

# **PILOT EVALUATION OF THE CATALYTIC OXIDATION OF MERCURY FOR ENHANCED REMOVAL IN WET FGD SYSTEMS**

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## **ABSTRACT**

The U.S. DOE National Energy Technology Laboratory (NETL) and EPRI are funding the development of a catalytic oxidation process to enhance mercury removal in wet flue gas desulfurization (FGD) systems. In the process, solid catalysts oxidize vapor-phase elemental mercury ( $\text{Hg}^0$ ) in flue gas and the oxidized mercury is removed in downstream FGD absorbers.

An earlier NETL/EPRI co-funded project identified materials active for catalyzing oxidization of  $\text{Hg}^0$  in flue gas. Promising catalysts were evaluated in small, fixed sand-bed reactors for periods up to six months each at three utility plants. Based on promising results from that testing, NETL funded the current project to conduct pilot-scale testing of catalyst materials on honeycomb structures. Tests are being conducted for 14-months each on 8000-acfm flue gas slipstreams at two power plants. One fires North Dakota lignite and the other fires Powder River Basin (PRB) coal. Four different catalyst materials will each treat one-fourth of the flue gas slipstream at each site.

The pilot unit has been designed and constructed, and will be installed at the lignite-fired plant in August 2002. Laboratory tests have been completed to select catalysts for pilot testing and to size honeycomb structures for each. This paper describes the pilot unit and test program, and summarizes the laboratory results.

## **INTRODUCTION**

The objective of this project is to demonstrate at pilot scale the use of solid honeycomb catalysts to promote the oxidation of  $\text{Hg}^0$  in the flue gas from coal combustion. The project is being funded by NETL under Cooperative Agreement DE-FC26-01NT41185. EPRI, Great River Energy (GRE), City Public Service (CPS) of San Antonio, and the North Dakota Industrial Council are project co-funders. URS is the prime contractor.

The mercury catalytic oxidation process under development will use catalyst materials on honeycomb substrates to promote the oxidation of  $\text{Hg}^0$  in the flue gas from coal-fired power plants that have wet lime or limestone FGD systems. The oxidizing species are already present in the flue gas, and could include chlorine, hydrochloric acid (HCl), and/or other species. Oxidized mercury is removed in the wet FGD absorbers and typically co-precipitates with the byproducts from the FGD system.

The catalytic oxidation process was initially developed in an earlier DOE/EPRI co-funded project that began in 1995. In the second phase of that project, candidate catalyst materials were evaluated for up to six months each in the actual flue gas from three different coal- or lignite-fired plants in small, slipstream sand-bed reactors.<sup>1,2,3</sup> Pilot-scale testing in the current project represents a logical progression towards commercialization of the process. The pilot-scale tests are to evaluate candidate materials at a larger scale and in a commercial form, to provide engineering data for future full-scale designs, and will continue for up to 14 months at each of two sites to provide longer-term catalyst life data.

Based on information collected during the U.S. EPA Mercury Information Collection Request (ICR), the technology for catalytic oxidation of  $\text{Hg}^0$  would have the greatest effect on the flue gas from subbituminous coal or lignite, where most of the mercury is present in the elemental form. There are currently over 30,000 MW of scrubbed capacity firing these fuels, with more systems planned.

The two utility team members are providing co-funding, technical input, and host sites for testing. GRE will host the first test site at their Coal Creek Station (CCS), which fires a North Dakota lignite; and CPS will host the second site at their J.K. Spruce Plant, which fires PRB. These two host sites each have existing wet FGD systems downstream of high-efficiency particulate control devices, an ESP at CCS and a reverse-gas fabric filter at Spruce. Each has substantial concentrations of  $\text{Hg}^0$  in their flue gas.

To date, a test plan has been prepared, the pilot unit has been designed and constructed, and laboratory tests have been conducted to select and size the catalysts for evaluation in the pilot unit. The test plan and pilot unit are described in the Experimental section below, and the results of the laboratory investigation are summarized in the Results section. Although the catalyst honeycomb dimensions selected for the pilot unit are presented in the Results section, there are no pilot-scale performance results yet available.

## **EXPERIMENTAL**

This section provides an overall project description and describes the catalytic oxidation pilot unit and test plan for this project. Each is described in a separate subsection below.

### **Project Description**

Figure 1 is a process flow diagram for the catalytic oxidation process under development. The process is very straightforward with no “moving parts.” A catalyst material in a honeycomb form is inserted into the flue gas path upstream of the FGD system. Downstream

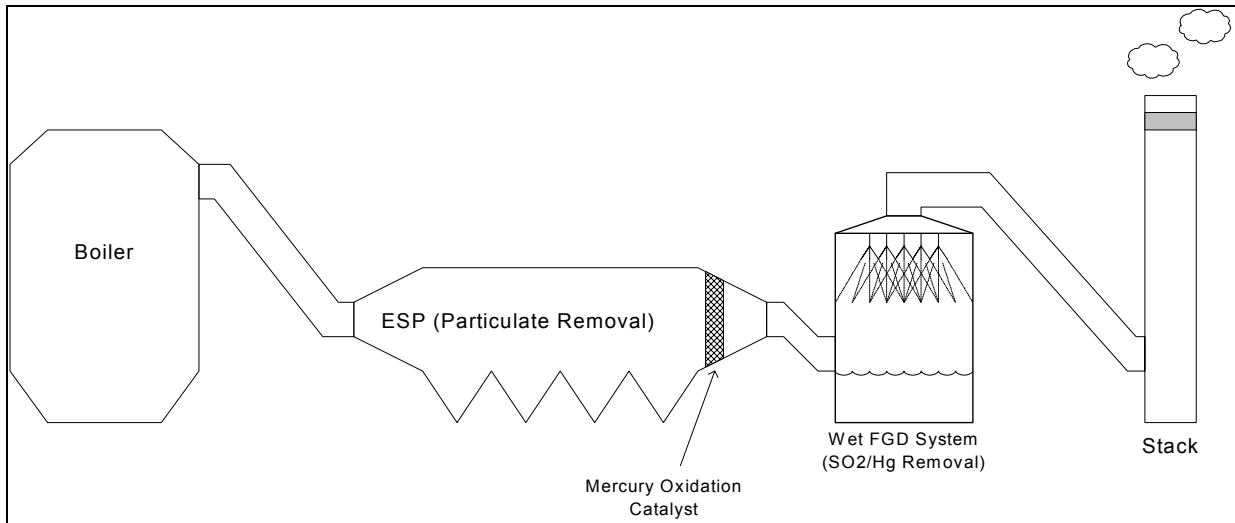


Fig. 1. General Process Flow Diagram for the Mercury Catalytic Oxidation Process.

of the catalyst, the oxidized mercury is scrubbed in the FGD absorber, and co-precipitates with the calcium sulfite or gypsum byproduct from lime or limestone wet FGD systems.

Development has focused on the outlet of a plant's cold-side particulate control device as the most likely location for such a catalyst, for two reasons. One is that the flue gas velocity is typically low as it exits a particulate control device (e.g., about 5 ft/sec as it exits an ESP) making it an ideal spot to operate a catalyst at longer residence time and lower pressure drop. The other is that with the flue gas being relatively particulate free at this location, a close-pitched catalyst can be used. This allows for a high surface area per volume of catalyst relative to "dirty gas" operation, such as in most SCR systems, and allows a smaller catalytic reactor to be used. Based on data collected in the previous Mega-PRDA project, it is estimated that only a 6-inch depth of a close-pitched palladium-based catalyst would be adequate to achieve 90% or greater oxidation of the  $Hg^0$  present in the flue gas at this location.

Preliminary economic estimates show that a catalytic process, if installed upstream of a wet FGD system, should allow plants so equipped to achieve 90% overall mercury control at a lower cost than by injecting activated carbon. However, the actual cost of the catalytic process will depend largely on the catalyst life and required catalyst volume. This project will collect data that will allow these two parameters to be evaluated for two coal types. Four different catalysts will be tested in parallel in a divided reactor that is sized to treat approximately 8000 acfm of flue gas (approximately 2000 acfm to each catalyst). This should allow enough catalyst in each reactor to avoid "wall effects" that can significantly affect results.

### **Pilot Unit Design and Construction**

Figure 2 shows a sketch of the pilot reactor concept. The pilot reactor will treat the low-dust flue gas from downstream of the existing ESP (CCS) or baghouse (Spruce), extracted

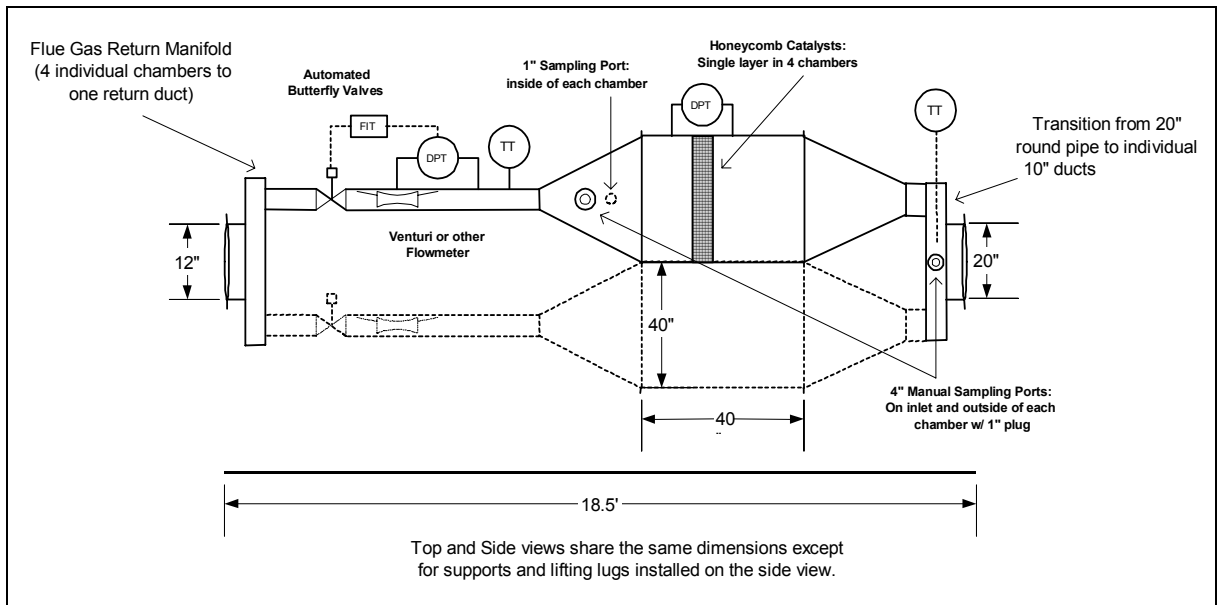


Fig. 2. Simplified Pilot Unit Design (side view).

isokinetically from the host unit’s ID fan outlet. The treated flue gas will return to the host unit’s ID fan inlet, thus avoiding the need for a dedicated fan. The amount of flue gas treated by the pilot unit is a very small fraction of the overall host unit flue gas flow, so this small amount of flue gas “recycle” will not cause any adverse effects.

The operating concept will be to install the pilot unit in the location described, with catalyst materials in each of the four catalyst chambers, and operate the unit in automated mode for approximately 14 months. The pilot unit will be heat-traced to maintain temperature (negating heat losses from the relatively small-scale equipment) and will be instrumented for temperature, pressure drop, and flow rate measurement for each catalyst bed. The gas flow rate through each catalyst bed will be controlled independently. The four pilot unit reactor chambers are each approximately 40 inches in height, width, and depth.

Figure 3 shows a simplified piping and instrument diagram (P&ID) for the pilot unit. The pilot unit is actually more complex than a future full-scale implementation would be because it has provisions to control flow rate and temperature in four separate catalyst chambers and to measure gas differential pressure and mercury species concentrations across each chamber. The full-scale implementation would not likely require any instrumentation or controls other than monitoring gas pressure drop across the catalyst.

Figures 4 and 5 are photographs of the completed pilot unit prior to being shipped to the CCS site. The pilot unit will be shipped in early August, and installation at CCS should be completed by mid- to late August 2002.

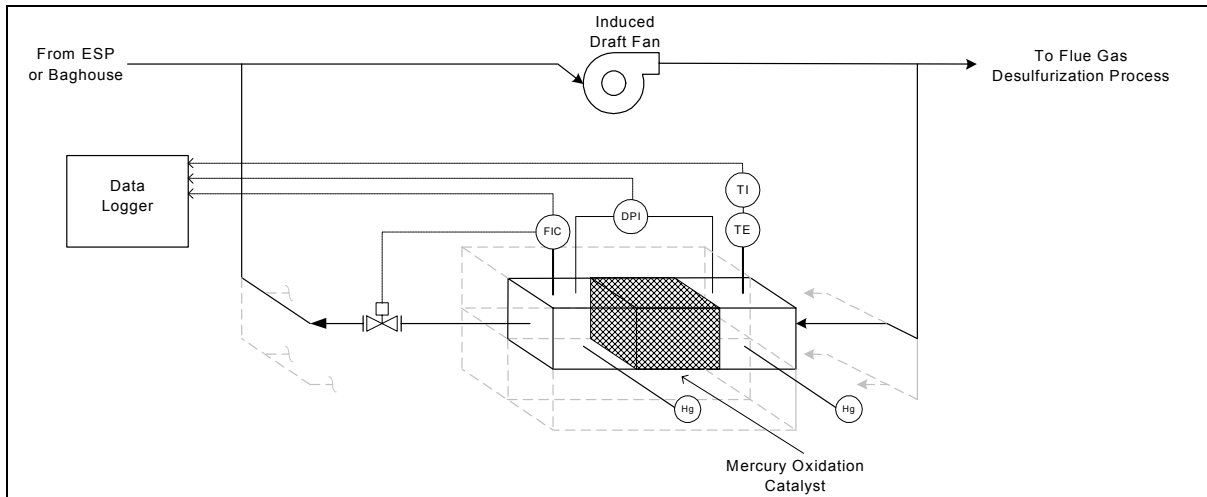


Fig. 3. Simplified P&ID for the Proposed Pilot Unit (1 of 4 catalyst chambers shown).



Fig. 4. Photograph of Side View of Completed Pilot Unit.

EPRI has funded the construction of a second pilot unit, so that catalyst tests can be conducted at the two sites in parallel. However, it has been decided to start up the pilot unit and conduct initial catalyst tests at the CCS site before the construction of the second pilot unit for Spruce Plant begins. This will allow “lessons learned” in the startup and initial operation of the first pilot unit to be reflected in the design and construction of the second.



Fig. 5. Photograph of Discharge End of Completed Pilot Unit.

### Test Plan

The pilot unit will be installed by GRE at the outlet of one of four ID fans on Unit 1. After startup and check out of the pilot unit, the four catalysts will be loaded into the pilot unit. Flue gas flow rates and temperatures will be established and placed in automatic control, and the unit will be left to operate unattended for at least a week to allow the catalyst materials to come to mercury adsorption equilibrium. After that period, initial catalyst performance data will be collected for each of the four materials. The flue gas flow rate through each chamber will be varied, to allow a determination of catalyst performance vs. area/space velocity. The results of these tests will be used to determine the ideal area velocity for testing each catalyst. The ideal area velocity should be low enough to achieve high elemental mercury oxidation efficiency (>90%) but not so low that a large excess of catalyst is present (i.e., 100% Hg<sup>0</sup> oxidation).

Once the optimum flue gas flow rate is established for each catalyst, the pilot unit will be left in operation for approximately 14 months. Telemetry equipment will be used to allow pilot unit flow rate, temperature, and pressure drop data to be monitored off site. The pilot unit will be automated for all operations except mercury analyses.

An EPRI semi-continuous mercury analyzer will be operated by URS team members while on site. About once per month, project team members will travel to the site and use the EPRI semi-continuous mercury analyzer to measure total and elemental mercury concentrations upstream and downstream of each catalyst, to determine oxidation activity. The EPRI semi-continuous mercury analyzer is illustrated in Figure 6. The analyzer is based on the amalgamation of Hg<sup>0</sup> with gold, and the cold-vapor atomic absorption of Hg<sup>0</sup> that is

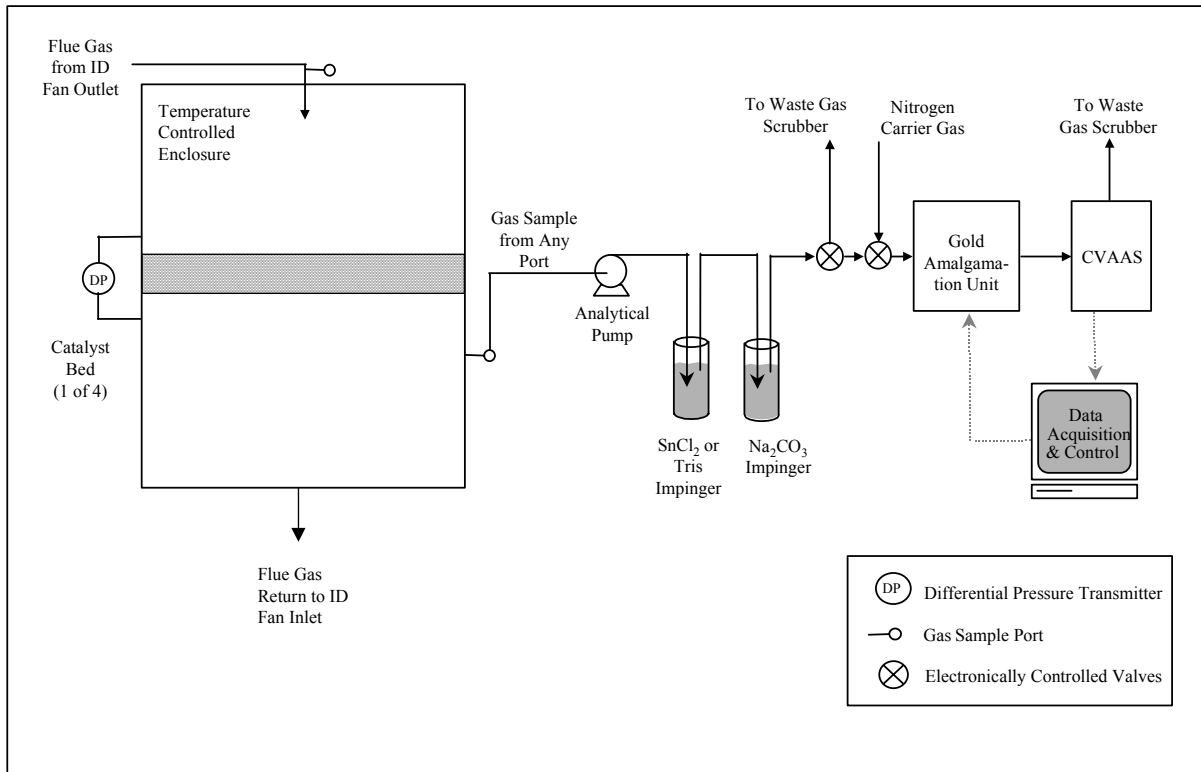


Fig. 6. Schematic of the EPRI Semi-continuous Mercury Analyzer.

subsequently thermally desorbed from the gold. The analyzer can determine total mercury concentrations by reducing all of the oxidized mercury to the elemental form with stannous chloride upstream of the gold, or only Hg<sup>0</sup> concentrations by using either a tris(hydroxymethyl)aminomethane (Tris) or potassium chloride (KCl) solution to capture oxidized mercury while allowing Hg<sup>0</sup> to pass through without being altered.

At least three times over the test period, manual measurements will be conducted using the draft Ontario Hydro method to verify results from the semi-continuous analyzer. Also, periodic measurements will be made for sulfuric acid and NO<sub>2</sub> concentrations upstream and downstream of each catalyst to quantify whether the catalysts also oxidize any of the flue gas SO<sub>2</sub> or NO (an undesirable effect).

At the beginning and end of testing at each site, a thorough characterization of the flue gas at the host site ID fan outlet will be conducted. This will include mercury concentrations and speciation by the Ontario Hydro method; flue gas sulfuric acid concentration by the Controlled Condensation method; HCl, chlorine, HF, and fluorine by Method 26a; and trace metals by Method 29. Ontario Hydro method measurements will also be conducted at the host unit FGD outlet, so the removal of mercury by species across the existing scrubber can be quantified.

During these flue gas characterization periods, samples of the coal, fly ash, scrubber liquor, and scrubber byproduct will be collected and subsequently analyzed for mercury content.

This will allow a mercury balance to be calculated across the host site, and will verify the fate of mercury absorbed in the host site FGD system. Also, the FGD byproduct will be evaluated in the laboratory for mercury stability. Both leaching of mercury into groundwater and volatilization of mercury into the air from the byproduct will be investigated.

The two host sites selected for this project, GRE's CCS and CPS' Spruce Plant, each fire relatively low-sulfur, low-chloride coals. CCS fires a North Dakota lignite with about a 6300 Btu/lb heat content, 0.7 wt% sulfur content, 0.1 ppm mercury content, and 100 ppm chloride content. Spruce fires a Wyoming, PRB subbituminous coal with about an 8400 Btu/lb heat content and a 0.5 wt% sulfur content. The mercury and chloride contents of the Spruce coal are similar to those at CCS (0.1 ppm mercury and 100 ppm chloride). Spruce occasionally co-fires some petroleum coke along with the PRB coal.

Notwithstanding fuel differences, the two sites are somewhat similar in equipment configuration. Both have tangentially-fired boilers rated at about 550 net MW (two units at CCS, only one at Spruce). CCS has a relatively large, cold-side ESP for particulate control, followed by ID fans and an Alstom (was ABB, and previously CE) wet lime FGD system. The FGD absorbers (open spray towers) operate at greater than 90% SO<sub>2</sub> removal efficiency. About 25% of the flue gas at the ID fan exit bypasses the FGD system, producing a reheat effect. At Spruce, a reverse-gas fabric filter is used for particulate control, followed by ID fans and an Alstom wet limestone FGD system. As at CCS, the absorbers are designed to achieve greater than 90% SO<sub>2</sub> removal, and a portion of the ID fan exit gas bypasses the FGD system. The CCS FGD system produces a calcium sulfite byproduct that is landfilled, while the Spruce FGD system produces a gypsum byproduct that is sold for wallboard or cement production.

Both of the host sites have made previous flue gas mercury concentration measurements. GRE's CCS was one of the sites selected for gas testing in the EPA Mercury ICR. Those tests showed the Hg<sup>0</sup> concentration in the flue gas downstream of the ESP and upstream of the FGD system to be about 8 µg/Nm<sup>3</sup>. The second site, CPS' Spruce Plant, had stack mercury tests conducted shortly after startup in 1993. Those measurements averaged about 9 µg/Nm<sup>3</sup> of total mercury. Based on assumed values of 80% of the flue gas being scrubbed, 30% mercury oxidation, and 90% net removal of oxidized mercury across the FGD absorbers, the Hg<sup>0</sup> concentration at the fabric filter outlet is estimated to be between 7 and 8 µg/Nm<sup>3</sup>. As mentioned above, the Hg<sup>0</sup> content of these two host sites will be verified by Ontario Hydro measurements at the beginning and end of the test period at each site.

### **Laboratory-scale Support Tests**

The pilot effort is being supported by offsite laboratory-scale testing using synthetic flue gases that are blended to match flue gas compositions at the host sites. Initial laboratory-scale support tests have been conducted to screen candidate catalysts for activity prior to procuring and installing them in the pilot unit. A description of these tests is provided below.

The catalyst materials used to prepare the pilot test samples were shown in the previous Mega-PRDA testing to effectively oxidize Hg<sup>0</sup> in flue gas. Additional tests were needed, however, to verify that the processes used to prepare the pilot-scale test samples (applying material to honeycomb substrates) do not alter their performance. Tests of prepared



honeycomb cores were conducted in URS' Austin laboratories to screen test samples for activity and to estimate the quantity of catalyst required to achieve high  $\text{Hg}^0$  oxidation across the pilot reactor. Tests involved exposing the catalysts to simulated flue gas at conditions similar to the CCS test site and measuring the extent of  $\text{Hg}^0$  oxidation across each catalyst as a function of simulation gas flow rate.

Figure 7 illustrates the laboratory test unit. A simulated flue gas is prepared by mixing heated nitrogen gas streams containing  $\text{SO}_2$ ,  $\text{HCl}$ ,  $\text{NO}_x$ ,  $\text{CO}_2$ ,  $\text{O}_2$ , and water. The gas composition is varied by appropriately adjusting the various gas rates. Mercury is injected into the gas by contacting nitrogen carrier gas with an elemental mercury permeation tube (VICI Metronics) in a mercury saturation vessel. The mercury concentration is controlled by the temperature of and the nitrogen flow rate through the mercury saturator. All gas mixing, water saturation, and mercury injection occur within a closed, temperature-controlled box designed to prevent water condensation, which can affect the behavior of mercury and the gas concentrations in the flow lines.

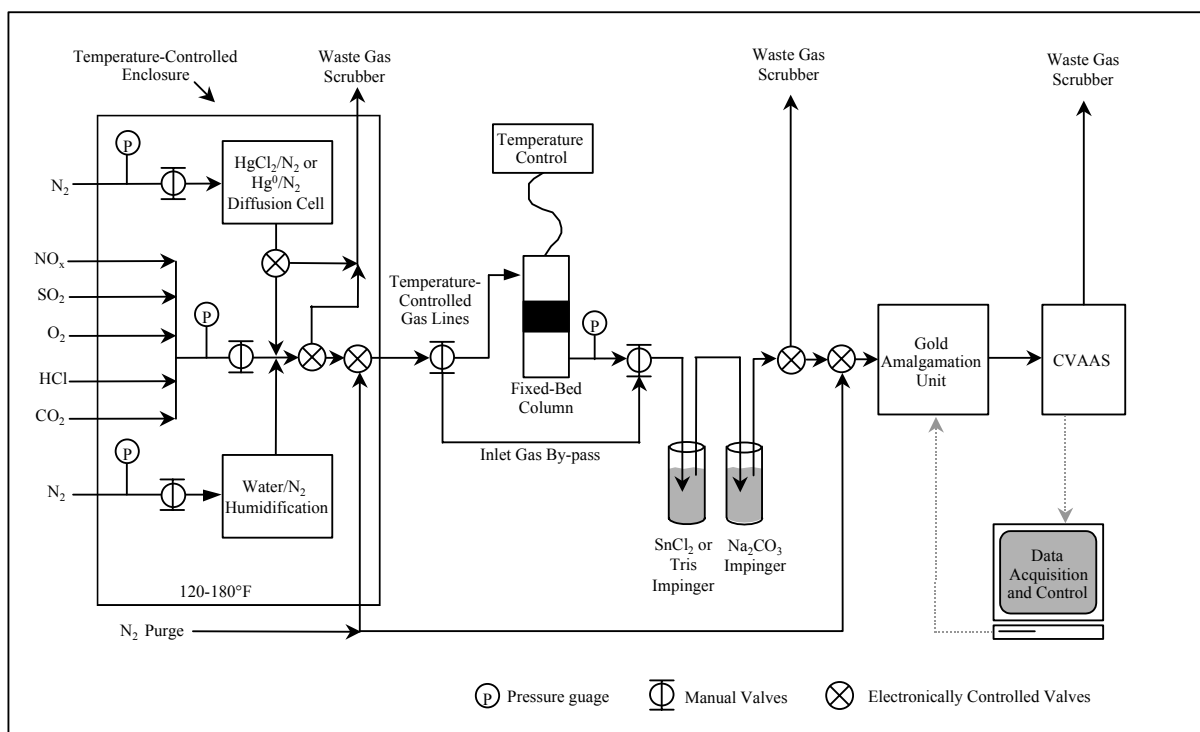


Fig. 7. Bench-scale Mercury Oxidation Test Unit.

The reaction gas flows at a desired rate (1 to 2 standard L/min) through heated Teflon lines ( $120^\circ\text{C}$ ) to a temperature-controlled Pyrex column (1.27-4.5 cm ID) containing the catalyst to be studied. The catalyst tube is configured vertically with the honeycomb held in place by a glass support located near the bottom of the tube. The column temperature is controlled based on the reading from an internally mounted thermocouple that is shielded from the gas with a glass sheath. The reaction gas flows downward through the column. Oxidation tests start by initiating flue gas flow across the heated catalysts. Tests are run until adsorption

equilibrium is attained (e.g., inlet total mercury equals the outlet total mercury). The  $Hg^0$  at the outlet of the catalyst is then measured to determine oxidation performance.

The composition of the laboratory flue gas is intended to simulate that of the host sites. Table 1 lists the flue gas composition used for the CCS screening tests, which simulate the flue gas from North Dakota lignite although with two exceptions. One is that the laboratory apparatus is not configured to simulate a flue gas with as high a moisture content as would be expected with North Dakota lignite (approximately 16-17%), so the simulation gas moisture content is limited to 9 to 10%. The other is that the simulation gas has a higher mercury content than is expected in the lignite flue gas. This is typical practice for our laboratory simulations, with the higher concentration being employed to improve the accuracy of mercury oxidation percentage measurements. CCS simulation tests are performed at the 350°F temperature expected at the outlet to the ESP (i.e., the location where the pilot unit will be installed).

**Table 1. Site 1 Simulated Flue Gas Composition used in Laboratory Tests.**

<b>Gas Component</b>	<b>Concentration</b>
SO <sub>2</sub>	500 ppm
HCl	6 ppm
NO <sub>x</sub>	200 ppm
H <sub>2</sub> O	9%
CO <sub>2</sub>	12%
O <sub>2</sub>	5%
Hg <sup>0</sup>	20-40 µg/Nm <sup>3</sup>
N <sub>2</sub>	Balance

Table 2 lists the catalyst materials tested. The catalyst materials include a commercially available palladium-based material, an SCR-type catalyst (titanium/vanadium on alumina), an experimental activated carbon on alumina, and a fly-ash-based material fixed on alumina. The experimental carbon and fly ash honeycomb catalysts are being prepared by Applied Ceramics.

**Table 2. Catalyst Materials for Evaluation in Pilot Mercury Oxidation Tests.**

<b>Catalyst Name</b>	<b>Description</b>
Pd #1	Commercial palladium catalyst on alumina substrate
SCR	Siemens commercial NO <sub>x</sub> catalyst; titanium-vanadium based honeycomb
Carbon #6 (C #6)	Activated carbon on alumina substrate
SBA #5	Active fly ash on alumina substrate

## **RESULTS AND DISCUSSION**

The only results available to date are for the laboratory simulations with the candidate catalysts. These results have been used to size the four catalysts to be tested in the pilot unit. Both the laboratory results and the pilot unit catalyst sizing are discussed below. Pilot unit results are not expected to start being available until sometime in September.

## Laboratory Simulation Results

The results of the laboratory simulation runs are plotted as a function of catalyst honeycomb area velocity in Figures 8 and 9. Note that the oxidation results shown in the figures were all measured after the catalysts had reached mercury adsorption equilibrium, so the performance shown truly represents the oxidation of  $\text{Hg}^0$  across the honeycomb sample and is not biased by  $\text{Hg}^0$  adsorption. In Figure 8, the results show that the C #6 carbon, SBA #5 fly ash, and the Siemens SCR catalysts have very similar activity for  $\text{Hg}^0$  oxidation. For the longer core length samples, approximately 95% oxidation was achieved at area velocities in the range of 20 to 40 standard ft/hr for all three materials.

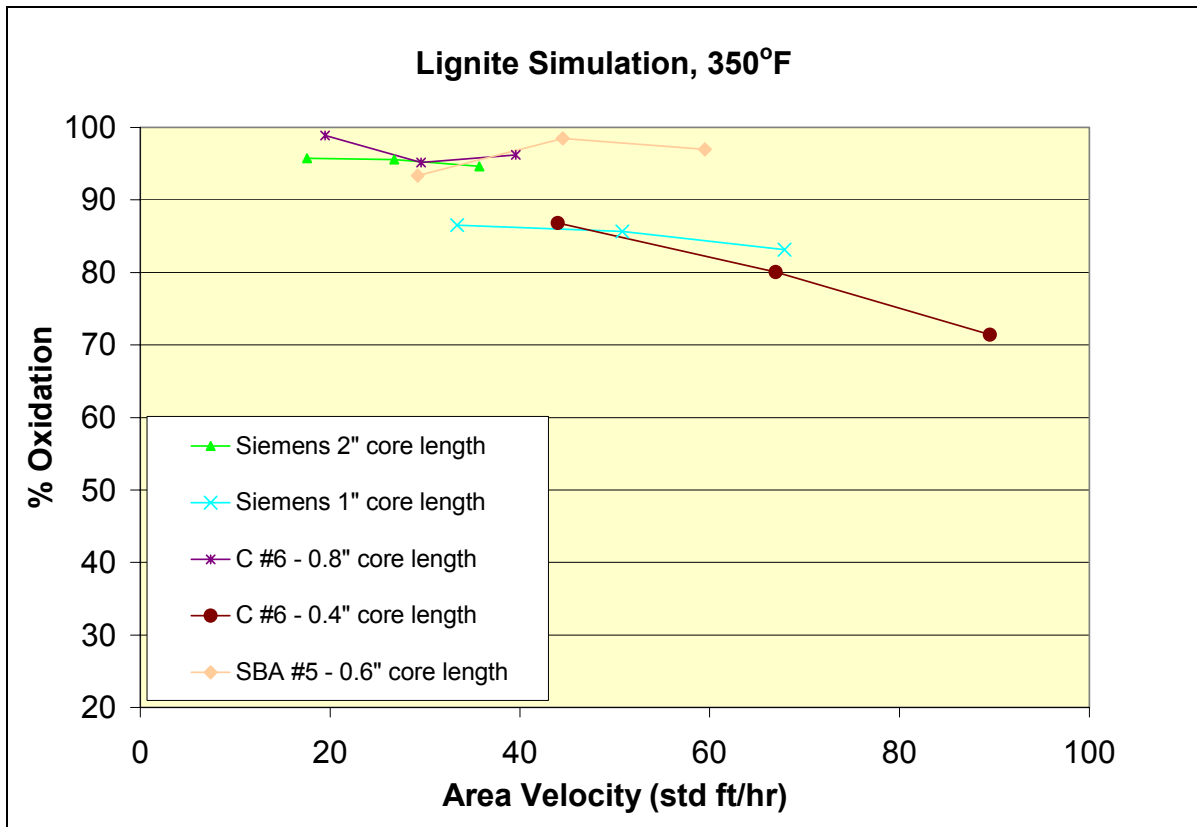


Fig. 8. Effect of Area Velocity on Oxidation of  $\text{Hg}^0$  for Three Catalysts.

When the Siemens SCR and C #6 catalysts were tested at shorter core lengths (1 inch vs. 2 inch for the Siemens SCR catalyst, 0.4 inch vs. 0.8 inch for the C #6 catalyst) to allow operation at higher area velocities, the oxidation performance of each dropped off considerably. It is speculated that at these shorter core lengths, the gas distribution across the face of the honeycomb was adversely affected, lowering the overall oxidation activity from what would have been realized with a more ideal gas distribution.

This speculation illustrates a limitation of the laboratory simulation setup. The catalyst core testing is being conducted with an existing apparatus that was set up to test mercury sorbent and catalyst materials in a sand-bed reactor. To test honeycomb catalyst cores, the sand bed

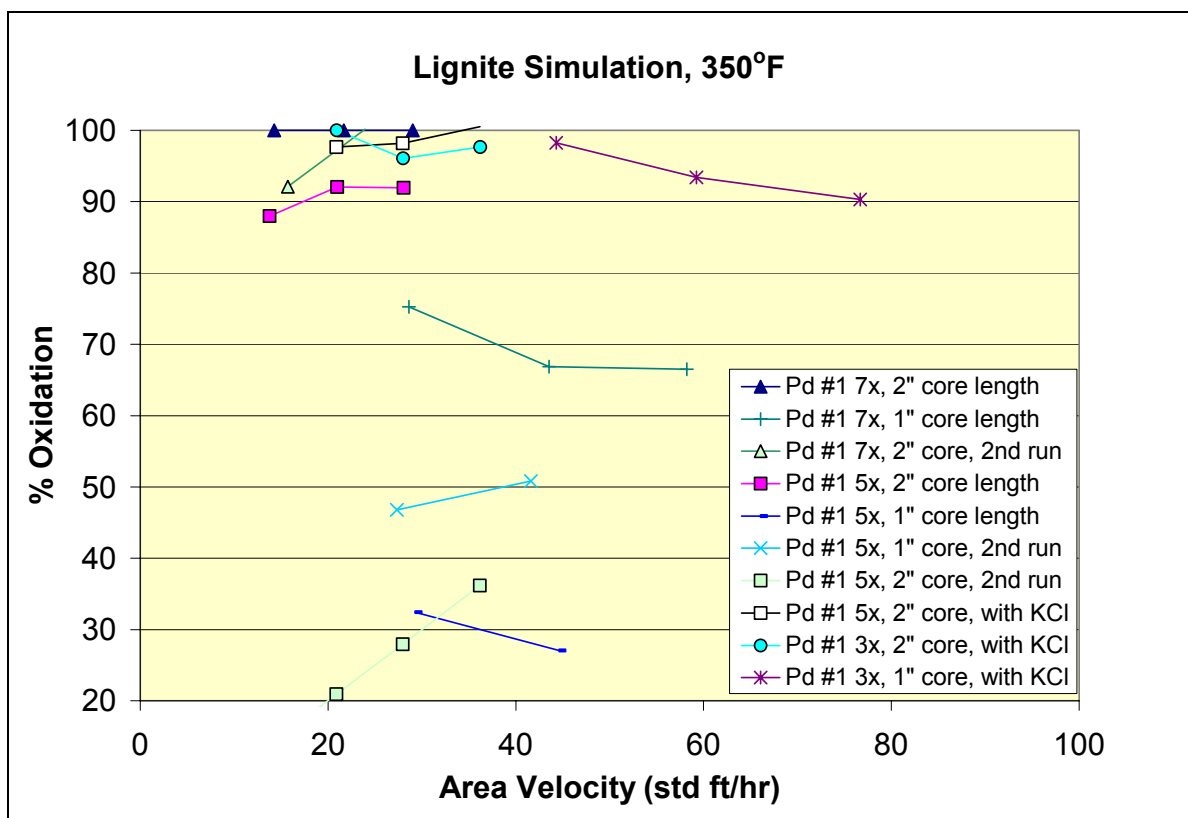


Figure 9. Effect of Area Velocity on Oxidation of  $Hg^0$  for Pd #1 Catalyst.

is replaced with a catalyst core of round cross section, typically about 5/8-inch diameter. The catalysts tested were acquired from the various vendors in whatever pitch was available or could be readily prepared. With the diameter of the catalyst core holder set at 5/8-inch, the core pitch being set, and the gas mixing flow rate of the existing apparatus limited to less than 2 L/min, the only variable that can be adjusted to increase area velocity is the core length. As can be seen in Figure 8, to achieve higher area velocities (approaching 100 standard ft/hr) the corresponding core lengths became very short, particularly for the C #6 and SBA #5 catalyst cores which were available in a very small cell pitch. In spite of this limitation, the data for the longer core lengths for the SCR catalyst and the C #6 catalyst should be adequate for determining catalyst quantities for the pilot unit.

The results for Pd#1 are plotted in Figure 9, for palladium applied at three different loadings on the honeycomb and for two core lengths. These results show a great deal of scatter. Two factors appear to cause this data scatter. One is the core length issue as described above. The 2-inch core length data are probably more representative than the 1-inch data. Another factor is that there appears to have been an interference between some component in the sample gas exiting the Pd #1 cores and the Tris solution that was used to remove oxidized mercury from the sample gas during most of the tests. This apparent interference caused a high degree of variability in the measured catalyst outlet  $Hg^0$  concentrations, and thus caused some catalyst performance results in Figure 9 to be suspect.

A number of Pd #1 tests were subsequently repeated using KCl rather than Tris solution to remove oxidized mercury from the sample gas going to the laboratory analyzer when measuring for Hg<sup>0</sup> concentration. These results are also plotted in Figure 9, and show that for the 3x and 5x palladium loadings (referenced as a multiplier times a baseline palladium loading) showed very high Hg<sup>0</sup> oxidation percentages (>95%) at area velocities up to 40 standard ft/hr. These results are nearly equal to the initial results with the palladium at a 7x loading, which are the first data shown in the legend. The oxidation across a 1-inch core of the palladium at a 3x loading remained above 90% up to an area velocity of nearly 80 standard ft/hr. Based the high oxidation percentage seen with the 3x loadings using the KCl impinger in the analyzer sampling train, it was decided that the 3x palladium loading would be adequate to achieve high Hg<sup>0</sup> oxidation in the pilot unit catalyst.

### Catalyst Sizing for Pilot Unit

The results plotted in Figures 8 and 9 were used to estimate the amount of each of the four catalysts required to achieve high Hg<sup>0</sup> oxidation percentages in the pilot unit. Table 3 shows the planned honeycomb pitch for each catalyst and the overall catalyst dimensions required for the pilot unit as predicted by a mass transfer model previously developed by URS.

**Table 3. Honeycomb Dimensions for Tested in Laboratory and Planned for Pilot Unit.**

Catalyst Type	Cores Tested in Laboratory		Pilot Unit Catalyst (sized for 95% Hg <sup>0</sup> oxidation)				
	Cell Pitch (mm)	CPSI (cells per in <sup>2</sup> )	Cell Pitch (mm)	CPSI (cells per in <sup>2</sup> )	Catalyst Cross-section (in x in)	Catalyst Length (in)	Area Velocity (std ft/hr)
Siemens SCR	4.2	37	4.2	37	35.4 x 35.4	19.7	21
Carbon #6	1.8	211	3.2	64	36 x 36	12	19
SBA #5	1.7	217	3.2	64	36 x 36	12	25
Pd #1	3.2	64	3.2	64	30 x 30	9	49

The model predicts mercury oxidation performance based on a simplifying assumption that mercury oxidation is instantaneous once the mercury has diffused to the catalyst surface. The actual performance of each catalyst in the laboratory simulations was compared to model predictions. If the actual performance was close to the model prediction, this is an indicator of very rapid catalytic oxidation at the catalyst surface; if the actual performance was well below the model prediction, it is an indicator of a slower surface reaction. The actual versus predicted oxidation is compared on the basis of “number of transfer units” (NTU), a mass transfer performance term, rather than on the basis of percent oxidation. This comparison is used to develop an empirical correlation between the lab results and the model prediction, expressing the actual NTU as a percentage of the “theoretical” NTU. This correlation accounts for the differing surface activity among the catalyst types.

The empirically adjusted model was then used to predict the performance of each catalyst at different cell pitches, catalyst cross-sectional area, and catalyst depth. These predictions were used along with the dimensions of the manufacturers’ standard catalyst block sizes to predict

the amount of each catalyst that would be required to achieve 90% Hg<sup>0</sup> oxidation across the pilot reactor.

Table 3 shows the planned cell pitch dimensions for each catalyst and the overall dimensions of each as required for the pilot unit. Note that the planned catalyst sizing is considered to be conservative (e.g., relatively low area velocities). It is hoped that the field performance for these catalysts will be well above 95% oxidation of Hg<sup>0</sup> at design flue gas flow rate conditions. If so, it will be possible to operate the catalysts at somewhat higher flue gas flow rates through the individual catalyst chambers, and hence allow them to operate at higher area velocity values. The flue gas superficial velocities corresponding with the catalyst sizing shown in Table 3 are in the range of typical ESP velocities, or 4 to 6 ft/sec.

Table 3 also includes the dimensions of the honeycomb core samples tested in the laboratory. These data illustrate why the catalyst performance was plotted in Figures 8 and 9 as a function of area velocity rather than space velocity. The cores supplied for lab testing were on substrates with varying cell pitches and wall thickness, and thus varied in active surface area per unit volume. Space velocity is defined as the standard gas flow rate divided by the catalyst volume. At similar space velocities the smaller-pitched catalysts would have more active surface areas than the coarser-pitched catalysts. The area velocity is defined as the standard gas flow rate divided by the catalyst external surface area, and thus better accounts for varied cell pitch and wall thickness. Since any of the catalyst materials could conceivably be applied on any substrate, area velocity is the more equitable parameter for catalyst comparison and is less influenced by the cell pitch at which the particular samples were available for testing.

These core sample data also illustrate the utility of the mass transfer model used to predict catalyst size requirements for the pilot unit. Without such a model, it would have been difficult to project performance results from one cell pitch and core length to the conditions desired for the pilot unit.

Based on results shown in Table 3, Siemens has proceeded with the preparation of a block of SCR catalyst. The completed catalyst module will be shipped to CCS in August. Also in August, Applied Ceramics will begin preparation of 6-inch by 6-inch by 3-inch deep catalyst blocks of the SBA #5 extrusion at the desired cell pitch. They will prepare blocks of this size for both the SBA #5 and C #6 to produce composite catalyst modules for each in the overall dimensions shown in Table 3.

## **CONCLUSIONS**

The planned pilot-scale evaluation of Hg<sup>0</sup> oxidation catalysts has not yet begun, so there are no conclusions to be made from that effort. However, based on laboratory evaluations of sample cores, a number of conclusions can be made about these Hg<sup>0</sup> oxidation catalysts.

The carbon-based and fly-ash-based materials, that had previously been tested only as powders in sand-bed reactors, can apparently be successfully combined with alumina and extruded and fired to produce a reactive honeycomb monolith. The activities of these two

catalyst materials for Hg<sup>0</sup> oxidation are about equal to that of the commercially available SCR catalyst.

The palladium-based catalyst is apparently more active than the other candidate materials. The projected area velocity for the Pd #1 catalyst is about twice that of the other three, indicating that half as much Pd #1 catalyst surface area would be required to achieve similar Hg<sup>0</sup> oxidation percentages compared to the other three catalysts.

The laboratory screening test results suggest that catalyst thicknesses of 6 to 12 inches will be required to achieve high oxidation percentages of Hg<sup>0</sup> in flue gas with these candidate catalyst materials. The upcoming pilot-scale tests with these catalyst materials will confirm their relative activities and lives for Hg<sup>0</sup> oxidation.

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