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Quarterly Progress Report:
High Temperature Membranes for
H₂S and SO₂ Separations
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by

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Purpose and Goals of Research

High temperature membrane separation techniques have been applied to gas mixtures involved in coal utilization. For coal gasification, H_2S has been removed from the syn-gas stream, split into hydrogen, which enriches the syn-gas, and sulfur, which can be condensed from an inert gas sweep stream. For coal combustion, SO_2 has been separated from the flue gas, with concentrated SO_3 produced as a by-product.

Both processes appear economically viable but each requires fundamental improvements: both the H_2S cell and the SO_2 cell require more efficient membranes and the H_2S cell needs a more efficient anode. Membranes will be fabricated by either hot-pressing, impregnation of sintered bodies or tape casting. Research conducted during the present quarter is highlighted, with an emphasis on progress towards these goals.

I. H₂S Removal Cell

Introduction

This electrochemical removal cell cleans a product gas of H₂S by electronating the most easily reduced component of the stream. In coal synthesis gas or natural gas, this is H₂S:

A membrane which contains sulfide ions in a molten state will act to transport sulfide across to the anode. If the membrane is capable of preventing diffusion of hydrogen from the cathode side, an inert sweep gas such as N₂ can be used at the anode to carry away oxidized sulfide ions as vaporous S₂:

Research Summary

We have demonstrated that tape casting technology can be used as an industrial method for membrane manufacture. Work has recently concentrated on the application of this process to polishing H₂S from 100 ppm down to below the analytical limits of the gas chromatograph (approximately 2 ppm) for use with Coal Gasification technology. Use of Molten Carbonate Fuel Cell (MCFC) materials has been examined, and use of more industrial suited cell housing materials, namely 316 stainless steel, has been examined.

A. Experimental Run Results

Run 44

This was our first attempt at the application of this technology to coal synthesis gas H_2S polishing. The tapes were manufactured using established tape casting methods. 16.5 grams of MgO , 83.3 grams of Metoramics binder system K565-4, 1 gram of Metoramics releasing agent M-1111, and 1 gram of Metoramics surfactant M-1114 were ball milled for 24 hours and then poured out on M-1111 coated glass. After the solvent had volatilized out, eight 3 inch diameter membranes were cut from the tape, each with a thickness of approximately 0.03 cm.

The actual membrane used in this experimental run was a combination of three of these membranes layered together with the acrylic binder burned out under an O_2 atmosphere. Eutectic composition Li/K carbonate (62% Li, 38% K) was then wicked into the membrane in-situ. Both the anode and cathode were lithiated NiO .

The composition of the gas fed to the cell was 13.2% CO_2 , 33.2% H_2 , 48.8% CO , 4.8% H_2O , and 100 ppm H_2S (assuming that the gas has gone to its equilibrium composition). This corresponded to an electrolyte equilibrium composition of 99.6% carbonate and 0.4% sulfide at an operating temperature of 650°C. Since sulfide is the current-carrying species, this low equilibrium sulfide level brought up the question of whether the process limiting current density would still be determined by gas phase mass transport of H_2S to the electrode/electrolyte interface (as in all previous experiments) or if it would now be determined by transport of sulfide ions through

the molten electrolyte. A theoretical comparison of the transport in the molten electrolyte to transport in the gas phase was performed in the following manner:

Current carried through the electrolyte by sulfide ions is described by,

$$i_d = 2FD\rho\Delta C/\Delta x \quad (7)$$

where ρ is the molar density of the electrolyte in the ceramic membrane (0.0208 gmole/cc), ΔC is the sulfide concentration gradient through the electrolyte (0.0044) assuming equilibrium sulfide concentration at the cathode and essentially zero concentration at the anode. Δx is the membrane thickness (0.1 cm), D is the sulfide diffusivity (approximated at 10^{-5} cm²/sec), and F is Faraday's constant 96500 coulombs/equivalent. These conditions give a membrane limiting current density of 1.8 mA/cm².

Current carried by mass transport of H₂S to the electrode surface through the gas phase is calculated using the following equation:

$$i_d = 2Fkp(y_{in} - y_{out})/\ln(y_{in}/y_{out}) \quad (8)$$

where k is the mass transfer coefficient and is approximated based on channel geometry and gas temperature, density, and viscosity ($k = 12.9$ cm/sec), ρ is the molar density of the gas phase (1.32×10^{-5} gmole/cm³), $y_{in} = 9.91 \times 10^{-5}$, and $y_{out} = 1 \times 10^{-5}$. These conditions give a gas phase limiting current of 1.3 mA/cm². Thus, sulfide transport limitations in the gas phase are seen to be comparable to those in the electrolyte.

The run itself was terminated before any removal data was collected. Several factors contributed to the decision to terminate the experiment. First of all, the

control relay on the furnace froze shut, causing the temperature to overshoot the acrylic burnout temperature of 350°C by over 300°. This damaged the membrane before the electrolyte was added and allowed process gas to cross over and mix with the sweep gas. The gas chromatograph results also became highly variable. It was found that a piece of the injection port septum broken off and fallen into the injection chamber, interfering with carrier gas flow. The GC was repaired and the run terminated due to gas cross-over through the damaged membrane.

The cell was shut down after 14.8 hrs of operation.

Run 48

A properly operating cell will transport carbonate in the absence of H₂S. It was decided to test for CO₂ transport in this experiment. Transport of CO₂ across the cell takes place by the following reactions:

at the cathode and:

at the anode. Thus, CO₂ removed at the cathode and produced at the anode should stoichiometrically match the applied current. An alternative current carrying mechanism was proposed in our July 1, 1991 to Sept. 30, 1991 report. That mechanism is outlined below:

If this mechanism was taking place parallel to carbonate transport then CO_2 would not be removed stoichiometrically with two Faradays of current.

This experimental run was the fourth run using coal gas and used two tapes of MgO , manufactured in a manner identical to that used in Run 44, and one mat of zirconia cloth as the membrane matrix material. The electrodes were both lithiated NiO . The acrylic binders were burned out under an O_2 atmosphere and the Li/K eutectic composition electrolyte was added with the cell at run temperature.

$\text{CO}_2/\text{H}_2\text{O}$ species removal from the process gas as a function of applied current is shown in Figure 1. Examination of this figure shows that the removal of species is stoichiometric across the range of applied currents examined. Figure 2 shows the exit anode gas CO_2 level as a function of applied current. Examination of this figure shows that CO_2 removed from the cathode process gas was transported across the cell and produced at the anode. The calculated CO_2 levels, based on current carried by the oxidation of the CO_3^{2-} ion, tracks the observed CO_2 values very closely. Figure 3 shows the overpotential of the cell with applied current. Examination of this data shows that the cross-cell potential crosses the -1 volt cathode to anode threshold potential required for carbonate transport at lower currents than previously seen with high levels of H_2S present (>1000 ppm). We see that in the higher CO_2 concentrations found in coal gasification synthesis gas, CO_2 removal is stoichiometric with applied current.

When these results are compared to the CO₂ removal study done in Run 42 (July 1, 1991 to Sept. 30, 1991 report), we see that at lower CO₂ levels (1.4% CO₂ in natural gas as compared to 13% CO₂ in coal gas) the possibility of O²⁻ transport may exist. The higher CO₂ content in coal gas provides enough CO₂ to the electrolyte-/electrode interface that the highly basic O²⁻ ion is attacked by CO₂ to form CO₃²⁻. Thus, even if NiO is acting as a catalytic species for production of the O²⁻ ion as suggested by the results from run 42, in this application the higher CO₂ levels cause the formation of CO₃²⁻ and CO₂ is still transported stoichiometrically with applied current.

Instead of adding eutectic Li/K carbonate as a powder and allowing it to melt down the reference electrode hole as in previous runs, electrolyte was melted external to the cell and poured directly down the reference electrode hole in this experiment. Cell seals were initially excellent on both sides of the cell, but flow out the anode side of the cell eventually stopped. Shortly after the anode flow stopped, cell cross-flow was observed and the cell run was terminated. A post mortem examination of the cell showed that electrolyte had run out of the membrane and into the anode side flow channel and down the tube. Once gas flow had forced a bubble of electrolyte outside the furnace it froze and blocked the tube, clogging the flow. The run was then terminated after 6.75 hours.

Figure 1.
CO₂/H₂O Species Removal vs Current.

Figure 2.
Anode CO₂ Production vs Applied Current.

Figure 3.
Cell Overpotential vs Applied Current.

Summary

Marginal success with the application of tape casting technology for the manufacture of membranes for the polishing of H_2S from simulated coal gasification process streams has been shown. No evidence was found to show the presence of the O^{2-} transport mechanism proposed in the previous report at these CO_2 concentration levels.

Planned Work for Next Quarter

Work will continue on the application of tape cast membranes to the polishing of coal gasification process streams. Attempts at achieving equilibrium sulfide levels in the electrolyte will be made in an effort to show the full electrochemical removal effects of this technology. Steady state operation of a tape cast cell membrane within a stainless steel housing will be the main goal.

II. SO₂ Removal Cell

Introduction

The key to successful application of this technique is the development of a membrane capable of supporting a current density of 50 mA/cm² at total voltages near 1 Volt. Flooding of electrode pores by electrolyte has been identified as a problem, leading to increased polarizations over time. To reduce flooding, new materials and fabrication methods will be investigated to produce an homogeneous, theoretically dense membrane. The matrix material must have the proper particle size distribution to develop sufficient capillary forces to prevent the electrodes from withdrawing the electrolyte.

Three techniques will be used for construction of the membranes or 'tiles'. The first is 'hot-pressing' in which powders of matrix material and electrolyte are intimately mixed in a dry atmosphere. The second technique will involve manufacturing a sintered ceramic matrix without electrolyte present and then 'wicking' the molten electrolyte into the matrix voids by capillary action. In the third method, tape-casting, the matrix material, dispersed along with an organic binder in a liquid, is continuously formed into a dried tape on a substrate, maintained at constant thickness by an overhead doctor blade. The flexible, 'green' tape is cut to the desired size, the binder is burned out, and the matrix is sintered, leaving a rigid body of the porosity needed for the final tile. Electrolyte is then infiltrated at operating temperature.

Research Summary

Work during the quarter concentrated on full cell testing to determine SO_2 and SO_3 removal capability and polarization performance. An effort was made to determine the possibility of using stainless steel housings and flow tubes as removal cell components. A minor emphasis was placed on electrode pore characterization by way of mercury porosimetry.

To evaluate stainless steel for the housings, a galvanostatic test was performed between two electrodes of SS 316, one pre-oxidized at 525°C , in a crucible of molten 5 wt.% V_2O_5 in $\text{K}_2\text{S}_2\text{O}_7$ to ascertain if SS 316 housings would survive the aggressive environment of electrolyte. Both pieces had about 2 cm^2 exposed to the electrolyte and handled up to 1000 mA with 12 V between them. At this extremely high current density, no pitting or electrical conductivity decrease was observed, but crystals of a sulfide did precipitate on the cathode. It is possible to reduce pyrosulfate to sulfide, but only under extreme voltage/current conditions such that oxygen or sulfur trioxide cannot get to the electrode surface. This phenomenon has never been seen in full cell studies with gas flowing past the electrodes. These new housings provide increased electrical contact with the electrodes along the $1/8$ " wide baffles, compared to the 0.010" (0.25 mm) Pt wire coil.

A full cell test performed during October, 1991, used a 49 volume percent silicon nitride tape cast matrix with 10 wt.% V_2O_5 in $\text{K}_2\text{S}_2\text{O}_7$ electrolyte, lithiated NiO electrodes and Macor housings with platinum leads. Some difficulty was encountered during start-up with insufficient electrolyte loading, but was corrected with

electrolyte additions through the reference electrode port in the top housing. Also, an excessive pressure drop was detected through the pre-oxidation catalyst bed at high flow rates.

Figure 4 shows that the cathode performs well at 100 mA applied current (5 mA/cm²), but the anode shows an extreme performance decline over time. At the same time that the anodic overpotential increased, so did the anode to reference electrolyte resistance. This increase in resistance may be due to a loss of contact area between the electrode and lead wire, but was fully reversible at open circuit. Figure 5 shows that when electrolyte is added while current is applied, the overpotential at the anode drops significantly, confirming the loss of contact area as the cause of performance decline. The loss is due to either consumption of electrolyte or the formation of a frozen sulfate film between the electrode and lead or electrode and membrane. In either case, the current density is highest at this contact point, since the platinum leads are 0.25 mm in diameter. Later runs with the same cell configuration showed a much lower rate and magnitude of anodic polarization increase, most likely due to proper electrolyte content.

However, operation with 50 or 100 mA did produce some SO₂ at the cathode outlet, a possibility from the electrochemical reduction of pyrosulfate:

Generation of SO₂ was 11% at 50 mA and varied from 17% to 24% at 100 mA, relative to inlet flow rate. For SO₃, removal (relative to equilibrium conversion of inlet

Overpotentials

100 mA; no SO₂ detected

Figure 4. Polarization performance of NiO electrodes with 49 v/o ceramic membrane, 400° C.

Overpotentials

100791; 100 mA
 Figure 5. Polarization of electrodes, showing effect of electrolyte addition during applied current.

SO₂) at 50 mA was 89% and 80% to 83% at 100 mA for a flow rate equal to 90% stoichiometric removal. One of the electrodes was kept in a vacuum desiccator and later analyzed with Electron Surface Characterization Analysis (ESCA). Investigation of the resulting peaks showed no Ni-S bonds present, negating the possibility of nickel sulfides or sulfates as corrosion products.

Once the new SS 316 housings had been machined, they were subjected to full cell evaluation. Inlet plumbing connections were made with Swagelok fittings and 1/4" SS 304 tubing. This run and the next one used a plug of platinized silica gel for the SO₂ pre-oxidation catalyst and suffered from excessive pressure drop at high flow rates. The outlet tubing consisted of a special flexible stainless bellows tube fused to a glass outlet line. This special fitting was required lest the effluent SO₂ would condense outside of the oven and possibly corrode the tubing, causing failure of the mass balance. Resistance measurements between the inlet tube and electrode were of the order of 3 ohms at room temperature, showing an exceptional conductivity.

A tape cast membrane was placed between the electrodes and slowly heated to 400° C in oxygen to burn out the polymer. Unfortunately, the membrane cracked, leading to excessive crossover of gases. Several other attempts identified that the polymer was softening during heat-up and sticking to the housings. This difficulty was fixed by pre-heating the tape on a glass plate to 200° C to remove residual solvent and a portion of the polymer.

IR-corrected voltage - current characteristics at low currents were obtained and are shown in Figure 6. A linear extrapolation of the slope for the lower overpotential

Polarization Performance

Figure 6. Polarization behavior of lithiated NiO after 15 mins. applied current.

portion of the curve showed that the exchange current density, i_0 , at 400° C,

$$i_0 = \frac{RT}{nF} \left(\frac{di}{d\eta} \right) = 0.029V \left(\frac{di}{d\eta} \right) \quad (2)$$

was 0.072 mA/cm² at the cathode and 0.063 mA/cm² at the anode. These figures compare favorably with the exchange current densities of Franke¹, who found a value of 0.030 mA/cm² in his full cell studies with perovskite electrodes and MgO-based membranes. Free electrolyte studies by Scott² determined an i_0 value of 30 mA/cm². Although there is some modest improvement with the present configuration, work is still required to boost the exchange current density to an adequate level.

The last full cell test of the quarter used a modified cell housing. A 1/4" diam. hole was drilled through the top housing into the gas flow channel. The modification was made so that molten electrolyte would be added onto the top electrode and distribute itself across the electrode and then into the membrane. The previous full cell experiment showed that when electrolyte was added through the reference port, some of the electrolyte flowed out from between the housings, resulting in lower electrolyte loadings.

The runs recounted above had high pressure drops through the pre-oxidation catalyst beds, so to fix this problem, the 1/4" tubing was cut and fitted to a 3/8" tube which contained the catalyst. This increase in diameter resulted in a 2.9x increase in cross-sectional area and a substantially decreased pressure drop at all flow rates investigated. However, some SO₂ bypassed the catalyst at the higher flows. A test of

bypass versus flow rate for the 1.745 g of 2% platinum-on-silica-gel used is shown in Figure 7. The curve shows a definite mass transfer limitation of SO_2 to the catalyst surface. The asymptotic rate of reaction approaches 95×10^{-6} moles/min for this loading or 54×10^{-6} moles/min/g catalyst. Further experiments will use a higher mass of catalyst to increase conversion at higher flow rates.

A graph of instantaneous ohmic resistance in the electrolyte versus applied current for this cell, Figure 8, showed a dependence of ohmic loss with current. This can be attributed to capacitive effects at both electrodes. The "capacitors" in the electrolyte discharge, creating larger voltage drops. These measured voltage drops are then divided by the value of the previously applied current to give the resistance.

Plots of SO_3 removal and SO_2 generation rates versus applied current show that as current increases, removal deviates significantly from stoichiometric levels. Figure 9 shows that SO_3 removal is approaching a mass transfer limit. The curve for SO_2 generation accounts for SO_2 slip through the pre-oxidation catalyst.

The decrease in SO_3 removal seems to be due to a mass transfer limitation at the electrode. Calculation of the mass transfer of SO_3 from the gas to the electrode, using a standard correlation for this cell configuration³

$$N_{Sh} = \frac{k_m D_{eq}}{D_{AB}} = 5.60 \quad (3)$$

shows that the gas-phase mass transfer limiting current is 50 mA/cm². This value is higher than the maximum applied current of 300 mA (15 mA/cm²), superficial area

SO2 Bypass

Mass of catalyst = 1.745g
 Figure 7. Bypass of SO₂ through pre-oxidation catalyst bed; 400° C; 0.3% SO₂, 3.0% O₂ in N₂.

Figure 8. Instantaneous ohmic resistance versus applied current, measured from cathode to anode.

Figure 3. SO₂ generation and SO₃ removal rates as a function of applied current; 400° C. 0.3% SO₂, 3.0% O₂ in N₂ at a flow rate such that current equals 90% removal of SO₂.

basis) and therefore the cell should be able to support these high currents without modification. Some other cause, such as limited transport of sulfate ions to the gas/electrolyte interface, is responsible for the SO_3 removal limitation.

The increase in SO_2 generation is due to a combination of factors. First, SO_2 generation is directly proportional to applied current, but does not follow the same slope in Figure 9. Second, as current increases, so does the gas flow rate, which means any gas escaping the membrane has a lower residence time and therefore lower contact time with the V_2O_5 in the electrolyte, however, the slope again does not follow that of the stoichiometric curve in Figure 9. Third, V^{5+} in the vanadium complex is reduced by SO_2 to V^{4+} , which has limited solubility in the melt. If this complex is not re-oxidized by gaseous oxygen, precipitation occurs, reducing the amount of catalyst available for SO_2 oxidation. This appears to be the case with the data.

Calculations of the rates of absorption of SO_2 and O_2 into vanadia-pyrosulfate melts and the re-oxidation of V^{4+} were performed to see if the melt chemistry is limiting removal performance. The work of Holroyd and Kenney⁴ for SO_2 absorption into films of molten $\text{V}_2\text{O}_5/\text{K}_2\text{S}_2\text{O}_7$ can be applied to the present process to show that the initial rate of absorption is 36×10^{-8} gmol/cm²/min. At 100 mA of current, SO_2 is electrochemically generated at the rate of 15.5×10^{-6} gmol/min. Complete absorption, assuming total escape of generated SO_2 , would require a melt surface area of 430 cm², or an electrode surface area enhancement of 21.5. These numbers increase linearly with increasing current. Holroyd and Kenney also investigated the reoxidation of V^{4+}

in the same melts⁵ and found that V^{4+} has a maximum solubility of 4% at 400° C. From their reaction constants, the rate of reoxidation in this system can be calculated as 14.5×10^{-6} gmol/cm²/min for 4% V^{4+} . With the rate of reoxidation being orders of magnitude faster than the absorption of SO_2 , this step cannot be limiting.

However, Mars and Maessen⁶ determined that there was a deviation from their normal kinetic behavior at lower temperatures ($T < 415^\circ \text{C}$) for sulfuric acid catalyst pellets. They attributed this to either an increase in the rate-retarding effect of SO_3 or the formation of sulfovanadates, which would stabilize V^{4+} , reducing the rate of reoxidation. Both the rate of SO_2 absorption and the rate of reoxidation could be limiting and therefore an experiment will be conducted next quarter to determine the conversion of SO_2 to SO_3 bubbled through molten 10 wt.% $V_2O_5/K_2S_2O_7$.

In addition to these runs with sulfur oxides, the cell was also used with NO and NO_2 gas mixtures. The Thermo Electron Model 10 NO_x Analyzer was calibrated with a known concentration of NO and then used to analyze the other gas cylinders. The NO mix used contained 989 ppm NO and 2 ppm NO_2 in N_2 , while the NO_2 mix was 7 ppm NO and 806 ppm NO_2 in N_2 . These gases were passed through the platinized silica gel catalyst and then over the working electrode. Outlet gas was diluted with dry air, analyzed and the reading corrected for dilution effects. Current was applied to the working electrode, either in reducing or oxidizing fashion.

The first NO study used 78 cc/min of SO_2 mix blended with 235 cc/min of NO mix, such that the reduction reactions would have some SO_x reagent available. However, the gas was set to handle 25 mA, but the current reached levels of 200 mA

to see if NO could be electrochemically reduced also. Figure 10 shows that NO and NO_x levels increase to a steady-state value of about 660 ppm, with or without reducing current, a removal of 10%. The initial lower concentrations are due to adsorption of NO and NO₂ on the surfaces of the flow system. This trend was confirmed in another run using 233 cc/min NO mix and 78 cc/min SO₂ mix, documented in Figure 11. A similar progression of reducing currents was used with 94 cc/min SO₂ mix and 233 cc/min NO₂ mix, as shown in Figure 12. Again, reducing current had no effect on NO or NO_x concentrations after adsorption reached equilibrium, with a net removal of 15%.

With oxidizing current applied to the working electrode, little change was seen until the SO₂ mix was turned off. At this time in Figure 13, the NO_x levels jumped due to a simple concentration effect. When the oxidizing current increased from 200 mA to 500 mA, the NO and NO_x concentrations dropped immediately to 40 ppm NO and 75 ppm total NO_x, a removal of 91% of NO_x. The outlet tubes were cleaned in the middle of this precipitous drop to ensure that removal was due to chemistry, not absorption of NO_x into condensed SO₃. After the levels were constant, the polarity of the current was switched to reducing, which caused an immediate increase in NO/NO_x concentrations. Under these conditions, removal was still 15%.

These findings again confirm that the cell has no obvious electrochemical effect on NO or NO_x. However, all studies showed removal of 10% to 15%, with or without current. When oxidizing current was applied, dramatic decreases in NO and NO_x concentrations were seen. Previous NO_x tests have demonstrated 12% removal over

Inlet NO = 742 ppm

Figure 10. Overpotentials and NO/NO_x levels with reducing current and NO mix. Current applied at 25, 50, 100 and 200 mA.

Inlet NO = 745 ppm

Figure 11. Overpotentials and NO/NO_x levels with reducing current and NO mix. Current applied at 25, 100 and 200 mA.

Initial = 580 ppm NOx

Figure 12. Overpotential and NO/NO_x levels with reducing current and NO_x mix. Current applied at 25, 50, 100, 200 and 300 mA.

Inlet NO₂ = 600 ppm, then 820 ppm
 Figure 13. Overpotentials and NO/NO_x levels with oxidizing and reducing current.

perovskite electrodes without current.

Mercury intrusion porosimetry was performed on the ERC and Fibrex electrode materials. The details of the ERC material are unpublishable due to a secrecy agreement. A plot of volume % v. pore radius, Figure 14, shows that the vast majority (87%) of the pores are greater than 1 micron in diameter ($r = 0.55\mu$). Some of compiled data, Table I, also showed that 57% of the pores were $\geq 4\mu$ in diameter, resulting in a good match with the electrolyte membrane structure.

Final experimentation of this quarter included a determination of the feasibility of using commercial sulfuric acid catalyst (Haldor-Topsoe VK38) in the gas channels of the flow cell to convert the electrochemically generated SO_2 to SO_3 for removal. A tests was performed using thin cylinders of catalyst trimmed from 6 mm x 6mm catalyst pellets. This test showed a mass transfer limitation to the reaction rate (the horizontal asymptote in Figure 15) with a maximum rate of 10×10^{-6} moles SO_2/min or 30×10^{-6} moles $\text{SO}_2/\text{min/g}$ catalyst at 400°C . At 375°C , this reaction was 5% slower, Figure 16.

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Figure 14. Mercury Porosimetry Curve for Lithiated and Oxidized Fibrex 50/50 mat.

Table I. Data for Mercury Porosimetry of Fibrex 50/50 mat.

Pore Radius, μm	Pore Diameter, μm	Intruded Volume, cc/g	% Volume Intruded
4.727	9.454	0.0139	3.8
1.947	3.894	0.2069	56.63
1.241	2.482	0.2579	70.58
0.8944	1.7888	0.2883	78.89
0.6942	1.3884	0.3052	83.52
0.5546	1.1092	0.3177	86.93
0.4554	0.9108	0.3264	89.33
0.3878	0.7756	0.3319	90.84

SO₂ Reaction Rate for VK38 Pellets in Flow Channels; 400°C

0.358 g VK38 (thin cylinders)

Figure 15. Rate and percent conversion of SO₂ over thin cylinders of VK38 catalyst at 400°C.

SO₂ Reaction Rate for VK38 Pellets in Flow Channels; 375°C

0.358 g VK38 (thin cylinders)
Figure 16. Rate and percent conversion of SO₂ over thin cylinders of VK38 catalyst at 375°C.

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