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MODIFICATION AND OPERATION OF THE ATMOSPHERIC ASH-AGGLOMERATING GASIFIER ON COAL

Project 9020 Monthly Progress Report

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ABSTRACT

Three gasifier tests (Runs 108, 109, and 110) were started during September, but only Run 108 had subbituminous coal fed to the gasifier. The remaining two tests were terminated due to bed defluidization while still in the coke start-up phase. We have noticed many coarse particles present in the recently received "fine breeze" coke supplied by the vendor. For three years this supplier has been delivering fine breeze coke for this and the earlier test program. Until this month, the coarse fraction (+6 mesh) had never exceeded 10% to 15% of the coke. Its presence in the coke of up to 40% has raised the superficial fluidizing velocity required, to a level above the current system's capability.

Nonetheless, Run 108 did yield, with continuous coal feed, a 58-hour gasifier test run, operated entirely by air gasification. The temperature of 1650°F achieved in Run 106 was duplicated. This long operating period produced significant results and verified the data obtained from Run 106, that the elutriated fines contained most of the ash fed. There were again no accumulations of ash or tar in any section of the reactor, quenchers, or dust collection equipment.

OPEN ITEMS

Because our supply of subbituminous coal is down to 30 tons, an order for 3 cars (approximately 175 tons) of a Utah bituminous coal has been placed, with delivery expected by October 15. It is possible that the subbituminous test work may be completed and that the reactor will be ready sometime before the new coal arrives.

ACHIEVEMENT OF PROJECT OBJECTIVES

The objectives of Task II of this program include determining the conditions required and obtaining reliable data for the gasification and the ash agglomerating of various classes of coal.

- Operating data have been obtained with steam-oxygen, air, and steam-air gasification of a Montana subbituminous coal, the first coal to be tested on this project.
- The high reactivity and moisture content exhibited by this subbituminous coal have limited the maximum observable bed temperature to 1650°F.
    1. This 1650°F temperature was obtained through air gasification of this coal after surface drying the material to 20% moisture. The use of either steam-air or steam-oxygen gasification limited the operating temperature to 1500°F.
    2. Data and samples on subbituminous coal gasification in this reactor reveal that approximately 85% of the coal ash is elutriated to the two external cyclones.
    3. The remaining test work with this coal will include fines recirculation to the gasifier for recovery and agglomeration of its ash content.

DETAILED DESCRIPTION OF MONTHLY TECHNICAL PROGRESS

Task 2. AAG Test Operations

Work Accomplished

Three tests were started during September, but only during one, Run 108, was subbituminous coal fed to the reactor. In both of the other two, Runs 109 and 110, the tests had to be terminated due to defluidization because the start-up coke received from the supplier contained excessive amounts of oversize particles and thereby exceeded the reactor's fluidization velocity capabilities.

However, Run 108 confirmed that an operable maximum temperature of 1700°F was obtainable when feeding this 20% moisture-bearing Montana subbituminous coal to the gasifier. Under these conditions, the coal ash constituent was essentially elutriated from the gasifier with the char fines.

The test plans for Run 108 were the following:

1. Conduct a steady-state test for at least 100 hours to yield data for heat and mass balances.
2. Operate under air gasification conditions at a superficial velocity of 2 ft/s, and monitor bed ash content. Withdraw bed material if ash agglomerates do not form and bed ash composition exceeds 40%.
3. Increase the velocity of 2.5 ft/s and ascertain the bed temperature rise achieved and the presence of ash agglomerates.
4. Operate at 2.0 ft/s and add oxygen. Ascertain the bed temperature rise and the presence of ash agglomerates.

The actual operation of Run 108 consisted of the following:

1. Because of an electrical malfunction in the feeder scale system, operating time was limited to 58 hours.
2. Sufficient operating time at 2 ft/s was obtained for heat and mass balances. It was not necessary to withdraw bed material.
3. The velocity was increased to 2.5 ft/s and the bed temperature rise was minimal with no ash agglomerates.
4. Oxygen was added but the result was the same as above.

Run 108 Operation

Test 108 was begun on September 11 and terminated on September 15. Continuous, uninterrupted feeding by L-valve directly to the fluidized bed was conducted for 58 hours using the 3/4x0-inch Montana subbituminous coal as a feedstock. The test was terminated only because of an electrical malfunction in the weigh scale feed system that necessitated special parts. Operation was by air gasification to maximize the bed temperature; this temperature did reach 1700°F but averaged near 1650°F during coal feeding.

Start-up was by nonvolatile bearing metallurgical coke. Coke operation was continued until ash agglomerates heavily predominated in the venturi discharge stream. Coal feed was substituted for coke at noon on September 13. Immediately the bed temperature dropped stabilizing near 1630°F but then rising and averaging 1670°F for the first 24-hour coal feeding period. The venturi discharge quantity diminished from over 30 to 10 lb/hr as the ash agglomerates present when feeding coke disappeared from this venturi stream after 48 hours of coal feeding.

After 24 hours of coal feeding, an auxiliary supply of plant air (about 600 lb/hr) was added to our Joy compressor output of 2000 lb/hr. The net result was a 30% increase in air supply to the reactor, which required a 30% increase in feed rate to sustain the bed level and netted a 30°F rise in reactor bed temperature. Ash agglomerates did not appear and fines elutriation to the external cyclones also increased by 25%. After a 24-hour period of adding extra air, the feed system for the reactor was running at maximum capacity, but because the higher bed temperature did not yield ash agglomerates, the extra air was withdrawn. A final attempt to raise the operating bed temperature was made shortly before the test was terminated by adding oxygen at the venturi jet line. The result of this oxygen addition was essentially the same as when extra air was added, increased feed needed to maintain the bed, along with additional fines elutriation and an increase in product gas quantity. The added oxygen also produced a minimal bed temperature rise (25°F) with no evidence of ash-agglomerate formation. As stated, due to an electrical relay failure in the weigh feed system, Run 108 was terminated and the reactor bed was emptied from the gasifier. A visual examination of the reactor revealed no accumulation of ash or slag on the walls, grid, or internal cyclone; also, there was no damage to any of the internal components.

Runs 109 and 110

Both Tests 109 and 110 developed the same defluidization problems. For more than 3 years, we have been purchasing coke from the Mid Continent Coal and Coke Company for use on both the previous HYGAS® supported AAG project and this low-Btu project. The initial shakedown runs on the earlier project resulted in modifications to the grid design when bed sintering developed. Upon making these grid modifications, we had never found it necessary to screen the coke to achieve its complete fluidization in the AAG reactor. The material is referred to as fine breeze and is specified by the vendor as being −1/4 inch. Our earlier checks on this material showed a minimum fluidization velocity of 0.4 ft/s and that 1.6 ft/s or better produced complete sustainable fluidization. We now find that the most recently received coke contains a larger fraction above +6 mesh and that the minimum fluidization as received has risen to 1.0 ft/s and complete fluidization has risen to over 4.0 ft/s.

This defluidization problem with coke was not detected in either Runs 109 or 110 until the ash-agglomerating stage was reached. Until then, there was no evidence of bed defluidization apparent at the bed view port, bed density, bed height chart traces, or by bed thermocouple temperature excursions. As ash agglomerates formed, both bed depth and density chart traces gave evidence of defluidization, which was then confirmed by view port observation. Consequently, the test start-up was terminated.

The recent problems of off-specification start-up coke will be eliminated in future test work as the coke will be screened so that the reactor feed will be a –6 mesh fraction. This unexpected occurrence did illustrate, however, that while fluidization at 2 ft/s is sustained when feeding –3/4-inch subbituminous coal, the start-up coke must be finer than 1/4 inch in size so that this same velocity is adequate to sustain its fluidization.

The test results obtained from Run 108 consist of the following:

1. Run 108 operated with continuous coal fed directly to the bed for 58 hours.

2. All volatile matter was gasified during Run 108, as evidenced by no accumulations of organic condensibles in any portion of the reactor system.

3. During Run 108, a high percentage (85%) of the ash in the coal fed to the gasifier elutriated from the reactor to the second and third stage cyclones. This result emphasizes the need for conducting fines recirculation work in the next test.

4. Ash agglomerates were not produced during this test or earlier ones using this Montana subbituminous coal due to its inherently high reactivity with both oxygen and steam. Accordingly, not only is the $2C + O_2 \quad 2CO$ reaction favored over the formation of CO_2, but reaction of the coal char with steam is highly favored compared with coke. This particular feedstock contains considerable "bound" water not removed by our surface drying system. Therefore, in our operation with air only, gasification, actually due to the water in the coal, had a near equivalent of 12% moisture in the process air.

5. Heat and material balances have been prepared for Test 108 covering a 20-hour period of relatively steady-state operation. The operating data for this test is listed in Table 1, with Table 2 containing the test data. The heat and material balance is tabulated on the flow sheet, shown as Figure 1.

6. Calculations have been made based upon kinetic data available and its relation to the product gas compositions experienced in this project when gasifying subbituminous coal. The results illustrate the effect that the moisture inherent in this coal has upon the obtainable operating temperatures. These results, given in Figure 2, reveal the reactor operating conditions necessary to yield higher bed temperatures. However, this AAG system is not designed to handle the large gas and solid flows that would be necessary to achieve these temperatures.

Table 1. ASH-AGGLOMERATING GASIFIER OPERATING DATA FOR RUN 108

Run No. 108 Date 9/13-9/14/77 Time Period 1600 to 1200

Unit Size: 4-ft diameter x 21-ft high

Venturi Diameter 3 inches Feed Material Subbituminous Coal

Cyclone Dust Collectors 3 Internal 1 External 2

1	Superficial Velocity, ft/s	2.04
2	Average Fluidized-Bed Temperature, °F	1648
3	Average Reactor Pressure, psig	5.15
4	Efficiency = Carbon in Prod. Gas/Carbon in Feed, %	76.6
5	Venturi Pressure Drop, psi	1.0
6	Venturi Velocity, ft/s	75.5
7	Grid Gas, mol/hr	49.95
8	Grid Gas Composition: %N_2	76.49
9	%O_2	20.33
10	%Steam	3.18
11	Venturi Gas, mol/hr	29.99
12	Venturi Gas Composition: %N_2	79.26
13	%O_2	17.94
14	%Steam	2.80
15	Carbon/O_2 (weight basis)	1.06
16	Carbon/Steam (weight basis)	12.05
17	Fluidized Bed Density, lb/CF	37.84
18	Fluidized Bed Height, ft	4.12
19	Feed Composition, %C/%ash	68.25/8.91
20	Average Fluidized Bed Composition, %C/%ash	75.3/23.2
21	Venturi Discharge Composition, %C/%ash	42.7/56.8
22	2nd Stage Dust Composition, %C/%ash	70.8/25.8
23	3rd Stage Dust Composition, %C/%ash	60.1/36.2
24	Superficial Velocity/Minimum Fluidization Velocity	5.1
25	Freeboard Velocity, ft/s	3.06

Mesh Size	3	6	12	20	40	70	100	140	200	270	40μ	20μ	10μ	5μ	−5μ
Feed	1.1	10.7	22.5	23.7	19.0	11.9	-	8.2	-	2.9					
Bed	2.3	13.8	20.3	15.1	14.8	14.7	-	16.5	-	2.5					
Ash	35.6	43.3	16.6	2.1	0.6	1.8									
2nd Stage Dust									34.2	7.7	20.2	21.4	15.3	1.2	
3rd Stage Dust											10.0	3.5	54.5	32.0	

Revised 8/22/77

Table 2, Part 1. ASH-AGGLOMERATING GASIFIER — DATA SUMMARY FOR RUN 108

Date: 9/13 - 9/14/77 Run No. 108

Time	1400	1600	1800	2000	2200	2400	9/14 → 0200	0400
TE 9206.1 Gas Temp entering venturi, °F	475	447	453	479	487	495	482	488
TE 9206.2 Gas temp entering grid, °F	733	742	744	743	737	729	717	711
TE 9206.3 Bed temperature, °F	1629	1606	1616	1631	1638	1630	1652	1660
TE 9206.16 Gas temp entering quencher, °F	1547	1565	1568	1577	1669	1573	1582	1591
FR 903 ________ flow to grid, mols/hr	0	0	0	0	0	0	0	0
FR 904 ________ flow to air heater, mols/hr	0	0	0	0	0	0	0	0
FR 905 Air flow to grid, mols/hr	49.37	49.34	49.28	49.21	49.22	49.91	50.23	51.11
FR 906 Air flow to venturi, mols/hr	23.50	23.82	23.46	23.42	23.77	23.12	23.54	23.79
FR 909 Sweep gas, mols/hr (Air)	2.98	3.05	3.04	2.97	2.90	2.85	2.89	2.75
FR 938 ________ oxygen to air heater, mols/hr	0	0	0	0	0	0	0	0
FR 9303 Aux. N_2 to venturi, mols/hr	3.56	3.53	3.52	3.53	3.53	3.55	3.57	3.50
FR 937 Atomizing N_2, mols/hr	4.56	4.50	4.50	4.56	4.56	4.50	4.63	4.56
FR 918 (1st stage quencher) Rate of quench water, mols/hr	26.19	27.40	26.19	24.58	24.58	25.79	25.79	26.19
FR 9330 (2nd stage quencher) Rate of quench water, mols/hr	0	0	0	0	0	0	0	0
FR 995 Air to air heater, mols/hr	81.92	81.53	81.58	81.23	81.30	77.07	82.84	81.61
FR 9304 Freeboard ________, mols/hr	0	0	0	0	0	0	0	0
Nitrogen purge to reactor, mols/hr	6.17	6.17	6.17	6.17	5.94	5.94	5.94	5.94
Bed density, lbs/cu ft	37.7	37.7	37.7	37.7	37.7	37.7	37.7	37.7
Bed height, ft	4.20	4.39	4.28	4.13	4.13	4.05	4.13	4.13
Reactor gas composition (dry), %								
N_2		56.6		61.1				57.2
CO		18.3		17.5				20.4
CO_2		9.3		7.8				8.7
H_2		14.1		12.2				12.9
CH_4		1.1		0.8				0.82
Coal feed rate, wet (avg), lbs/hr	937.5	937.5	937.5	937.5	937.5	937.5	937.5	937.5
C-4 discharge rate, lbs/hr	20.0	11.0	7.0	8.5	7.0	6.5	11.5	8.5
2nd stage dust removal rate, lbs/hr	140.6	140.6	140.6	140.6	140.6	140.6	140.6	140.6
3rd stage dust removal rate, lbs/hr	15.35	15.35	15.35	15.35	15.35	15.35	15.35	15.35
Feed composition, %C				67.9		67.9		
%Ash				7.1		7.1		
Reactor bed composition, %C				76.4				
%Ash				22.0				
Ash composition, %C				29.9		53.7		
%Ash				69.5		45.4		
2nd stage dust, %C				71.7		71.5		
%Ash				24.8		25.3		
3rd stage dust, %C				61.8		59.8		
%Ash				34.3		36.7		

Table 2, Part 2. ASH-AGGLOMERATING GASIFIER — DATA SUMMARY FOR RUN 108

Date: 9/14/77 Run No. 108

Time	0600	0800	1000	1200	1400	1600	1800	2000
TE 9206.1 Gas Temp entering venturi, °F	482	493	505	508	516	516	494	514
TE 9206.2 Gas temp entering grid, °F	707	709	657	732	569	565	583	591
TE 9206.3 Bed temperature, °F	1674	1675	1669	1678	1707	1688	1664	1632
TE 9206.16 Gas temp entering quencher, °F	1599	1608	1604	1607	1639	1618	1589	1568
FR 903 Air flow to grid, mols/hr	0	0	0	0	22.38	22.44	23.01	22.59
FR 904 ____ flow to air heater, mols/hr	0	0	0	0	0	0	0	0
FR 905 Air flow to grid, mols/hr	50.50	50.49	50.11	50.04	50.41	50.27	49.19	49.71
FR 906 Air flow to venturi, mols/hr	23.84	23.76	23.58	23.59	23.79	23.49	24.77	23.61
FR 909 Sweep gas, mols/hr (Air)	2.68	2.63	2.73	2.87	3.07	3.07	2.84	2.68
FR 938 ____ oxygen to air heater, mols/hr	0	0	0	0	0	0	0	0
FR 9303 Aux. N_2 to venturi, mols/hr	3.50	3.57	3.55	3.53	3.53	3.53	3.54	3.54
FR 937 Atomizing N_2, mols/hr	4.63	4.63	4.63	4.63	4.38	4.44	4.38	4.38
FR 918 (1st stage quencher) Rate of quench water, mols/hr	21.36	24.99	26.19	26.19	31.43	32.24	34.25	24.18
FR 9330 (2nd stage quencher) Rate of quench water, mols/hr	0	0	0	0	0	10.00	10.00	10.00
FR 995 Air to air heater, mols/hr	81.61	83.92	82.74	82.67	81.90	82.63	81.78	81.98
FR 9304 Freeboard ____, mols/hr	0	0	0	0	0	0	0	0
Nitrogen purge to reactor, mols/hr	5.94	5.94	5.94	5.94	5.94	5.94	5.94	5.94
Bed density, lbs/cu ft	37.7	37.7	39.28	37.7	36.13	36.13	36.13	32.99
Bed height, ft	4.13	4.09	3.89	3.97	3.91	3.83	3.03	2.97
Reactor gas composition (dry), %								
N_2		57.7		57.0				54.4
CO		19.8		20.4				32.8
CO_2		8.4		8.4				10.7
H_2		12.6		12.8				15.8
CH_4		0.78		0.75				1.2
Coal feed rate, wet (avg), lbs/hr	937.5	937.5	937.5	937.5	1185	1185	1185	1185
C-4 discharge rate, lbs/hr	8.5	8.0	11.0	9.5	9.5	16.0	18.0	13.0
2nd stage dust removal rate, lbs/hr	140.6	140.6	140.6	140.6	167.5	167.5	167.5	167.5
3rd stage dust removal rate, lbs/hr	15.35	15.35	15.35	15.35	26.0	26.0	26.0	26.0
Feed composition, %C		68.6						
%Ash		7.5						
Reactor bed composition, %C		74 2						
%Ash		24.3						
Ash composition, %C		44.6						
%Ash		55.5						
2nd stage dust, %C		69.2						
%Ash		27.3						
3rd stage dust, %C		58.8						
%Ash		37.6						

Table 2, Part 3. ASH-AGGLOMERATING GASIFIER — DATA SUMMARY FOR RUN 108

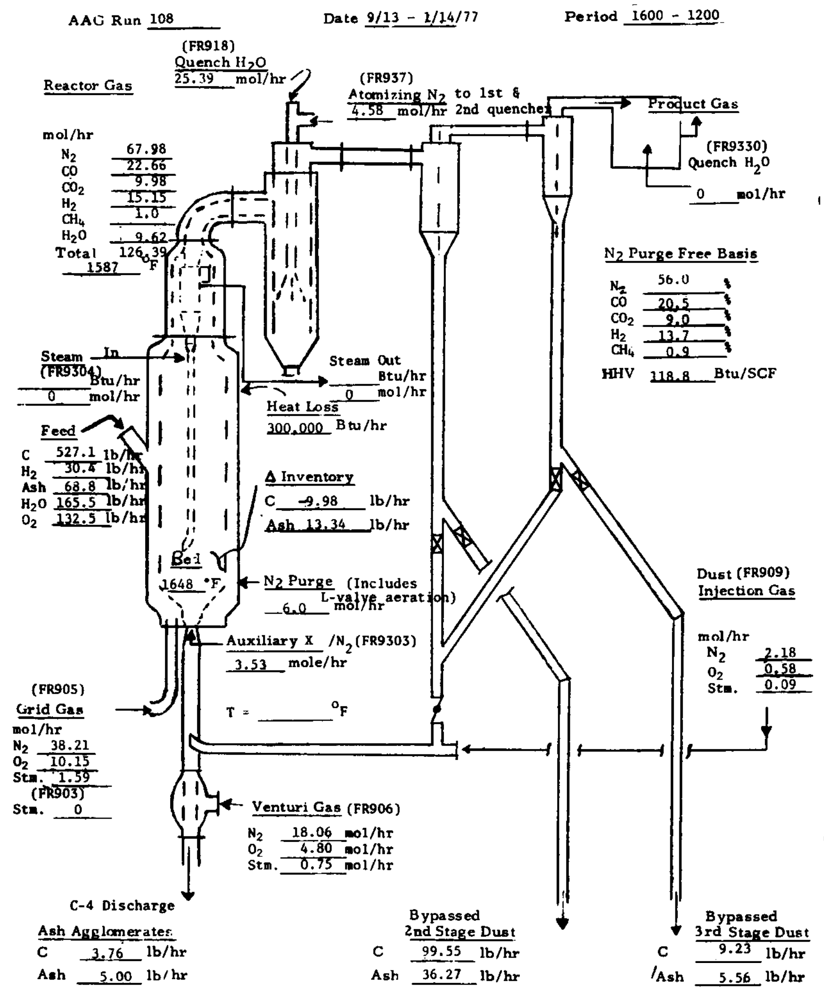
Date: 9/14 - 9/15/77 Run No. 108

Time	2200	2400	9/15→ 0200	0400	0600	0800	1000	1200
TE 9206.1 Gas Temp entering venturi, °F	508	529	551	543	511	514	503	501
TE 9206.2 Gas temp entering grid, °F	620	696	690	679	683	695	688	688
TE 9206.3 Bed temperature, °F	1623	1574	1546	1607	1551	1533	1570	1574
TE 9206.16 Gas temp entering quencher, °F	1551	1525	1502	1510	1528	1502	1530	1551
FR 903 Air flow to grid, mols/hr	16.70	9.99	12.35	13.03	9.52	7.93	7.52	7.59
FR 904 ________ flow to air heater, mols/hr	0	0	0	0	0	0	0	0
FR 905 Air flow to grid, mols/hr	45.39	37.91	37.87	51.81	46.79	47.31	52.61	52.45
FR 906 Air flow to venturi, mols/hr	23.15	22.75	21.17	21.57	21.37	21.83	21.61	21.65
FR 909 Sweep gas, mols/hr (Air)	3.01	2.99	3.11	3.19	3.16	2.91	2.82	2.82
FR 938 ________ oxygen to air heater, mols/hr	0	0	0	0	0	0	0	0
FR 9303 Aux. N_2 to venturi, mols/hr	3.54	3.40	3.41	3.44	3.44	3.55	3.55	3.55
FR 937 Atomizing N_2, mols/hr	4.44	4.63	4.56	4.56	4.50	4.56	4.56	4.56
FR 918 (1st stage quencher) Rate of quench water, mols/hr	28.21	25.79	25.39	24.99	28.21	24.99	27.40	28.21
FR 9330 (2nd stage quencher) Rate of quench water, mols/hr	10.00	0	0	0	0	0	0	0
FR 995 Air to air heater, mols/hr	75.68	67.38	66.28	81.71	77.87	77.74	83.05	82.98
FR 9304 Freeboard ________, mols/hr	0	0	0	0	0	0	0	0
Nitrogen purge to reactor, mols/hr	5.94	5.94	5.94	5.94	5.94	5.94	5.94	5.94
Bed density, lbs/cu ft	34.56	39.28	36.13	37.7	37.7	36.13	36.13	36.13
Bed height, ft	2.50	2.35	3.35	3.97	3.67	4.39	4.31	3.99
Reactor gas composition (dry), %								
N_2		55.9		56.1		55.2		94.8
CO		16.0		16.8		17.0		1.9
CO_2		10.7		10.5		11.2		2.1
H_2		15.5		14.8		15.3		1.2
CH_4		1.2		1.1		1.3		0.07
Coal feed rate, wet (avg), lbs/hr	1185	1185	1185	1185	1185	1185	1185	1185
C-4 discharge rate, lbs/hr	13.0	9.5	8.5	10.5	12.5	12.5	13.0	17.0
2nd stage dust removal rate, lbs/hr	167.5	167.5	167.5	167.5	167.5	167.5	167.5	167.5
3rd stage dust removal rate, lbs/hr	26.0	26.0	26.0	26.0	26.0	2.60	26.0	26.0
Feed composition, %C								
%Ash								
Reactor bed composition, %C								
%Ash								
Ash composition, %C								
%Ash								
2nd stage dust, %C								
%Ash								
3rd stage dust, %C								
%Ash								

Table 2, Part 4. ASH-AGGLOMERATING GASIFIER – DATA SUMMARY FOR RUN 108

Date: 9/15/77 Run No. 108

Time	1400	1600	1800					
TE 9206.1 Gas Temp entering venturi, °F	540	623	567					
TE 9206.2 Gas temp entering grid, °F	690	781	781					
TE 9206.3 Bed temperature, °F	1595	1677	1544					
TE 9206.16 Gas temp entering quencher, °F	1570	1628	1551					
FR 903 Air flow to grid, mols/hr	7.83	0	0					
FR 904 ____ flow to air heater, mols/hr	0	0	0					
FR 905 Air flow to grid, mols/hr	52.64	46.37	51.16					
FR 906 Air flow to venturi, mols/hr	21.60	21.72	21.66					
FR 909 Sweep gas, mols/hr (Air)	2.88	3.01	3.31					
FR 938 ____ oxygen to air heater, mols/hr	0	0	0					
FR 9303 Aux. ____ to venturi, mols/hr	3.54 N_2	7.9 O_2	3.04 N_2					
FR 937 Atomizing N_2, mols/hr	4.5	4.44	4.44					
FR 918 (1st stage quencher) Rate of quench water, mols/hr	28.21	28.21	28.21					
FR 9330 (2nd stage quencher) Rate of quench water, mols/hr	0	0	0					
FR 995 Air to air heater, mols/hr	82.25	75.87	81.06					
FR 9304 Freeboard ____, mols/hr	0	0	0					
Nitrogen purge to reactor, mols/hr	5.94	5.94	5.94					
Bed density, lbs/cu ft	36.13	36.13	40.85					
Bed height, ft	3.75	3.27	2.75					
Reactor gas composition (dry), %								
N_2		34.2						
CO		18.9						
CO_2		25.4						
H_2		18.4						
CH_4		3.1						
Coal feed rate, wet (avg), lbs/hr	1185	1185	1185					
C-4 discharge rate, lbs/hr	17.0	14.5	13.5					
2nd stage dust removal rate, lbs/hr	167.5	167.5	167.5					
3rd stage dust removal rate, lbs/hr	26.0	26.0	26.0					
Feed composition, %C								
%Ash								
Reactor bed composition, %C								
%Ash								
Ash composition, %C								
%Ash								
2nd stage dust, %C								
%Ash								
3rd stage dust, %C								
%Ash								



* See the explanation.

	Ash	Carbon	Nitrogen	Oxygen	Hydrogen	Btu/hr
Feed Subbituminou Coal	68.8	527.1	0	279.6	48.79	−1,891.457
To venturi and grid	0	0	1735.44	535.84	4.86	127.089
Purges/Freeboard	0	0	168.0	0	0	0
Total Input, lb/hr	68.80	527.1	1903.44	815.44	53.65	−1,764.368
Wall losses	0	0	0	0	0	300.000
Venturi discharge	5.0	3.76	0	0	0	4.874
Reactor gas	0	403.68	1903.44	835.84	53.54	−2,600.411
2nd stage dust	36.27	99.55	0	0	0	73.585
3rd stage dust	5.26	9.23	0	0	0	8.034
Bed inventory change	13.34	−9.98	0	0	0	−5.528
Total Output, lb/hr	60.17	506.24	1903.44	835.84	53.54	−1,908.229
Net (Output − Input)	−8.63	−20.86	0	+20.5	−0.11	−143.861
% Balance, Output/Input	87.5	96.0	100	102.5	100	108.2

Figure 1. HEAT AND MATERIAL BALANCE FOR RUN 108

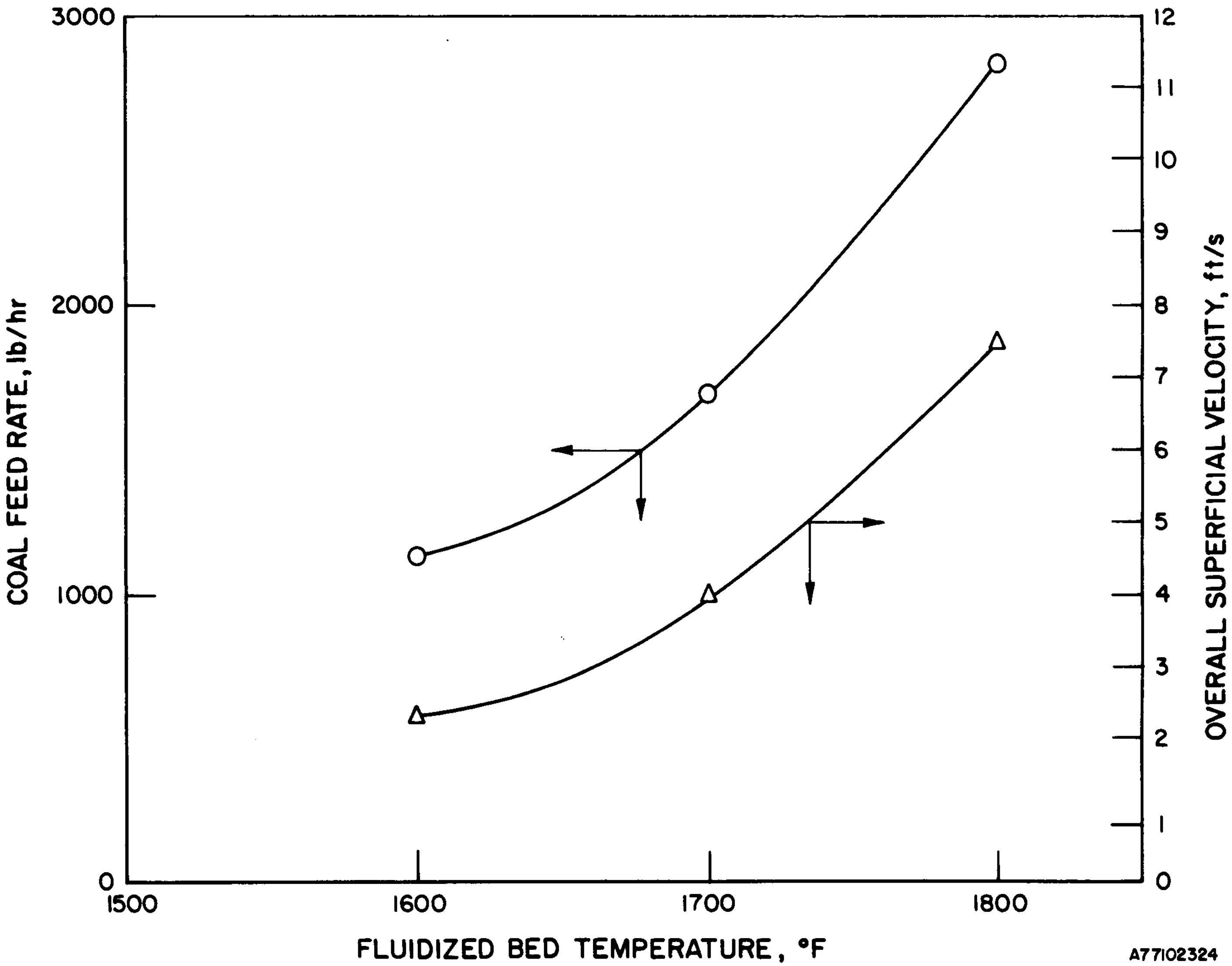


Figure 2. PROJECTED PILOT PLANT GASIFIER CONDITIONS USING AIR GASIFICATION (Montana Subbituminous Coal [18% H_2O])

Cold Model Test Work

Work has resumed on the 4-foot cold model using the new 3 inch venturi. The objectives are primarily to finalize a discharge equation for use in reactor design by comparing the results with the previously tested 2-inch venturi. To date, data have been obtained with actual char and glass spheres. The glass spheres can be obtained in a more uniform size cut than the alumina previously used. Work will also be done with previously used acrylic plastic and alumina spheres. For the graphs enclosed as Figures 3 and 4, discharge rates were obtained for 1/4-inch diameter glass spheres in a bed of char obtained from actual coal gasifier tests. The following behavior was observed:

1. The rate of sphere discharge increased directly with the weight fraction of spheres present in the bed, over the range 0 to 0.7, with air velocity through the venturi throat at 46.5 ft/sec and superficial bed velocity at 2.03 ft/sec.

2. The rate of sphere discharge decreased as air velocity through the venturi throat increased over the range 39 to 59 ft/sec.

3. The bed height showed no statistical effect on rate of sphere discharge over the range 12 inches to 21 inches above the grid.

Work Forecast for October 1977

- Test operation start-up for Run 111 will now use the –6 mesh fraction of the coke for which bench-scale fluidization tests show a 1.5 ft/s complete fluidization velocity, well within system capability.
- After attaining operating temperatures of approximately 1850°F, coke fines will be reinjected at 1000°F.
- After stabilization is reached, operations will be switched to a coal feed.
- Stable operating conditions will be sought by changing operating temperatures and/or process input compositions as necessary.
- If stability cannot be achieved the test will be terminated and mechanical changes made to introduce the fines directly to the bed for the succeeding test run.
- The method and objectives for Run 112 will be the same as for Run 111.
- Work will begin on Utah bituminous coal when it arrives. It is anticipated that this higher grade (11,500 vs. 9500 Btu/lb) of coal will have a lower reactivity; a lower reactivity should enable higher bed temperatures to be reached than those attained with the Montana subbituminous coal.

Task 3. A Design and Cost Estimate for an Atmospheric Pressure Commercial Gasifier

Work Accomplished

No design work has begun.

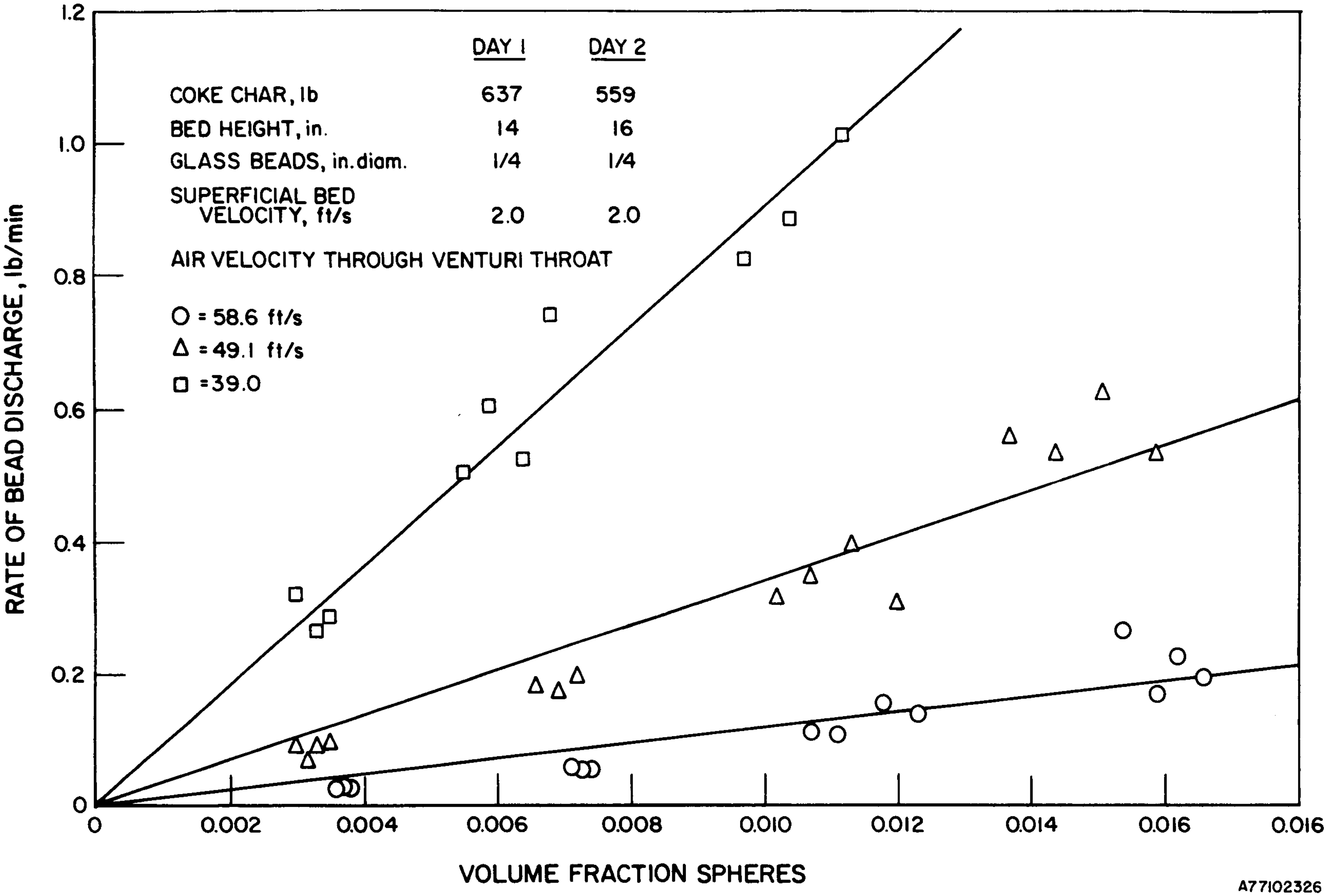


Figure 3. AAG COLD MODEL RUN 33
(3-in. Venturi Throat, 2-day Run)

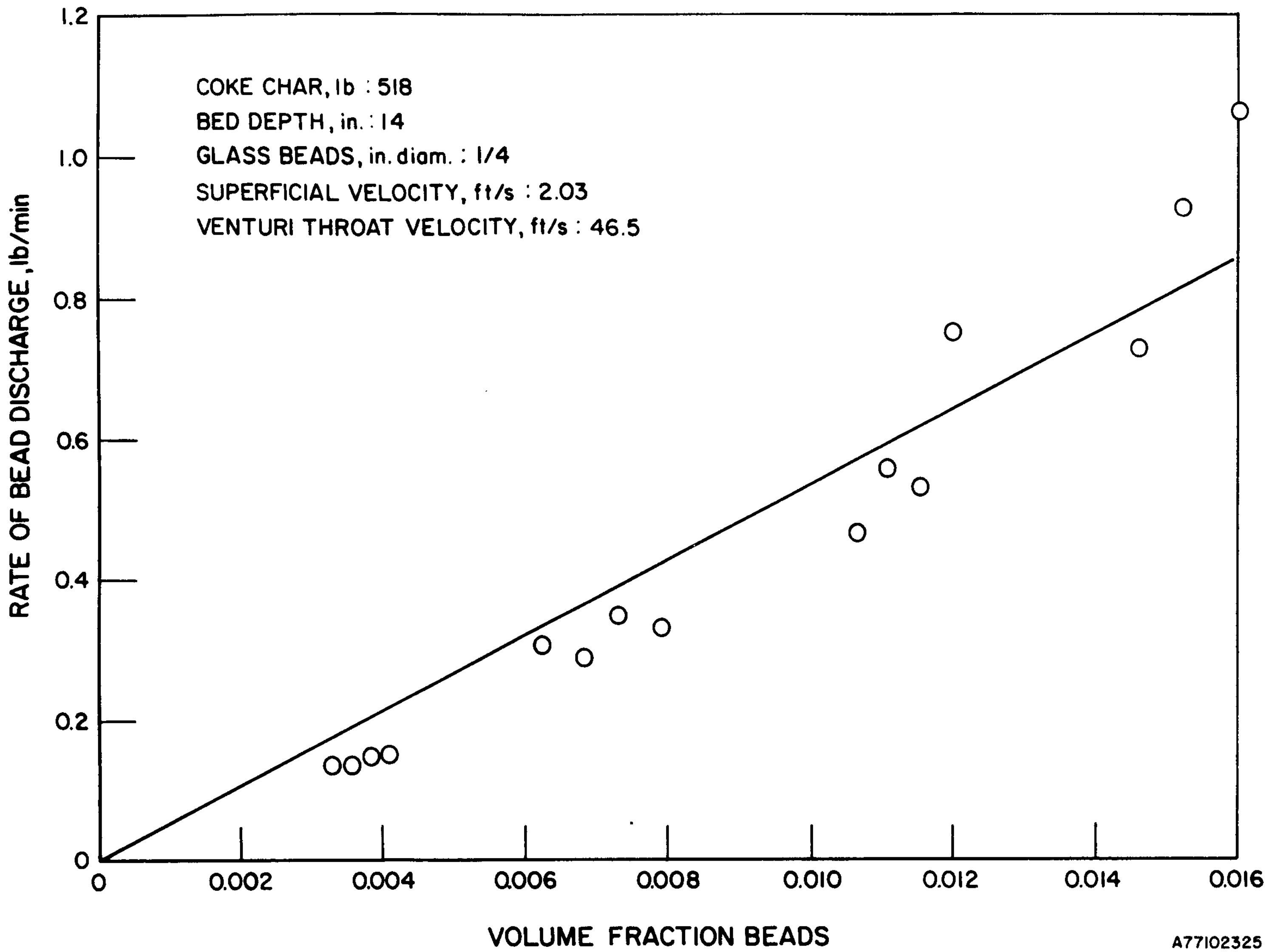


Figure 4. AAG COLD MODEL RUN 34
(3-in. Venturi Throat)

Work Forecast

The design work will not begin in October 1977, but will begin as soon as sufficient data are obtained from Task 2.

Task 4. A Design of a Pressurized Pilot Plant Gasifier

Work Accomplished

No design work has begun.

Work Forecast

The design work will begin in November 1977 or as soon as sufficient data are obtained from Task 2.

Project Plan and Progress Report

Figure 5 shows the progress to date on Tasks 1 through 4.

Project Fund Expenditure Schedule

Figure 6 illustrates the current and projected rates of funds expenditure of this AAG low-Btu modification, operation, and process design.

PSG/LF

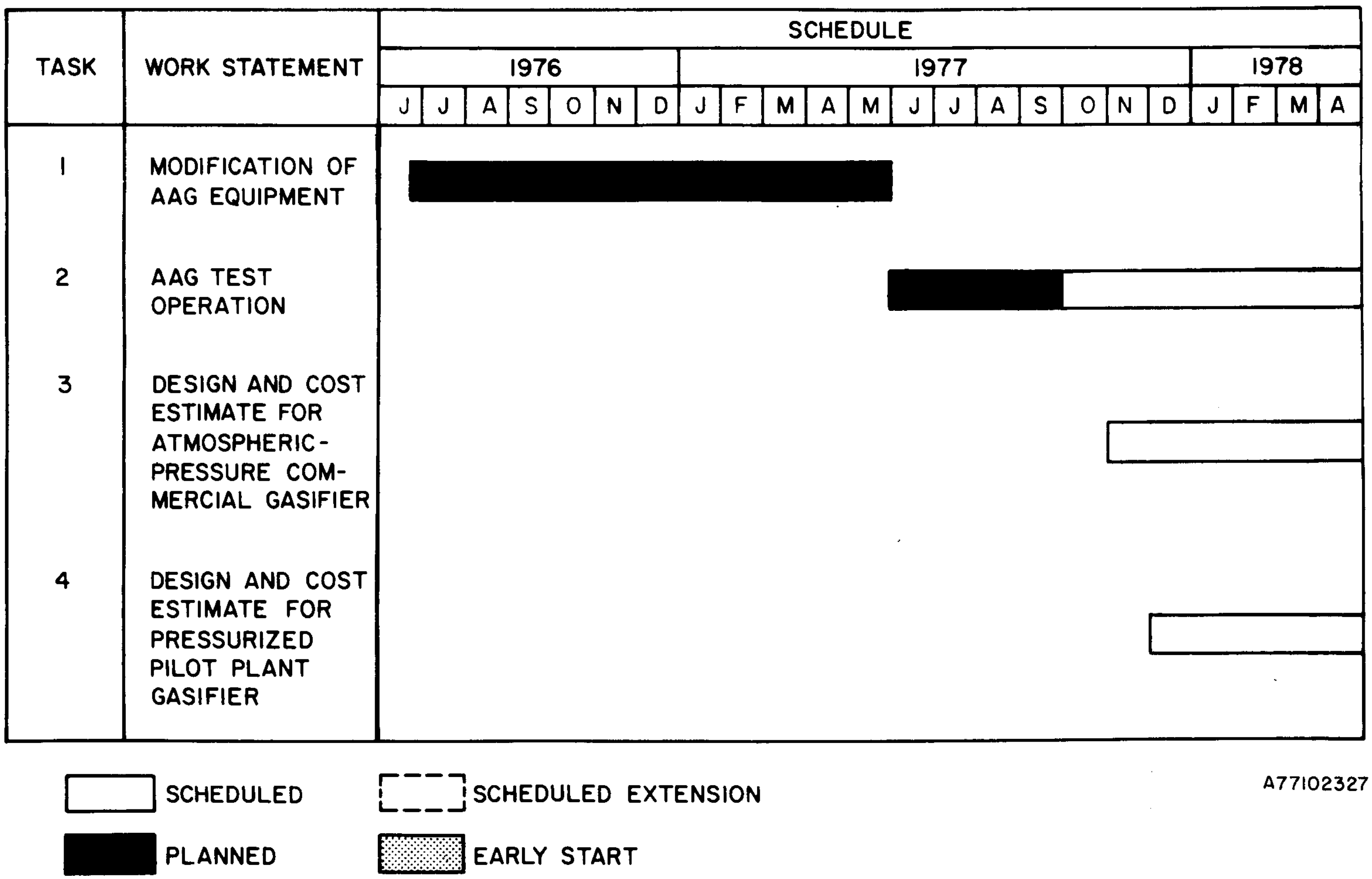


Figure 5. PROGRESS REPORT

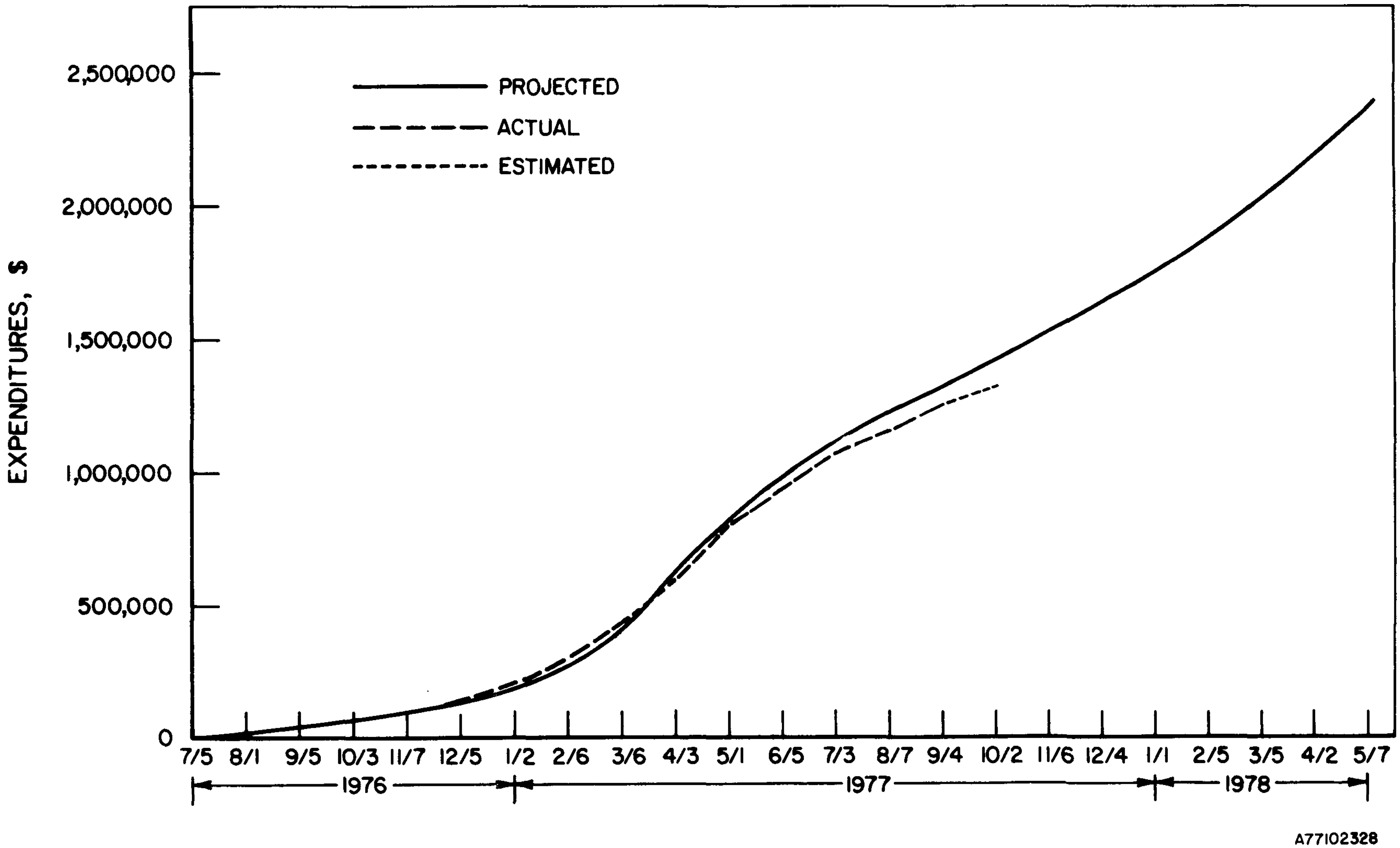


Figure 6. PROJECT FUNDS EXPENDITURE SCHEDULE