IV. PROJECT PLANNING FOR PHASES II - IV

This section describes the proposed workscope and task plans for Phases II through IV of the program. Phase II, Technology Development, comprises additional laboratory- and pilot-scale testing, as well as expanded thermodynamic and kinetics calculations. Phase III, Technology Validation, is focused on design, assembly, and checkout of individual subsystems. Phase IV, Demonstration of Scale-Up, will demonstrate SCWG system operation with the fully integrated 5 MT/day pilot plant to verify generation of hydrogen from biomass during extended tests.

IV.A PHASE II - TECHNOLOGY DEVELOPMENT

IV.A.1 Work Scope and Task Plans

During Phase II, further development of SCWG technology will be performed to resolve knowledge gaps and other critical issues, and to define performance requirements and system interfaces for the Phase III, Technology Validation. The technical issues requiring further development were discussed in Sec. IIB, and summarized in Table II-8.

Task 100 - Feed preparation and Pumping. Demonstrate drying and heating/liquefaction/ash separation/pressurizing/pumping of simulant (corn flakes) and sewage sludge.

1. Design feed liquefaction and pumping system and ash removal system
2. Procure sludge dryer, dried sludge feeder, and pump upgrades
3. Perform pilot-scale tests at GA with the existing, adapted sewage sludge pump
4. Analyze solid feed, liquefied feed, and ash, and document results
5. Key decision point: feasibility of feeding 40 wt% liquefied sewage sludge

Task 200 - Extended Testing. Evaluate alternate catalysts, verify higher H_2 production, and demonstrate extended operation (>8 hr).

1. Perform lab-scale catalyst tests
2. Run short-term pilot-scale tests with alternate catalysts
3. Incorporate liquefied feed system into pilot plant and test
4. Analyze data and document results
5. Decision point: reliable production of 40% to 50% hydrogen (mole fraction)

Task 300 - Process Analyses. Define performance requirements and system interfaces.

1. Perform chemical equilibrium calculations
2. Compare results to lab- and pilot-scale test results
3. Analyze data and document results

Task 400 - Performance/Interface Requirements. Define performance requirements and system interfaces (e.g., sewage sludge feed, utilities, product disposition) for Phase III testing.

1. Prepare Phase III PFD and M&EBs for subsystems and integrated system, and verify projections for economically viable, large-scale systems.
2. Define interface requirements

Task 500 - Project management - Manage Phase II technical effort, budget, and schedule, and prepare Phase II report.

IV.A.2 Schedules, Milestones, and Decision Points

Fig. IV-1 presents the schedule, milestones and decision points for proposed Phase II activities. Contract award is assumed to occur April 1, 1998.

IV.B PHASE III - TECHNOLOGY VALIDATION

IV.B.1 Work Scope and Task Plans

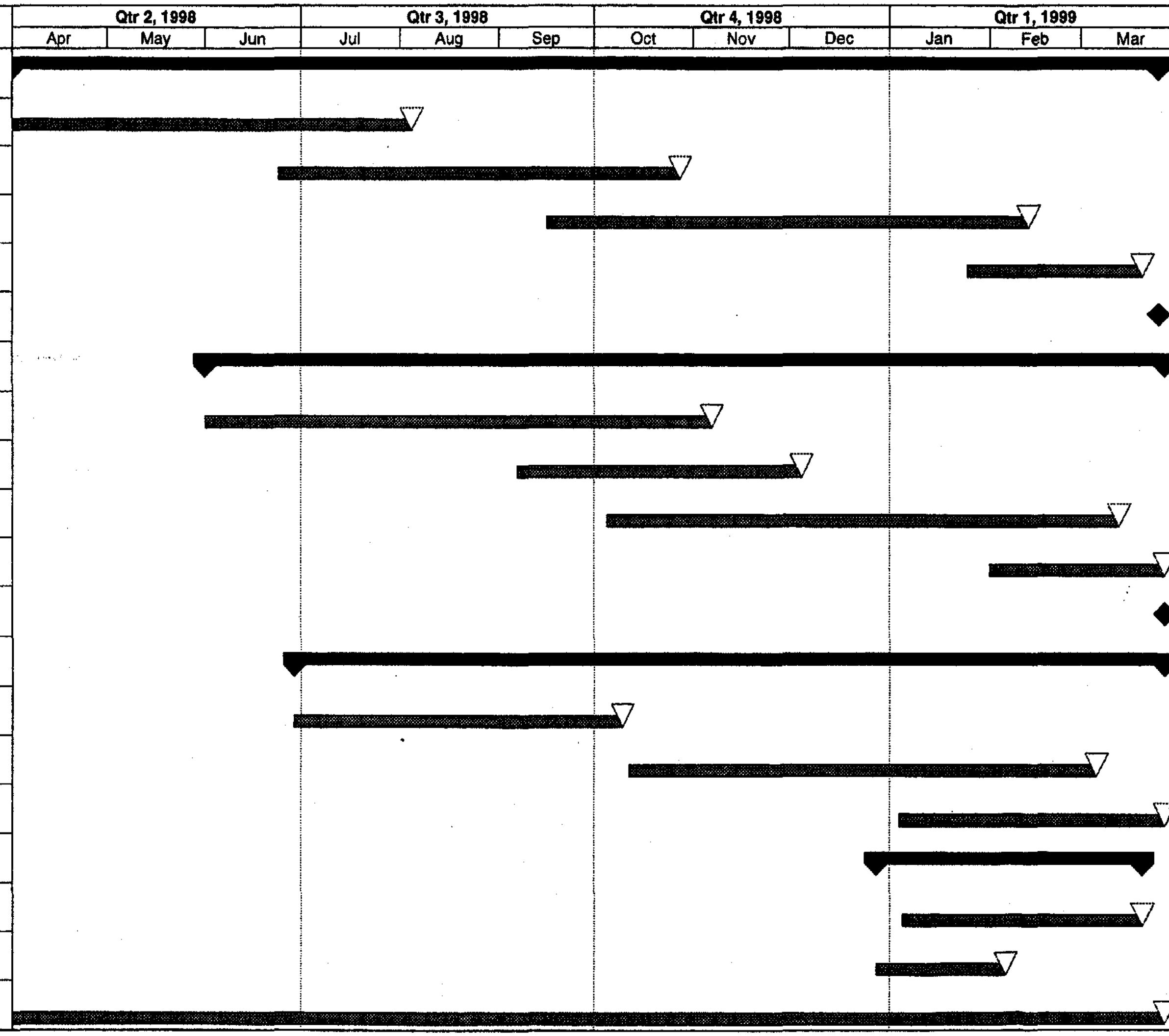
Phase III will consist of a subsystem validation effort including design, procurement and assembly of skid-mounted subsystems. Each subsystem will then undergo shakedown testing as a prelude to the Phase IV Integrated Pilot-Scale Demonstration. Safety analyses, reliability and maintainability studies, permitting studies, process control definition and updated life-cycle-cost analyses will also be performed.

Task 100 - Systems Analyses. Prepare systems analysis studies.

1. Perform safety analyses. Define hazards and hazard categories and design changes to mitigate unacceptable hazards.
2. Perform reliability, availability, and maintainability (RAM) analysis
3. Perform permitting study
4. Prepare updated life-cycle cost analysis
5. Decision point: identify potential barriers to commercialization

FIG. IV-1
SCHEDULE, MILESTONES AND DECISION POINTS FOR PHASE II

ID	Task Name
1	**TASK 100 - FEED PREPARATION AND PUMPING**
2	- Design feed liquefaction and pumping systems
3	- Procure equipment
4	- Perform pilot plant tests
5	- Analyze data and document results
6	**- Decision Point: feasibility of pumping 40 wt% liquefied sewage sludge**
7	**TASK 200 - EXTENDED TESTING**
8	- Perform lab-sacle catalyst studies
9	- Run short-term pilot plant tests with alternate catalysts
10	- Incorporate liquefied feed system into pilot plant and test
11	- Analyze data and document results
12	**- Decision Point: reliable production of 40% to 50% hydrogen (mole fraction]**
13	**TASK 300 - PROCESS ANALYSES**
14	- Perform chemical equilibrium analyses
15	- Compare analyses to lab- and pilot-scale results
16	- Analyze and document results
17	**TASK 400 - PERFORMANCE/INTERFACE REQUIREMENTS**
18	- Prepare Phase III PFD and M&EBs
19	- Prepare interface requirements
20	**TASK 500 PROJECT MANAGEMENT**



GENERAL ATOMICS

Task 200 - Systems Design. Prepare 5 MT/day equipment drawings and specifications, and define facility and support requirements

1. Define P&IDs and control logic diagrams
2. Prepare equipment drawings and specifications
3. Define vendor-supplied equipment and components, including membrane separators, PSA, etc.
4. Define pilot plant test area upgrades and support needs
5. Prepare fabrication/installation drawings
6. Decision point: verify availability of materials/equipment/components

Task 300 - Equipment Procurement/Assembly/Test. Procure 5 MT/day equipment and assemble as skid-mounted subsystems. Test subsystems.

1. Procure/refurbish SCWG components
2. Prepare skids and assemble subsystem components.
3. Perform subsystems testing.
4. Acquire support components for Phase IV (membrane unit, PSA, etc.)

Task 400 - Project Management

IV.B.2 Schedules, Milestones, and Decision Points

Fig. VI-2 presents the schedule, milestones and decision points for proposed Phase III activities. Contract award is assumed to occur April 1, 1999.

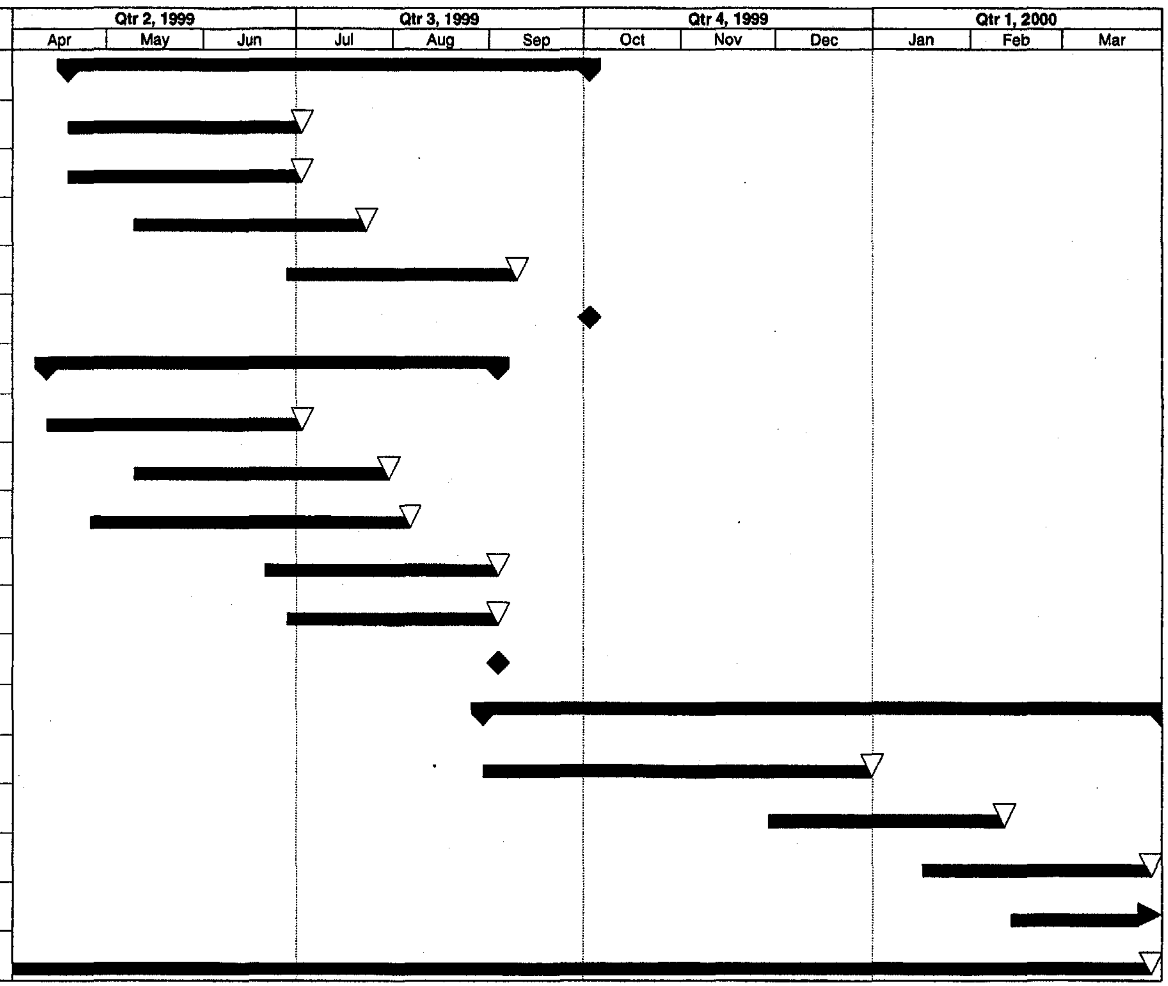
IV.C PHASE IV - DEMONSTRATION OF SCALE-UP

IV.C.1 Work Scope and Task Plans

The Phase IV effort will be directed at demonstrating gasification in the integrated pilot-scale SCWG system. Plant throughput will be approximately 10 times that of the existing GA SCWG pilot plant, but some existing GA-owned SCWO components may be used to reduce the costs for a 5 MT/day system. The skid-mounted subsystem modules, including the feed system, preheat system, gasifier, letdown system, and gas/liquid separator, will be combined with leased membrane separation and PSA units. Hydrogen storage and fuel cells are an optional component of this phase. The fuel cell size will be about 200 kW, a size available from a number of manufacturers.

FIG. IV-2
SCHEDULE, MILESTONES AND DECISION POINTS FOR PHASE III

ID	Task Name
1	**TASK 100 - SYSTEMS ANALYSES**
2	- Perform safety study
3	- Perform RAM analysis
4	- Perform permitting study
5	- Prepare updated life cycle cost analyses
6	**- Decision Point: identify potential barriers to commercialization**
7	**TASK 200 - SYSTEM DESIGN**
8	- Prepare P&IDs and control logic diagrams
9	- Prepare equipment drawings and specifications
10	- Define vendor-specific equipment and components (e.g., PSA)
11	- Define pilot plant facility upgrades and support needs
12	- Prepare fabrication/installation drawings
13	**- Decision Point: verify availability of materials, equipment and components**
14	**TASK 300 - EQUIPMENT PROCUREMEN/ASSEMBLY/TEST**
15	- Procure/refurbish SCWG components
16	- Prepare skids and assemble subsystem components
17	- Perform subsystem tests
18	- Acquire support components for Phase IV (e.g., membrane unit)
19	**TASK 400 PROJECT MANAGEMENT**



Task 100 - Procurement. Procure remaining system components and equipment

1. Acquire balance-of-plant equipment and materials

Task 200 - System Assembly. Integrate the SCWG subsystems and prepare facility.

1. Assemble subsystems for integrated operation.
2. Program software for integrated operations.
3. Prepare pilot plant facility

Task 300 - System Checkout. Begin testing of major system components

1. Perform SCWG checkout tests with simulants
2. Perform SCWG checkout tests with sewage sludge

Task 400 - Integrated Testing. Complete testing of integrated SCWG system verifying hydrogen production and system reliability.

1. Perform SCWG integrated tests with simulants
2. Perform SCWG integrated tests with sewage sludge
3. Prepare and issue final report

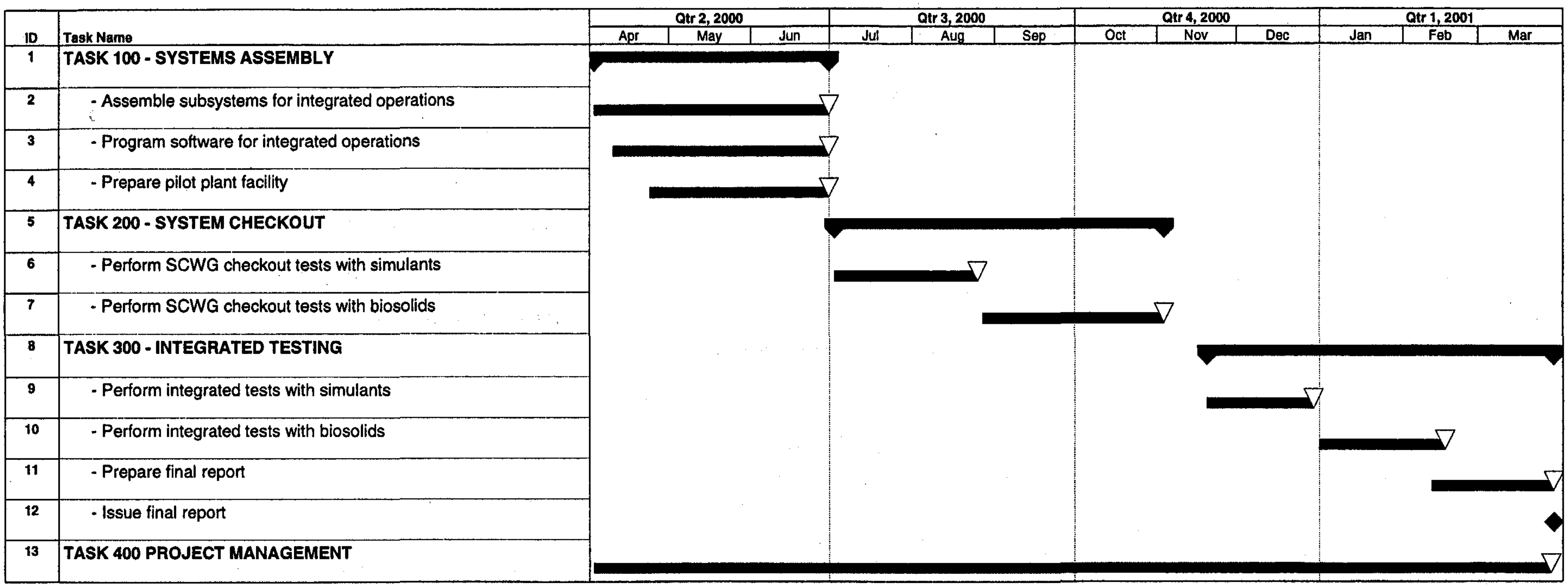
Task 500 - Project Management

IV.C.2 Schedules, Milestones, and Decision Points

Fig. presents the schedule, milestones and decision points for proposed Phase III activities. Contract award is assumed to occur April 1, 2000.

FIG. IV-3

SCHEDULE, MILESTONES AND DECISION POINTS FOR PHASE IV



GENERAL ATOMICS

V. TEAMING AGREEMENTS FOR PHASES II - IV

GA will continue as the team leader for the follow-on phases. In addition, the Encina Wastewater Authority will continue the role it played in Phase I as an interested party to the development of SCWG for hydrogen production. The following sections describe the GA team, its capabilities, qualifications and experience, and facilities and equipment.

V.A TEAM MEMBERS AND RATIONALE FOR SELECTION

GA is one of the leading advanced technology companies in the U.S. with over 40 years of experience in science-oriented research and development as well as engineering development. GA facilities include engineering, test, manufacturing, and advanced technology laboratories. Phase 1 pilot plant studies were performed at the GA SCW facility. The proposed work for Phases II through IV will also be performed at GA. We have extensive experience in the use of our facilities in the design and testing of pilot- and full-scale equipment in accordance with DoD and other Government standards. The SCWG project is being conducted by the Advanced Process Systems (APS) Division of the Advanced Technologies Group. The project will continue to receive high visibility within the GA corporate organization. The Program Manager will continue to report directly to the Director of the APS Division, who, in turn, reports directly to the Senior Vice President for Advanced Technologies.

GA continues its major SCWO program, with over $20 million in contracts over the past five years. GA brings demonstrated experience in the management and execution of SCWO, and hazardous waste activities, including development and demonstration of concepts, technologies, and leading edge hardware.

Encina Wastewater Authority, located just north of GA, has been an important contributor to Phase I of the project and maintains a continuing interest in the effort to commercial sewage sludge gasification and hydrogen production. They operate a state-of-the-art facility serving over 225,000 residents in a location directly adjacent to the Pacific Ocean. They have an excellent reputation in the wastewater treatment industry as a well-run, progressive facility. In addition, they have exceptionally good relations with their customers, a fact that is especially noteworthy given high population density surrounding the plant and environmental sensitivity of nearby residents.

V.B TEAM MEMBER CAPABILITIES

GA is one of the largest privately owned centers for diversified research, development, and engineering in the world. GA is engaged in broad scope research, development, and production, with activities embracing research and development programs for power generation systems, energy conversion systems, waste management, environmental restoration, DoD and DOE programs, and other science-based technologies. Personnel with many years of experience in advanced science and engineering programs make up the various technical groups. Over half of the U.S.-based staff hold technical degrees; of these, more than half have advanced degrees.

GA currently has four Government-sponsored projects underway in SCWO of toxic and hazardous materials, two for DARPA and two for the U.S. Air Force. Completed SCWO projects include one for the Defense Advanced Research Projects Agency (DARPA), several for DOE and one for the National Aeronautics and Space Administration (NASA). GA has also licensed its SCWO technology to two other firms. GA has a well-qualified staff of engineers and technicians needed to design, build, and test SCWG systems of any size from bench-scale test rigs to commercial systems.

Encina provides a reliable source of primary and secondary sludge, skilled personnel, and the infrastructure needed to support an on-site demonstration of the SCWG system. Other municipal waste water treatment facilities have similar capabilities.

V.C QUALIFICATION AND EXPERIENCE OF KEY PERSONNEL

Key personnel in Phase I of the SCWG effort will continue throughout Phases II through IV. Dr. Dan Jensen will remain as the Project Manager. He brings over 25 years experience in the science- and engineering-based research and development, and over 10 years experience in managing large tasks and projects. He was the Deputy Project Manager for the initial $6.8 million DARPA contract to develop SCWO for the treatment of chemical warfare agent and was the on-site project manager for the design, equipment procurement, construction and checkout of a comprehensive conventional munitions disposal facility located south of Berlin, Germany

Dr. David Hazlebeck is a key technical lead on all of GA's SCWO projects and will continue in this role with the SCWG program. He brings over 10 years of experience in the design and testing of chemical process equipment and SCW systems. He is currently the Project Manager for the DARPA-sponsored effort leading to a modular, highly compact SCWO system for the treatment of excess hazardous materials onboard Naval vessels.

Mr. Kevin Downey was the lead process engineer during Phase I SCWO and SCWG testing and will continue in this role in subsequent phases. He brings over 15 years experience in the design and testing of systems for the treatment and disposal of waste materials and the startup of advanced chemical process systems.

Dr. Glenn Hong, a consultant for GA, has been a key member of the Phase I SCWG design and analysis team. He brings over 20 years experience derived from his pioneering role in helping develop SCWO as a near-commercial business, including the receipt of numerous SCWO-related patents. He will continue to provide his first-hand knowledge of the chemical process industry to the Phase II through IV effort as analysis of test data and design of the pilot plant evolve.

At present, none of the Encina staff will be expected to play a key role in the program. The primary interface will continue to be Mr. Paul Bushee.

V.D TEAM MEMBER FACILITIES AND EQUIPMENT

GA provides the complete spectrum of facilities to design, build, and test SCWG equipment utilized during Phase I of the project. GA has a materials engineering facility with extensive capabilities for subcritical and supercritical fluid systems, corrosion and solid/fluid flow testing, metals and ceramics, research and joining/fabrication technology along with complete familiarity with all associated specifications, codes and standards. Analyses of stress corrosion, erosion, and high temperature gaseous corrosion can be performed with the aid of metallurgical diagnostics.

GA has manufacturing facilities used to fabricate specialized components and systems for military applications to ASME codes, including a documented quality assurance system. We also has developed a network of manufacturing subcontractors throughout Southern California capable of performing any processes required in the construction of advanced, high-technology systems.

Phase I SCWG studies were performed in the pilot-scale facility located in San Diego previously used for a broad range of SCWO tests. This system will continue to be used for pilot-scale testing during proposed Phase II testing. During Phases III and IV, the pilot plant will be reconfigured as needed to accommodate the larger throughput planned for these stages of testing. This will include a larger gasifier vessel and related components as needed.

Encina has facilities that could be used to house the SCWG systems. Space is available in several buildings near the sludge processing area that could house the SCWG equipment and interface with the existing sludge treatment works.

V.E TEAM MEMBER STATEMENTS OF COMMITMENT

GA views renewable energy sources as vital to power generation in the decades to come and is committed to their development for the generation of hydrogen. In addition, Encina Wastewater Authority has indicated its continuing interest in the program for hydrogen generation via SCWG of sewage sludge. We look forward to the opportunity to continuing this effort together with the DOE in the development of SCWG for production of hydrogen from biomass fuels.

ENCINA WASTEWATER AUTHORITY

Serving North San Diego County

ENCINA WASTEWATER AUTHORITY

A Public Agency

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(760) 431-7493 (Admin)

December 17, 1997

Ref: 3518

General Atomics
P.O. Box 85608
San Diego, CA 92138-9784

Attention: Dr. Dan Jensen

SUBJECT: Supercritical Water Gasification of Wastewater Biosolids

Earlier in the year the Encina Wastewater Authority (EWA) provided primary and secondary biosolids for use in General Atomics' (GA) pilot plant supercritical waste oxidation (SCWO) and SCWG studies. We recently met to discuss the results for both the SCWO workup tests and SCWG runs for hydrogen production.

The EWA's facility uses a variety of advanced treatment technologies to process the wastewater from over 225,000 residents in our service area, while recovering energy where possible. In the future, we believe that SCWG could potentially provide both an economical means of treating wastewater in an environmentally friendly manner and serve as a reliable means of producing hydrogen for on-site power production or off-site sale.

In light of this, we would like to express our continuing interest in the development of SCWG as a potential alternate means of processing both biosolids feed and solids effluent. If you have any questions please do not hesitate to contact me at (760) 727-3614 or E
-mail me at "PAUL@Encinajpa.com".

Very truly yours,

Paul Bushee

Paul Bushee
Resource Reclamation Specialist

PB:lc

xc: John Murk, EWA General Manager
Mike Hogan, EWA Director of Operations
Mike Fileccia, EWA Technical Services Director

VI. RESOURCE REQUIREMENTS FOR PHASES II - IV

Resource requirements for Phases II through IV were developed from the proposed project planning described in Sec. IV. Personnel, equipment, materials, supplies and other requirements were defined. These were then integrated into budgets for each phase and funding requirements defined.

VI.A PERSONNEL

Table VI-1 shows the personnel staffing requirements for each phase of the project. Key personnel discussed in Sec. VC will be assisted by staff personnel at GA.

Table VI-1. Personnel Requirements for Phases II through IV

Position Title	Phase II Man-hours	Phase III Man-hours	Phase IV Man-hours	Total Man-hours
Technical Staff	7040	15,920	5760	28,720
Project Management	480	1440	480	2400

VI.B EQUIPMENT, MATERIALS, AND SUPPLIES

Table VI-2 presents a list of equipment, materials and supplies required for Phases II through IV.

VI.C OTHER RESOURCE REQUIREMENTS

Other resources include the GA pilot plant and supporting facilities, utilities, computer control system, sewage sludge receiving/holding equipment and related items.

Table VI-2. Required Equipment, Materials and Supplies

Phase	Item
II	Sewage sludge press
	Sewage sludge liquefaction components
	Updated pumping system components
	Laboratory supplies
	Analytical services
III	Updated liquefaction/pumping system components
	Gas-fired trim heater
	New or altered GA SCW gasifier
	Heat recovery heat exchanger and waste heat boiler
	Refurbished gas/solid separator
	Analytical services
IV	Leased membrane and PSA units
	Analytical services

VI.D TOTAL BUDGET ESTIMATE

Table VI-3 presents the budgetary estimate for Phases II through IV based on the proposed scope of work are as follows

Table VI-3. Budgetary Estimate for Phases II through IV

Phase	Budget ($)
Phase II, Technology Development	1,287,750
Phase III, Technology Validation	3,364,235
Phase IV, Demonstration of Scale-Up	993,697
Total for all phases	5,645,682

VI.E DOE FUNDING REQUIREMENTS AND CONSORTIUM COST SHARE

Table VI-4 presents the DOE and GA Team funding requirements for Phases II through IV.

Table VI-4. DOE Funding Requirements and GA Cost Share

Phase	Budget ($)	DOE Share ($)	GA Share ($)
Phase II, Technology Development:	1,287,750	1,030,200	257,550
Phase III, Technology Validation:	3,364,235	2,691,388	672,847
Phase IV, Demonstration of Scale-Up	993,697	496,849	496,848
Total for all phases	5,645,682	4,218,437	1,427,245

VII. REFERENCES

1. Antal, M.J. and X. Xu, "Total, Catalytic, Supercritical Steam Reforming of Biomass", Univ. of Hawaii, manuscript draft, 1997a.
2. Antal, M.J., S. Xu and J. Arteiro, "Monthly Progress Report for Coop. Agreement DE-FG36-94AL85804", September, 1997b.
3. Antal, M.J., "Catalytic Supercritical Gasification of Wet Biomass", International Patent Publication No. WO 96/30464, 1996.
4. Appleby, A.J., "Fuel Cells and Hydrogen Fuel", in *Hydrogen Energy Progress IX,* Societe des Ingenieurs et Scientifiques de France, Paris, 1992.
5. Bannister, R.L., D.J. Huber, R.A. Newby and J.A. Paffenbarger, "Hydrogen-Fueled Combustion Turbine Cycles", ASME Paper 96-GT-247, 1996a.
6. Bannister, R.L. and R.A. Newby, "Natural Gas-Fired Combustion Turbines for the 21st Century", citation needed, 1996b.
7. Barnes, C.M. and C.H. Oh, "Chemical Reactor Design for Supercritical Water Oxidation of U.S. DOE Waste", First Intl. Conf. on Advanced Oxidation Technologies for Water and Air Remediation, London, Ontario, Canada, 1994.
8. Bastian, R.K., "The Biosolids (Sludge) Treatment, Beneficial Use, and Disposal Situation in the USA", *European Water Pollution Control,* 7:62-79, 1997.
9. Blosser, R.O. and R.A. Miner, "Sludge Handling and Disposal Practices in the U.S. Pulp and Paper Industry", Environmental Protection in the 90's EUCEPA Symposium, Helsinki,1986.
10. Buchner, H., "Technology for Gaseous Hydrogen Production", *Hydrogen and Other Alternative Fuels for Air and Ground Transportation,* Pohl, H.W. ed., John Wiley & Sons, New York, 1995.
11. Dillon, A.C. et al., "Carbon Nanotube Materials for Hydrogen Storage", in *Proceedings of the 1995 U.S. DOE Hydrogen Program Review,* NREL/CP-430-20036, available from NTIS, 1995.
12. Elliott, D.C., L.J. Sealock, Jr. and E.G. Baker, "Method for the Catalytic Conversion of Organic Materials Into a Product Gas", U.S. Pat. No. 5,616,514, 1997.
13. Elvers, B., ed., *Ullman's Encyclopedia of Industrial Chemistry,* VCH Verlagsgesellschaft mbH, Weinheim, Germany, 1989.
14. "Test Report: Biosolids Gasification and Oxidation in Supewrcritical Water", GA Doc. 3264902 N/C, December 1997.
15. Golden, T., "Application of Activated Carbon in Hydrogen Pressure Swing Adsorption", Activated Carbon Conference, 1997.

16. Haussinger, P., R. Lohmuller and A.M. Watson, "Hydrogen", in *Ullmann's Encyclopedia of Industrial Chemistry*, 1985.
17. Hong, G.T., B. Borchers, S. Pisharody and J. Fisher, "A Supercritical Water Oxidation Unit for Treating Waste Biomass", SAE Paper No. 961560, 26th International Conference on Environmental Systems, Monterey, California, 1996.
18. Hong, G.T., "Process for Oxidation of Materials in Water at Supercritical Temperatures and Subcritical Pressures", U.S. Pat. No. 5,106,513, 1992.
19. Hynek, S. et al., "Hydrogen Storage by Carbon Sorption", in *Hydrogen Energy Progress X*, Florida Solar Energy Center, University of Central Florida, 1994.
20. Katofsky, R.E., *The Production of Fluid Fuels From Biomass*, Princeton University Center for Energy and Environmental Studies Report No. 279, 1993.
21. Kordesch, K. and J.C.T. Olivera, *Int. J. Hydrogen Energy* 13:411, 1988.
22. Kroschwitz, J.I., ed., *Kirk-Othmer Encyclopedia of Chemical Technology*, John Wiley & Sons, New York, 1995.
23. Mann, M.K., "Technical and Economic Assessment of Producing Hydrogen by Reforming Syngas from the Battelle Indirectly Heated Biomass Gasifier", NREL/TP-431-8143, August, 1995.
24. Marion, C.P. and b. Reynolds, "Synthesis Gas Generation", U.S. Pat. No. 3,743,606, 1973.
25. McMahon, M.A., R.M. Suggitt, R.J. McKeon and A. Brent, "Partial Oxidation of Sewage Sludge", U.S. Pat. No. 4,933,086, 1990.
26. McMahon, M.A., R.M. Suggitt, R.J. McKeon and A. Brent, "Partial Oxidation of Sewage Sludge", U.S. Pat. No. 4,983,296, 1991.
27. Michael, Keith P., "Industrial Gas: Surveying Onsite Supply Options", Chemical Engineering, January, 1997.
28. Miller, G.Q. and J. Stoecker, "Selection of a Hydrogen Separation Process", National Petroleum Refiners Association Annual Meeting, San Francisco, 1989.
29. Modell, M., "Treatment of Pulp Mill Sludges by Supercritical Water Oxidation", Final report, DOE Contract No. FG05-90CE40914, July, 1990.
30. Modell, M., "Gasification and Liquefaction of Forest Products in Supercritical Water", *Fundamentals of Thermochemical Biomass Conversion*, R.P. Overend, T.A. Milne, and L.K. Mudge, eds., Elsevier Applied Science, London, 1985.
31. Modell, M., R.C. reid and S.I. Amin, "Gasification Process", U.S. Pat. No. 4,113,446, 1978.
32. Najjar, M.S. and W.C. Gates, Jr., "Partial Oxidation of Solid Carbonaceous Fuel-Water Slurries", U.S. Pat. No. 4,666,464, 1987.

33. Parker, W.G. and F.P. Bevc, "SureCell™ Integrated Solid Oxide Fuel Cell/Combustion Turbine Power Plants for Distributed Power Applications", PowerGen 1995 - Americas, Anaheim, 1995.

34. Pelloux-Gervais, P., "Technology for Cryofuel Production", *Hydrogen and Other Alternative Fuels for Air and Ground Transportation*, Pohl, H.W. ed., John Wiley & Sons, New York, 1995.

35. Pope, W., Permea Division, Monsanto Corp., personal communication, October, 1997.

36. Rarig, D., Air Products and Chemical, Inc., personal communication, October, 1997.

37. Sealock, L.J. Jr. and D.C. Elliott, "Method for Catalytic Destruction of Organic Materials", U.S. Pat. No. 5,019,135, 1991.

38. Sealock, L.J. Jr., E.G. Baker and D.C. Elliott, "Method for the Catalytic Conversion of Lignocellulosic Materials", U.S. Pat. No. 5,630,854, 1997.

39. Sinor, J.E., *Technical Economics, Synfuels, and Coal Energy*, pp. 103-109, ASME, 1989.

40. Weibe, R., "The Binary System Carbon Dioxide-Water Under Pressure", *Chem. Rev.* 29:475-81, 1941.

41. Weitkamp, J. et al., "Zeolites as Media for Hydrogen Storage", *Int. J. of Hydrogen Energy*, 20:967-970, 1995.

42. Wilcher, F.P. et al., "Technologies for the Separation and Recovery of Hydrogen from Refinery Streams", Proceedings of the 210th National Meeting, ACS, Chicago, 1995.

43. Xu, X and M.J. Antal, "Gasification of Sewage Sludge and Organics in Supercritical Water", AIChE Annual Meeting, Los Angeles, 1997.

44. Xu, X., Y. Matsumura, J. Stenberg and M.J. Antal, Jr., "Carbon-Catalyzed Gasification of Organic Feedstocks in Supercritical Water", *Ind. Eng. Chem. Res.* 35:2522-30, 1996.

45. Yokoyama, S. Y., et al., "Application of Thermochemical Liquidization of Sewage Sludge" in Developments inThermochemical Biomass Converson (Bridgewater, A.V.; Boocock, D.G.B., eds.), Blackie Academic & Professional, London, 1997, pp. 328-338.

46. Yu, D., M. Aihara and M.J. Antal, Jr., "Hydrogen Production by Steam Reforming Glucose in Supercritical Water", *Energy and Fuels*, 7:574-77, 1993.

47. Zimmermann, F.J., "New Waste Disposal Process", *Chemical Engineering* pp. 117-120, August 25, 1958.

48. Zoborowski, M.E. and D.E. Radloff (Zimpro, Inc.), "Transportaion of Slurried Minerals by a High Pressure Cylindrical Diaphragm Pump", *3rd Int. Conf. on Slurry Transportation*, Las Vegas, 1978.

APPENDIX A
M&EBS FOR 20 AND 40 WT% BIOSOLIDS

MASS AND ENERGY BALANCE FOR 27 MT/DAY SCWG SYSTEM, 20% BIOSOLIDS, HIGHER HYDROGEN YIELDS

SCWG Mass and Energy Balance				**Plant size =**		**30**	**tpd combustible biosolids**										
Liquefied sludge solids, wt%		**20**				**39.6**	**tpd total biosolids**										
Stream No.	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**	**15**	**16**	**17**
Stream Name	Biosolids Feed	Liquefier Ash Purge	Liquefier Sludge	Pre-heated Sludge	Reactor Feed	Reactor Effluent	Partially Cooled Effluent	Cooled Effluent	Liquid + Solid Effluent	High Pressure Gas	Medium Pressure Gas	Membrane Fuel Gas	Membrane H2	PSA Fuel Gas	Mixed Fuel Gas	PSA H2	Storage H2
Parameter:																	
Temperature, C	25	200	200	444	650	650	296	40	40	40	25	25	25	25	25	25	25
Pressure, psia	14.7	3400	3400	3400	3400	3400	3400	3400	3400	3400	1950	20	500	20	20	500	200
Mass flow, kg/sec	1.8	0.2	1.6	1.6	1.6	1.6	1.6	1.6	1.2	0.4	0.4	0.3	0.2	0.2	0.4	0.0	0.0
Heat flow, MWatts	0.0	0.0	1.1	2.7	0.94	0.0	-2.71	-1.3	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Solids, kg/sec	0.42	0.10	0.32	0.32	0.32	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H2O, kg/sec	1.36	0.10	1.26	1.26	1.26	1.08	1.08	1.08	1.08	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H2, kg/sec	0.00	0.00	0.00	0.00	0.00	0.03	0.03	0.03	0.00	0.03	0.03	0.00	0.03	0.01	0.01	0.02	0.02
CO, kg/sec	0.00	0.00	0.00	0.00	0.00	0.02	0.02	0.02	0.00	0.02	0.02	0.01	0.01	0.01	0.02	0.00	0.00
CO2, kg/sec	0.00	0.00	0.00	0.00	0.00	0.40	0.40	0.40	0.06	0.34	0.34	0.20	0.13	0.13	0.34	0.00	0.00
CH4, kg/sec	0.00	0.00	0.00	0.00	0.00	0.04	0.04	0.04	0.00	0.04	0.04	0.03	0.00	0.00	0.04	0.00	0.00
								Regen HX balance					Stream 8 CO2			Stream 15 mol%	
Assumptions:								T guess	dH(G&S)	H(H2O)	MW		partial pressure			H2	27.1
For gaseous reaction products, use Antal 9/97 yields on poplar/corn starch.								296	-173.6	1311.8	-2.709		34	mol%		CO	5.4
Reaction assumed to be thermally neutral.													79	atm		CO2	51.4
All available fuel gas burned in trim heater.								Lower Heating Value of Fuel Gas								CH4	16.2
Heat losses from reactor and lines ignored.								Gas	Hc, kcal/mol	MW			Stream 11 mol%			Total	100.0
Heat capacity of nonwater constituents approximated as 1 J/g/K.								H2	57.8	1.0			H2	57.0			
Gas fired heater efficiency, %				30				CO	67.6	0.2			CO	3.2		H2 MW Equivalent	
Concentrated sludge solids, wt%				23	(before liquefier)			CH4	192	1.9			CO2	30.3		2.4	
Noncombustible content of sludge solids, wt%				25	(before liquefier)			Total		3.1			CH4	9.5			
Liquefied sludge solids, wt%				20	(after liquefier)								Total	100.0			
Noncombustible content of sludge solids, wt%				1	(after liquefier)			Fired heater balance									
Ash purge from liquefier is 50% solids.								T guess	dH(G&S)	H(H2O)	MW		Stream 13 mol%				
								444	0.1	2954.6	0.940		H2	78.2			
													CO	1.9			
								Excess enthalpy in steam loop					CO2	18.4			
								0.27	MW				CH4	1.5			
													Total	100.0			
								H2 Production									
								772,010	scfd								
								770,000	scfd for this size plant in M.K. Mann study of Batelle gasifier (woody biomass)								

GENERAL ATOMICS

MASS AND ENERGY BALANCE FOR 27 MT/DAY SCWG SYSTEM, 40% BIOSOLIDS, HIGHER HYDROGEN YIELDS

SCWG Mass and Energy Balance				**Plant size =**		**30**	**tpd combustible biosolids**										
Liquefied sludge solids, wt%		**40**				**39.6**	**tpd total biosolids**										
Stream No.	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**	**15**	**16**	**17**
Stream Name	Biosolids Feed	Liquefier Ash Purge	Liquefier Sludge	Pre-heated Sludge	Reactor Feed	Reactor Effluent	Partially Cooled Effluent	Cooled Effluent	Liquid + Solid Effluent	High Pressure Gas	Medium Pressure Gas	Membrane Fuel Gas	Membrane H2	PSA Fuel Gas	Mixed Fuel Gas	PSA H2	Storage H2
Parameter:																	
Temperature, C	25	200	200	372	650	650	383	40	40	40	25	25	25	25	25	25	25
Pressure, psia	14.7	3400	3400	3400	3400	3400	3400	3400	3400	3400	1950	20	500	20	20	500	200
Mass flow, kg/sec	1.0	0.2	0.8	0.8	0.8	0.8	0.8	0.8	0.3	0.5	0.5	0.3	0.2	0.2	0.4	0.0	0.0
Heat flow, MWatts	0.0	0.0	0.5	0.5	0.94	0.0	-0.52	-0.8	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Solids, kg/sec	0.42	0.10	0.32	0.32	0.32	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H2O, kg/sec	0.57	0.10	0.47	0.47	0.47	0.30	0.30	0.30	0.30	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H2, kg/sec	0.00	0.00	0.00	0.00	0.00	0.03	0.03	0.03	0.00	0.03	0.03	0.00	0.03	0.01	0.01	0.02	0.02
CO, kg/sec	0.00	0.00	0.00	0.00	0.00	0.02	0.02	0.02	0.00	0.02	0.02	0.01	0.01	0.01	0.02	0.00	0.00
CO2, kg/sec	0.00	0.00	0.00	0.00	0.00	0.40	0.40	0.40	0.02	0.38	0.38	0.23	0.15	0.15	0.38	0.00	0.00
CH4, kg/sec	0.00	0.00	0.00	0.00	0.00	0.04	0.04	0.04	0.00	0.04	0.04	0.03	0.00	0.00	0.04	0.00	0.00
	*	*	*														
								Regen HX balance					Stream 8 CO2			Stream 15 mol%	
Assumptions:								T guess	dH(G&S)	H(H2O)	MW		partial pressure			H2	25.3
For gaseous reaction products, use Antal 9/97 yields on poplar/corn starch.								383	-130.9	2327.7	-0.524		34	mol%		CO	5.1
Reaction assumed to be thermally neutral.													79	atm		CO2	54.5
All available fuel gas burned in trim heater.								Lower Heating Value of Fuel Gas								CH4	15.1
Heat losses from reactor and lines ignored.								Gas	Hc, kcal/mol	MW			Stream 11 mol%			Total	100.0
Heat capacity of nonwater constituents approximated as 1 J/g/K.								H2	57.8	1.0			H2	54.8			
Gas fired heater efficiency, %				30				CO	67.6	0.2			CO	3.1		H2 MW Equivalent	
Concentrated sludge solids, wt%				42	(before liquefier)			CH4	192	1.9			CO2	33.0		2.4	
Noncombustible content of sludge solids, wt%				25	(before liquefier)			Total		3.1	*		CH4	9.2			
Liquefied sludge solids, wt%				40	(after liquefier)								Total	100.0			
Noncombustible content of sludge solids, wt%				1	(after liquefier)			Fired heater balance									
Ash purge from liquefier is 50% solids.								T guess	dH(G&S)	H(H2O)	MW		Stream 13 mol%				
								372	0.1	1841	0.942		H2	76.3			
													CO	1.9			
								Excess enthalpy in steam loop					CO2	20.4			
								0.31	MW				CH4	1.4			
													Total	100.0			
								H2 Production									
								772,010	scfd								
								770,000	scfd for this size plant in M.K. Mann study of Batelle gasifier (woody biomass)								

STEAM BALANCES FOR 27 MT/DAY SYSTEM, HIGHER HYDROGEN YIELDS

SCWG Steam Production		Plant size =	30	tpd municipal sewage sludge solids			18:12	30-Dec-97	GTH		
1200 psig steam						DDJ					
Tsat = 569F or 298C	Case	20%	40%								
Liquefier loop steam recovery:				Trim heater gas analysis		Air In		Partial P	40C Pvap		
Available MW		0.27	0.31	Gas	Fuel Gas In	(100% excess)	Off Gas	psia	psia		
1200 psig steam, kg/sec		0.10	0.11	H2O	0.00		0.16	1.31	1.07		
1200 psig steam, kg/kg dry biomass		0.32	0.36	H2	0.01						
				CO	0.02						
Flue Gas steam recovery				CO2	0.38		0.52				
Available MW		2.14	2.14	CH4	0.04						
1200 psig steam, kg/sec		0.77	0.77	O2		0.46	0.23				
1200 psig steam, kg/kg dry biomass		2.46	2.46	N2		1.98	1.98				
				Total kg/sec	0.45	2.44	2.89				
Total 1200 psig steam recovery:		2.77	2.82	MW, 25C to 40C			0.05				
MMBtu/day		198	201								
Annual credit		$ 294,020	$ 298,476								
M. Mann Steam, kg/kg dry biomass			Heating stream	Heating stream	Heating stream	Heating stream	Steam	Steam	Steam	Steam	
Location	psig	Amount	Initial T, C	Final T, C	Initial T, F	Final T, F	Initial T, C	Final T, C	Initial T, F	Final T, F	
Between shift reactors	500	0.32	434.8	200.0	815	392	17.3	254.5	63	490	
Fuel cell	100	1.26	203.3		398						
Air compression	100	0.12	212.7	189.3	415	373					
Combustor flue gas	100	0.43	238.3	27.8	461	82	15.4	182.2	60	360	
Gas to PSA	100	0.85	221.0	23.9	430	75	15.4	204.2	60	400	
Checking M.Mann Steam Generation Calcs					276.564						
Location	psig	Btu/hr/lb of wood	Steam Btu/lb	lb steam/lb wood							
Between shift reactors	500	365.5	1194	0.31	46.094						
Syngas compression	100	1530	1178	1.30	92.188						
Air compression	100	131.3	1178	0.11							
Combustor flue gas	100	485.3	1178	0.41							
Gas to PSA	100	967.7	1199	0.81							
Steam credits											
M. Mann											
500 psi	$3.57	/1000 lb	$ 4.73	/MMBtu							
100 psi	$2.35	/1000 lb	$ 3.12	/MMBtu							
Modell 1990											
1200 psi	$5	/MMBtu									
600 psi	$4	/MMBtu									
150 psi	$3	/MMBtu									

MASS AND ENRGY BALANCE FOR 27 MT/DAY SCWG SYSTEM, 20% BIOSOLIDS, LOWER HYDROGEN YIELDS

SCWG Mass and Energy Balance				Plant size =		30	tpd combustible biosolids										
Liquefied sludge solids, wt%		20				39.6	tpd total biosolids										
Stream No.	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17
Stream Name	Biosolids Feed	Liquefier Ash Purge	Liquefier Sludge	Pre-heated Sludge	Reactor Feed	Reactor Effluent	Partially Cooled Effluent	Cooled Effluent	Liquid + Solid Effluent	High Pressure Gas	Medium Pressure Gas	Membrane Fuel Gas	Membrane H2	PSA Fuel Gas	Mixed Fuel Gas	PSA H2	Storage H2
Parameter:																	
Temperature, C	25	200	200	445	650	650	327	40	40	40	25	25	25	25	25	25	25
Pressure, psia	14.7	3400	3400	3400	3400	3400	3400	3400	3400	3400	1950	20	500	20	20	500	200
Mass flow, kg/sec	1.8	0.2	1.6	1.6	1.6	1.6	1.6	1.6	1.3	0.3	0.3	0.2	0.1	0.1	0.3	0.0	0.0
Heat flow, MWatts	0.0	0.0	1.1	2.7	0.93	0.0	-2.72	-1.7	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Solids, kg/sec	0.42	0.10	0.32	0.32	0.32	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H2O, kg/sec	1.36	0.10	1.26	1.26	1.26	1.20	1.20	1.20	1.20	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H2, kg/sec	0.00	0.00	0.00	0.00	0.00	0.02	0.02	0.02	0.00	0.02	0.02	0.00	0.01	0.00	0.00	0.01	0.01
CO, kg/sec	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO2, kg/sec	0.00	0.00	0.00	0.00	0.00	0.30	0.30	0.30	0.07	0.23	0.23	0.14	0.09	0.09	0.23	0.00	0.00
CH4, kg/sec	0.00	0.00	0.00	0.00	0.00	0.05	0.05	0.05	0.00	0.05	0.05	0.05	0.01	0.01	0.05	0.00	0.00
					Same as 100 MTD SCWO Pla			Regen HX balance					Stream 8 CO2			Stream 15 mol%	
Assumptions:								T guess	dH(G&S)	H(H2O)	MW		partial pressure			H2	19.9
For gaseous reaction products, use Antal 9/97 yields on poplar/corn starch.								327	-121.1	1482.1	-2.722		39	mol%		CO	0.8
Reaction assumed to be thermally neutral.													89	atm		CO2	49.5
All available fuel gas burned in trim heater.								Lower Heating Value of Fuel Gas								CH4	29.8
Heat losses from reactor and lines ignored.								Gas	Hc, kcal/mol	MW			Stream 11 mol%			Total	100.0
Heat capacity of nonwater constituents approximated as 1 J/g/K.								H2	57.8	0.5			H2	47.0			
Gas fired heater efficiency, %				30				CO	67.6	0.0			CO	0.5		H2 MW Equivalent	
Concentrated sludge solids, wt%				23	(before liquefier)			CH4	192	2.6			CO2	32.7		1.3	
Noncombustible content of sludge solids, wt%				25	(before liquefier)			Total		3.1			CH4	19.7			
Liquefied sludge solids, wt%				20	(after liquefier)								Total	100.0			
Noncombustible content of sludge solids, wt%				1	(after liquefier)			Fired heater balance									
Ash purge from liquefier is 50% solids.								T guess	dH(G&S)	H(H2O)	MW		Stream 13 mol%				
								445	0.1	2959.6	0.933		H2	73.5			
													CO	0.4			
								Excess enthalpy in steam loop					CO2	22.7			
								0.59	MW				CH4	3.4			
													Total	100.0			
								H2 Production									
								411,179	scfd								
								770,000	scfd for this size plant in M.K. Mann study of Batelle gasifier (woody biomass)								

MASS AND ENERGY BALANCE FOR 27 MT/DAY SCWG SYSTEM, 40% BIOSOLIDS, LOWER HYDROGEN YIELDS

SCWG Mass and Energy Balance				Plant size =		30	tpd combustible biosolids					17:56					
Liquefied sludge solids, wt%		40				39.6	tpd total biosolids										
Stream No.	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17
Stream Name	Biosolids Feed	Liquefier Ash Purge	Liquefier Sludge	Pre-heated Sludge	Reactor Feed	Reactor Effluent	Partially Cooled Effluent	Cooled Effluent	Liquid + Solid Effluent	High Pressure Gas	Medium Pressure Gas	Membrane Fuel Gas	Membrane H2	PSA Fuel Gas	Mixed Fuel Gas	PSA H2	Storage H2
Parameter:																	
Temperature, C	25	200	200	373	650	650	395	40	40	40	25	25	25	25	25	25	25
Pressure, psia	14.7	3400	3400	3400	3400	3400	3400	3400	3400	3400	1950	20	500	20	20	500	200
Mass flow, kg/sec	1.0	0.2	0.8	0.8	0.8	0.8	0.8	0.8	0.44	0.3	0.3	0.2	0.1	0.1	0.3	0.0	0.0
Heat flow, MWatts	0.0	0.0	0.5	0.5	0.934	0.0	-0.525	-1.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Solids, kg/sec	0.42	0.10	0.32	0.32	0.32	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H2O, kg/sec	0.57	0.10	0.47	0.47	0.47	0.41	0.41	0.41	0.41	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H2, kg/sec	0.00	0.00	0.00	0.00	0.00	0.02	0.02	0.02	0.00	0.02	0.02	0.00	0.01	0.00	0.00	0.01	0.01
CO, kg/sec	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO2, kg/sec	0.00	0.00	0.00	0.00	0.00	0.30	0.30	0.30	0.02	0.28	0.28	0.17	0.11	0.11	0.28	0.00	0.00
CH4, kg/sec	0.00	0.00	0.00	0.00	0.00	0.05	0.05	0.05	0.00	0.05	0.05	0.05	0.01	0.01	0.05	0.00	0.00
	*	*	*														
				Same as 100 MTD SCWO Pla				Regen HX balance					Stream 8 CO2			Stream 15 mol%	
Assumptions:								T guess	dH(G&S)	H(H2O)	MW		partial pressure			H2	18.2
For gaseous reaction products, use Antal 9/97 yields on poplar/corn starch.								395	-95.6	2606.9	-0.526		39	mol%		CO	0.8
Reaction assumed to be thermally neutral.													89	atm		CO2	53.9
All available fuel gas burned in trim heater.								Lower Heating Value of Fuel Gas								CH4	27.2
Heat losses from reactor and lines ignored.								Gas	Hc, kcal/mol	MW			Stream 11 mol%			Total	100.0
Heat capacity of nonwater constituents approximated as 1 J/g/K.								H2	57.8	0.5			H2	44.2			
Gas fired heater efficiency, %				30				CO	67.6	0.0			CO	0.5		H2 MW Equivalent	
Concentrated sludge solids, wt%				42	(before liquefier)			CH4	192	2.6			CO2	36.7		1.3	
Noncombustible content of sludge solids, wt%				25	(before liquefier)			Total		3.1	*		CH4	18.5			
Liquefied sludge solids, wt%				40	(after liquefier)								Total	100.0			
Noncombustible content of sludge solids, wt%				1	(after liquefier)			Fired heater balance									
Ash purge from liquefier is 50% solids.								T guess	dH(G&S)	H(H2O)	MW		Stream 13 mol%				
								373	0.1	1857.1	0.934		H2	70.4			
													CO	0.4			
								Excess enthalpy in steam loop					CO2	26.0			
								0.64	MW				CH4	3.3			
													Total	100.0			
								H2 Production									
								411,179	scfd								
								770,000	scfd for this size plant in M.K. Mann study of Batelle gasifier (woody biomass)								

GENERAL ATOMICS

STEAM BALANCES FOR 27 MT/DAY SCWG SYSTEM, LOWER HYDROGEN YIELDS

SCWG Steam Production		Plant size =	30	tpd municipal sewage sludge solids			18:02	30-Dec-97	GTH	
1200 psig steam						DDJ				
Tsat = 569F or 298C	Case	20%	40%							
Liquefier loop steam recovery:				Trim heater gas analysis		Air In		Partial P	40C Pvap	
Available MW		0.59	0.64	Gas	Fuel Gas In	(100% excess)	Off Gas	psia	psia	
1200 psig steam, kg/sec		0.21	0.23	H2O	0.00		0.16	1.31	1.07	
1200 psig steam, kg/kg dry biomass		0.68	0.73	H2	0.01					
				CO	0.02					
Flue Gas steam recovery				CO2	0.38		0.52			
Available MW		-0.70	-0.70	CH4	0.04	Not required				
Available MW		2.13	2.13	CH4	0.04					
1200 psig steam, kg/sec		0.77	0.77	O2		0.46	0.23			
1200 psig steam, kg/kg dry biomass		2.45	2.45	N2		1.98	1.98			
				Total kg/sec	0.49	2.44	2.89			
Total 1200 psig steam recovery:		3.12	3.18	MW, 25C to 40C			0.05			
MMBtu/day		223	227							
Annual credit		$ 331,295	$ 337,075							
						Same as 100 MTD SCWO Plant				
M. Mann Steam, kg/kg dry biomass			Heating stream	Heating stream	Heating stream	Heating stream	Steam	Steam	Steam	Steam
Location	psig	Amount	Initial T, C	Final T, C	Initial T, F	Final T, F	Initial T, C	Final T, C	Initial T, F	Final T, F
Between shift reactors	500	0.32	434.8	200.0	815	392	17.3	254.5	63	490
Fuel cell	100	1.26	203.3		398					
Air compression	100	0.12	212.7	189.3	415	373				
Combustor flue gas	100	0.43	238.3	27.8	461	82	15.4	182.2	60	360
Gas to PSA	100	0.85	221.0	23.9	430	75	15.4	204.2	60	400
Checking M.Mann Steam Generation Calcs					276.564					
Location	psig	Btu/hr/lb of wood	Steam Btu/lb	lb steam/lb wood						
Between shift reactors	500	365.5	1194	0.31	46.094					
Syngas compression	100	1530	1178	1.30	92.188					
Air compression	100	131.3	1178	0.11						
Combustor flue gas	100	485.3	1178	0.41						
Gas to PSA	100	967.7	1199	0.81						
Steam credits										
M. Mann										
500 psi	$3.57	/1000 lb	$ 4.73	/MMBtu						
100 psi	$2.35	/1000 lb	$ 3.12	/MMBtu						
Modell 1990										
1200 psi	$5	/MMBtu								
600 psi	$4	/MMBtu								
150 psi	$3	/MMBtu								

GENERAL ATOMICS

APPENDIX B
TEST REPORT: SEWAGE SLUDGE GASIFICATION AND OXIDATION IN SUPERCRITICAL WATER

GENERAL ATOMICS

TEST REPORT: SEWAGE SLUDGE GASIFICATION IN SUPERCRITICAL WATER

GENERAL ATOMICS
P. O. BOX 85608
SAN DIEGO, CA 92186-9784

DECEMBER 1997

PREPARED FOR THE UNITED STATES
DEPARTMENT OF ENERGY
Under Cooperative Agreement
No. DE-FC36-97GO10216

LIST OF ACRONYMS

BOD	Biological oxygen demand
COD	Chemical oxygen demand
GA	General Atomics
HC	hydrocarbon
HCl	hydrochloric acid
MT	metric ton
NO_x	Nitrogen oxides
pg/g	picograms/gram
ppm	parts per million (1 part in 10^6)
SCW	Supercritical water
SCWG	Supercritical water gasification
SCWO	Supercritical water oxidation
SO_x	Sulfur oxides
TCLP	Toxicity Characteristic Leaching Procedure
TOC	Total organic carbon
TS	Total solids
TSS	Total suspended solid
VSS	Volatile suspended solids
VTS	Volatile total solids

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FIGURES

TABLES

1. INTRODUCTION

This report presents the results of sewage sludge gasification testing conducted at the General Atomics (GA) supercritical water (SCW) pilot plant during the period of April - November of 1997. Specific activities included the characterization of local municipal sewage sludge, preparation and performance of pumping tests, modification of the pilot plant feed system (to remove grit and to grind/emulsify the sewage sludge), definition and implementation of sewage sludge handling methods and devices (for personnel safety and odor control), performance of supercritical water oxidation (SCWO) workup tests with full and then partial oxidation of sewage sludge in combination with heat recovery and/or auxiliary fuel addition, and finally the production of a hydrogen-rich synthesis gas at high pressures and temperatures.

The supercritical water gasification (SCWG) tests generally verified the performance of laboratory tests conducted at the University of Hawaii at Manoa using a carbon catalyst bed, although the pilot plant tests yielded somewhat lower hydrogen concentrations and higher volatile hydrocarbon concentrations.

Section 2 presents the summary and conclusions for the test series. Section 3 discusses the tests that were conducted, the technical conclusions that were drawn from the data, and the technical uncertainties that still exist.

2. SUMMARY AND CONCLUSIONS

The conclusions GA has drawn from the test data are summarized below. In general, there are three qualifications that apply to all of the technical conclusions that should be kept in mind when the data are being interpreted.

1. The tests were generally of short duration (i.e., <8 hr for SCWO and <1.5 hr for SCWG), so long term reliability data were not obtained.
2. The test program included very few repeat tests, so the formal statistics are weak.
3. The test facility is of a reasonable size (up to 4 l/min) and provides good flexibility for the treatment of various feedstocks. However, it does have system limitations that prevented optimum sewage sludge gasification.

These qualifications imply that the data base that was developed, while valid and significant, is still incomplete, and all conclusions drawn should recognize this.

1. Sewage sludge with up to 10% solids could be pumped to operating pressure in a repeatable manner. The technique was to first macerate the feed and then use a proprietary high pressure pumping system to feed it to the SCW reactor. This combination worked well for 8-hr SCWO workup tests (performed to verify methods of sewage sludge preparation, sludge pumping and feeding to the reactor, and pressure letdown) and during shorter-duration SCWG tests.
2. Sewage sludge could be successfully pumped through the preheater piping to temperatures as high as 650°C. No signs of plugging were observed. Preheating was successfully demonstrated both with electric heat and with reactor exit heat recovery.
3. While the preheaters were capable of heating the sewage sludge feed to 650°C, the feed rate was limited to ~0.48 kg/min due to heater power constraints. Either more heater power or, more likely, some degree of heat recovery is required for higher throughputs.
4. Sewage sludge could be injected into the reactor through an existing GA-designed nozzle for extended periods without plugging.
5. Although no quantification was attempted, preheat of the sewage sludge feed to gasification temperatures (>600°C) will likely produce some degree of char that may

eventually cause plugging of the catalyst bed. Upon inspection of the bed following the gasification test of 11/24/97, some fine, char-like material was present at the inlet of the catalyst bed within the top 1 to 3 in.

6. Because of heat loss to the environment, the addition of tape heaters to the external reactor wall was necessary in order to maintain required temperatures.
7. The coconut shell carbon bed material selected for pilot plant testing was somewhat friable and prone to attrition and compaction. Relatively small reactor pressure fluctuations caused small fragments of bed material to fall through the bed support screen. Over time, the accumulation of bed material at the inlet of the pressure letdown valve resulted in a loss of system pressure control. A more robust bed material is needed.
8. The pressure letdown system functioned satisfactorily, but material erosion was a problem during extended SCWO workup tests that required certain parts to be replaced frequently. A more durable material or material coating should provide better performance. Removal of ash from the feed will also reduce erosion.
9. Liquid effluent TOC levels were approximately 1500 to 1700 ppm, indicating a TOC destruction of ~94% from the initial concentration in the sewage sludge of ~26,500 ppm. Further process optimization is required for complete TOC destruction.
10. The hydrogen concentration in the gaseous effluent was approximately 25 volume %, somewhat lower than laboratory-scale tests which yielded concentrations of 33% (Ref. 1). The difference is due mostly to the presence of significant quantities of higher molecular weight hydrocarbons (generally C_2 to C_6) in the pilot plant tests that were not found in high concentration in the laboratory tests.
11. The estimated conversion of carbon to volatile carbon-containing species was 94% (by wt%). The estimated conversion of hydrogen to H_2 gas was 28.5%. Much of the hydrogen remained bonded in organic species, principally methane, ethane, and propane.

3. TEST RESULTS

A series of SCWO workup tests and SCWG tests were carried out with sewage sludge in GA's SCW pilot plant. The purpose of the tests was to verify laboratory-scale results, define pilot plant design operating conditions, determine areas of technical uncertainty, and establish parameters for the economic assessment of commercial-scale units. Table 3-1 summarizes the key features of the tests. Feed material and effluent sampling and analyses were performed during workup tests and during all sewage sludge feed tests. Operating data were collected during all tests with the pilot plant automated data acquisition system.

Figure 3-1 shows as-received and blended sewage sludge feed, SCWO reactor effluent immediately after discharge, and effluent after ~1 to 2 hr to allow time for settling of particulates. Figure 3-2 shows similar materials for the SCWG tests. These results are typical of those found during tests performed at optimized conditions.

The following sections describe the waste feed characteristics and key findings of the tests.

3.1. FEED CHARACTERISTICS OF SEWAGE SLUDGE

Primary and secondary sewage sludge was provided by the Encina Wastewater Authority in Carlsbad, California, located approximately 10 miles north of GA. The Encina plant treats the sewage for a population base of approximately 225,000 people and generates 90-100 MT/day of treated secondary waste at ~17 to 18 wt% solids. Table 3-3 presents minimum, mean and maximum values for various batches of the Encina primary and secondary sludge. Data are presented for TOC, total solids (TS), total suspended solids (TSS), volatile total solids (VTS), volatile suspended solids (VSS), and heavy metals content.

TABLE 3-1
TEST MATRIX FOR SCWO WORKUP TESTS

Test No.	Feed Type	Sludge Feed Rate (kg/min)	Temperature (°C)	Pressure (psi)	Test Date	Comments
1	Sludge at 0.5 wt%	0.32	650	23.4	4/21/97	Initial test of sludge processing
2	Sludge at 0.5 wt%	0.62	300-400	23.4	5/1/97	No oxidant feed
3	Sludge at 6 wt%	0.6	625	23.4	5/6/97	Workup tests at 6 wt%
4	Sludge at 6 wt%	0.75	610	20.8	5/7/97	
5	Sludge at 6 wt%	0.74	615	13.8	5/7/97	Performed to determine effect of pressure
6	Sludge at 6 wt%	0.66	605	7.72	5/13/97	
7	Sludge at 6 wt%	0.77	505	23.4	5/14/97	
8	Sludge at 6 wt%	0.85	590	23.4	5/9/97	Performed to determine effect of temperature
9	Sludge at 6 wt%	0.85	545	23.4	5/9/97	
10	Sludge at 6 wt%	0.85	595-635	23.4	5/9/97	Performed to determine effect of flow
11	Sludge at 6 wt%	0.76	520	23.4	5/14/97	
12	Sludge at 10 wt%	0.65	580	23.4	5/16/97	Performed to determine pressure and
13	Sludge at 10 wt%	0.67	505	23.4	5/16/97	temperature effects with 10% sludge.
14	Sludge at 10 wt%	0.82	500	2220	5/16/97	
15	Sludge at 10 wt%	0.65	550-610	23.4	5/23/97	Initial heat recovery heat exchanger test.

TABLE 3-2
TEST MATRIX FOR SCWG TESTS

Test No.	Feed Type	Sludge Feed Rate (kg/min)	Reactor Inlet Temp. (°C)	Pressure (MPa)	Test Date	Run Duration (min)	Comments
1a	Sludge at 0.5 wt%	0.62	400	23.4	5/1/97	60	Performed at very low concentration as a workup trial and to verify no significant solids deposition in preheater piping. No catalyst bed used.
1b	Sludge at 0.5 wt%	0.62	300	23.4	5/1/97	20	See above.
2	Sludge at 6 wt%	0.48	450-475	23.4	5/14/97	48	No catalyst bed used.
3	Sludge at 7.5 wt%	0.46	525	23.4	7/17/97	84	Coconut shell carbon catalyst bed used.
4	Sludge at 4.1 wt%	0.48	640-660	23.4	11/24/97	109	Coconut shell carbon catalyst bed used.

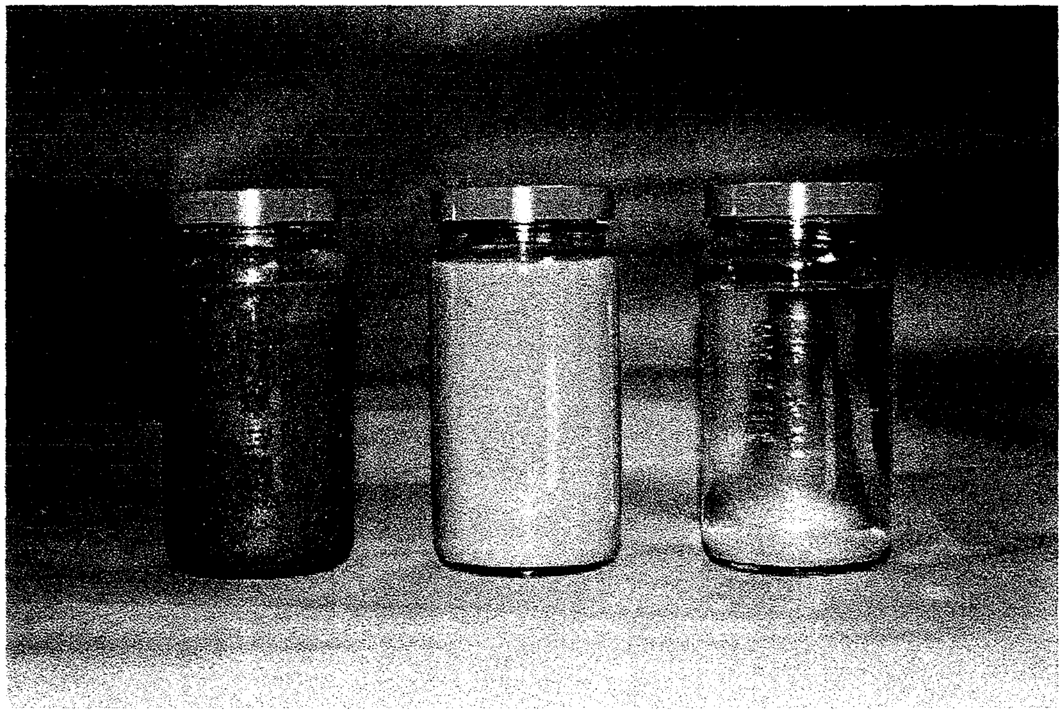


Fig. 3-1. Biosolids feed and SCWO effluent

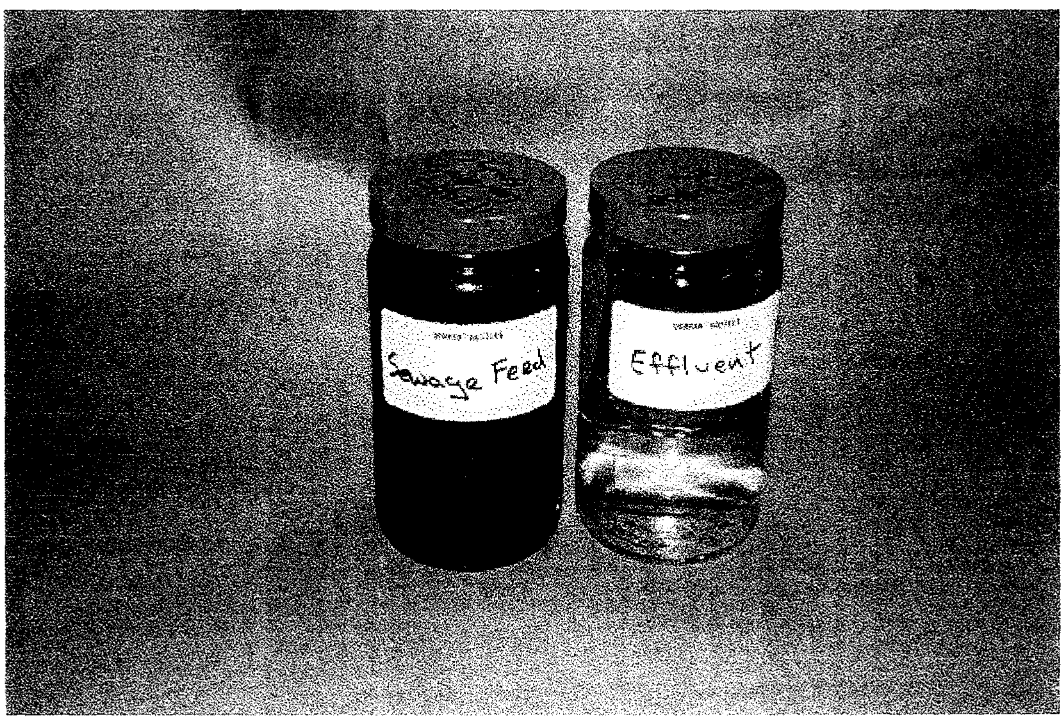


Fig. 3-2. Biosolids feed and SCWG effluent

**TABLE 3-3
MINIMUM, MEAN, AND MAXIMUM VALUES
OF SEWAGE SLUDGE FEED PARAMETERS**

	Primary Sludge			Secondary Sludge		
Parameter	Minimum	Mean	Maximum	Minimum	Mean	Maximum
TOC (ppm)	14,000	21,350	33,500	5300	12,320	16,000
TS (ppm)	49,700	85,790	152,000	20,000	39,910	64,800
TSS (ppm)	48,800	77,178	135,000	24,000	33,344	61,300
VTS (ppm)	10,300	64,760	131,000	3920	31,442	59,200
VSS (ppm)	32,000	63,727	113,000	20,400	27,989	51,100
As (ppm)	5.0	5.2	5.3	5.0	6.9	18.8
Cd (ppm)	0.4	0.74	1.1	0.4	0.94	1.3
Cr (ppm)	1.3	3.1	7.9	1.0	1.5	1.9
Cu (ppm)	9.5	16.7	38.5	5.4	9.1	15.8
Fe (ppm)	194	415	575	75.8	117.8	158
Ni (ppm)	1.0	1.1	1.1	-	-	-
Pb (ppm)	1.0	4.6	11.3	1.0	4.0	8.6
Hg (ppm)	0.1	0.34	0.5	0.1	0.35	0.5
Mo (ppm)	0.8	1.4	5.0	0.8	2.3	12.0
Se (ppm)	5.0	6.4	7.5	5.0	6.6	7.5
Zn (ppm)	34.1	60.1	95.1	18.6	27.2	38.5

Total carbon, hydrogen, and nitrogen content and heating values were also measured for two mixtures of ~10 wt% macerated mixture of 50% primary and 50% secondary sludge. Average concentrations (on a dry basis) were 43.75% carbon, 6.46% hydrogen, and 3.86% nitrogen. The average heating value (on a wet basis) was 1.90 MJ/kg (820 Btu/lb). Concentrations of sludge in the 50/50 mixture were also calculated for five feeds batches: May 14 - 7.4%; May 16 - 10.7%; May 23 - 4.6%; June 5 - 9.1%; and June 9 - 8.8%. Dioxin/furan levels of 10.2 picograms/gram (pg/g) were also measured.

3.2. SIZE REDUCTION AND PUMPING OF SEWAGE SLUDGE

A key objective of the test program was to verify that concentrated sewage sludge could be reliably pumped to the reactor for treatment. Following shakedown of equipment during simulant testing, sewage sludge was successfully pumped under all design conditions.

Sewage sludge pumping tests were performed to demonstrate that GA's proprietary pumping system could be used to pump sludge concentrations up to approximately 10 wt%. Testing involved two primary process steps: feed particle size reduction (performed as a feed pretreatment) and pumping.

Prior experience with pumping solids-containing streams indicated that the as-received sewage sludge would require size-reduction/maceration prior to use in the pilot plant. Several different size-reduction options were evaluated with simulated sludge (a mixture of breakfast cereal, seeds, and paper). An existing GA Gorator® macerator/pump combination, operating in a continuous recycle mode, was found to be the most effective of the size-reduction options tested and was, therefore, used throughout the test program (see Fig. 3-3). The Gorator® macerator was capable of producing particle sizes of <0.5 mm with sewage sludge concentrations up to ~10 wt%. At higher concentrations, plugging at the grinder entrance occurred. Plugging is less likely to occur in larger, commercial-scale grinders designed specifically for sewage sludge size reduction.

Upon receipt of sewage sludge from the Encina plant, the primary and secondary fractions were combined on an equal weight basis and mixed. If 4 to 6 wt% sludge was required for testing, the material was immediately size-reduced in the macerator. Macerator processing times were generally 15 to 30 minutes. If thickened sludge (~10 wt%) was required for testing, the mixed sludge was first treated with a polymer thickening agent which agglomerated the solids fraction and allowed water to be removed via draining through a filter (see Fig. 3.4). The thickened sludge was then processed through the Gorator® macerator. Once adequately size-reduced, the sludge was pumped into a barrel and then stored in a refrigerator until needed. (Note: a continuous process for size-reduction of sewage sludge is more appropriate for a commercial SCWG facility rather than the batch processing employed during pilot plant testing.)

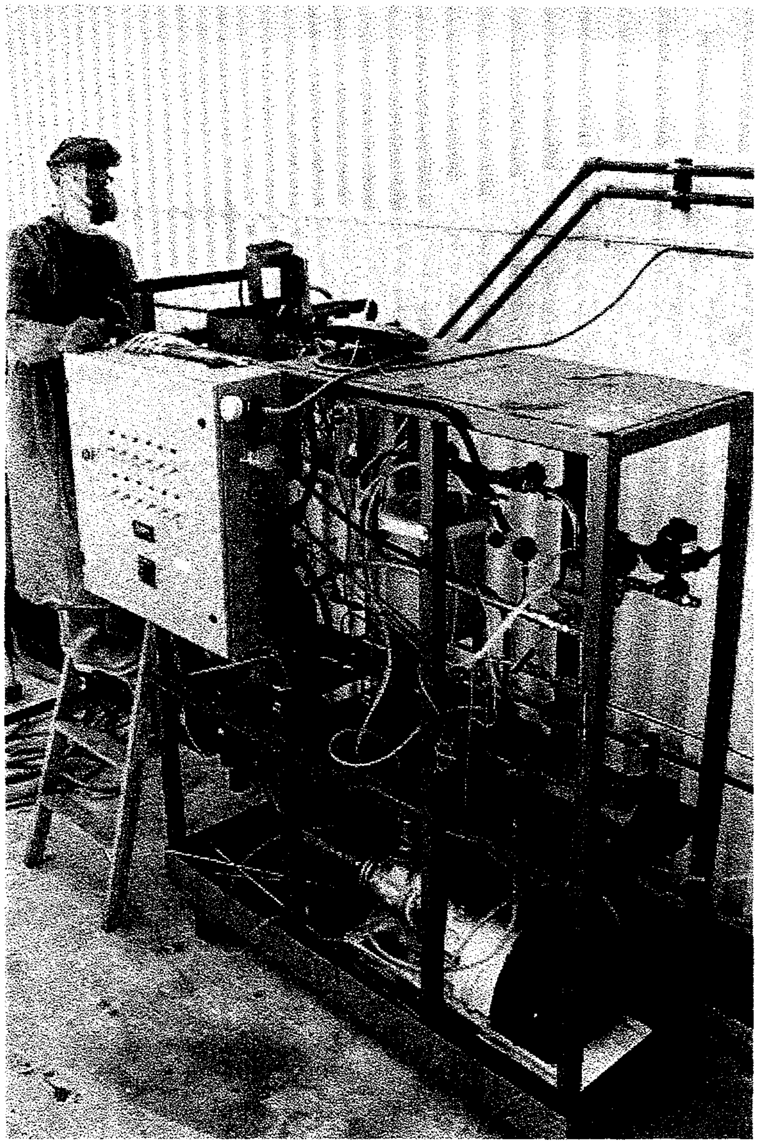
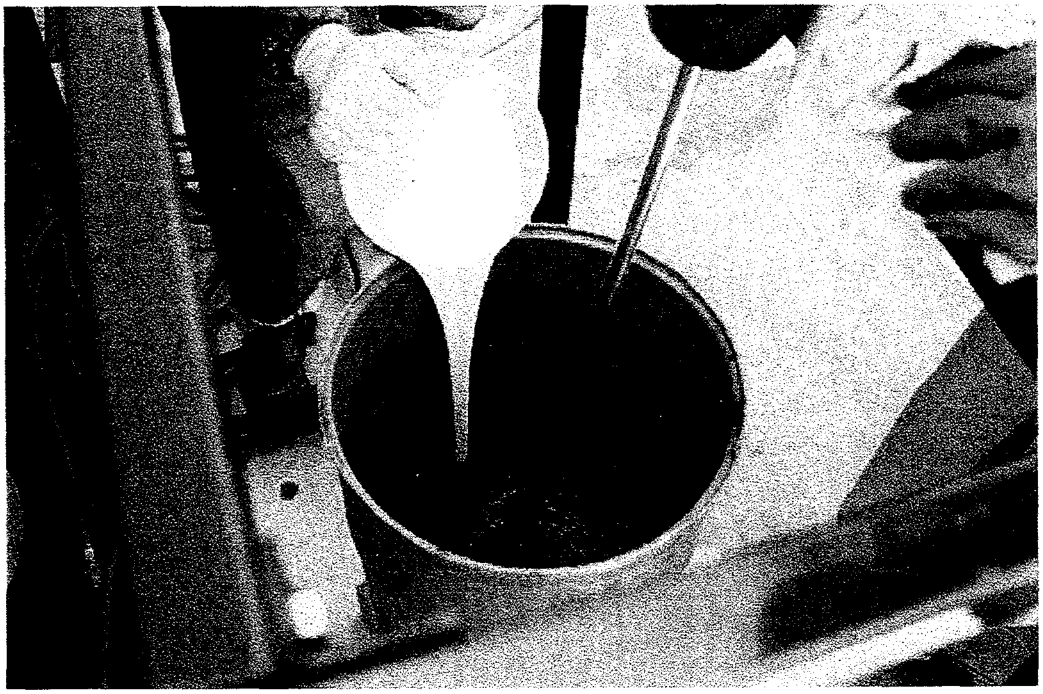


Fig. 3.3. Gorator® macerator and pump

(a)



(b)

Fig. 3.4. Biosolids filtering (a) and thickening (b)

The GA proprietary sewage sludge pumping system was demonstrated to be very effective in pumping sludge with solids contents to up 10.7 wt%. No additive to the feed was required to facilitate pumping. The system was first tested with a simulant in manual mode. After confirming that sewage sludge could be successfully pumped at the desired feed rates (up to ~1 kg/min), pump operations were fully automated and integrated into the pilot plant process control system. During subsequent testing, the pump was found to operate with a high degree of reliability, with little or no evidence of plugging or degradation. This same pump configuration can be utilized for a commercial-scale operation, although more long-term operational data are required to establish pump reliability.

3.3. TEST DESCRIPTION

All testing was performed in the GA pilot plant. Figure 3-5 shows a photograph of the pilot plant reactor/gasifier skid, and Fig. 3-6 shows a simplified process flow diagram. Use of the pilot plant for gasification testing imposed several limitations on test conditions. For example, electrical preheat of the feed to >600°C prior to entering the reactor limited the maximum sewage sludge feed rate due to heater power limits, and reactor wall materials limited maximum reactor operating temperatures. A carbon catalyst bed height of about 1/3 of the available reactor length was chosen to minimize abrasion of the top of the bed due to the feed injection methods employed. Therefore, the test conditions chosen for use during the final optimized test of 11/24/97 were:

Catalyst bed weight:	2.0 kg (~19 in. depth)
Pressure:	23.4 MPa (3400 psig)
Temperature:	600-650°C
Sewage sludge feed rate:	0.45-0.50 kg/min
Test duration:	>1 hr

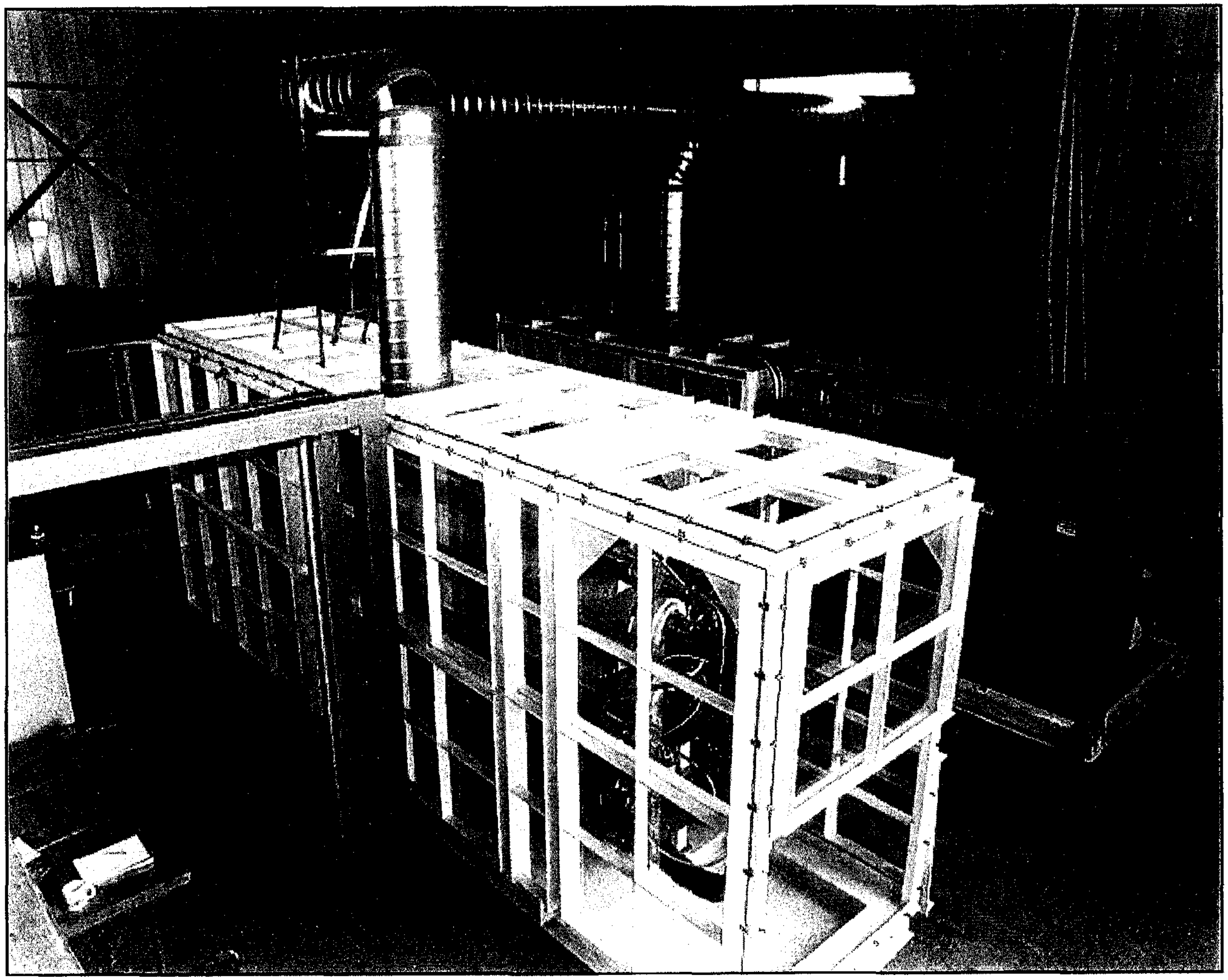


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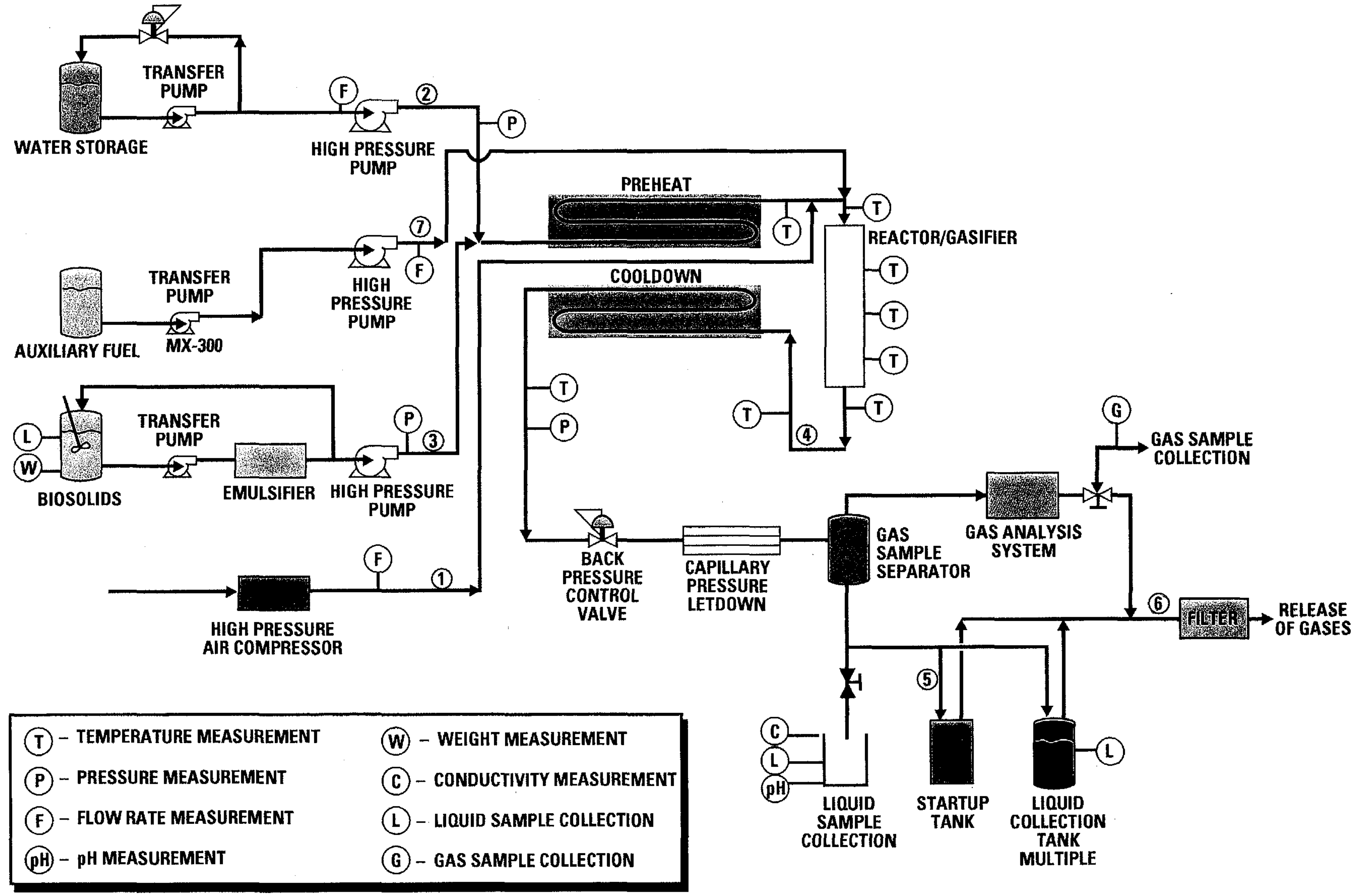
Fig. 3-5. GA SCW PILOT PLANT REACTOR SKID



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12-16-97

FIG. 3-6. PROCESS FLOW DIAGRAM FOR BIOSOLIDS SCWO/SCWG TESTS

The following is a simplified description of the procedures involved in performing the SCWG test of 11/24 /97. Procedures for a SCWO test will not be described herein.

1. **Prepare reactor.** Preparations included insertion of coconut carbon catalyst and mesh screen support into the reactor and addition of tape heaters along the reactor external surface.

2. **Heat reactor.** Following equilibration at the target pressure (23.4 MPa), reactor heatup was accomplished via a combination of preheat of reactor feed water and control of external tape heater power level. The water flow rate during heatup was 0.45-0.5 kg/min, and the reactor internal and external temperatures were 600-650°C.

3. **Begin feed of sewage sludge.** Sewage sludge feed from the GA proprietary pumping system was begun at a low flow rate. The flow rate of startup feed water was slowly reduced as the sewage sludge feed rate was increased, such that the system temperature remained relatively constant. This continued until an undiluted sewage sludge feed rate of 0.48 kg/min was attained.

4. **Maintain Conditions.** Temperature, pressure, and sewage sludge flow rate were maintained at target conditions of 600-650°C, 23.4 MPa, and 0.48 kg/min, respectively. Sewage sludge feed continued for approximately 1.25 hr.

5. **Collect Samples.** Liquid and gaseous effluent samples were collected every 5 to 10 min during sewage sludge feed. Liquid samples were collected in 250-ml glass bottles, and gas samples were collected in 3-l sample bags.

6. **Shut Down System.** Sewage sludge feed was terminated, and water flow was initiated. Reactor tape heaters were turned off.

Figure 3-7 presents a plot of the reactor internal and wall temperatures observed during the SCWG test of 11/24/97. The internal temperature at 40 in. was located just above the top of the carbon catalyst bed. The reactor pressure is also included. To aid in review of the plot, sewage sludge feed began at 1723 hours, reached full flow of 0.48 kg/min at 1756 hours, and ended at 1912 hours.

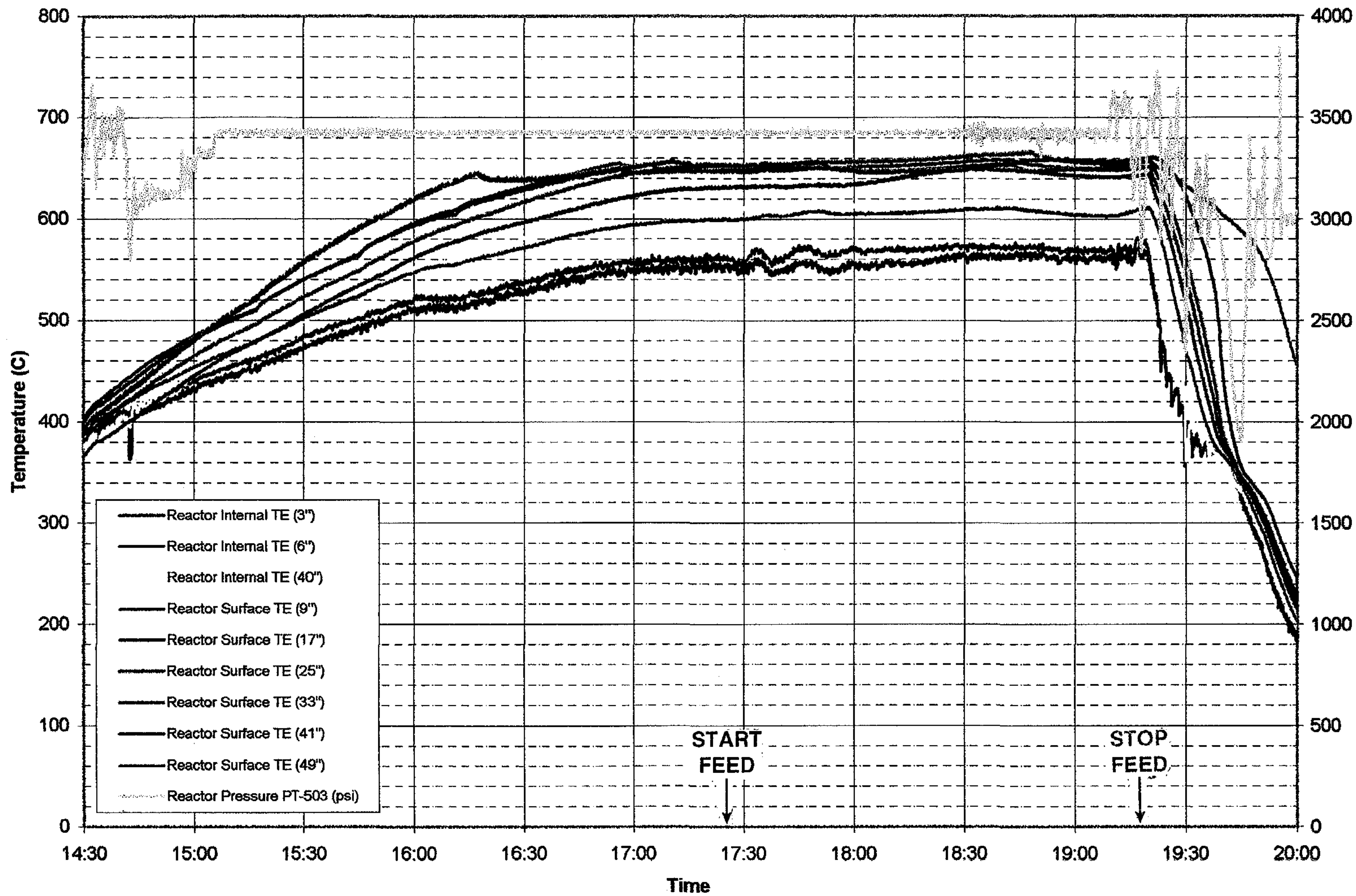

Fig. 3-7. REACTOR TEMPERATURE PROFILE FOR SCWG RUN OF 11/24/97

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3.4. ANALYTICAL RESULTS

During pilot plant testing, liquid and gaseous effluent samples are routinely collected for later analysis. Typical liquid analyses may include TOC, solids or ash content, anions, metals, and pH, while gas analyses will typically include component concentrations such as CO or CH_4. For the SCWG tests included in Table 3-2, liquid analyses were limited to TOC. More detailed analyses were performed on the gas samples, including analysis for H_2, CH_4, CO, CO_2, O_2, N_2, and a host of higher molecular weight hydrocarbons. Not all analyses were performed for all tests. The results of the liquid and gas sample analyses are presented below in Tables 3-4 and 3-5.

Typical sewage sludge feeds contain a small percentage of nonoxidizable or nongasifiable solids such as metal oxides or sand. For the gasification test of 11/24, for example, the ash content of the feed was measured at 0.8%. Because of the presence of the carbon catalyst bed, which tended to retain ash solids, no solids were contained in the liquid effluent during this run, at least during the early portion of the run when pressure control was optimum (see Section 3.6). Therefore no characterization of solids formed during the SCWG treatment of sewage sludge was attempted. For general comparison purposes, however, Fig. 3-8 shows the results of a particle size distribution analysis performed for a solid collected during SCWO treatment of sewage sludge.

Based on TOC analyses for the test of 11/24/97, the organic carbon content in the feed was approximately 2.65%. Assuming an average liquid effluent TOC concentration of 1600 ppm, the gasification efficiency for carbon (to volatile carbon-containing species) was 94%. No hydrogen analysis was performed on the feed for this test. However, assuming a carbon-to-hydrogen weight ratio in the feed of 6.7 to 1 (typical of prior analyses), the hydrogen gasification efficiency (for H_2 formation only) was estimated at 28.5%. These analyses neglect potential organic material holdup in the bed. However, based on post-test analysis, holdup was small.

3.5. HEAT RECOVERY

The hot reactor effluent from SCWO and SCWG processes can be used for heat recovery in two different ways. The energy can be used to preheat the feed material, thereby reducing electrical or gas-fired heater requirements, or the energy can be used for driving external processes such as turning a turbine for electricity generation. During pilot plant testing, only the use of heat recovery for feed preheat was utilized.

TABLE 3-4
LIQUID EFFLUENT ANALYSIS RESULTS

Test No.	Test Date	Sample No.	Sample Time	TOC (ppm)
1a	5/1/97	None	---	---
2	5/14/97	E31	1705	530
"	"	E34	1720	7960
3	7/17/97	E2	1345	589
"	"	E4	1418	786
"	"	E6	1438	746
4	11/24/97	Feed	---	26500
"	"	E2	1749	608
"	"	E3	1802	1160
"	"	E5	1815	1340
"	"	E6	1824	1560
"	"	E7	1829	1510
"	"	E9	1838	1570
"	"	E11	1854	1740

For low heating value feeds, such as sewage sludge, heat recovery is important. In order to maintain adequate reactor temperature to ensure full oxidation or gasification, heat input is required through feed preheat, auxiliary fuel addition, or a combination of both. The degree of feed preheat that can be used, at least for oxidation conditions, may be limited by pyrolysis. If the temperature of the sewage sludge feed reaches ~400°C, in the absence of oxygen, char formation may result, and char is difficult to fully oxidize. Under gasification conditions, heating to 600-650°C will undoubtedly produce some char, although the char produced presents more of a potential plugging problem than an inherent limitation on hydrogen production.

TABLE 3-5
GASEOUS EFFLUENT ANALYSIS RESULTS

Test No.	Test Date	Sample No.	Sample Time	H_2 (vol %)	CH_4 (vol %)	Non-CH_4HCs (vol %)	CO (vol %)	CO_2 (vol %)	O_2[(a)] (vol %)	N_2 (vol %)
1a	5/1/97	G1	1245	0.35	0.06	~0.25[(b)]	1.1	11	19	N/A[(c)]
2	5/14/97	G4	1648	0.04	0.003	N/A	0.004	1.2	16	N/A
"	"	G5	1655	0.2	0.04	N/A	0.06	0.68	13	N/A
"	"	G6	1703	1.3	0.89	N/A	1.1	4.5	10	N/A
"	"	G7	1721	3.3	3.08	N/A	2.5	7.4	13	N/A
3	7/17/97	G3	1433	18	21	>18.0[(4d]	22	19	0.2	1
"	"	G4	1452	17	20	>18.6[(d)]	22	21	0.1	0.4
4	11/24/97	G2	1759	24.3	23.1	5.6[(e)]	7.85	22.0	1.06	16.2
"	"	G4	1820	25.3	28.5	8.4[(e)]	10.3	24.4	0.51	2.70
"	"	G6	1840	24.5	28.4	8.9[(e)]	10.9	24.2	0.60	2.47

(a) The presence of significant oxygen concentrations in the gaseous effluent indicates a system leak whereby air is contaminating the sample. Measured concentrations above must be adjusted to compensate for dilution. (b) Sum of measured concentrations of multiple volatile organics (e.g., butane, butene, pentadiene, etc.).

(c) N/A = not available, analysis not performed.

(d) Sum of multiple volatile organics. Actual value slightly greater due to several volatile species beyond calibrated concentration.

(e) Sum of multiple volatile organics.

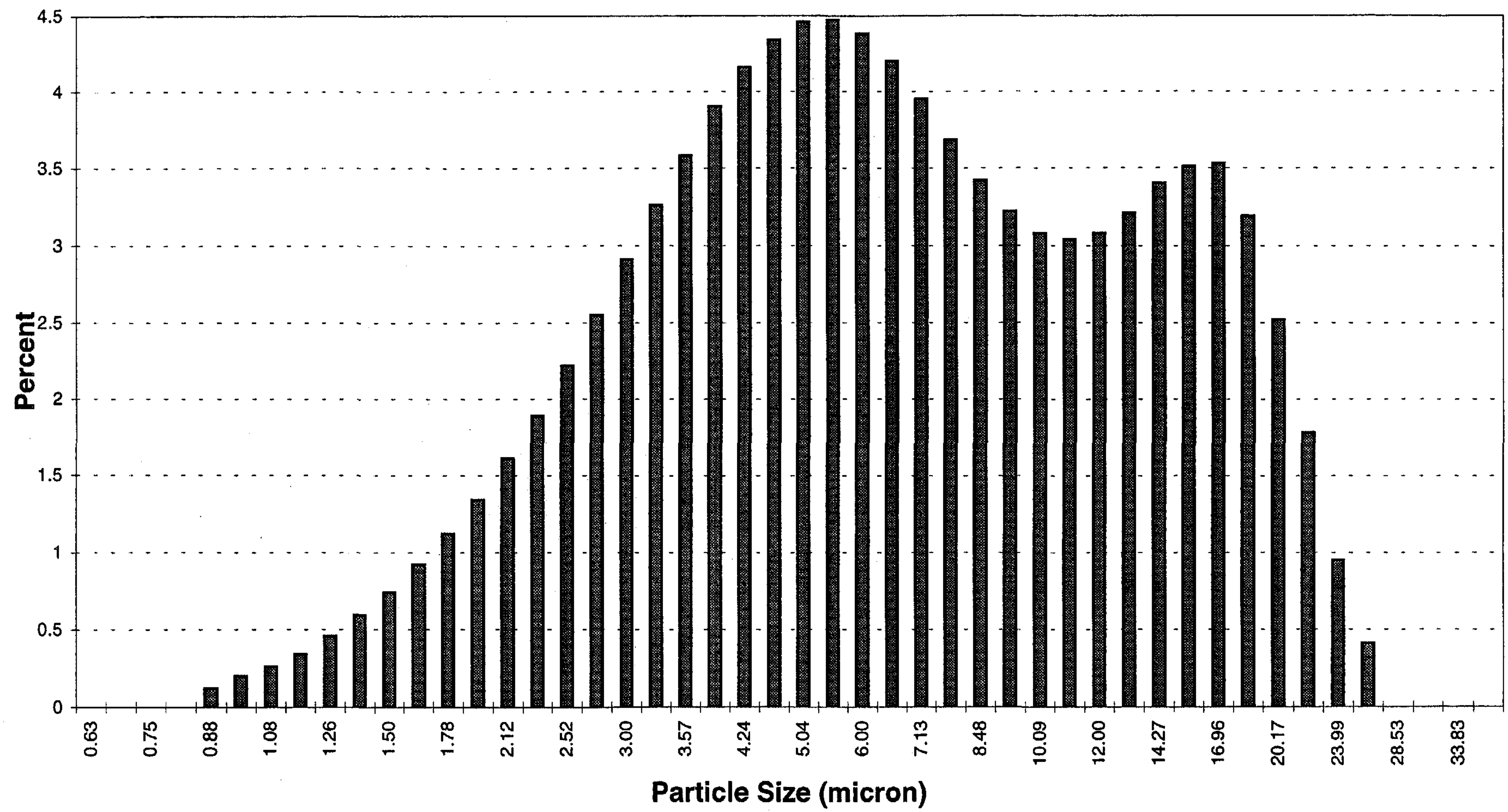

FIG. 3-8. SOLID EFFLUENT PARTICLE SIZE DISTRIBUTION FOR SCWO OF BIOSOLIDS

During SCWO workup testing, the feasibility of utilizing heat recovery was tested in three different ways: (1) simulated heat recovery using electrical preheat, (2) actual heat recovery heat exchange in combination with low level electrical preheat, and (3) actual heat recovery heat exchange alone. During tests utilizing electrical preheat, the preheater outlet control temperature was 300-400°C. Some solids accumulation in the preheater was observed, especially at lower flow rates. When the heat recovery heat exchanger was on line, exchanger outlet temperatures of 330-380°C were used. No signs of solids accumulation within the heat recovery heat exchanger were observed, probably due to the significantly higher velocities employed in the heat recovery heat exchanger relative to the preheater. Since the preheat temperature in all three cases was $\leq$400°C, the use of auxiliary fuel was required in order to achieve the desired reactor operating temperature of 550-650°C.

For SCWG processes, some degree of char formation is probably unavoidable. However, as long as overall char production is reasonably low and it does not present a plugging problem downstream, higher heat recovery temperatures can be used, lessening the need for external preheat.

3.6. PRESSURE EFFECTS

The coconut carbon catalyst chosen for use was quite friable. Thus, during both pilot plant tests utilizing the carbon catalyst bed, bed attrition was a problem. For example, during the test of 11/24/97, pressure control was excellent for about the first 70 minutes of sewage sludge feed (of ~110 minutes total sewage sludge feed time). Over this time, the liquid effluent samples were clear. Then, small bed particles began to pass through the bed support and collect in the downstream pressure control valve, thus worsening pressure control. With worse pressure control, the pressure fluctuations in the reactor began to increase, resulting in buffeting of the bed. This buffeting led to even more rapid bed attrition and even worse pressure control, eventually necessitating test termination. Clearly, bed attrition must be reduced for a commercial-scale SCWG process. Bed attrition can be reduced in several ways, including: (1) more robust catalyst materials, (2) improved pressure control methods to reduce reactor pressure fluctuations, (3) lower throughputs to reduce velocities through the bed, and (4) higher operating pressures to increase process fluid densities, thereby also reducing velocities through the bed. Additional testing is needed to identify the best method or methods for commercial-scale application.

3.7. TEMPERATURE EFFECTS

For typical SCWO homogeneous gas phase kinetics, the reaction rate for organic destruction is temperature dependent according to the Arrhenius equation. Relatively small increases in temperature will, therefore, yield relatively large increases in organic destruction. The kinetics effects are reasonably well understood and will not be discussed herein. To determine the temperature effects for SCWG applications, temperatures ranging from 300°C to 660°C were investigated as shown in Table 3-2. Unfortunately, for the early low-temperature tests of 5/1/97 and 5/14/97, no carbon catalyst bed was used, so the temperature effects are obscured by the catalytic effects of the bed. Therefore, the best runs for determination of the qualitative effects of temperature on hydrogen production are the runs of 7/17/97 and 11/24/97. Both runs utilized a 2.0 kg carbon catalyst bed (~19 in. deep within the reactor) with a similar sludge flow rate. Aside from temperature, the only other significant difference between the runs was the solids concentration of the feed, 7.5 wt% for the run of 7/17/97 and 4.1 wt% for the run of 11/24/97.

The Run of 7/17 97 was performed at a maximum reactor temperature of approximately 540°C. Based on reactor wall temperature data, the carbon catalyst bed temperature reached only about 525°C. The target temperature of ~600°C could not be achieved due to excessive heat loss from the reactor. The heat tapes added to the reactor wall to counteract heat loss had insufficient heating capability. For the test of 11/24/97, the reactor insulation and external reactor heat tape power were improved significantly such that the reactor and carbon catalyst bed could be uniformly maintained at ~650°C. At a catalyst bed temperature of ~525°C, gaseous effluent analyses show a hydrogen concentration of approximately 18 vol %, with methane and other higher molecular weight hydrocarbons totaling almost 40 vol %. At a catalyst bed temperature of ~650°C, the hydrogen concentration increases to about 25 vol %, while the methane/hydrocarbon concentration decreases only slightly. The major difference appears to be in the relative CO concentrations. One possible explanation is that at higher temperatures some steam reforming of the higher molecular weight hydrocarbons to form CO and hydrogen is taking place (1), while some of the CO is then being converted to CO_2 and hydrogen via the water-gas shift reaction (2). While some improvement in hydrogen production was realized by increasing the operating temperature from 525°C to 650°C, the improvement was only moderate, and the hydrogen concentration still falls somewhat below the laboratory-scale test value of 33 vol % (Ref. 1).

$$(1)\ C_nH_m + nH_2O + heat \rightarrow nCO + (m/2 + n)H_2$$

$$(2)\ CO + H_2O \rightarrow CO_2 + H_2 + heat$$

The liquid effluent TOC values were shown previously in Table 3-4 for the gasification runs of 7/17/97 and 11/24/97. The TOC values for the ~650°C run of 11/24/97 were significantly higher than those measured for the ~525°C run of 7/17/97. Generally, at least under oxidizing conditions, the TOC value is expected to decrease rapidly with an increase in temperature. One possible explanation for this could be if the above steam reforming and water gas-shift reactions are the rate limiting steps in hydrogen production from sewage sludge. At higher operating temperature, the pyrolysis of sewage sludge and the production of soluble organic species may be increased, and this increase will yield a comparable increase in the liquid phase TOC level unless the organics are themselves consumed by the steam reforming and water gas-shift reaction pathways. For the test of 11/24/97, the TOC concentration of the sewage sludge feed was 26,500 ppm. The effluent samples were generally about 1600 ppm, which yields a TOC destruction/conversion of 94%, neglecting any organic-containing solids which may have collected on the carbon catalyst bed.

3.8. PRESSURE LETDOWN

Pressure letdown was accomplished through a combination of pressure control valves and capillaries. The fine pressure control was performed by a single control valve. This valve was intended to take only a portion of the required system pressure drop in order to reduce valve wear. The remainder of the pressure drop was taken over a control valve and a capillary connected in parallel. This arrangement proved to work well for a portion of the testing. As long as bed material was not abraded and passed through the mesh support screen, pressure control was excellent, and the liquid effluent samples were clear. After about 1 hr of sewage sludge feed, however, some solid particles began to appear in the effluent, and pressure control began to degrade. This degradation of control led to some pressure cycling in the reactor which further abraded the bed material, which worsened pressure control. This cycle continued for about another 40 min, after which pressure control was poor enough to necessitate run termination. The solids exiting the reactor were black and appeared to be crushed carbon catalyst bed material. Another possibility is that the solids were char from pyrolysis of the sewage sludge feed which had slowly worked their way down through the catalyst bed. A post-test inspection, however, showed some fine carbon slurry present in the top 1-3 in. of the bed, probably a result of feed char. No signs of a similar carbon slurry were found elsewhere in the bed.

For a full-scale plant design, several system modifications could be made to improve pressure control, including use of more robust catalysts, better catalyst support methods, and larger, more forgiving control valves.

3.9. EFFECTIVENESS OF THE CARBON CATALYST

Laboratory-scale SCWG testing has shown near stoichiometric yields of hydrogen production for tests utilizing a carbon catalyst bed (Ref. 2). During pilot plant testing, tests were conducted both with and without a catalyst bed in an effort to shed further light on the bed effectiveness for gasification. The first two series of tests, as shown previously in Table 3-2, used maximum operating temperatures of only 300-475°C. During these tests, the maximum hydrogen concentration in the effluent was only 3.3 vol % (see Table 3-5). The presence of significant quantities of oxygen imply a system leak, but even accounting for the leak yields a maximum hydrogen concentration of only ~8.7 vol%.

The final two series of pilot plant tests utilized a carbon catalyst bed. The carbon used was a coconut shell carbon, supplied in 4-mm pellets from Barnabey and Sutcliffe. The size of the individual carbon particles was significantly greater than that used during laboratory-scale studies. Approximately 2 kg of carbon catalyst was used for each test. As shown in Table 3-2, these tests used significantly higher reactor temperatures (525°C and ~650°C) than the earlier tests without a catalyst. Very low oxygen concentrations were observed in the gaseous effluent, thus indicating that significant inleakage of air into the effluent samples did not occur. At a bed temperature of 525°C, the maximum hydrogen concentration in the effluent was measured at 18 vol %, about twice that observed at 475°C without a bed. At a reactor temperature of 640-660°C, the hydrogen concentration in the effluent increased to about 25 vol%. The increase in hydrogen concentration in the gaseous effluent showed a dependence on temperature as seen in the test results at 525°C and ~650°C, with higher temperatures appearing to be beneficial. The benefit of using a carbon catalyst bed is not quite as clear. Unfortunately, due to funding limitations, no test was performed at a prototypic temperature (600 to 650°C) without a carbon catalyst bed. Since the early runs performed without a carbon catalyst bed also utilized low temperatures, the temperature effect could not be separated from the bed effect.

3.10. COMPARISON WITH LABORATORY-SCALE DATA

Table 3-6 shows a comparison between the results of the pilot plant SCWG run of 11/24/97 and a laboratory test on sewage sludge (without feed enhancement additives)

(Ref. 1). The results compare relatively well. The laboratory data show somewhat higher hydrogen levels, while the pilot data show higher CO and hydrocarbon concentrations. It thus appears that the steam reformation and water gas-shift reactions discussed in Section 3.7 did not progress to the same extent in the pilot-scale testing as in the laboratory-scale work. More testing is required for process optimization.

TABLE 3-6
COMPARISON BETWEEN PILOT-SCALE AND LABORATORY-SCALE SCWG DATA FOR SEWAGE SLUDGE FEED

Parameter/Component	Pilot Plant Test of 11/24/97[a]	Laboratory-Scale Test
Temperature (°C)	600 to 650	600
Pressure (MPa)	23.4	34.4
WHSV[b] [(g/hr)/g]	0.6	0.5
Liquid TOC (ppm)	~1600	280
H_2 (vol %)	24.9	33
CO (vol %)	10.6	2.9
CO_2 (vol %)	24.3	36
CH_4 (vol %)	28.5	24
C_2 and above (vol %)	8.7	6.78

(a) Gas concentrations given are an average of the G4 and G6 analyses (see Table 3-5).
(b) WHSV = weight hourly space velocity (feed concentration x feed flow rate / amount of catalyst).

3.11. REFERENCES

1. Antal, M., X. Xu, Y. Matsumura, and J. Stenberg, "Hydrogen Production from High-Moisture Content Biomass in Supercritical Water", Proceedings of the 1995 U. S. DOE Hydrogen Program Review, NREL/CP-430-20036, Volume II, pp. 757-795, September 1995.

2. Xu, X., Y. Matsumura, J. Stenberg, and M. Antal, Jr., "Carbon-Catalyzed Gasification of Organic Feedstocks in Supercritical Water", Ind. Eng. Chem. Res., 1996, Vol. 35, pp. 2522-2530.

APPENDIX C
BACKUP FOR SYSTEM COSTING

GENERAL ATOMICS

COSTS FOR 27 MT/DAY SCWG SYSTEM, 20% BIOSOLIDS, HIGHER HYDROGEN YIELDS

PLANT SIZE, T/DAY BIOSOLIDS	30					
PRETREATED BIOLSOLIDS, wt%	20					
CAPITAL COSTS						
Item	**Size**	**Units**	**Size Basis**	**Materials**	**Cost**	**Cost Basis**
Biosolids storage tank	8111	gal	4 hours holdup +25% head space	Concrete	$17,000	Engineering estimate
Biosolids transfer pump	135	gpm	4x progressive cavity pump	SS	$11,000	0.6 exponent scale from 100 MTD SCWO Plant
Emulsifier macerator	135	gpm	4x progressive cavity pump	SS	$23,000	0.6 exponent scale from 100 MTD SCWO Plant
Progressive cavity pump	34	gpm	28 gpm required	SS	$78,000	0.6 exponent scale from 100 MTD SCWO Plant
Filter press	27	MT/day	M&EB	CS	$0	Not required
Liquefier/pump	1146	gal	30 minutes residence time	Ti/steel	$397,000	Estimated from Zimpro vessel
Heat recovery heat exchanger	2.71	MW	M&EB	C276	$810,000	0.6 exponent scale from 100 MT SCWO Plant
Gas-fired heater	3.1	MW	M&EB		$311,000	0.6 exponent scale from 100 MTD SCWO Plant
Flue gas heat exchanger	2.2	MW	M&EB	SS	$156,000	Half of fired heater
Reactor	1187	gal	1.5 minutes residence time	Alloy 718	$617,000	0.6 exponent scale from MODAR vessel x 2/3
Waste heat boiler	1.3	MW	M&EB	C276	$286,000	0.6 exponent scale from 100 MTD SCWO Plant
Gas/Liquid separator	172	gal	5 min liquid RT, 50% head space	SS	$68,000	0.6 exponent scale from 33 gal vessel
Liquid letdown valve	21	gpm	M&EB	TiN/SS	$16,000	Same as 100 MTD SCWO Plant
System pressure control valve	27	gpm	M&EB	SS	$19,000	0.6 exponent scale from liquid letdown valve
Membrane separator	772,010	scfd	M&EB		$200,000	Phone quote
PSA module	772,010	scfd	M&EB		$200,000	Mann report
Hydrogen storage tank			M&EB		$100,000	Engineering judgment.
Total major equipment					$3,309,000	
Bulk items factor					$4,467,000	1.35 times major equipment cost
Design & fab labor factor					$1,985,000	0.6 times major equipment cost
Control system					$80,000	GA SCWO systems
Facilities					$331,000	0.1 times major equipment cost
Startup cost					$662,000	0.2 times major equipment cost
TOTAL INSTALLED COSTS					$10,834,000	
Notes:						
1. Overall factor on major equipment	3.3					
2. All equipment sizing is at least 20% excess capacity over requirement.						
3. Antal high H2 yields assumed.						
OPERATING COST, $/YR @ 330 ANNUAL OPERATING DAYS						
Item	**Assumption**				**Cost**	**Cost Basis**
Labor	4 operators				$150,000	$18.75/hr
Utilities, etc.	1% of capital cost				$108,000	0.5 SCWO utility costs
Hydrogen credit	$10/GJ; 0.1184 GJ (lower)/kg				$ (697,000)	Mann report values
Steam credit	M&EB values				$ (298,000)	Mann report values
Feed credit	$90/bone dry ton				$ (891,000)	0.75 of Encina disposal cost
TOTAL					($1,628,000)	
IRR @ $120/ton avoided disposal cost						
- Initial investment	($10,834,000)					
- Year 1	$1,628,000	-85%				
- Year 2	$1,628,000	-53%				
- Year 3	$1,628,000	-31%				
- Year 4	$1,628,000	-18%				
- Year 5	$1,628,000	-9%				
- Year 6	$1,628,000	-3%				
- Year 7	$1,628,000	1%				
- Year 8	$1,628,000	4%				
- Year 9	$1,628,000	7%				
- Year 10	$1,628,000	8%				

GENERAL ATOMICS

COSTS FOR 27 MT/DAY SCWG SYSTEM, 40% BIOSOLIDS, HIGHER HYDROGEN YIELDS

PLANT SIZE, T/DAY BIOSOLIDS	30					
PRETREATED BIOLSOLIDS, wt%	40					
CAPITAL COSTS						
Item	**Size**	**Units**	**Size Basis**	**Materials**	**Cost**	**Cost Basis**
Biosolids storage tank	4516	gal	4 hours holdup +25% head space	Concrete	$12,000	Engineering estimate
Biosolids transfer pump	75	gpm	4x progressive cavity pump	SS	$8,000	0.6 exponent scale from 100 MTD SCWO Plant
Emulsifier macerator	75	gpm	4x progressive cavity pump	SS	$17,000	0.6 exponent scale from 100 MTD SCWO Plant
Progressive cavity pump	19	gpm	15 gpm required	SS	$54,000	0.6 exponent scale from 100 MTD SCWO Plant
Filter press	27	MT/day	M&EB	CS	$125,000	Vendor quote/engineering judgment
Liquefier/pump	638	gal	30 minutes residence time	Ti/steel	$279,000	Estimated from Zimpro vessel
Heat recovery heat exchanger	0.52	MW	M&EB	C276	$104,000	0.6 exponent scale from 10 MTD SCWO Plant
Gas-fired heater	3.1	MW	M&EB		$311,000	0.6 exponent scale from 100 MTD SCWO Plant
Flue gas heat exchanger	2.2	MW	M&EB	SS	$156,000	Half of fired heater
Reactor	621	gal	1.5 minutes residence time	Alloy 718	$418,000	0.6 exponent scale from MODAR vessel x 2/3
Waste heat boiler	0.8	MW	M&EB	C276	$85,000	0.6 exponent scale from 10 MTD SCWO Plant
Gas/Liquid separator	47	gal	5 min liquid RT, 50% head space	SS	$31,000	0.6 exponent scale from 33 gal vessel
Liquid letdown valve	6	gpm	M&EB	TiN/SS	$16,000	Same as 100 MTD SCWO Plant
System pressure control valve	30	gpm	M&EB	SS	$42,000	0.6 exponent scale from liquid letdown valve
Membrane separator	772,010	scfd	M&EB		$200,000	Phone quote
PSA module	772,010	scfd	M&EB		$200,000	Mann report
Hydrogen storage tank			M&EB		$100,000	Engineering judgment.
Total major equipment					$2,158,000	
Bulk items factor					$2,913,000	1.35 times major equipment cost
Design & fab labor factor					$1,295,000	0.6 times major equipment cost
Control system					$80,000	GA SCWO systems
Facilities					$216,000	0.1 times major equipment cost
Startup cost					$432,000	0.2 times major equipment cost
TOTAL INSTALLED COSTS					$7,094,000	
Notes:						
1. Overall factor on major equipment	3.3					
2. All equipment sizing is at least 20% excess capacity over requirement.						
3. Antal high H2 yields assumed.						
OPERATING COST, $/YR @ 330 ANNUAL OPERATING DAYS						
Item	**Assumption**				**Cost**	**Cost Basis**
Labor	4 operators				$150,000	$18.75/hr
Utilities, etc.	1% of capital cost				$71,000	0.5 SCWO utility costs
Hydrogen credit	$10/GJ; 0.1184 GJ (lower)/kg				$ (697,000)	Mann report values
Steam credit	M&EB values				$ (298,000)	Mann report values
Feed credit	$90/bone dry ton				$ (891,000)	0.75 of Encina disposal cost
TOTAL					($1,665,000)	
IRR @ $120/ton avoided disposal cost						
- Initial investment	($7,094,000)					
- Year 1	$1,665,000	-77%				
- Year 2	$1,665,000	-38%				
- Year 3	$1,665,000	-16%				
- Year 4	$1,665,000	-2%				
- Year 5	$1,665,000	6%				
- Year 6	$1,665,000	11%				
- Year 7	$1,665,000	14%				
- Year 8	$1,665,000	17%				
- Year 9	$1,665,000	18%				
- Year 10	$1,665,000	20%				

ANALYSIS OF BCL COST ESTIMATES BY M. MANN

Line No.	30 tpd	Scheme 2					Closest Mann	
	Equipment				SCWG 40% Costs		Categories	
1	Uninstalled capital	$572,940			Total major equipment	$2,158,000	$572,940	line 1
2	Other equipment	$521,042	91%	of line 1	Bulk items + control system	$2,993,000	$1,560,325	lines 2+6+7+8
3	Total equipment	$1,093,982			Design & fab labor	$1,295,000	$1,323,719	lines 4+13
4	Installation cost	$514,172	47%	of line 3	Facilities	$216,000	$1,137,741	lines 9+10+11+12
5	Total installed cost	$1,608,154		line 3 + line 4	Startup cost	$432,000	$459,472	line 14?
					TOTAL INSTALLED COSTS	$7,094,000	$5,054,197	line 16
	Other fixed capital							
6	Instrumentation	$196,917	18%	of line 3				
7	Piping	$722,028	66%	of line 3				
8	Electrical	$120,338	11%	of line 3				
9	Buildings	$196,917	18%	of line 3				
10	Yard improvements	$109,398	10%	of line 3				
11	Service facilities	$765,787	70%	of line 3				
12	Land	$65,639	6%	of line 3				
13	Engineering and construction	$809,547	74%	of line 3				
14	Contingencies	$459,472	42%	of line 3				
15	Total	$3,446,043						
16	**Total Fixed Capital Investment**	$5,054,197		line 5 + line 15				
	Overall Factor on Equipment	3.14		line 16/line 5				
	Burlington Gasifier							
	200	tpd						
	$375	per kWe						
	Scale down to 30 tpd							
	$1,171	per kWe						
	40%	electrical efficiency						
	$468	per kW fuel value						
	8500	Btu/lb biomass heating value						
	60,000	lb biomass/day						
	21,250,000	Btu/hr						
	6226	kW						
	$2,915,164	Installed capital cost of gasifier alone						
	$615,339	M. Mann installed cost of gasifier alone						

COSTS FOR 27 MT/DAY SCWG SYSTEM, 20% BIOSOLIDS, LOWER HYDROGEN YIELD

PLANT SIZE, T/DAY BIOSOLIDS	30					
PRETREATED BIOLSOLIDS, wt%	20					
CAPITAL COSTS						
Item	**Size**	**Units**	**Size Basis**	**Materials**	**Cost**	**Cost Basis**
Biosolids storage tank	8111	gal	4 hours holdup +25% head space	Concrete	$17,000	Engineering estimate
Biosolids transfer pump	135	gpm	4x progressive cavity pump	SS	$11,000	0.6 exponent scale from 100 MTD SCWO Plant
Emulsifier macerator	135	gpm	4x progressive cavity pump	SS	$23,000	0.6 exponent scale from 100 MTD SCWO Plant
Progressive cavity pump	34	gpm	28 gpm required	SS	$78,000	0.6 exponent scale from 100 MTD SCWO Plant
Filter press	27	MT/day	M&EB	CS	$0	Not required
Liquefier/pump	1146	gal	30 minutes residence time	Ti/steel	$397,000	Estimated from Zimpro vessel
Heat recovery heat exchanger	2.72	MW	M&EB	C276	$811,000	0.6 exponent scale from 100 MT SCWO Plant
Gas-fired heater	3.1	MW	M&EB		$310,000	0.6 exponent scale from 100 MTD SCWO Plant
Flue gas heat exchanger	2.2	MW	M&EB	SS	$155,000	Half of fired heater
Reactor	1187	gal	1.5 minutes residence time	Alloy 718	$617,000	0.6 exponent scale from MODAR vessel x 2/3
Waste heat boiler	1.3	MW	M&EB	C276	$286,000	0.6 exponent scale from 100 MTD SCWO Plant
Gas/Liquid separator	172	gal	5 min liquid RT, 50% head space	SS	$68,000	0.6 exponent scale from 33 gal vessel
Liquid letdown valve	21	gpm	M&EB	TiN/SS	$16,000	Same as 100 MTD SCWO Plant
System pressure control valve	27	gpm	M&EB	SS	$19,000	0.6 exponent scale from liquid letdown valve
Membrane separator	411,179	scfd	M&EB		$200,000	Phone quote
PSA module	411,179	scfd	M&EB		$200,000	Mann report
Hydrogen storage tank			M&EB		$100,000	Engineering judgment.
Total major equipment					$3,308,000	
Bulk items factor					$4,466,000	1.35 times major equipment cost
Design & fab labor factor					$1,985,000	0.6 times major equipment cost
Control system					$80,000	GA SCWO systems
Facilities					$331,000	0.1 times major equipment cost
Startup cost					$662,000	0.2 times major equipment cost
TOTAL INSTALLED COSTS					$10,832,000	
Notes:						
1. Overall factor on major equipment	3.3					
2. All equipment sizing is at least 20% excess capacity over requirement.						
3. Antal high H2 yields assumed.						
OPERATING COST, $/YR @ 330 ANNUAL OPERATING DAYS						
Item	**Assumption**				**Cost**	**Cost Basis**
Labor	4 operators				$150,000	$18.75/hr
Utilities, etc.	1% of capital cost				$108,000	0.5 SCWO utility costs
Hydrogen credit	$10/GJ; 0.1184 GJ (lower)/kg				$ (371,000)	Mann report values
Steam credit	M&EB values				$ (337,000)	Mann report values
Feed credit	$90/bone dry ton				$ (891,000)	0.75 of Encina disposal cost
TOTAL					($1,341,000)	
IRR @ $120/ton avoided disposal cost						
- Initial investment	($10,832,000)					
- Year 1	$1,341,000	-88%				
- Year 2	$1,341,000	-58%				
- Year 3	$1,341,000	-37%				
- Year 4	$1,341,000	-23%				
- Year 5	$1,341,000	-14%				
- Year 6	$1,341,000	-8%				
- Year 7	$1,341,000	-3%				
- Year 8	$1,341,000	0%				
- Year 9	$1,341,000	2%				
- Year 10	$1,341,000	4%				

GENERAL ATOMICS

COSTS FOR 27 MT/DAY SCWG SYSTEM, 40% BIOSOLIDS, LOWER HYDROGEN YIELDS

PLANT SIZE, T/DAY BIOSOLIDS	30					
PRETREATED BIOLSOLIDS, wt%	40					
CAPITAL COSTS						
Item	**Size**	**Units**	**Size Basis**	**Materials**	**Cost**	**Cost Basis**
Biosolids storage tank	4516	gal	4 hours holdup +25% head space	Concrete	$12,000	Engineering estimate
Biosolids transfer pump	75	gpm	4x progressive cavity pump	SS	$8,000	0.6 exponent scale from 100 MTD SCWO Plant
Emulsifier macerator	75	gpm	4x progressive cavity pump	SS	$17,000	0.6 exponent scale from 100 MTD SCWO Plant
Progressive cavity pump	19	gpm	15 gpm required	SS	$54,000	0.6 exponent scale from 100 MTD SCWO Plant
Filter press	27	MT/day	M&EB	CS	$125,000	Vendor quote/engineering judgment
Liquefier/pump	638	*gal*	*30 minutes residence time*	*Ti/steel*	*$279,000*	*Estimated from Zimpro vessel*
Heat recovery heat exchanger	0.52	MW	M&EB	C276	$105,000	0.6 exponent scale from 10 MTD SCWO Plant
Gas-fired heater	3.1	MW	M&EB		$310,000	0.6 exponent scale from 100 MTD SCWO Plant
Flue gas heat exchanger	2.2	MW	M&EB	SS	$155,000	Half of fired heater
Reactor	571	gal	1.5 minutes residence time	Alloy 718	$398,000	0.6 exponent scale from MODAR vessel x 2/3
Waste heat boiler	1.1	MW	M&EB	C276	$105,000	0.6 exponent scale from 10 MTD SCWO Plant
Gas/Liquid separator	65	gal	5 min liquid RT, 50% head space	SS	$38,000	0.6 exponent scale from 33 gal vessel
Liquid letdown valve	9	gpm	M&EB	TiN/SS	$16,000	Same as 100 MTD SCWO Plant
System pressure control valve	22	gpm	M&EB	SS	$28,000	0.6 exponent scale from liquid letdown valve
Membrane separator	411,179	scfd	M&EB		$200,000	Phone quote
PSA module	411,179	scfd	M&EB		$200,000	Mann report
Hydrogen storage tank			M&EB		$100,000	Engineering judgment.
Total major equipment					$2,150,000	
Bulk items factor					$2,903,000	1.35 times major equipment cost
Design & fab labor factor					$1,290,000	0.6 times major equipment cost
Control system					$80,000	GA SCWO systems
Facilities					$215,000	0.1 times major equipment cost
Startup cost					$430,000	0.2 times major equipment cost
TOTAL INSTALLED COSTS					$7,068,000	
Notes:						
1. Overall factor on major equipment	3.3					
2. All equipment sizing is at least 20% excess capacity over requirement.						
3. Antal high H2 yields assumed.						
OPERATING COST, $/YR @ 330 ANNUAL OPERATING DAYS						
Item	**Assumption**				**Cost**	**Cost Basis**
Labor	4 operators				$150,000	$18.75/hr
Utilities, etc.	1% of capital cost				$71,000	0.5 SCWO utility costs
Hydrogen credit	$10/GJ; 0.1184 GJ (lower)/kg				$ (371,000)	Mann report values
Steam credit	M&EB values				$ (337,000)	Mann report values
Feed credit	$90/bone dry ton				$ (891,000)	0.75 of Encina disposal cost
TOTAL					($1,378,000)	
IRR @ $120/ton avoided disposal cost						
- Initial investment	($7,068,000)					
- Year 1	$1,378,000	-81%				
- Year 2	$1,378,000	-45%				
- Year 3	$1,378,000	-23%				
- Year 4	$1,378,000	-9%				
- Year 5	$1,378,000	-1%				
- Year 6	$1,378,000	5%				
- Year 7	$1,378,000	8%				
- Year 8	$1,378,000	11%				
- Year 9	$1,378,000	13%				
- Year 10	$1,378,000	14%				