drop and/or spray droplet carryover potential;
- Modification or implementation of spray headers, absorber sieve trays and wall rings to improve gas-liquid contact; and
- Changes to reaction tank air sparger design and placement to improve sulfite oxidation without contributing to air-induced pump cavitation, compromising the ability to maintain solids suspension, or increasing the potential for higher mercury re-emissions.



*Figure 4: Optimization of wet FGD sieve tray design to improve flow profile and gas/liquid contact (velocity profile just above tray shown).*

CFD has also optimized the performance of other air pollution control processes, including selective catalytic reduction systems for nitrogen oxide control, sorbent injection processes for acid gas and mercury control, and particulate control devices, to name a few.

**SUMMARY**

CFD offers an efficient means of adeptly addressing the pressing needs of the modern-day coal-fired power plant. Through skilled engineering and the use of state-of-the art hardware and software, CFD advances R&D, ensures a process is properly designed, diagnoses known problems, resolves operating issues, and optimizes performance.

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Claire.schmit@aecom.com*

**BIOGRAPHY**



Scott Miller is a Professional Engineer who utilizes over 20 years of process engineering, project execution, and pollution control experience to bring innovative and cost competitive solutions to the marketplace. Skilled in research and development, technology commercialization and capital project execution, Scott has helped deliver

innovative solutions to the coal-fired power sector to mitigate hazardous air pollutant emissions. Scott is focused on helping coal plants retain environmental compliance while reducing capital investments and operating costs to improve their competitiveness with alternate sources of power.



Claire Schmit is a Principal Process Engineer for AECOM Process Technologies with nearly 15 years' experience in air pollution control and wastewater treatment particularly in the coal fired power industry. She has executed a number of capital projects associated with upgrade and process improvement

projects. She has extensive experience with process analysis and design, process troubleshooting and optimization, project scope development, budget and schedule development, and engineering design coordination.



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# The Real-Time Measurement of Sulfur Bearing Vaporous Compounds (“SO<sub>3</sub>”, sulfuric acid, ammonium bisulfate) and Its Application for Use for Process and Environmental Control

Written by Daniel T. Menniti; Breen Energy Solutions

## ABSTRACT

In the hydrated lime injection process, the measurement of “SO<sub>3</sub>” is a key parameter to quantify to allow the plant operations to control the Hydrate injection rate. In addition to the measurement of “SO<sub>3</sub>” it is important to know and understand the impact of the concentrations of any vaporous compound within the plant ducts or equipment. The net value is a result of two main components:

- Steady state real-time formation of the vaporous compound from combustion or other plant processes and,
- Release or storage of the compound vapors from surface areas due to increasing or decreasing temperature transients, respectively.

Understanding this, one can utilize a measurement of such in addition to the “SO<sub>3</sub>” value to understand the transition of species of the vaporous compounds and employ this knowledge to establish true Minimum Operating Temperature (MOT) of an SCR system, mitigate air heater fouling and control mercury and/or visible emissions (blue plume). This article outlines the application of the measurement technology application in various utilities.

## INTRODUCTION

Within multiple industries, there are fossil fuel fired boilers and process heaters that are used for power and steam generation, as well as process heating. Within these generating units there are multiple process locations where the level of “SO<sub>3</sub>” and flue gas dewpoint is of interest. With the proliferation of SCR’s and Dry Sorbent Injection Systems it has become more critical than ever to be able to measure and control the various forms of “SO<sub>3</sub>”.

The term “SO<sub>3</sub>” is used loosely across the

power industry and it typically is referring to the resultant species created when SO<sub>3</sub> combines with water and other chemical compounds, typically ammonia (NH<sub>3</sub>) or sodium (Na). Of particular interest within the industry is ammonium bisulfate and sulfuric acid. When sulfur bearing fuel is combusted, these condensable species that are found in the combustion process are

- **SO<sub>3</sub> - Gas**

SO<sub>3</sub> Gas is non-condensable and exists in the process as a gas at various percentages from - 800°F to 387°F

- **H<sub>2</sub>SO<sub>4</sub> Vapor**

Sulfuric Acid Vapor, in the absence of other sulfate species will exist in the combustion gas at varying percentages of conversion, beginning at 800°F until “all” of the SO<sub>3</sub> is converted to H<sub>2</sub>SO<sub>4</sub> by 387°F. Figure 5 shows the conversion ratio of SO<sub>3</sub>/H<sub>2</sub>SO<sub>4</sub>.

For post combustion NO<sub>x</sub> removal, ammonia is introduced into the flue gas and the excess ammonia combines with H<sub>2</sub>SO<sub>4</sub> combines to form ammonium bisulfate. Ammonium bisulfate typically condenses (forms) and evaporates be-

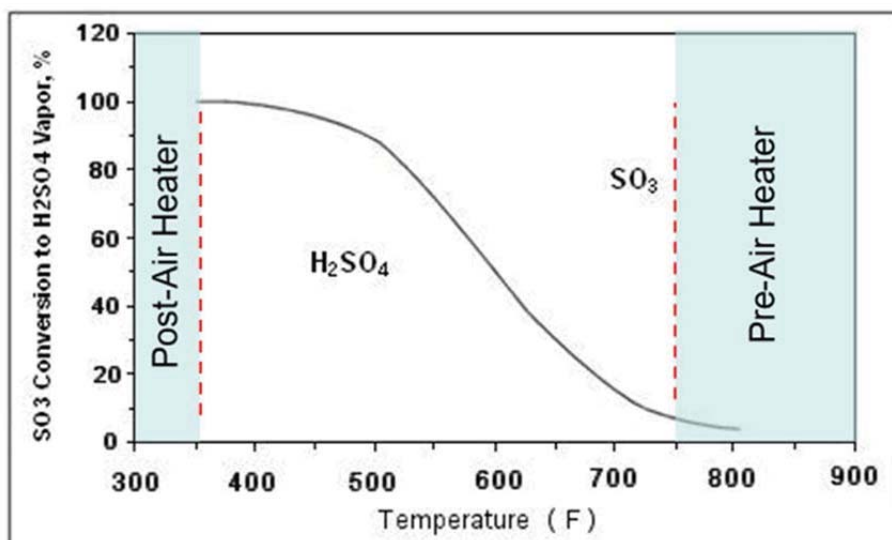


Figure 5: SO<sub>3</sub>/H<sub>2</sub>SO<sub>4</sub> Vs Flue Gas Temperature

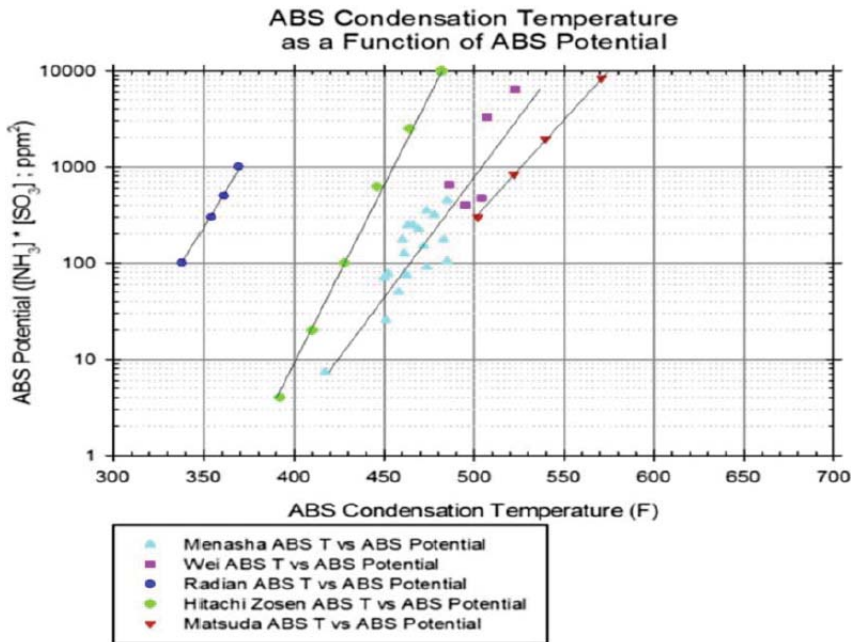


Figure 6: Summary from previous literature on ABS formation as a function of temperature<sup>1</sup>

tween 325°F to 550°F (this general temperature range does not take into consideration the capillary dewpoint temperature within the catalyst pores). Figure 6 shows the summary of ABS dew points from previous literature.

**FLUE GAS SULFUR CONDENSABLE MEASUREMENT**

The sensor tip has a glass surface with a thermocouple and a circular electrode embedded in the center of the glass. Compressed air is directed from the flanged end of the probe, through an internal tube to cool the sensor tip in a controlled manner (Figure 7).

Referencing Figure 8, the upper portion presents two typical measurement cycles. As the probe is cooled, conductivity of the liquid material condensing on the probe cell is detected as current (green line). As the inset shows, when current is

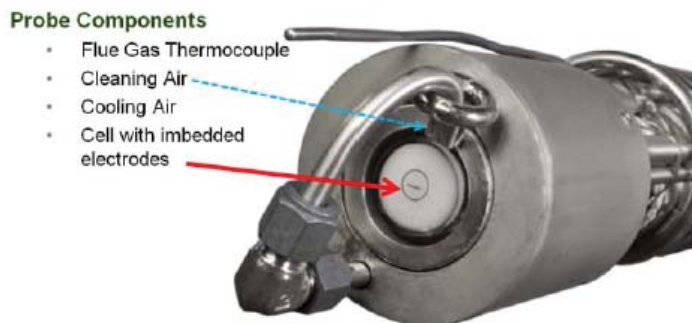


Figure 7: Sulfur Condensable Measurement Sensor

first detected, the probe temperature is recorded as the Formation Temperature.

The probe logic will then reverse the cooling process and begin controlled heating of the probe. During this period, probe current may continue to climb indicating condensable material continues to form on the probe cell faster than it is being evaporated. Conversely, on the back side of the current wave form, the current is decreasing indicating condensable material is evaporating faster than it is condensing.

The point at which the current peaks is the equilibrium point where the condensation and evaporation rates are the same. This is the definition of equilibrium and defines the Dew Point Temperature. When the current returns to the original value near the end of its cycle, the probe temperature is recorded as the Evaporation Temperature.

The Formation and Evaporation temperatures are used to identify the species of condensable material. The Formation Temperature reflects the temperature where condensation is greater than evaporation and hence allows buildup of the condensable on process surfaces. Conversely, the Evaporation temperature reflects the temperatures where evaporation is greater than condensation and is the temperature at which the condensable will self evaporate.

Specifically, this paper will discuss the measurement of Sulfuric Acid and Ammonium bisulfate species and the use of such to optimize balance of plant operations:

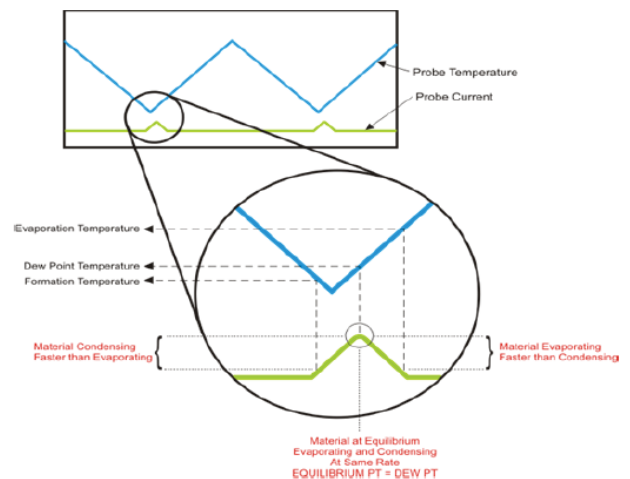


Figure 8: Measurement Cycle

1. In units with SCR  $\text{NO}_x$  reduction, ammonia combines with the free  $\text{SO}_3$  and water to form ammonium bisulfate (ABS). Knowing the temperature at which the vaporous condensable will form on a surface, is critical to understanding the safe operation of an SCR below its minimum operating temperature (MOT).

In addition to the MOT question, the knowledge of this formation temperature also provides information that allows the plant operator to control, or mitigate, the levels of ABS formation to avoid deposition on air heater tubes and baskets, where it can cause plugging.

2. In addition to cold end corrosion, the level of sulfur condensables downstream of the air preheater affects the efficiency of sorbents used to remove mercury. A condensables measurement is essential to control the mitigation of  $\text{SO}_3/\text{H}_2\text{SO}_4$  in this location. Once the condensable levels are known it can be monitored to reduce these sulfur-based species prior to an Activated Carbon Injection (ACI), thus enhancing the capture of mercury by the Powder Activated Carbon (PAC).
3. In processes with a Wet Flue Gas Desulfurization, Sulfuric Acid Mist (" $\text{SO}_3$ ") levels above  $>5\text{ppm}$  have been known to form "blue plumes". There are well known algorithms to convert Sulfuric Acid Dewpoint Temperature to  $\text{SO}_3$  or  $\text{H}_2\text{SO}_4$  ppm, thus allowing for the control and mitigation of the plume acid levels to  $<5\text{ppm}$ .

In the interest of brevity, the following presents examples of the use of the measurement for the control of the condensable species ahead of the air heater and its use for controlling  $\text{SO}_3$  mitigation with Hydrate to optimize PAC injection.

### SCR REAL MINIMUM OPERATING TEMPERATURE

Typically, a generating unit SCR has a unique Minimum Operating Temperature (MOT) supplied by the SCR or catalyst supplier. If operating below the MOT, the operator risks irreversibly fouling or plugging the catalyst pores with ammonium bisulfate (ABS). With the onset of load cycling on large coal fired units, the industry has been driven to find a way by which the real MOT can be verified, thus allowing the plant to run below the manufacturers MOT over short periods of time, and safely burn off the condensed ABS, to protect the catalyst from plugging.

To establish the MOT of an SCR system utilizing a measurement system, the sensors should be installed at the inlet and outlet ducts of each SCR reactor. Each test run begins after the unit has been above a load that is consistent with the established MOT + 50 degrees for a minimum of 3 hours. The test run begins with a load reduction to a target load while maintaining a prescribed ramp rate. The unit will be maintained at the target load for a prescribed minimum duration. After the target load operating period, SCR ammonia will be taken out of service prior to the increasing load. After the SCR is taken out of service, load is increased, again at a prescribed rate, until the unit is again above the load consistent with established MOT + 50 degrees. The test run concludes when load has been maintained above this value for a minimum of 3 hours.

On the load increase, probe data will reveal whether true MOT had been attained during the load reduction. Attaining the true MOT will be evident by an observation that the outlet duct condensable levels during and following the load ramp will be aggravated by the release of stored ABS on the catalyst. Further, once the ABS formed and stored on the catalyst is depleted, conditions at the outlet duct will diminish to the steady state real-time values. If not, the true MOT had not been attained during the load reduction and a new reduced target load should be selected for the next test run. This process is repeated until the true MOT is revealed.

The following graph is taken from a paper presented by Haldor Topsoe on the subject of catalyst ABS formation below MOT. In this case study the plant uses a MOT setting of  $320\text{C}$  for normal operation of the SCR. Figure 9 shows the load duration at temperatures below MOT and shows the theoretical (red trace) and Breen condensables probe (blue trace) acid release on load ramp.

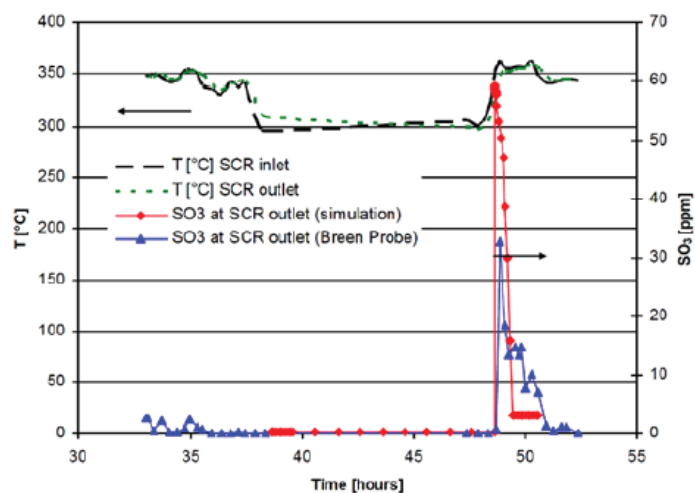


Figure 9: Haldor Topsoe ABS Model<sup>2</sup>

If the SCR had been operated at a load consistent with SCR inlet temperatures above 320C, the blue trace would have been much smaller in magnitude.

As reference in the protocol above, it is important to maintain consistency in unit load and duration during low temperature operation, mill Configuration, fuel sulfur content and unit ramp rate from low load to high load operation. A flow chart for the Test Run sequence is illustrated below:

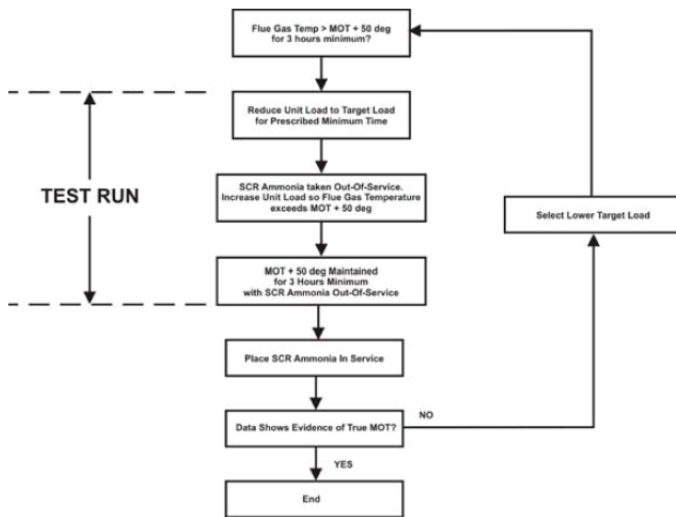


Figure 10: SCR real MOT determination protocol

In practice the further below MOT the unit operates and the longer it operates at this condition both contribute to the magnitude of the measured release spike on the measurement probe. Additional data gathered from such operation is seen in Figure 11.

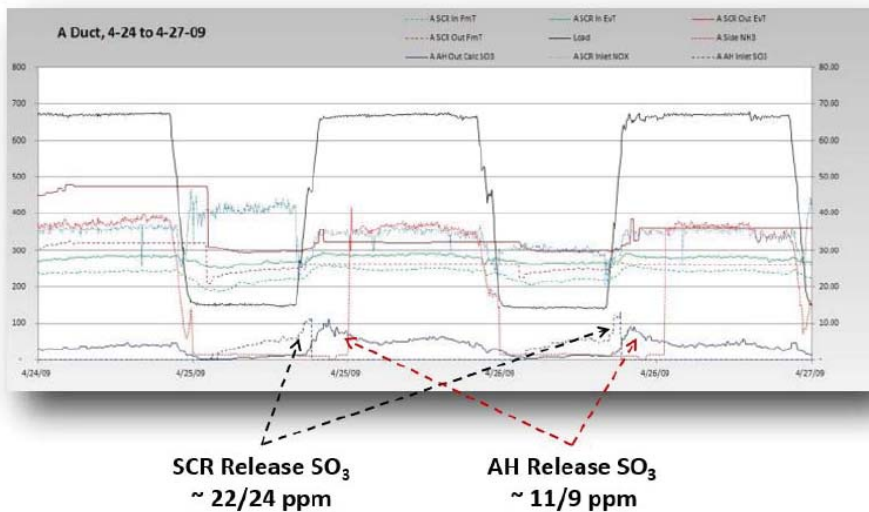


Figure 11: Volitization of sulfur condensables from low load operation

REAL TIME USE FOR CONTROL OF CONDENSABLE SPECIES

To mitigate air heater fouling, the transition of condensable species from a fouling condensable (ABS) to a less severe condensable (Sulfuric Acid) is one of the main objectives for upstream hydrate injection. The lack of sufficient sorbent will result in fouling material forming on the upper levels of the air heater, requiring the operator to raise the air heater metal temperatures through the use of steam coils. Lacking steam coils, this fouling would result in increased differential pressures forcing a unit de-rate or outage. By using a condensables measurement system, the plant will know the level and species of condensable and is able to optimize its sorbent injection rate to mitigate the fouling material.

Figures 12 and 13 on page 9 were generated at a plant burning Illinois basin coal. Baseline condensable species were observed forming and evaporating in the ABS range when the unit was at high load. The plant’s hydrated lime process, targeted at 500 lb/hr, was over-treating at low loads and under-treating at high loads. In an effort to minimize sorbent usage at low loads and to reduce the severity of deposits at high loads, the plant moved to a practice of 750 lb/hr at high load and 300 lb/hr at low load. The trend in figure 12 shows that these feed rates were not sufficiently optimized.

Optimally the formation and evaporation temperature of the detected condensable species should reside in the orange band associated with sulfuric acid. At low load the flue gas exhibits no detectable condensable material; at high load the condensable is firmly in the ABS range (nominally 330°F to approximately 500°F). Once the plant began varying feed rates they found high level ABS below 300 lb/hr of hydrated lime, mixed acid and ABS between 300 lb/hr and 900 lb/hr of hydrate, and low level acid above 1,100 lb/hr of hydrate (Figure 13).

By using the condensable measurement for feedback, the plant increased hydrated lime to reduce the air heater inlet SO<sub>3</sub>/ABS, increasing the full load hydrate injection rate by 350 lb/hr, to 1,100 lb/hr. They were also able to reduce the low load hydrate injection rate by 200 lb/hr, to 100 lb/hr. This mitigated the fouling of their air heaters, allowing for the elimination of the use of steam coils, resulting in an immediate cash flow improvement 2-3 times the cost of the increased sorbent usage.

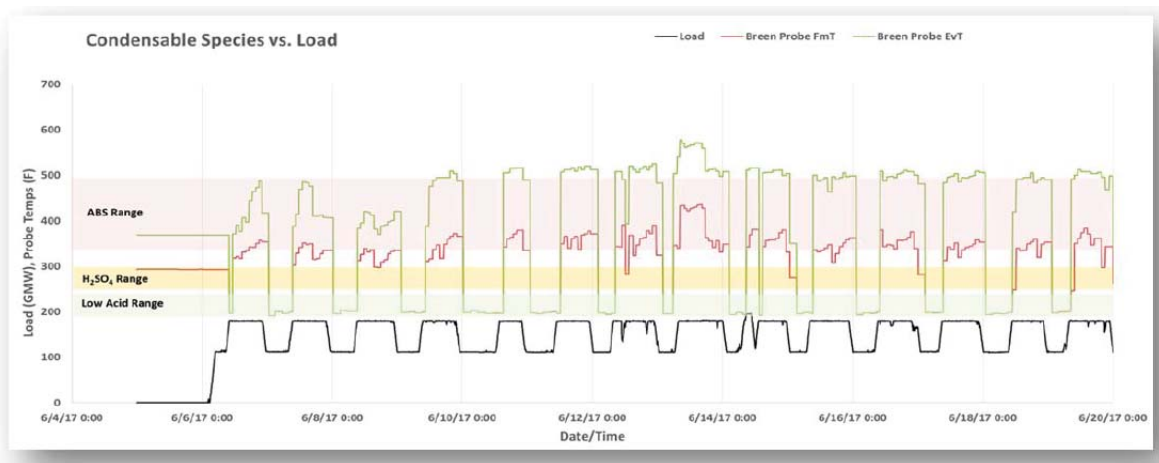


Figure 12: Baseline Condensables Species Trend

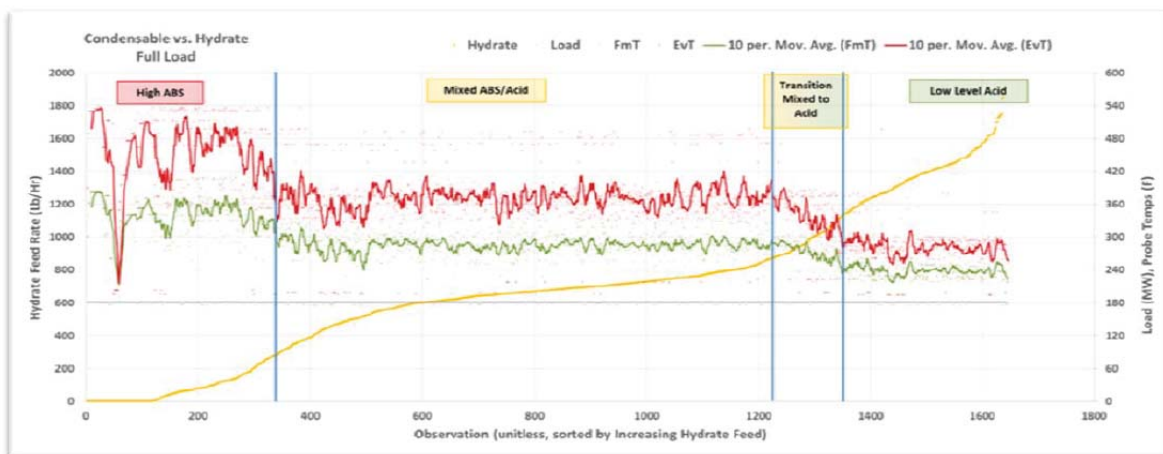


Figure 13: Transition of Species with Increasing Hydrate Injection Rates

**DUAL SORBENT OPTIMIZATION**

The presence of SO<sub>3</sub> upstream of and Activated Carbon Injection System can significantly reduce the effectiveness of the Powder Activated Carbon (PAC). Consequently, some plants over inject both Hydrate and/or PAC, to avoid exceeding their mercury emissions limit. The utilization of a condensables measurement is essential for the optimization and control of hydrate and PAC injection rates.

The condensable measurement system was installed on a boiler burning high sulfur coal. The Unit had been inject-

ing hydrate and PAC to meet their MATS compliance limits. The objective of the demonstration was to optimize the sorbents to meet the MATS compliance limit, while minimizing sorbent cost. Testing was performed at both high and low loads. The process flow one-line diagram can be seen in Figure 14.

After monitoring effectiveness of sorbents at baseline conditions, sorbent optimization was undertaken and at both load conditions the plant was able to reduce the usage of PAC with slight increases in the hydrated lime injection rates

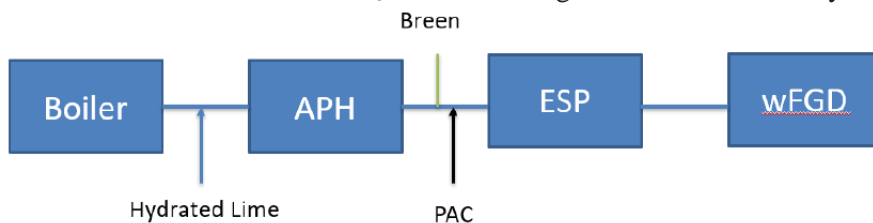


Figure 14: Combustion Gas Flow One Line Diagram

(Figure 15). During the demonstration, a Breen probe was used upstream of PAC injection. Output from the probe on acid condensable content provided guidance on hydrated lime feed rates necessary to reach low level SO<sub>3</sub> content before the PAC injection point. The probe was used to optimize feed rate of the hydrated lime so we could minimize PAC feed rates while maintaining acceptable Hg removal.

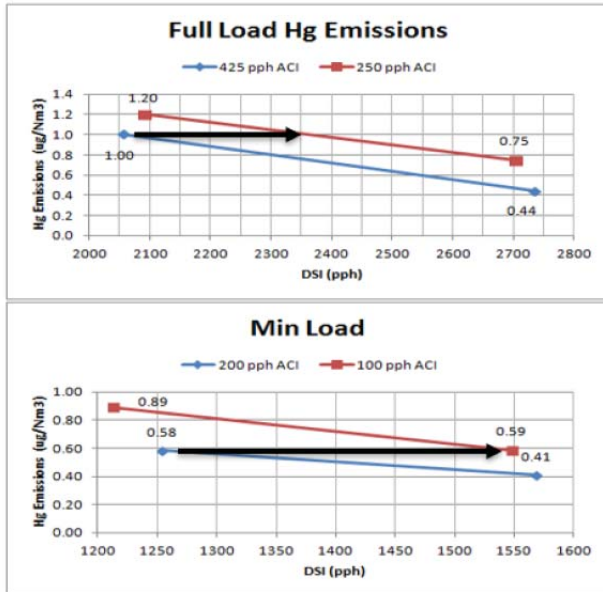


Figure 15: Sorbent Usage Vs Mercury Emissions at high and low load

The optimization of hydrated lime feed to minimize ACI while maintaining Hg emission targets, results in potential savings to plant operations by approximately \$700,000 annually (Figure 16).

**SUMMARY AND FURTHER WORK**

Over the past year there has been a progression in the industry to better understand the real MOT of an SCR, reduce air heater outlet temperatures by removing steam coil usage and optimize sorbent usage and cost through measurement and control. There is also recognition that understanding the sorbent effectiveness and distribution, with the reliable and repeatable capability to measure various sulfur condensable species, is critical to this effort. There is a significant amount

Load	ACI lbs/hr	Hydrated Lime lbs/hr	Total Sorbent Usage Savings (\$)*
High	425	2,060	---
High	250	2,350	\$470,000
Low	200	1,250	---
Low	100	1,550	\$230,000

\*Assumptions: 70% serv factor, 50/50 split high and low load. PAC @ \$1/lb

Figure 16: Total Sorbent Cost Savings after Dual Sorbent Optimization

of work being done with regards to these efforts and we expect to be able to report the results over the coming year.

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2. Jones, Wayne S.; Haldor Topsoe; Reinhold Conference, "ABS from Sub MOT Operation", 2-16-13 pp. 31

For further information contact Daniel Menniti at dmenniti@breenes.com

**BIOGRAPHY**



Daniel Menniti is currently The Senior Director of Sales and Marketing for Breen, a Mississippi Lime FGT Business Unit Company. He is responsible for the Business Development and Application of the Breen Probe Measurement systems for ABS/SO<sub>3</sub>/Acid Dewpoint. Over the last 20 years, Daniel has experience with the application of analytical technologies in combustion, environmental and process control in the Power and Industrial Markets. Most recently, working closely with the US Coal Fired Utilities to measure and optimize the balance of plant effects of sulfur based condensables and hydrated lime injection.

Prior to Joining Breen, Daniel earned his BS in Chemical Engineering from Penn State University and an MBA from Farleigh Dickinson University. He also worked with the NJDEP and held various positions in analytical instrumentation companies, including Global Product Management, VP of International Sales and Director of Customer Service.



# Maintaining and Operating an Electrostatic Precipitator in a World of Cycling Coal Assets

*Written by Jesse D. Price, Southern Environmental Inc.*

## ABSTRACT

Traditionally, coal-fired electric generating units (EGUs) and their associated air pollution controls have been designed to be base-loaded, operating at steady-state full-load conditions for a period of time lasting up to a year or more. In today's generation mix, however, these coal-fired EGUs are often used as peaking units, cycling on and off at a much greater frequency than designed. This places stresses and strains on the entire boiler plant, including the air pollution control equipment. This is especially true of the particulate emissions control devices, namely the dry electrostatic precipitator (ESP) or the fabric filter (FF), which are the focus of this article.

## ARTICLE

A coal-fired EGU dispatched as a peaking unit, or any other non-base-loaded utilization, may not necessarily be dispatched at full load. This can result in the process operating conditions of the flue gas entering the ESP or FF deviating significantly from the design process operating conditions, leading to performance and maintenance issues. While beyond the scope of this article, it should be noted that a complete study of the boiler, its upper and lower capacity limits, and the resulting process conditions at each piece of downstream equipment, as well as the effect of the new process conditions on each piece of downstream equipment, should be undertaken before a base-loaded coal-fired EGU boiler plant is dispatched in any other utilization than base-loaded.

Because coal-fired boilers take a lot of time to startup and shutdown, dispatching a coal-fired boiler as a peaking unit requires advance notice. To shorten the time required to bring an EGU on line, some operators idle the boiler by firing with gas at a heat input high enough to maintain a minimum output of steam from the boiler. Whether the boiler is started from a cold start, or from a warm start, the ESP or FF will experience up and down temperature cycles as the EGU is brought on-line or taken off-line.

As the EGU boiler heats up or cools down as it cycles, the temperature of the flue gas flowing through the ESP or fabric filter rises or falls accordingly. As the temperature of the flue gas entering the ESP or FF rises and falls during these boiler

cycles, it passes through two (2) separate dewpoint temperatures. The higher temperature dewpoint, the acid dewpoint, is the temperature at which the sulfuric acid vapor ( $H_2SO_4$ ) in the flue gas becomes saturated and condenses as liquid sulfuric acid on the interior surfaces of the ESP or FF. Sulfuric acid vapor is formed in the flue gas at temperatures below 1000 degF (540 degC) as the sulfur trioxide ( $SO_3$ ) present in the flue gas reacts with the water vapor present in the flue gas. The lower temperature dewpoint is the temperature at which the moisture present in the flue gas becomes saturated and condenses. Both of these dewpoint temperatures are highly dependent on the chemistry and pressure of the flue gas. The chemistry of the flue gas is directly related to the chemistry of the fuel and the inherent moisture in the combustion air. Since all coals commercially suitable for power generation have some sulfur in their chemical makeup, there will always be a small amount of  $SO_3$ , and thus  $H_2SO_4$ , in the flue gas, unless removed upstream of the ESP or FF by injection of a wet or dry chemical reagent or sorbent such as lime or trona. Typical acid dewpoint temperatures range from 250 degF (120 degC) for low sulfur coals to 300 degF (150 degC) for higher sulfur coals. The dewpoint temperature for the flue gas moisture can range from 95 to 130 degF (35 to 55 degC), depending on flue gas moisture concentration and pressure.

The number one maintenance issue in a modern ESP is corrosion of the carbon steel internal components and casing. Corrosion is also a major contributor to maintenance issues in a FF. The majority of this corrosion is a result of sulfuric acid condensation. However, condensation of the flue gas moisture can also be a significant source of corrosion in the ESP or FF. When under operation, the oxygen ( $O_2$ ) concentration of the flue gas passing through an ESP or FF is typically less than 8% on a wet basis. When the ESP or FF is not operating and has cooled down to ambient temperatures, the oxygen concentration can be much higher, contributing to further corrosion, since the oxidation of steel occurs faster when higher concentrations of oxygen are present along with moisture. One means of combating this corrosion is to draw a sufficient volume of air, heated by gas, through the ESP or FF to dry it out before it is allowed to cool down to ambient temperatures. If the EGU is dispatched on a frequent enough

basis, it may be prudent to keep the ESP or FF in a warmed-up state to prevent corrosion.

While it has been mentioned in this article that sulfuric acid condensation occurs when the boiler is cycling on and off, anytime the flue gas temperature drops below the sulfuric acid dewpoint temperature, sulfuric acid will condense and will cause corrosion in the ESP or FF. The three (3) most significant reasons for sulfuric acid condensation in an ESP or FF while it is operating are as follows: first, air in-leakage causing localized cooling of the flue gas; second, the flue gas entering the ESP is at or below the acid dewpoint temperature; and third, cold spots in the casing due to inadequate or damaged insulation. Most coal-fired EGUs are designed with an induced draft (ID) fan downstream of the particulate collection device, so the negative pressure inside the ESP or FF draws ambient air into the ESP or FF through holes, cracks, or other openings in the ESP or FF casing. When there is air in-leakage into the ESP or FF, ambient air leaking into the casing mixes with and cools the flue gas below the acid dewpoint temperature in the volume around the leak. The in-leaking air also cools the ESP or FF casing and internal parts nearest the leak, as well, resulting in the more localized corrosion found near doors, hatches, and other casing penetrations. If the air in-leakage is sufficient, it can be observed as a temperature drop, typically greater than 15 degF, between the inlet and outlet of the ESP or FF. Corrosion due to in-leakage is an issue in ESPs and FFs under all types of operating regimes, but it becomes more of an issue when the boiler is cycling on and off. The corrosion due to air in-leakage can be reduced or even eliminated by proper maintenance of door and hatch gaskets, as well as packing glands for any rotating shaft-type penetrations. It is also imperative that all casing leaks are eliminated, making sure that all welds at the seams are properly seal welded to eliminate leaks. Cold spots due to poor insulation can be detected with thermography. Repair or replacement of the insulation and lagging of the proper design will eliminate cold spots.

If the flue gas entering the ESP or FF is at or just below the acid dewpoint temperature, sulfuric acid will condense on all surfaces, including non-metallic surfaces. This causes corrosion of all steel surfaces inside the ESP or FF that are exposed to the flue gas. Even if the average flue gas temperature is above the acid dewpoint, there may exist a phenomenon known as stratification, where the temperature of the flue gas exiting one side of the air pre-heater is significantly higher than the temperature of the flue gas exiting the other side of the air pre-heater. This flue gas travels down the duct without mixing, so the flue gas entering one side of the ESP

or FF can be above the acid dewpoint, while the flue gas entering the other side of the ESP or FF will be below the acid dewpoint. This is generally due to leaking seals in the air pre-heater or a rotation speed that is a little too high. Operation and maintenance of the air pre-heater is beyond the scope of this article, but operators of coal-fired EGUs need to remember that proper operation of a well-maintained air pre-heater is vital to the proper functioning of a cold-side ESP or FF, especially in the less than full load conditions that may exist when the EGU is dispatched as a peaking unit. To address temperature stratification, it may also be necessary to install a static mixing device in the duct between the air pre-heater and ESP or FF.

While corrosion affects the steel parts of the ESP, the ceramic high voltage insulators are also affected by moisture, particularly the support insulators. Both condensed sulfuric acid and flue gas moisture are conductive, leading to tracking and electrical failure of the insulator if this condensation occurs on the insulator itself. Most high voltage support insulators are either an open-ended cylinder or an open-ended truncated cone, with the high voltage ESP discharge electrode system suspended from a steel shaft running through the center of the support insulator. The high voltage support shaft is supported by a steel plate covering the top of the support insulator. The inside of the support insulator is exposed to the flue gas, along with the dust the ESP is designed to collect, while the outside of the support insulator is exposed to the atmosphere of the insulator compartment or penthouse. To protect these support insulators from condensation, several methods have been traditionally used. A fairly simple method is to place an electric resistance heating element, or “ring” heater, of approximately 1000 W around the outside of insulator on stand-offs a few inches above the base of the insulator. Another method is to blow hot purge air from either a steam heater or electric resistance heater into the insulator compartment. The heated purge air exits the insulator compartment through orifice ports in the steel plate on top of the insulator.

This purge air method has the advantage over the “ring heater” approach of also helping to keep the inside of the insulator clean and free of dust buildup, if properly designed. Physical flow modelling has determined the optimal number and geometry of the orifice ports, as well as the optimal purge air volume flow through the ports, in order to keep the inside of the insulator, which is exposed to the flue gas and dust, clean and free of dust and condensation.

The temperature of this purge air should be sufficiently high

to prevent condensation of the moisture in the purge air from forming on the outside of the insulator, and also high enough to keep from cooling the flue gas below the acid dewpoint temperature in the zone where the purge air mixes with the flue gas. This is the major cause of failure of corona shields, which, typically, are hyperbolic shaped steel rings extending down from the base of the support insulator, the purpose of which is to screen the surface of the hot roof from the high voltage support shaft.

Additionally, the insulators need to be protected from thermal shock. The rate of temperature rise, as well as the differential temperature from the outside wall of the support insulator to the inside wall of the support insulator must be kept within the manufacturer's specification to prevent premature failure of the support insulator. As the purge air system must be started two (2) to eight (8) hours prior to the energization of the ESP and passage of dust laden flue gas through the ESP, staging of the heaters to provide a constant rate of temperature rise is necessary to prevent thermal shock from a cold start-up.

As most ESPs on coal-fired EGUs operate at negative pressure, many purge air systems in existence have a blower used to provide the required volume of heated air to the insulators in pre-heat mode, but which is turned off when under operation, allowing the negative pressure in the ESP created by the ID fan to draw the required volume of air through the heaters and subsequently through the insulators. This may work well when the boiler is being operated at full load, however when the boiler is operating at less than full load, there may be insufficient purge air volume flow to keep the insulators clean and dry. In this case, it may be necessary to run the blower during all modes of operation, and it may also be necessary to change the blower fan to a type which can provide the required purge air flow volume under varying pressure conditions in the ESP.

Another important issue with purge air systems is flow distribution. Whether the support insulators are housed in individual insulator compartments or in a single penthouse insulator compartment covering the entire roof of the ESP chamber, it is important that each support insulator receive purge air at the same volume flow and temperature as every other support insulator served by that particular purge air system. Often, with a penthouse insulator compartment, the purge air is simply blown into the penthouse, with a single horizontal deflector to distribute the purge air throughout the penthouse. This may be adequate for an ESP operated in a base-loaded situation; however, this design does not ad-

equately distribute the purge air to each insulator. In a recent physical scale model air flow study (unpublished), it was shown that the maldistribution to each side of the penthouse with the horizontal deflector plate was over 70%. By using a symmetrically branching duct inside the penthouse to distribute the purge air, this side-to-side maldistribution was reduced to 13%.

Some existing purge air systems may be inadequately designed to provide the necessary volume of heated air to each insulator for all modes of operation. These systems will need to be replaced with purge air systems engineered to provide the proper volume of heated purge air to each support insulator for all modes of operation. It may also be necessary to run the purge air system at all times, including when the boiler is offline and no flue gas is passing through the ESP.

While FFs do not have support insulators and purge air, there are some important considerations for operating a FF in cycling mode. The first consideration is that it is important that the filter bags be adequately cleaned by "pulsing them down" prior to allowing the FF to cool down. If the filter bags are not adequately cleaned and sufficient sulfuric acid or moisture condenses on them, the dust on the filter bags and the condensation will combine to form a mud, blinding the filter bags. It is nearly impossible to "un-blind" the filter bags blinded this way without removing the bags from the FF.

Other considerations are that important to both FFs and ESPs are the hoppers. As with the filter bags, it is important to remove the collected dust from the hopper prior to condensation happening. Once again, the condensation and the dust in the hopper can combine to form a mud, which can turn to clinker if it is re-heated. Most FF and cold-side ESPs have some form of hopper heating. It is important that the hopper skin remain above the acid dewpoint at all times there is dust present in the hoppers. This can be accomplished by making sure that the hopper heating system meets or exceeds the Institute of Electrical and Electronic Engineers (IEEE) standard 1069. As with the purge air system, it may be necessary to operate the hopper heating system at all times, even when the boiler is off-line. Ash removal systems are beyond the scope of this article, but should also be evaluated for proper operation and maintenance under boiler cycling.

Operating the boiler at less than full load can be beneficial to the performance of the ESP or FF due to a lower flue gas volume flow, hence higher specific collection area (SCA) or lower air to cloth ratio (ACR). However, if the flue gas vol-

ume flow drops a certain percentage below the full load volume flow, it may be necessary to isolate an ESP chamber or a FF module in order keep the flue gas properly distributed through the ESP or FF. In the case of the ESP, depending on the chemistry of the flyash, the resistivity of the flyash can increase with changing temperature, leading to a degradation of collection efficiency.

Operation of a coal-fired EGU in a cycling mode versus a base-loaded mode requires a complete evaluation of the particulate collection system. Different methods and frequencies of start-up and shutdown of the EGU boiler will dictate whether some systems, such as the purge air and hopper heating systems, are operated full-time, even when the boiler has been allowed to cool down.

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### BIOGRAPHY



Jesse Price received his Bachelor of Science in Electrical Engineering from Mississippi State University in 1991, and worked in the architect/engineer and construction industries before joining Southern Environmental, Inc., in 1996 as an electrical design engineer. He has worked as an applications engineer in SEI's pre-contract department since 1997, designing solutions, estimating project costs, and writing proposals for numerous domestic and international air pollution control projects, mostly involving new and rebuilt dry ESPs, but also including fabric filters and novel concepts. Jesse is licensed as a professional engineer in the State of Florida.



## Who We Are

The Worldwide Pollution Control Association (WPCA) has assembled a group of people and companies who are experts at some aspect of pollution control. In addition, the WPCA has organized a user advisory board who can give this group direction and assistance in performing service to pollution control business throughout the world.

## Our Mission

The mission of the WPCA is to enhance technical communication through seminars, technical journals and a website. The WPCA is a non-profit organization and our members and advisors need to be motivated by a desire to see the pollution control community make world wide technical progress through improved technical communication.

## Who Directs the WPCA?

The WPCA is a partnership which includes system/equipment/services suppliers, consultants and users. The WPCA President, Vice President and Advisory Committee are equipment users. The Corporate Sponsors and Board of Directors are suppliers. Together they develop annual seminars and events to achieve their goal of better technical communication for users of air pollution control systems.

## How do I become a Member of the WPCA?

In order to be a WPCA member, you must be an end user of pollution control equipment. When you register on-line for any WPCA sponsored seminar, you automatically become a member. If you would like to join, but cannot attend a seminar at this time, please download and send in the Registration Form at the top of the members list at [www.wpca.info](http://www.wpca.info). You will then be emailed regarding upcoming events and sent future copies of the WPCA News.

## EPA Re-Thinks OTM-37 for Condensable Particulate

*Written by Scott Evans, Clean Air Engineering*

Particulate matter from industrial stacks comes in two varieties – filterable and condensable. Filterable Particulate Matter (FPM) is in a solid phase in the stack and is captured on a filter during testing – hence the name. Condensable Particulate Matter (CPM), on the other hand is gaseous in the stack. At first, this may seem counter-intuitive. How can gas be particulate? While gaseous in the stack, certain compounds will condense to solids when cooled to ambient temperature. This is CPM. So Total PM = FPM + CPM.

EPA's specific definition of CPM is...

*“Material that is vapor phase at stack conditions, but condenses and/or reacts upon cooling and dilution in the ambient air to form solid or liquid PM immediately after discharge from the stack. Note that all condensable PM is assumed to be in the PM<sub>2.5</sub> size fraction.” 40 CFR 51*

It has been well established that accurate measurement of CPM has long been problematic. Back in 1977, EPA's initial proposal for Method 5, the standard FPM test method, included a “back-half” for CPM measurement. There was so much controversy over including CPM, that it was dropped from the final method. Method 5 is still strictly for FPM. It took EPA almost 15 years, until 1991, to issue a CPM test method. This was known as Method 202.

In its simplest description, Method 202 sampling involved bubbling stack gas through cold, water-filled impingers. Anything that condensed out was considered CPM. This approach is problematic in several ways.

1. If the stack gas contains SO<sub>2</sub> or NO<sub>x</sub>, these compounds will oxidize in the cold water to form sulfates and nitrates,
2. If the stack gas contains any water-soluble organic compounds, those will condense out, and
3. If there is ammonia in the gas stream (from SCR ammonia slip for example), that will be easily absorbed in the cold, water-filled impingers. This makes them very efficient caustic scrubbers that will remove any reactive gases such as SO<sub>2</sub> or HCl.

Bionda (2011) describes these effects in detail. A version of his paper was previously published in this newsletter.

The upshot of all of the above is the stunning revelation that if you bubble reactive gases through cold water... you get reactions. Had the gas been left to exit the stack of its own accord, much of it would have remained gaseous and never formed particulate (or so the theory goes). Much of the “particulate” measured by Method 202 was pseudo-particulate. Not real. Purely an artifact of the sampling system.

EPA was finally convinced that this method needed repair. Since the problem was bubbling the stack gas through cold water, they designed a method using “dry impingers” and a condenser in order to minimize gas/liquid contact. In 2011, a mere 20 years after Method 202 was issued, EPA issued a brand new method... Method 202. Yes. They kept the old name to avoid having to re-write thousands of air permits that called out Method 202. So, there was the old Method 202. Now we have the new Method 202.

This new Method 202 is, in fact, considerably less biased than the old Method 202. But it was a low bar. Many believe the new 202 is still significantly biased, particularly when ammonia is present in the gas stream along with SO<sub>2</sub>.

But remember how CPM is formed in the first place, through cooling in ambient air? What if there could be a test method that simulated what happened when a gas plume exits the stack and cools. If this were possible, then the whole issue of CPM disappears. At ambient conditions, everything relevant is solid. You could just collect it on a filter. Life would be simple.

As it turns out, EPA has been working on just such a method. Actually, they've been working on it for quite some time – since before the new Method 202 was issued. The shorthand name for this is the “dilution sampling method”. EPA's name is OTM-37. OTM for Other Test Method. EPA says that OTM methods are “potentially useful” test methods. In the past, it has also been known as CTM-039 (in a previous version). CTM for Conditional Test Method. I think OTM is a promotion from CTM.

Essentially, OTM-37 attempts to mimic the natural condensation of stack gas by mixing the gas with a known quantity of cooled air and allowing a bit of time for condensation to occur. Once this happens, everything solid is collected on a filter. With this method, Total PM = FPM. Since this method mimics what actually happens to a gas plume in the ambient air, EPA has called this method the “gold standard” of particulate measurement (Myers, 2007).

Much R&D testing has already been done with OTM-37. Time after time in paired tests against Method 202, OTM-37 has shown lower results, sometimes significantly lower. This would tend to support the idea that Method 202 is (still) biased high even with the dry impinger modifications.

So why aren't we all using OTM-37 at this point? First of all, it's not yet an official reference method. EPA is still conducting validation testing. States are free to use the method if they desire, but its use for determining compliance with federal requirements must be done through a SIP process (with public notice and comment) or approved as an alternative under 40 CFR 60.8 or 63.13. But according to EPA, this is the “gold standard,” right? EPA should certainly be on board to get this method approved.

Well, everyone thought the OTM-37 train was pulling out of the station. States were interested. Industries were interested. Trade groups such as the American Petroleum Institute (API) and the National Council of Air and Stream Improvement (NCASI) were backing the method. The dilution solution for measuring pollution was the future of particulate testing. And then...

Let's back up for a minute. Remember the base assumption regarding what is and is not CPM. The operating assumption has been that gas/liquid interactions in the Method 202 sampling train create artificial particulate that would not be created in the ambient air from the actual plume. Specifically, if gaseous ammonia and gaseous SO<sub>2</sub> were just left to their own devices, they would happily exit the stack as gas and remain so. At least during the timeframe of interest. They would still be pollutants but they would not be particulate. And then...

About a month ago, during the annual meeting of the Source Evaluation Society (SES), Dr. Dave Nash from EPA gave a presentation called “Condensable PM (CPM) Measurement Method Comparison: Understanding CPM Formation.” In his talk, Dr. Nash presented the early results of his CPM formation research. Essentially, he introduced ammonia vapor,

SO<sub>2</sub>, and water vapor into a flow tube which would simulate natural mixing and cooling similar to OTM-37. He also added measured amounts of particulate. He used an FTIR to measure concentrations of the gaseous components and a very impressive-looking piece of hardware called an aerosol mass spectrometer (AMS) to determine the chemical composition of the particulate.

What he found appeared to shake up some of our assumptions about the fate of gaseous compounds in ambient air. Perhaps “shake up” is too strong. Some of his conclusions have been known for some time, but they still raise questions. Remember our happy ammonia and SO<sub>2</sub> gases? Well, it turns out that in Dr. Nash's flow tube, they appear to form particulate. More specifically, they interact with the particulate Dr. Nash added and form solid salts – ammonium sulfate, bi-sulfate, etc.

These very preliminary results suggest that at least some of what we have called Method 202 bias may not, in fact, be bias. Perhaps Method 202 is still biased high, but is OTM-37 also biased low? If so, how do these balance out? Are there some circumstances where Method 202 may be the appropriate method and others where OTM-37 is appropriate? What is the role of water in these reactions? Would we see the same effect in an actual stack that we are seeing in the lab?

These are all questions that EPA (read that as Dave Nash) intends to answer during continuing research into this issue. Both API and NCASI are working with EPA to resolve these questions. Unfortunately, the pace of research is slow and it is likely that we will not know the answers to these questions for a least a year, possibly longer.

So in the meantime, what is the fate of OTM-37? There are reports that a few states have tentatively approved OTM-37 for some uses. EPA, however, seems to be holding off on endorsing the use of OTM-37 until the research is further along. States that are contacting EPA about the use of the method are definitely not getting a warm and fuzzy feeling at the moment. More of a let's wait and see approach.

This certainly is not the end of OTM-37. But the method that ultimately emerges at the end of this process will likely differ significantly from the current version and may be more restrictive in its application. Also, it is not likely that Method 202 will go away any time soon. The two methods will likely co-exist with each used for specific applications.

Forty-two years after Method 5 came into the world and we're still learning about measuring particulate. I guess that's what keeps it all interesting.

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### BIOGRAPHY



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## Evaluation of SCR Operation for a Gas Conversion Project

Written by Rod Beittel, Babcock Power

Many owners of coal-fired boilers are opting to convert them to co-fire gas, or to fire gas exclusively. Beyond the basic equipment supply of burners, ignitors, and gas flow control trains, a fuel conversion project requires detailed consideration of the fuel influence on steam generator performance, as well as “back end” equipment including the ID fans, air heaters, and emission control equipment. As an OEM supplying boilers, firing systems, FGD and SCR systems, Babcock Power Inc (BPI) has developed a significant business in these conversions.

We are currently working with the owner of a plant with four coal fired boilers to convert them for co-firing or 100% gas firing. All of the units are equipped with wet FGD systems. Three of the units include SNCR systems, while one includes an SCR installed by Riley Power Inc, a BPI company, in 2008. This article discusses the impact of gas conversion on the operation of the SCR. The SCR-equipped unit will fire up to 55% natural gas at full load, and 100% gas at up to 55% load. The unit may be further modified in the future to fire 100% gas at full load.

### ORIGINAL SCR DESIGN AND PERFORMANCE

The SCR system was installed in 2008 and underwent Performance Testing in January 2009. Selected system design

parameters and Performance Test results are shown in Figure 17.

A limitation on operating the SCR at reduced load is the catalyst minimum continuous operating temperature,  $T_{\text{mincot}}$ . The value for  $T_{\text{mincot}}$  is specified by the catalyst supplier and represents the temperature below which the catalyst activity will be reduced due to formation of ammonium bisulfate in the catalyst pores. While the value for  $T_{\text{mincot}}$  must be determined by the catalyst supplier for the specific catalyst formulation and gas conditions,  $T_{\text{mincot}}$  can be estimated based on the flue gas concentration of ammonia, water, and sulfur trioxide. Using design and performance test gas composition, the estimated values for  $T_{\text{mincot}}$  are shown in Figure 18 on page 19.

Based on the measured  $\text{SO}_3$  values in the performance test, the minimum catalyst temperature would have been significantly lower than the design value.

The design value for  $\text{SO}_3$  was based on 1% conversion of  $\text{SO}_2$  leaving the boiler. This conversion is considered an average rate. Actual  $\text{SO}_2$  to  $\text{SO}_3$  conversion varies with the specific boiler, operating conditions, and fuel. The conversion in the performance test was 0.16%.  $\text{SO}_3$  data are limited

	Unit	Design	Performance Report
Boiler Heat Input	MBtu/h	6,687	6,110
Output	MWg	730	715
Gas temperature	°F	660	658
Inlet $\text{NO}_x$	lb/MBtu	0.25	0.22
$\text{NO}_x$ Removal	%	84	90
Inlet $\text{SO}_2$	ppmvd	1,440	1,380
Inlet $\text{SO}_3$	ppmvd	14.4	2.2
Flue gas moisture	%	11.9	NR
Flue gas $\text{O}_2$	%dry	3.26	3.0
Catalyst Minimum Continuous Operating Temperature (Supplier)	°F	602	NA

Figure 17: Original SCR Design and Performance Results

	Design	2009 Performance Test
NH <sub>3</sub> , ppmv, wet	131	127
SO <sub>3</sub> , ppmv, wet	12.7	2.0
H <sub>2</sub> O, %vol.	11.9	11.9 (assumed)
T <sub>mincot</sub> , estimated, °F	604	575

**Figure 18: Catalyst Minimum Continuous Operating Temperature Based on Gas Composition**

but are consistent with a conversion well below 1% for this unit. If the boiler outlet SO<sub>3</sub> is currently less than the design value, the minimum operating temperature for 100% coal would be significantly less than the original value.

**CURRENT SCR OPERATIONS, COAL-FIRING**

Figure 19 shows the SCR temperature over the load range. In practice, the SCR is kept in service (that is, ammonia is injected, there is no SCR bypass) at low loads and temperatures during daily load cycles, with the understanding that operating time at higher temperatures reverses any short-term bisulfate deposition.

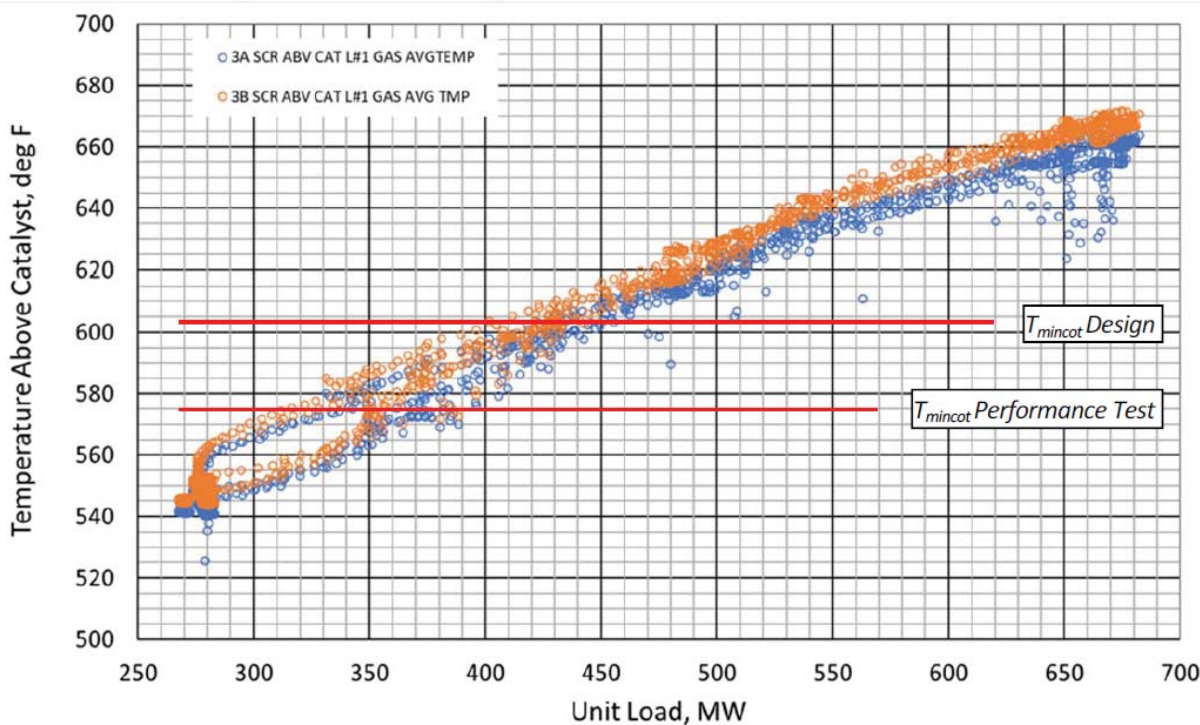
**SCR PERFORMANCE AFTER GAS ADDITION**

The boiler performance model cases, calibrated to the coal-fired operating data, were used as the basis for evaluating SCR performance:

- 40% Multi-pollutant Catalytic Reduction (MCR) firing natural gas only
- 47% MCR firing natural gas only.
- 55% MCR firing natural gas only.
- 109% MCR firing 55% natural gas and 45% coal, heat input basis.
- 109% MCR firing coal only.

Figure 20 lists parameters taken directly from the boiler model, followed by additional values defining the SCR conditions. Here are the assumptions used to develop the SCR cases:

- Boiler outlet NO<sub>x</sub> of 0.32 lb/MBtu for either coal or coal/gas co-firing at full load, based on the gas conversion project guarantee (greater of 0.32 lb/MBtu or pre-outage baseline).
- Boiler outlet NO<sub>x</sub> of 0.24 lb/MBtu for gas-only fir-



**Figure 19: Flue Gas Temperature at SCR Inlet versus Unit Load**

- ing, based on the gas conversion project guarantee.
- SCR inlet NH<sub>3</sub> calculated for 84% NO<sub>x</sub> removal, and typical assumption of NO/NO<sub>2</sub> speciation and stoichiometry (95% NO, NH<sub>3</sub>/NO=1, NH<sub>3</sub>/NO<sub>2</sub>=2).
- SCR inlet SO<sub>3</sub> equal to 1% of boiler outlet SO<sub>2</sub> from the boiler model, for coal or co-firing cases.
- SCR inlet SO<sub>3</sub> equal to 0.06 ppm for gas-only cases.
- Temperature, water vapor concentration from boiler model.
- Flue gas volume calculated from boiler model mass flow, composition, temperature, and duct pressure.

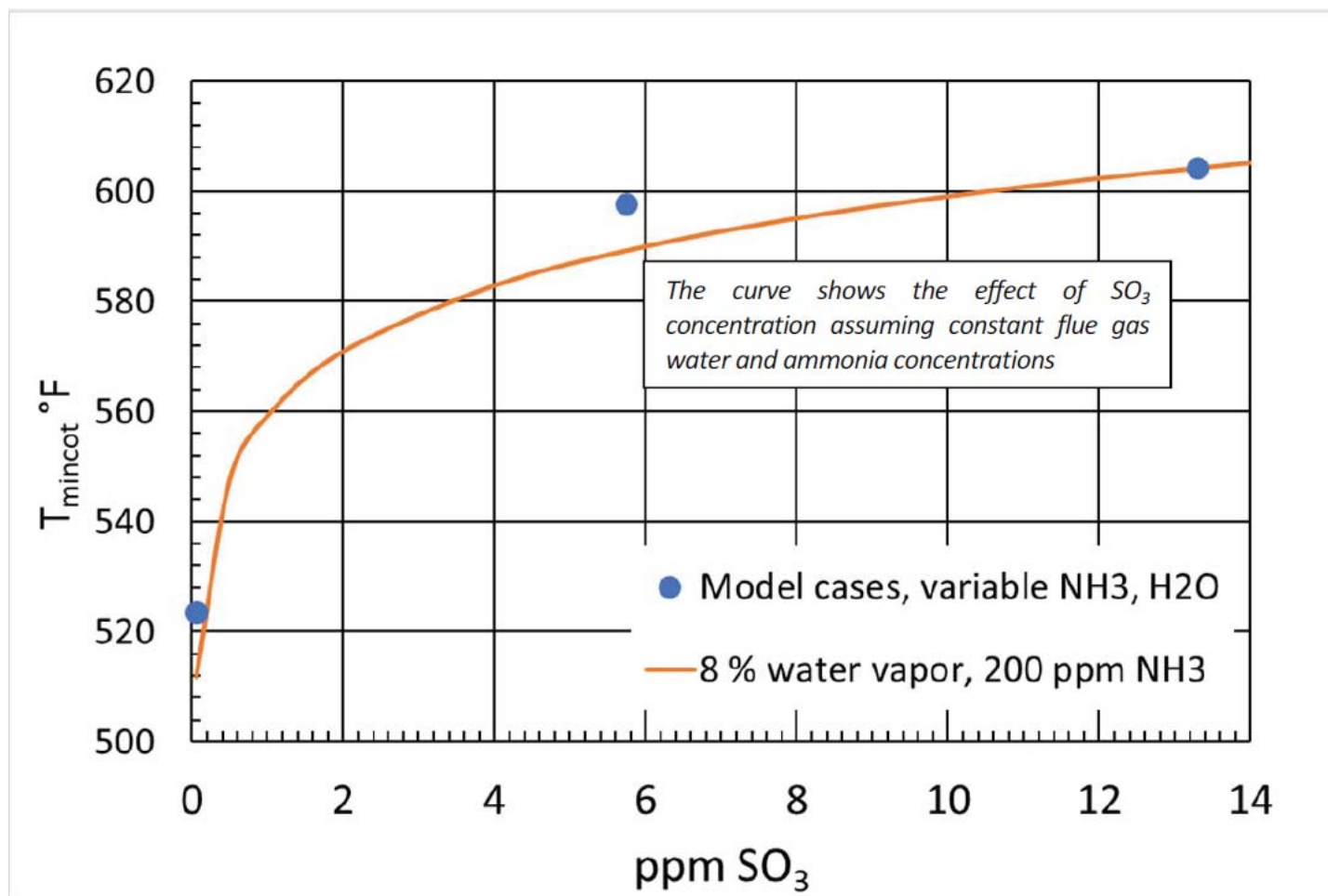
As discussed above, the available SO<sub>3</sub> data suggests that SO<sub>3</sub> is typically less than 1% of the SO<sub>2</sub> for this unit. While SO<sub>3</sub> has a large effect on T<sub>mincot</sub>, 1% conversion is assumed for comparison of coal-only and co-firing cases, and to provide

a conservative T<sub>mincot</sub> absent more data on actual SO<sub>3</sub>. The biggest change in SO<sub>3</sub> is with 100% gas-firing. The gas specification does not include any fuel sulfur analytical values. The fuel gas specification does show “less than” values for sulfur and H<sub>2</sub>S. As an upper limit, we calculated the equivalent total sulfur as 0.6 ppmvw SO<sub>2</sub> in the flue gas with 100% gas firing. At this low level, it is unlikely that the 1% conversion typical of coal-fired SO<sub>2</sub> levels would apply. An SO<sub>3</sub> value of 0.06 ppmvw, or 10% conversion of the total sulfur, was assumed for estimating T<sub>mincot</sub>.

Figure 21 on page 21 shows the variation of T<sub>mincot</sub> for the model cases. Co-firing with gas reduces the SO<sub>3</sub> proportional to the gas input. For 55% heat input from gas, T<sub>mincot</sub> would decrease to 589°F from 604°F with coal alone, based on sulfur reduction only. Due to the increased water vapor with gas

	40% MCR with 100% Gas	47% MCR with 100% Gas	55% MCR with 100% Gas	109% MCR with 55% Gas/45% Coal Input	109% MCR with 100% Coal Input
<b>Boiler Performance Model Parameters</b>					
Coal Heat Input (MMBtu/hr)		-	-	2,743	5,984
Natural Gas Heat Input (MMBtu/hr)	2,352	2,719	3,156	3,562	-
Total Heat Input (MMBtu/hr)	2,352	2,719	3,156	6,305	5,984
Gas Temperature Leaving Economizer (°F)	549	531	551	672	668
O <sub>2</sub> (%vol,dry)	2.5	2.5	2.5	4.1	4.3
SO <sub>2</sub> , ppmvw	0	0	0	574	1,330
H <sub>2</sub> O, vol%	18.5	18.5	18.5	13.4	8.2
<b>Additional Assumed or Calculated Values for SCR Performance</b>					
NO <sub>x</sub> , lb/MBtu	0.24	0.24	0.24	0.32	0.32
NO <sub>x</sub> , ppmvw	171	171	171	213	213
SO <sub>3</sub> , ppmvw	0.06	0.06	0.06	5.7	13.3
NH <sub>3</sub> , ppmvw	151	151	151	187	187
Dust, gr/acf	0	0	0	0.7	1.7
Gas flow (total), kpph	1,987	2,298	2,667	5,884	5,742
Gas flow (total), kacfm	887	1,008	1,196	2,924	2,737
Estimated T <sub>mincot</sub> , °F	524	524	524	598	604

Figure 20: Predicted Performance Summary for SCR Evaluation



**Figure 21: Estimated  $T_{\text{mincot}}$  for coal firing (13.3 ppm  $\text{SO}_3$ ), co-firing (5.7 ppm  $\text{SO}_3$ ) and gas-firing (assumed 0.06 ppm  $\text{SO}_3$ ) model cases.**

firing, the net effect with co-firing is a  $T_{\text{mincot}}$  of 598°F. As noted previously, the actual  $\text{SO}_3$  with 100% coal firing may vary as much or more than the reduction with gas firing, due to specific unit, fuel or firing characteristics.

For the assumed  $\text{SO}_3$  value of 0.06 ppm with 100% gas firing, the estimated  $T_{\text{mincot}}$  is much lower, about 524°F. There are two important caveats to the specific value. First, in addition to the potential trace sulfur represented by the assumed  $\text{SO}_3$  value, there will likely be residual sulfur in the boiler system, and especially in the catalyst, as firing shifts between coal, co-firing, and 100% gas. Second, the model used to estimate  $T_{\text{mincot}}$  is based on coal firing correlations and may not pertain to very low  $\text{SO}_3$  values. Despite these uncertainties, firing with 100% gas will clearly reduce the catalyst minimum operating temperature to values comparable to the flue gas temperature at minimum load. Therefore, flue gas temperature is not expected to limit SCR operation for 100% gas firing. For co-firing, the effect on  $T_{\text{mincot}}$  will be minimal, and SCR operation will need to observe either the minimum

temperature, or limited-duration operation at low loads, as now practiced for 100% coal.

#### CATALYST SUPPLIER ASSESSMENT

The model flue gas conditions were submitted to the original catalyst supplier for review and comment. Key items are summarized below. For the assumed gas compositions, the catalyst supplier  $T_{\text{mincot}}$  was similar to RPI's estimate, as shown in Figure 20 on page 20.

- Gas co-firing at full load was characterized as more challenging than 100% coal firing at full load. Factors included lower  $\text{O}_2$ , higher  $\text{H}_2\text{O}$ , lower  $\text{SO}_2$ , and slightly higher gas flow compared to the coal case. In addition, the same inlet  $\text{NO}_x$  value in lb/MBtu was assumed for both cases, and with the higher boiler heat input for co-firing, the inlet  $\text{NO}_x$  mass flow is higher.
- Conversely, the current 100% coal case is more favor-

	47% MCR with 100% Gas	55% MCR with 100% Gas	109% MCR with 55% Gas/45% Coal Input	109% MCR with 100% Coal Input
Gas Temperature Leaving Economizer (°F)	531	551	672	668
<b>Catalyst Supplier Predicted Performance Maintaining NOx Removal</b>				
T <sub>mincot</sub> , F	530	530	600	605
For constant NOx removal				
NOx removal, %	84	84	84	84
Ammonia slip, ppm	10	6	2	2
<b>Catalyst Supplier Predicted Performance Maintaining Ammonia Slip</b>				
Ammonia slip, ppm	2	2	2	2
NOx removal, %	53	66	84	84

*Figure 22: NOx Removal Rates Predicted for the Different Firing Cases*

able than the original design, due to the substantially lower gas flow compared to the original SCR design. The current design case gas flow for full load, 100% coal is consistent with the gas flow recorded in the Performance Test in 2009. Both the coal and co-firing cases are therefore more favorable than the design case. Both are expected to achieve 84% removal with <2 ppm ammonia slip, even with though the assumed inlet NOx of 0.32 lb/MBtu is higher than the original design of 0.25 lb/MBtu.

- There is an important factor for gas-firing beyond the change in T<sub>mincot</sub>: With 100% gas, and no flue gas SO<sub>2</sub>, the resulting desorption of sulfate from the catalyst can significantly reduce catalyst activity, in addition to the reduction due to lower temperature.
- The desorption effect may not be fully realized with daily cycling between 100% gas and co-firing. The supplier recommends laboratory testing of the catalyst to characterize desorption rates and performance reduction.

The supplier model for catalyst performance predicts an increased ammonia slip of 6-10 ppm for low-load, 100% gas firing at a constant NOx removal rate of 84%. The ammonia slip values are high relative to coal firing SCR operation. Higher ammonia slip is typically be acceptable for gas-fired SCR systems, since there is no SO<sub>3</sub> to form bisulfates in the catalyst. The NOx removal rates in the last row of Figure 22 were predicted assuming ammonia slip is limited to 2 ppm.

**AMMONIA INJECTION CONTROL**

The ammonia for this unit is produced on site in a urea hydrolyzer. The material balance for the original SCR design required 1,080 lb/hr of hydrolyzer vapor to each reactor. For the design operation using 50 wt% urea, the vapor contains 28.3% ammonia by weight (or 37.5% by volume), with the balance consisting of carbon dioxide and water vapor.

**Control valves**

The ammonia flow will vary with load, SCR inlet NOx, and outlet control point. The boiler model cases cover a unit MCR from 109% to 40% of rated steam flow. Following the gas conversion burner changes, guaranteed maximum inlet NOx is 0.32 lb/MBtu for coal or co-firing, and 0.24 lb/MBtu for gas-only firing. With a fixed outlet control point of 0.04 lb/MBtu, (original design), the flow rate of vapor will vary from 321 to 1,210 lb/hr per reactor to cover all cases. Controlling instead at 0.10 lb/MBtu outlet reduces the minimum vapor flow to 225 lb/hr per reactor. Load cycling data shows that the ammonia vapor flow is controllable and stable from under 200 lb/hr to over 1,200 lb/hr with the existing control valves and flow meters.

Gas firing and low load will likely reduce the NOx levels entering the SCR. For the lowest MCR case modeled, the ammonia flow rate will be in the controllable range with SCR inlet NOx as low as 0.12-0.15 lb/MBtu, and an outlet of 0.04 lb/MBtu.

**Hydrolyzer operation**

The hydrolyzer supplier states that the general system de-

sign should provide 10:1 turndown from the design point of 2,200 lb/hr total, or 110 lb/hr per reactor. In recent load cycling data, the skid supply pressure and temperature were stable across the full operating range, so the hydrolyzer does not appear to be limiting at the minimum injection rates recorded. It may be necessary to adjust set points, re-tune steam control valves, and possibly modify valve trims for optimum operation if lower production rates become the norm. The supplier's current design provides capacity control by controlling the condensate flow, rather than the steam flow directly as in the existing system. This updated control approach may be beneficial for low-rate operation.

### **Total flow meter**

The total flow to each reactor is measured with a Coriolis type flowmeter. The cycling data did not indicate issues in the 150-200 lb/hr minimum observed flow range. Based on the flow meter sizing software, the meter should be accurate to 2% in this range and should be able to measure as low as 60 lb/hr at 5% accuracy. Should accuracy or stability begin to impact flow control, the existing system can be optimized around the lower flow rates with valve and control loop tuning.

### **Lance flow meters**

The flow to each lance is measured with a vortex flow meter. These are expected to have less turndown capability than the total flow meter. In the cycling data, the sum of the individual flow meters agrees with the total (Coriolis) flow meter over most of the range except at 200-300 lb/hr and lower total flow where some lance flows begin showing zero. The lance flow meters are used primarily to aid in tuning the ammonia distribution at or near full load. They are not used in the control logic. Loss of accuracy or even loss of reading at low flows should not affect the controllability of the system.

### **Controls**

The original control logic tripped the ammonia flow below a temperature alarm point based on  $T_{\text{mincot}}$ . This has been modified to allow operation at lower temperatures, for limited periods followed by higher temperature operation. This could be extended to increase or remove the time limitations with 100% gas-firing. The catalyst supplier should be consulted to establish limits on low-temperature operation.

### **SUMMARY**

There does not appear to be an immediate need for SCR equipment or controls changes to accommodate gas addition. After actual low-load, gas-only NO<sub>x</sub> levels are determined, and operating experience is gained, it may be de-

sirable to optimize ammonia production and flow control logic. Gas-only firing for extended duration may deactivate the catalyst due to desulfation. Cycling between 100% gas and co-firing is expected to limit deactivation, however, the catalyst supplier should be engaged to confirm this through laboratory testing.

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*Rod Beittel is a Senior Principal Engineer with Babcock Power Environmental, where he is responsible for process engineering and technical support of emission control systems. He joined the company's R&D department in 1988, managing test facility operations in projects for burner development, fuel utilization, and emissions control.*

*Prior to joining Babcock Power, he developed and operated pilot scale combustion facilities at Southern Research Institute. Rod is a licensed professional chemical engineer with a B.S.Ch.E. from Lehigh University and an M.S.Ch.E. from the University of Arizona.*



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