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ADVANCED SEPARATION TECHNOLOGY FOR FLUE GAS CLEANUP

Abhoyjit S. Bhown
Neeraj Pakala
Tracy Riggs
Troy Tagg

Kamalesh K. Sirkar
Sudipto Majumdar
Debabrata Bhaumick

SRI International
333 Ravenswood Avenue
Menlo Park, CA 94025

New Jersey Institute of Technology
University Heights
Newark, NJ 07102

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CONTENTS

DISCLAIMER

FIGURES

TABLES

INTRODUCTION

The objective of this work is to develop a novel system for regenerable SO_2 and NO_x scrubbing of flue gas that focuses on (a) a novel method for regeneration of spent SO_2 scrubbing liquor and (b) novel chemistry for reversible absorption of NO_x. In addition, high efficiency hollow fiber contactors (HFC) are proposed as the devices for scrubbing the SO_2 and NO_x from the flue gas. The system will be designed to remove more than 95% of the SO_x and more than 75% of the NO_x from flue gases typical of pulverized coal-fired power plants at a cost that is at least 20% less than combined wet limestone scrubbing of SO_x and selective catalytic reduction of NO_x. In addition, the process will make only marketable byproducts, if any (no waste streams).

The major cost item in existing technology is capital investment. Therefore, our approach is to reduce the capital cost by using high efficiency hollow fiber devices for absorbing and desorbing the SO_2 and NO_x. We will also introduce new process chemistry to minimize traditionally well-known problems with SO_2 and NO_x absorption and desorption. For example, we will extract the SO_2 from the aqueous scrubbing liquor into an oligomer of dimethylaniline to avoid the problem of organic liquid losses in the regeneration of the organic liquid. Our novel chemistry for scrubbing NO_x will consist of water soluble phthalocyanine compounds invented by SRI and also of polymeric forms of Fe^{++} complexes similar to traditional NO_x scrubbing media described in the open literature. Our past work with the phthalocyanine compounds, used as sensors for NO and NO_2 in flue gases, shows that these compounds bind NO and NO_2 reversibly and with no interference from O_2, CO_2, SO_2, or other components of flue gas.

The final novelty of our approach is the arrangement of the absorbers in cassette (stackable) form so that the NO_x absorber can be on top of the SO_x absorber. This arrangement is possible only because of the high efficiency of the hollow fiber scrubbing devices, as indicated by our preliminary laboratory data. This cassette (stacked) arrangement makes it possible for the SO_2 and NO_x scrubbing chambers to be separate without incurring the large ducting and gas pressure drop costs necessary if a second conventional absorber vessel were used. Because we have separate scrubbers, we will have separate liquor loops and deconvolute the chemical complexity of simultaneous SO_2/NO_x scrubbing.

We will conduct our work in a 60-month period (5/92 to 4/97), encompassing 16 tasks (Table 1), beginning with studies of the fundamental chemistry and of the mass transfer characteristics of small HFC modules in the laboratory. We will then examine the most favorable method of SO_2 liquor regeneration, determine the ability of the HFC devices to withstand particulate matter,

and examine the behavior of scalable modules. In the final 15 months of the program, we will determine the fundamental mass transfer behavior of a subscale prototype system. Based on these data, a computational design model will be devised to guide further scaleup efforts that may follow.

Table 1

PROJECT TASKS AND SCHEDULE

Task Number	Title	Duration
1	Project Definition	5/92 – 8/92
2	Capacity, Reversibility and Lifetime	7/92 – 6/94
3	Chemical Synthesis	7/92 – 6/94
4	SO_2 Scrubbing with HFCs	7/92 – 9/93
5	NO_x Scrubbing with HFCs	2/93 – 3/94
6	SO_2 Liquor Regeneration	7/93 – 9/94
7	Particle Deposition	8/93 – 9/94
8	Integrated NO_x Life Tests	8/94 – 1/96
9	Scalable Modules	9/94 – 1/96
10	Computational Model	3/95 – 1/96
11	Construction of Subscale Prototype	2/96 – 4/96
12	Operation of Subscale Prototype	5/96 – 4/97
13	Refinement of Computational Model	1/97 – 4/97
14	Economic Evaluation	Various
15	Reporting	5/92 – 4/97
16	Chemical Synthesis for Process Scale-up	5/94 – 4/96

SUMMARY OF QUARTERLY PROGRESS

During the third quarter of 1995, we continued work on Task 8. We also obtained some mass transfer data on Task 9.

In Task 8, we gathered additional 400 hours of NO_x absorption/desorption data. We also presented the data on NO_x absorption/desorption behavior over cumulative time to date (650 hours). The performance indicates unchanged reversible characteristics of Co(II) phthalocyanine solution. Therefore, we believe that NO_x absorption/desorption chemistry is robust.

In Task 9, we experimentally observed the channeling of liquid flow due to poor design of previous rectangular modules. Newly obtained welded rectangular modules out-performed previous modules. We also presented SO_2 absorption data using different Na_2SO_3 concentrations. The SO_2 absorption seem to be dependent on liquid flow rate, a rather surprising result compared to earlier results. Apparatus for combined absorption/desorption of SO_2 in rectangular modules is also given in this report.

TASK 8: INTEGRATED NO$_X$ LIFE TESTS

Liquid scrubbing systems for NO$_x$ have traditionally been plagued by degradation of performance over time with subsequent need for convoluted liquor regeneration schemes. Therefore, it is essential to determine whether or to what extent there is any loss with time in the performance of the NO$_x$ absorption/desorption system. We have previously shown that Co(II)-phthalocyanine is a promising candidate for NO$_x$ absorption in the presence of O$_2$. We have also demonstrated the superior mass transfer characteristics of the Co(II)-phthalocyanine scrubbing system in a 300 fiber HFC. Consequently, the objective here is to devise a continuously operating system for determining if the NO$_x$ absorption/desorption chemistry has the potential to be commercially robust.

As described in the Quarterly Technical Report #13, we experienced problems with the NO$_x$ analyzer. These problems were fixed with the help of a Technical Service staff at SRI. After restarting the system, we noticed that the removal of NO from the feed stream by the HFC was very poor. Examination of the contactor revealed that solid precipitate had clogged fiber pores, resulting in lower mass transfer area. Cleaning of the contactor proved a much harder job than originally anticipated. After thorough cleaning, we noticed a water leak from the shell side to the tube side. This is either due to a broken fiber, due to hydrophylic pores. Subsequent testing of the contactor pointed to the change in characteristics of the fiber; the breakthrough pressure was 5-10 psig for water. If a fiber was broken, the breakthrough pressure would be much less than this. A new hollow fiber contactor was located in our lab and tested. The new contactor has 1155 fibers, and has slightly different liquid flow configuration. The breakthrough pressure of this contactor is approximately 70 psig. This contactor was installed into the NO$_x$ absorption/desorption system.

Meanwhile, we also synthesized about 40 gms of the 4-4'-4"-4''' tetra sodium salt of Co(II)-sulfophthalocyanine for our future requirements.

We leak tested the apparatus with a 1150 fiber HFC and determined that it was suitable for our purposes. We resumed testing of the NOx scrubbing solutions and logged an additional 120 hrs (see Figure 1). The horizontal (broken) line in the figure represents the NO concentration in the feed stream to HFC. The top and bottom curves represent NO concentration in the gas streams from absorption (HFC) and desorption devices, respectively. Note that at steady state the sum of NO concentrations exiting absorption and desorption devices must equal to feed concentration.

4
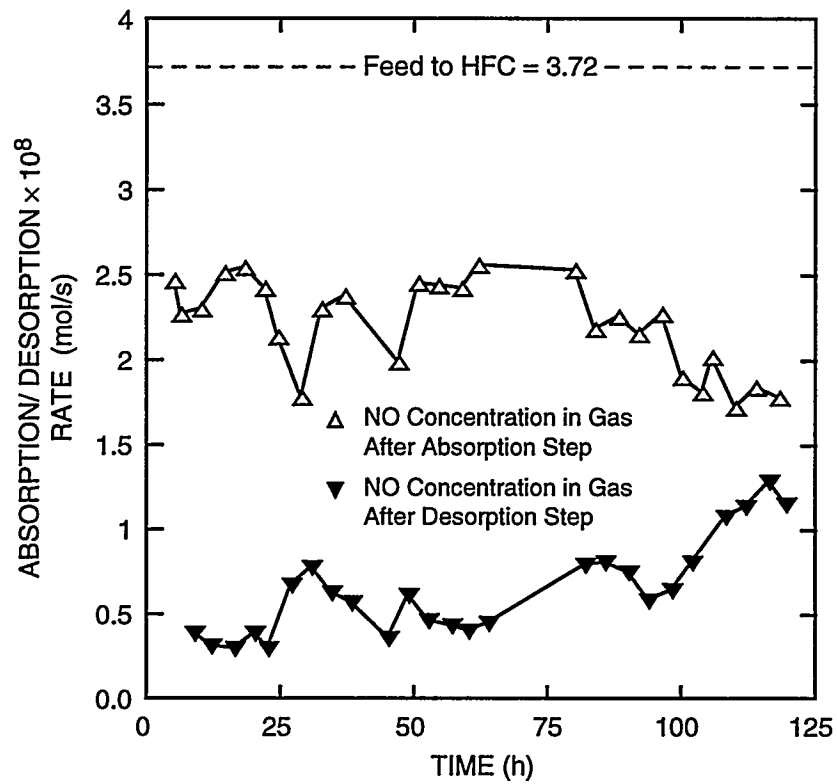
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Figure 1. NO$_x$ scrubbing lifetime studies (Aug. 1995).

5

Therefore, from the figure, it can be inferred that the process is at an unsteady condition and with increase in time, the scrubbing solution capacity for NO absorption is also increasing. The NO removal after 120 hours of operation is about 54%. This behavior demonstrates that the solution is not deactivating. So far, the current scrubbing solution has been on-stream for about 240-250 hours.

For the existing experimental system, we have installed some additional safety features (i.e. check-valves) that will allow us to run for longer continuous periods of time. We collected additional 400 hours of experimental data using the modified NO_x absorption/desorption system during the later parts of the reporting period. Figure 2 shows the overall lifetime of the solution to date. Figure 2 indicates the absorption rate in the HFC and also the desorption rate in the stripping device. Despite some noise due to either change of contactor or shut-off of the apparatus, the trend in absorption/desorption behavior is steady. Therefore, we believe that the Co(II)-phthalocyanine solution performance is unchanged during this test period (650 hours). The low removal rates can be attributed to the inefficiency of the stripping device; the rate limiting step. In summary, the lifetime studies on CO(II) phthalocyanine solution are very promising and give us the confidence to test the system in a bigger scale.
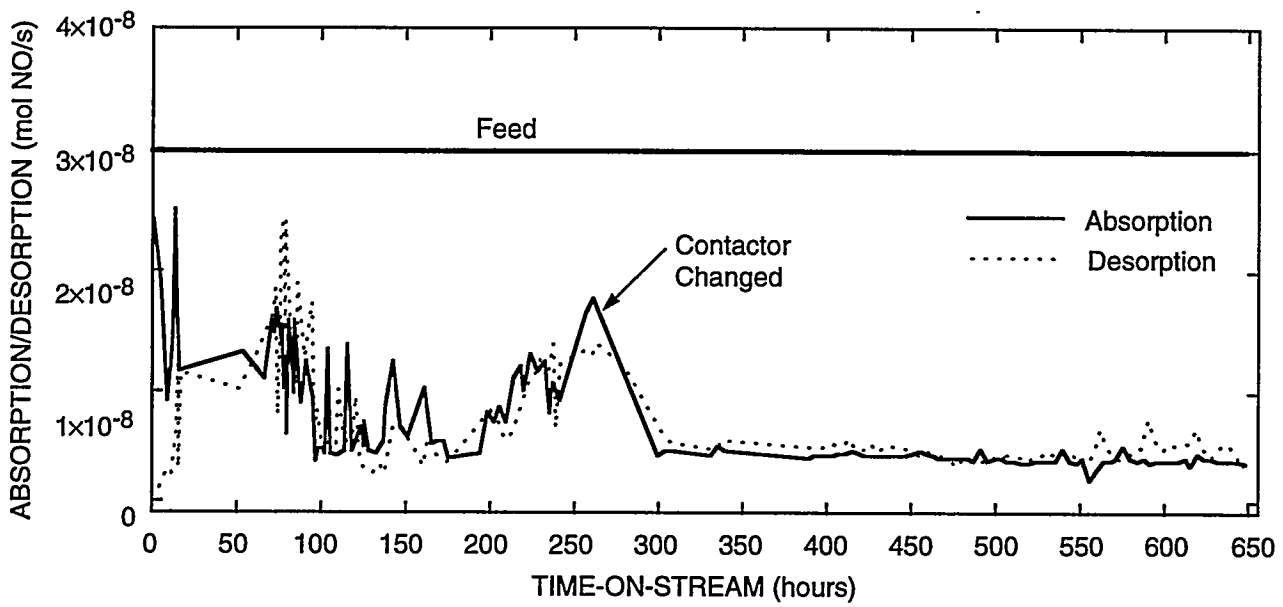
Figure 2.   NO$_x$ scrubbing solution lifetime data.

7

TASK 9: PERFORMANCE OF SCALABLE MODULES

Because of the need for billions of (approximately 30-cm long) fibers to treat the flue gas from a 500 MW(e) plant, it is critical to establish the mass transfer characteristics of a module that can be scaled up to a prototypical size. To appreciate this point, one must recognize that approximately 250,000 modules of 2" diameter would be required to provide one billion fibers. Such an arrangement would provide a ducting, plumbing, and maintenance nightmare in a full-scale plant and clearly would not be economical or workable. A new design concept, such as rectangular modules, is needed for a full-scale plant. Therefore, the objective of Task 9 is to develop the mass transfer fundamentals of rectangular modules.

In the beginning of this quarter, we tested the rectangular module using various sodium sulfite solutions as the scrubbing liquid for SO_2 removal. The results of these experiments are summarized next.

The experiments were performed in a similar manner to those using pure water as the scrubbing liquid (Monthly Technical Report #13). We conducted the first set of tests using a constant gas flow rate of 100 SLPM, varying liquid flow rates, and the following concentrations of sodium sulfite in the liquid: 0.01M, 0.05M, 0.2M, and 0.5M. The SO_2 concentration in the feed gas was 3000 ppm. Figure 3 shows the percent of SO_2 removed as a function of liquid flow rate for the various sodium sulfite solutions and for pure water. In the second set of experiments, the liquid flow rate was held constant at 1 L/min and the gas flow rate was varied. The sodium sulfite concentrations used for these experiments were 0.05M and 0.2M. The removal results obtained for the various gas flow rates are shown in Figure 4 (the data for pure water is not included because the experiments were performed at a different liquid flow rate). The highest SO_2 removal, about 98%, was obtained with a gas flow rate of 20 L/min, a liquid flow rate of 1 L/min, and a sodium sulfite concentration of 0.02M.

The mass transfer coefficients for these experiments were calculated in the same way as described for the pure water tests (see Quarterly Technical Report #13). The difference, however, lies in the calculation of Henry's coefficient. The value for the SO_2-water system was determined from experimental data, but for the SO_2-Na_2SO_3(aq) system, it was necessary to calculate the value based on both the physically dissolved SO_2 and the sulfite/bisulfite ions. By performing a charge balance for various SO_2 concentrations, we were able to construct equilibrium curves for the different sodium sulfite solutions, as shown in Figure 5. (We also used this method to calculate an SO_2-water equilibrium curve, which is also shown in Figure 5 along with the experimental data that was used for the SO_2-water calculations.) It is important to note that for each curve shown in the figure, there is a "threshold" value of the mole fraction SO_2 in the liquid
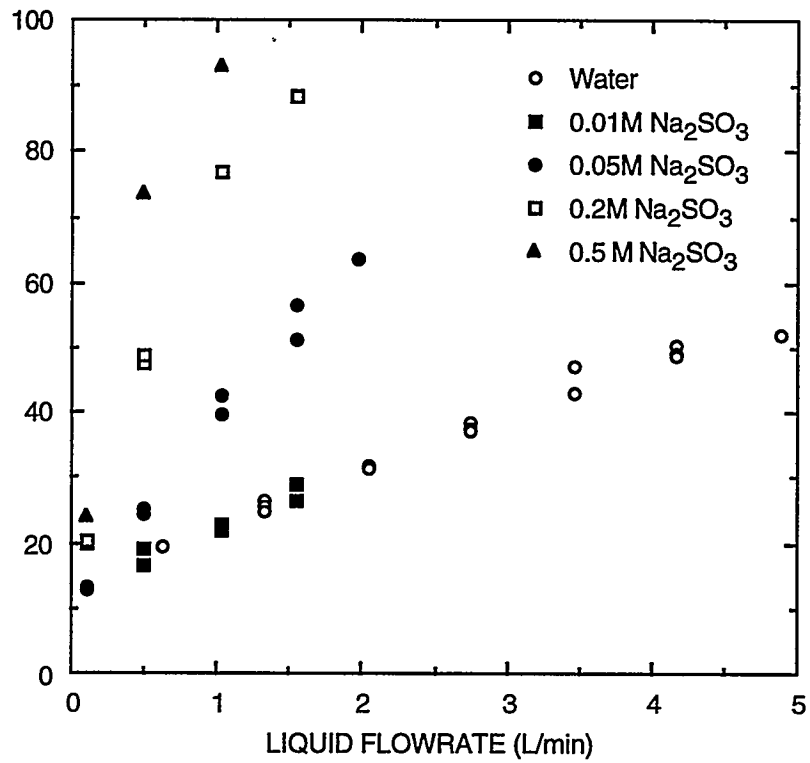
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Figure 3. Percent SO$_2$ removal versus liquid flow rate.
Gas flowrate = 100 SLPM.

9
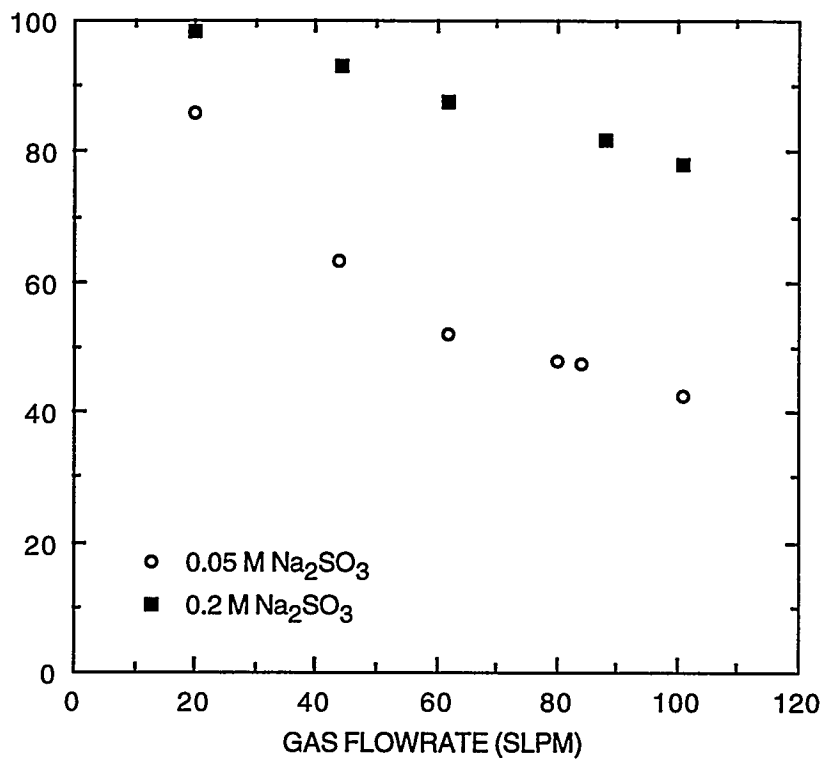
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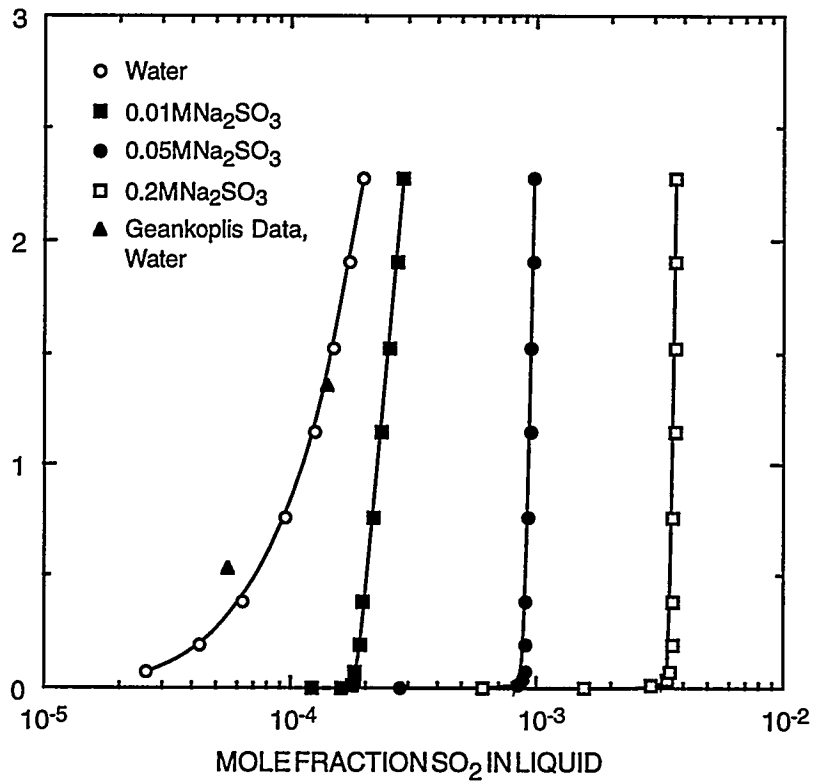Figure 4.   Percent SO_2 removal versus gas flow rate.
Liquid flowrate = 1 L/min.

10

Figure 5. SO_2 equilibrium curves in various sodium sulfite solutions.

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11

below which the partial pressure of SO_2 in the vapor is essentially zero below this value. In our experiments, the concentration of SO_2 absorbed in the liquid was always below this threshold value, so we assumed $Hx_{out} = 0$ when determining the log-mean partial pressure difference.

The calculated overall mass transfer coefficients as a function of the liquid flow rate and of the gas flow rate are shown in Figures 6 and 7, respectively. The graphs show that the mass transfer appears to depend upon the liquid flow rate, but is independent of the gas flow rate near 100 SLPM.

Although the fibers are not demonstrating removal percentages as high as expected, we believe this may be due to specific problems with the housing design, rather than the cross-flow geometry or the fibers themselves (see Quarterly Technical Report #13 for a description of the design). It appears that some of the liquid may be flowing around and outside of the submodules instead of through them. We have seen evidence of this flow pattern by performing tests with colored dye. In addition, the high packing density (85%) may also be causing flow channeling within the submodules, rather than allowing even flow around the fibers. To eliminate the possibility of flow channeling around the submodules, we have decided to test one submodule welded permanently into its own housing.

To examine the channeling aspect, we obtained one submodule (12.5" x 1.5" x 1.25") welded permanently into a polypropylene housing. We have performed one set of experiments with this submodule using pure water as the scrubbing liquid for a flue gas concentration of 3000 ppm SO_2. The liquid flow rate was varied while the gas flow rate was maintained constant at about 20 SLPM. Although this is slightly different than the previous flow rate of 25 SLPM submodule (total flow = 100 SLPM), we believe this is a reasonable comparison because the mass transfer is independent of gas flow at these flow rates (Task 9, Quarterly Technical Report #13).

Figure 8 shows the percentage of SO_2 removed as a function of the liquid flow rate for the single submodule. The graph shows that the highest removal achieved was 97% at a water flow rate of 1.7 L/min. Figure 9 compares the mass transfer coefficients measured for both the welded submodule and the previous device (housing with four submodules). The mass transfer coefficient is shown as a function of liquid flow rate because it should be independent of the gas flow rate. The mass transfer coefficients obtained with the single submodule are almost an order of magnitude higher than those measured with the four submodules in the housing. These results show that the problem was in fact due to the housing design, which allowed liquid to flow around the modules rather than through the fiber bundle.
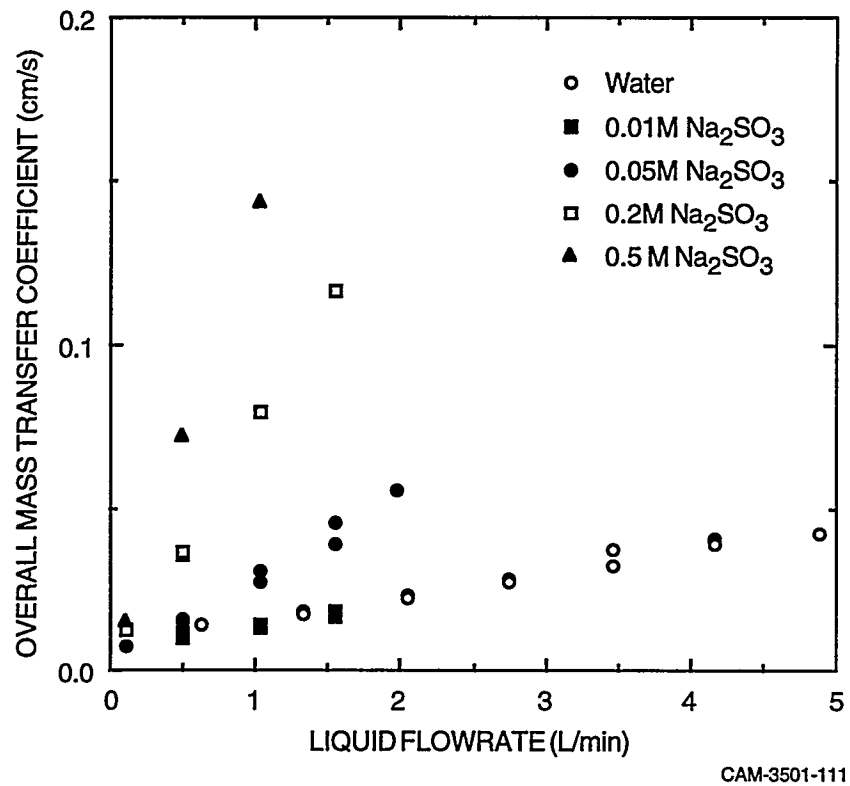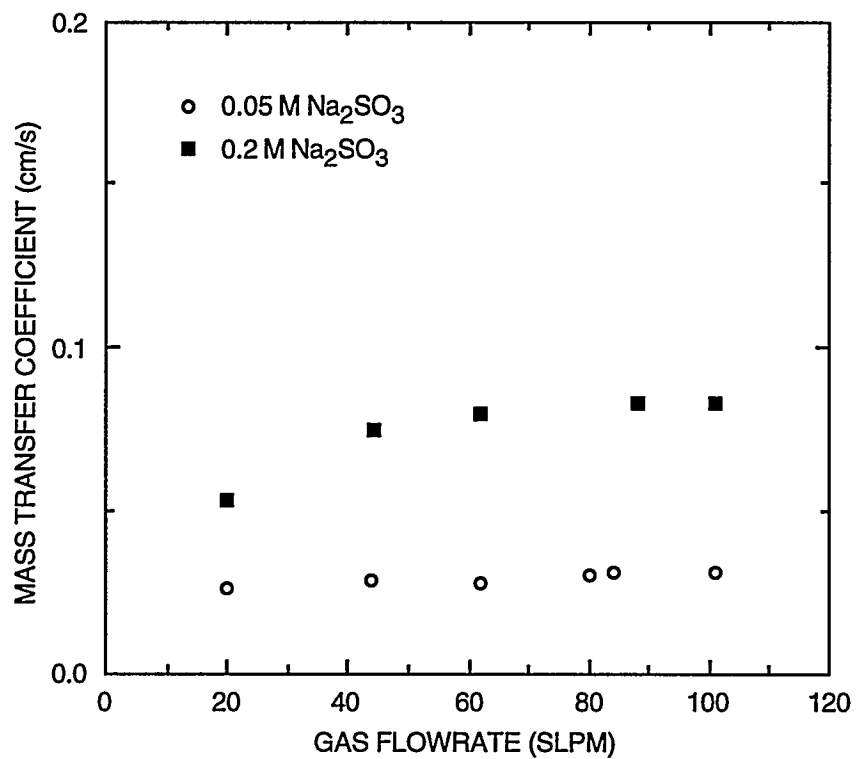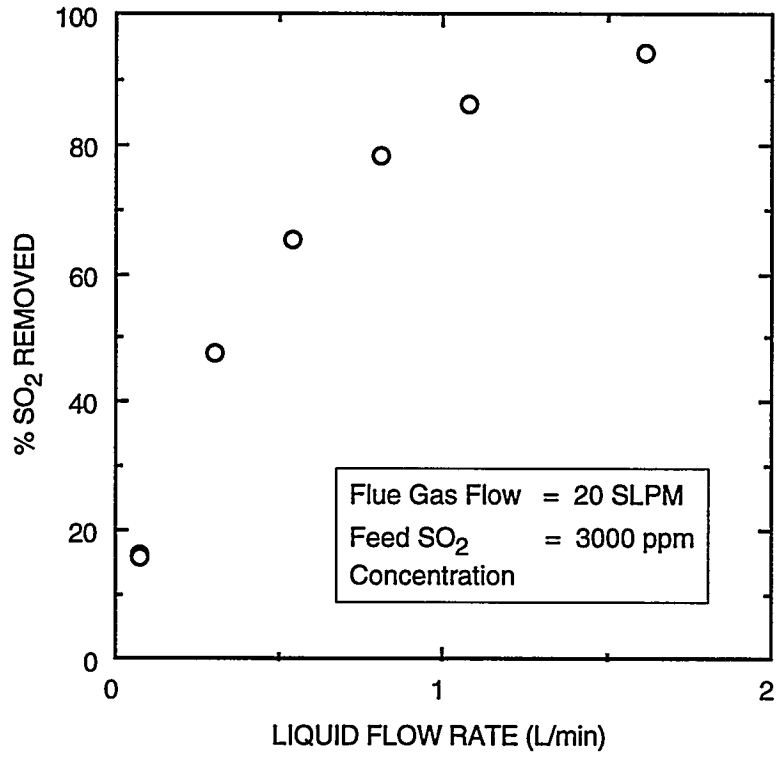
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Figure 6. Overall mass transfer coefficient versus liquid flowrate.
Gas flowrate = 100 SLPM.

13
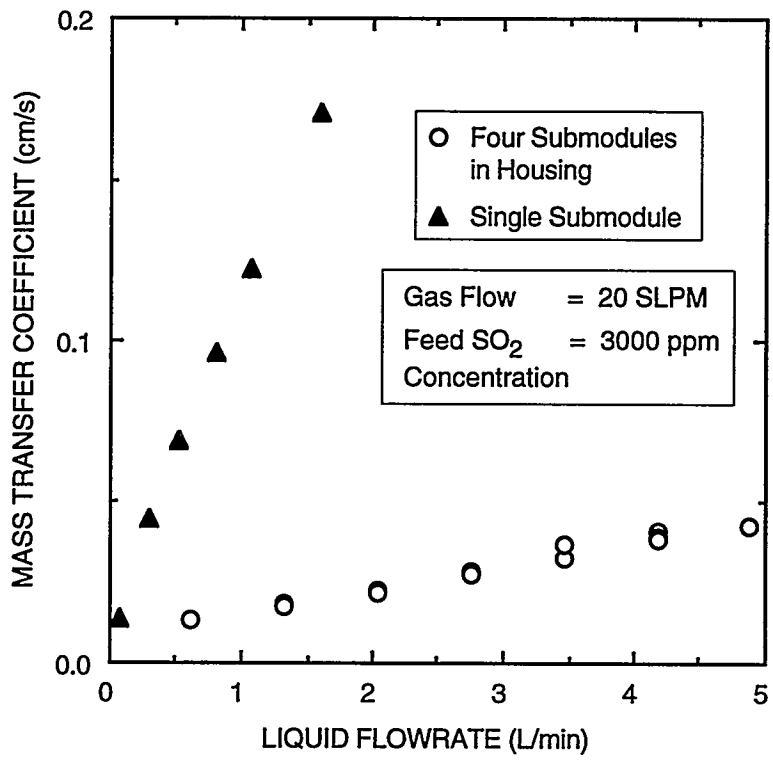
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Figure 7. Overall mass transfer coefficient versus gas flowrate.

Liquid flowrate = 1 L/min.

14

Figure 8. Percent removal versus water flow rate.

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Figure 9. Mass transfer coefficient versus water flow rate.

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We also plan to study combined SO_2 absorption/scrubbing liquor regeneration using the cross-flow modules and have designed an apparatus for these experiments (Figure 10). The simulated flue gas is generated in the same way as described in Management Status Report #34. Sodium sulfite solution is pumped from a reservoir through the first hollow fiber module to absorb SO_2 from the flue gas stream. The scrubbing liquor then passes into the second module, where the liquor is regenerated by the organic (d-DMA). The pH of the scrubbing liquor stream is monitored at the outlet of both modules. It is also necessary to continuously regenerate the d-DMA stream because of high synthesis costs involved. After passing through the second hollow fiber module, the SO_2-rich organic liquid is heated to about 125-150°C and the SO_2 is desorbed by a nitrogen stream bubbling into the liquid. The d-DMA is then pumped through a cooler to lower the temperature to about 25°C before entering the regenerator module. The feed gas, the gas exiting the scrubber, and the gas leaving the d-DMA stripper are all monitored by the SO_2 analyzer. We are currently awaiting arrival of equipment and completion of d-DMA synthesis.

We indicated above that a single submodule in a welded polypropylene housing out-performs the previous rectangular hollow fiber contactor because of channeling of liquid flow. During the later parts of this reporting period, we continued testing the single submodule using varying concentrations of sodium sulfite as the scrubbing liquid.

The results of the experiments are shown in Tables 2 to 5. In these experiments, we have studied the effect of gas flow rate, liquid flow rate, and sodium sulfite concentration on the performance of HFC for SO_2 removal. All the experiments were conducted at room temperature and the inlet flue gas contained about 3000 ppm of SO_2.

Table 2 presents data using water as the scrubbing medium. The WG runs represent experiments with constant gas flow rate and the WL series represent data using constant liquid flow rate. The highest removal achieved was 98% at a water flow rate of 1.7 L/min and a gas flow rate of 20 SLPM (Run No. WG22). Table 3 summarizes the results obtained by scrubbing with 0.01M Na_2SO_3 solution. We conducted experiments with constant gas flow, and we chose not to study this concentration extensively because the results were not that different compared to those obtained with water. We were, however, able to remove 99.9% of the SO_2 from a 20 SLPM flue gas stream with a liquid flow rate of 0.9 L/min (Run No. S01G8). Note that when removal rates are this high (~99.9%), we cannot calculate the mass transfer coefficient because of the log-mean concentration term involved in the calculations.

Figure 10. Apparatus for scalable module testing.

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18

Table 2
Results of SO₂ absorption in water

Run No.	Temp. (C°)	Liquid Flow (mL/min)	Gas Flow SLPM	Inlet SO₂ Composition (ppm)	Outlet SO₂ Composition (ppm)	Percent SO₂ Removed	Kog (cm/s)	Re (gas)	Re (liq)	Sh
WG1	24.6	72	20.3	2924	2441	17	0.0135	71.5	5.4	80
WG2	24.5	72	20.3	2914	2446	16	0.0130	71.5	5.4	76
WG3	24.5	302	20.3	2914	1521	48	0.0442	71.5	22.7	260
WG4	24.5	302	20.3	2914	1521	48	0.0442	71.5	22.7	260
WG5	24.5	538	20.3	2924	1018	65	0.0684	71.5	40.5	402
WG6	24.5	807	20.3	2924	635	78	0.0961	71.4	60.9	565
WG7	24.4	1081	20.3	2919	403	86	0.122	71.4	81.4	718
WG8	24.5	1615	20.3	2919	170	94	0.171	71.4	122	1006
WG9	23.5	96	20.3	3014	2198	27	0.0263	71.6	7.3	155
WG10	23.5	368	20.3	3024	1168	61	0.0660	71.6	27.7	388
WG11	23.5	591	20.3	3019	705	77	0.0962	71.6	44.5	566
WG12	23.5	846	20.3	3014	403	87	0.129	71.6	63.7	756
WG13	23.6	96	20.3	3029	2178	28	0.0282	71.6	7.3	166
WG14	23.7	368	20.3	3029	1212	60	0.0631	71.6	27.7	371
WG15	23.5	591	20.3	3033	789	74	0.0882	71.6	44.5	519
WG16	23.5	846	20.3	3038	412	86	0.178	71.6	63.7	751
WG17	26.6	77	20.2	3068	2317	25	0.0255	71.2	5.8	150
WG18	26.6	358	20.2	3068	1242	60	0.0628	71.2	27.0	370
WG19	26.7	590	20.2	3073	750	76	0.0934	71.2	44.5	550
WG20	26.8	877	20.2	3073	422	86	0.127	71.2	66.1	745
WG21	26.7	1169	20.2	3073	232	92	0.161	71.2	88.1	947
WG22	26.5	1740	20.2	3073	76	98	0.224	71.2	131	1317
WL1	26.7	590	20.2	3054	772	75	0.0900	71.2	44.5	529
WL2	26.9	590	15.3	3029	552	82	0.0827	53.9	44.5	486
WL3	26.9	590	12.2	3024	355	88	0.0817	42.9	44.5	481
WL4	27.1	590	9.6	3043	112	96	0.0987	33.8	44.5	581
WL5	27.3	590	22.1	3043	834	73	0.0935	77.8	44.5	550
WL6	24.5	590	20.3	3038	765	75	0.0900	71.4	44.5	530
WL7	24.7	590	15.3	3033	556	82	0.0819	54.0	44.5	482
WL8	24.7	590	12.2	3029	344	89	0.0823	42.8	44.5	484
WL9	25.1	590	9.7	3033	118	96	0.0973	34.1	44.5	572
WL10	23.2	590	20.3	3043	790	74	0.0877	71.3	44.5	516
WL11	23.2	590	16.1	3038	536	82	0.0889	56.9	44.5	523
WL12	23.3	590	13.2	3038	313	90	0.0947	46.5	44.5	557
WL13	23.5	590	9.7	3048	101	97	0.102	34.1	44.5	601

Table 3
Results of SO$_2$ absorption in 0.01M Na$_2$SO$_3$ solution

Run No.	Temp. (C°)	Liquid Flow (mL/min)	Gas Flow SLPM	Inlet SO$_2$ Composition (ppm)	Outlet SO$_2$ Composition (ppm)	Percent SO$_2$ Removed	Kog (cm/s)	Re (gas)	Re (liq)	Sh
S01G1	24.5	77	20.3	2974	1919	35	0.0241	71.4	5.8	142
S01G2	24.6	77	20.3	2984	2128	29	0.0186	71.4	5.8	109
S01G3	24.8	358	20.3	2994	894	70	0.0664	71.4	27.0	391
S01G4	24.8	590	20.3	2994	453	85	0.104	71.4	44.5	610
S01G5	24.1	77	20.3	3004	2028	32	0.0215	71.4	5.8	126
S01G6	22.9	77	20.3	3004	2098	30	0.0196	71.4	5.8	115
S01G7	23.0	77	20.3	3004	2011	33	0.0122	71.4	5.8	129
S01G8	23.0	877	20.3	3004	3.4	99.9	-	71.4	66.1	-
S01G9	23.2	77	20.3	2994	1984	34	0.0226	71.4	5.8	132
S01G10	23.3	590	20.3	2994	126	96	0.172	71.4	44.5	1013
S01G11	23.4	358	20.3	3004	620	79	0.0859	71.4	27.0	506

Table 4
Results of SO$_2$ absorption in 0.05M Na$_2$SO$_3$ solution

Run No.	Temp. (C°)	Liquid Flow (mL/min)	Gas Flow SLPM	Inlet SO$_2$ Composition (ppm)	Outlet SO$_2$ Composition (ppm)	Percent SO$_2$ Removed	Kog (cm/s)	Re (gas)	Re (liq)	Sh
S05G1	26.9	77	20.3	3004	645	79	0.0850	71.4	5.8	500
S05G2	27.3	358	20.3	3043	2.1	99.9	-	71.4	27.0	-
S05G3	27.3	590	20.3	3004	1.7	99.9	-	71.4	44.5	-
S05G4	27.1	77	20.3	3004	844	72	0.0703	71.4	5.8	413
S05G5	24.4	110	20.2	3003	522	83	0.0956	71.2	8.3	562
S05G6	24.6	20	20.2	3003	2336	22	0.0138	71.2	1.5	81
S05G7	24.8	43	20.2	3013	1759	42	0.0295	71.2	3.2	174
S05G8	24.7	65	20.2	3013	1271	58	0.0473	71.3	4.9	278
S05G9	24.7	88	20.2	3013	863	71	0.0685	71.3	6.6	403
S05L201	23.1	20	20.3	3043	2466	19	0.0115	71.3	1.5	68
S05L202	23.3	20	16.8	3033	1426	20	0.0102	59.2	1.5	60
S05L203	23.4	20	14.0	3003	2346	22	0.0094	49.3	1.5	55
SO5L204	23.6	20	12.0	3013	2366	21	0.0079	42.2	1.5	46
S05L205	23.7	20	14.0	3003	2441	19	0.0079	49.3	1.5	46
S05L206	23.8	20	9.7	3013	2326	23	0.0068	34.1	1.5	40
S05L207	23.8	20	20.2	2993	2585	14	0.0080	71.3	1.5	47
S05L208	24.0	20	16.8	3033	2495	18	0.0089	59.2	1.5	52
S05L651	23.0	65	20.3	3033	1322	56	0.0454	71.5	4.9	297
S05L652	23.2	65	20.3	3043	1342	56	0.0447	71.4	4.9	263
S05L653	23.2	65	17.0	3024	1098	64	0.0466	60.0	4.9	274
S05L654	23.4	65	14.1	3033	322	73	0.0499	49.7	4.9	294
S05L65.	23.4	65	12.0	3033	559	82	0.0550	42.2	4.9	323
S05L656	23.6	65	11.9	3029	623	79	0.0512	42.0	4.9	301
S05L657	23.8	65	9.7	3033	347	89	0.0575	34.3	4.9	338

Table 5
Results of SO$_2$ absorption in 0.2M Na$_2$SO$_3$ solution

Run No.	Temp. (C°)	Liquid Flow (mL/min)	Gas Flow SLPM	Inlet SO$_2$ Composition (ppm)	Outlet SO$_2$ Composition (ppm)	Percent SO$_2$ Removed	Kog (cm/s)	Re (gas)	Re (liq)	Sh
S2G1	23.3	77	20.3	3024	1.7	99.9	-	71.4	5.8	-
S2G2	23.3	77	20.3	3024	1.5	99.9	-	71.4	5.8	-
S2G3	23.2	358	20.3	3029	1.4	99.9	-	71.4	27.0	-
S2G4	23.1	77	20.3	3033	1.1	99.9	-	71.4	5.8	-
S2G5	26.1	110	20.2	2994	1.4	99.9	-	71.3	8.3	-
S2G6	25.9	20	20.2	3004	1097	63	0.0555	71.3	1.5	326
S2G7	25.7	43	20.2	3024	79	97	0.200	71.3	3.2	1176
S2G8	25.8	20	20.3	2914	1023	65	0.0577	71.4	1.5	339
S2G9	25.7	43	20.3	2914	68	98	0.206	71.4	3.2	1214
S2G10	25.5	43	20.2	3003	198	93	0.149	71.3	3.2	877
S2G11	25.1	20	20.2	3003	1251	58	0.0481	71.3	1.5	283
S2G12	25.4	54	20.2	3003	36	99	0.241	71.3	4.1	1418
S2G13	25.3	65	20.2	3003	2.8	99.9	-	71.3	4.9	-
S2G14	23.9	43	20.2	3063	140	95	0.168	71.3	3.2	990
S2G15	24.2	43	20.2	3013	136	96	0.169	71.3	3.2	995
S2G16	24.2	110	20.2	3013	1.1	99.9	-	71.3	8.3	-
S2G17	24.1	20	20.2	3018	1246	59	0.0484	71.3	1.5	285
S2L1	23.2	20	20.3	3043	1073	65	0.0570	71.4	1.5	335
S2L2	23.4	20	26.8	3014	760	75	0.0624	59.0	1.5	368
S2L3	23.8	20	23.7	3033	442	85	0.0718	48.4	1.5	422
S2L4	23.9	20	11.9	3033	260	91	0.0798	42.1	1.5	470
S2L5	24.2	20	9.7	3024	58	98	0.104	34.0	1.5	612

The data shown in Table 4 was obtained using 0.05M Na_2SO_3 scrubbing solution. At this concentration level, we studied two different constant liquid flow rates; 20 ml/min and 65 ml/min (represented by the S05L20 and S05L65 runs, respectively). SO_2 removal rate of 99.9% from a flue gas stream of 20 SLPM was obtained using 360 ml/min of liquid flow (Run No. S05G2). Table 5 shows data from experiments performed using 0.2M Na_2SO_3, the highest concentration studied, as the scrubbing medium. In these experiments, 99.9% removal was achieved with a gas flow rate of 20 SLPM and a liquid flow rate of 65 ml/min (Run No. S2G13).

In addition to the percent removal and mass transfer coefficients, we have also represented the data in terms of the dimensionless Reynolds number and Sherwood number in order to scaleup the data to Phase III of the program. The Reynolds number, calculated for both the gas and liquid, is given by:

$$Re = \frac{\rho v D}{\mu}$$

where ρ is the fluid density, μ is the fluid viscosity, D is the inner diameter of the fiber (gas side) or outer diameter (liquid side), and v is the fluid velocity. For the liquid side, the velocity was determined by calculating the smallest space between the fibers and determining the velocity through that area (the maximum velocity of the liquid). The Sherwood number is given by:

$$Sh = \frac{K_{og}D}{D_{AB}}$$

Where K_{og} is the overall mass transfer coefficient, D is the outer diameter of the fiber, and D_{AB} is the diffusion coefficient of SO_2 in water.

Figure 11 and 12 show variation of Sherwood number as a function of the liquid side Reynolds number for a constant gas flow rate. Figures 13 to 15 show Sherwood number vs. gas side Reynolds number for three different liquid flow rates. The graphs illustrates that the mass transfer coefficient is dependent primarily on the liquid flow rate rather than the gas flow rate. This is rather a surprising result for us, since our original laboratory work (Task 4) indicated the gas side dependence of mass transfer coefficient.

For the combined SO_2 absorption and scrubbing liquor regeneration step, we intend to use 0.05M Na_2SO_3 solution. We believe that 0.01M Na_2SO_3 solution will not allow enough absorption and that 0.2M Na_2SO_3 solution will be too difficult to regenerate. There may be some additional problems with 0.2M Na_2SO_3, including pore-wetting and oxidation of sulfites to form sulfates. We are investigating the latter possibility, which is likely a result of using house air in the
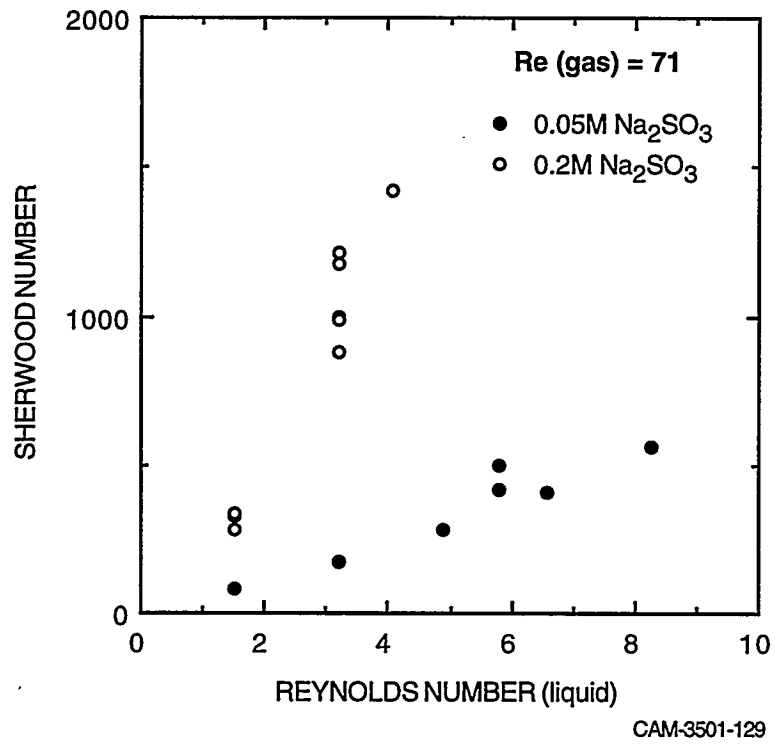
23

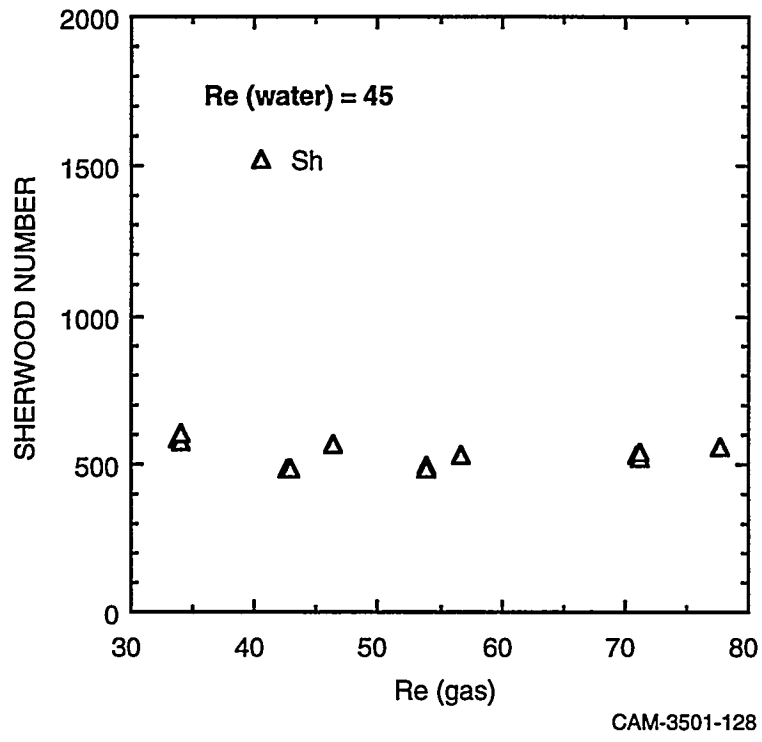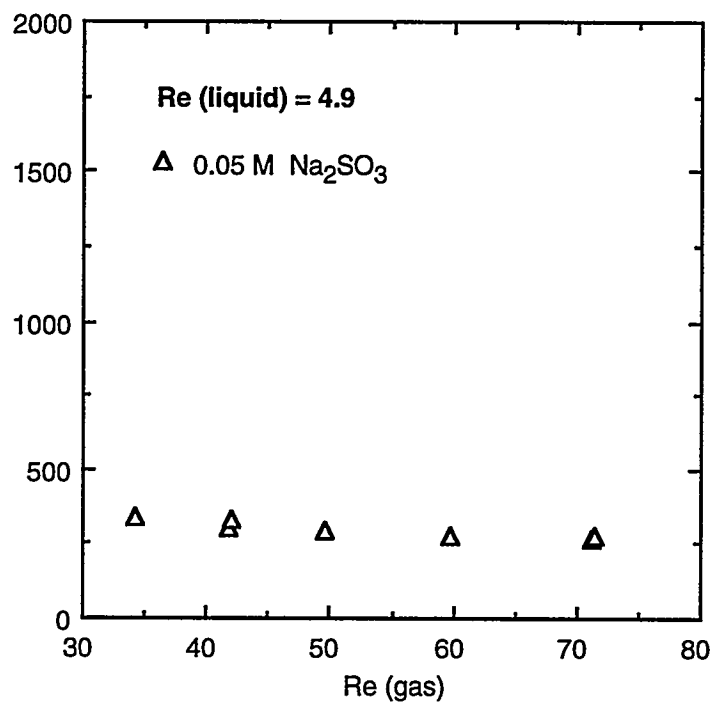CAM-3501-130

Figure 11. Sherwood Number vs. Reynolds Number (liquid).

24

Figure 12. Sherwood Number vs. Reynolds Number (liquid).

Figure 13. Sherwood number vs. Reynolds number (gas).

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26

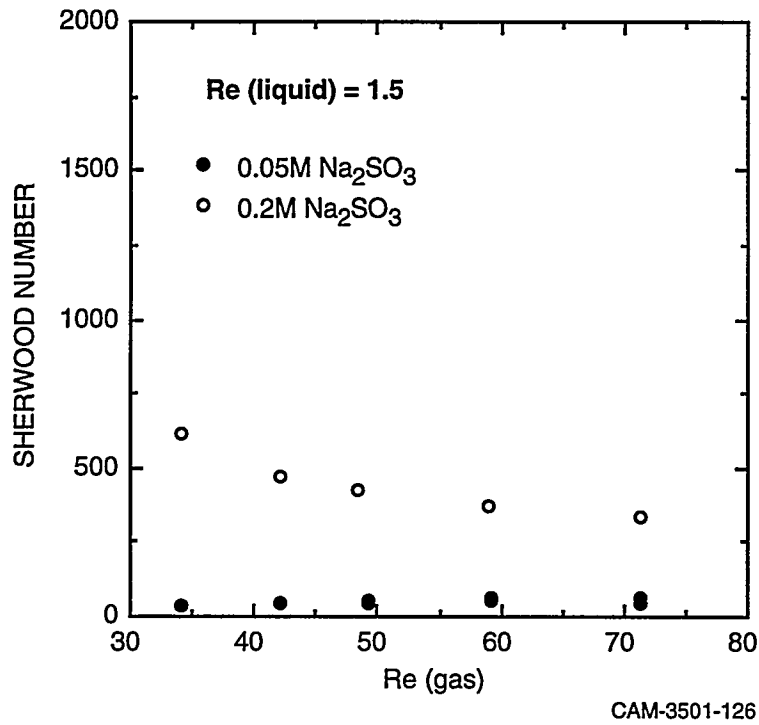Figure 14. Sherwood number vs. Reynolds number (gas).

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27

Figure 15. Sherwood number vs. Reynolds number (gas).

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28