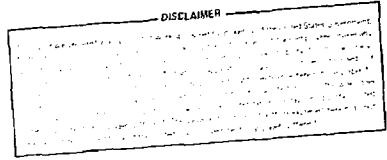


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THERMOCHEMICAL HYDROGEN PRODUCTION BASED ON
 MAGNETIC FUSION¹

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ABSTRACT

Conceptual design studies have been carried out on an integrated fusion/chemical plant system using a Tandem Mirror Reactor fusion energy source to drive the General Atomic Sulfur-Iodine Water-Splitting Cycle and produce hydrogen as a future feedstock for synthetic fuels. Blanket design studies for the Tandem Mirror Reactor show that several design alternatives are available for providing heat at sufficiently high temperatures to drive the General Atomic Cycle. The concept of a Joule-boosted decomposer is introduced in one of the systems investigated to provide heat electrically for the highest temperature step in the cycle (the SO₂ decomposition step), and thus lower blanket design requirements and costs. Flowsheeting and conceptual process designs have been developed for a complete fusion-driven hydrogen plant, and the information has been used to develop a plot plan for the plant and to estimate hydrogen production costs. Both public and private utility financing approaches have been used to obtain hydrogen production costs of \$12-14/GJ based on July 1980 dollars.

KEYWORDS

Magnetic fusion energy; tandem mirror reactor; blanket design; thermochemical hydrogen cycle; water-splitting; sulfur-iodine cycle; joule-boosted decomposer; hydrogen production costs.

INTRODUCTION

In view of the significant advances that have been made in recent years in magnetic fusion reactor designs, we have a reasonable expectation that magnetic fusion energy will emerge as a commercial reality in about 30-40 years. To date, the main motivation for developing fusion has been for the generation of electric power. However, as we look at our Nation's energy needs we find that some three-

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fourths of our needs are in portable and transportable fuels, as contrasted with one-fourth in electricity. We expect this proportionate need to continue, and as fossil fuel resources become depleted we need to look toward fusion to fulfill our needs for both synthetic fuels and electricity.

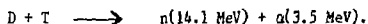
In addition to the fossil fuel depletion problem, we are faced with a potential problem of detrimental climatic changes that can result from increased carbon dioxide in the atmosphere as fossil fuels continue to be burned. This problem is particularly accelerated from the burning and gasification of low-hydrogen fuels such as coal. This adds to our motivation for developing processes for producing hydrogen and hydrogen-derived synthetic fuels from fusion.

Although 30-40 years for the emergence of fusion energy may seem like a long time, nonetheless the time for development of a technology for producing fuels from fusion is very limited if we are to meet this timeframe. In an initial effort to evaluate the technology and economics of producing hydrogen-derived synfuels based on magnetic fusion, the U. S. Department of Energy - Office of Fusion Energy has established scoping studies in the areas of (1) thermochemical hydrogen cycles, and (2) high temperature steam electrolysis. In the thermochemical hydrogen cycle area, a joint study has been underway for three years at the Lawrence Livermore National Laboratory (LLNL) and the University of Washington (UW); and for the past year and a half the General Atomic Company (GA) has been an active participant in the study. The thrust of this study is to develop a conceptual engineering design for using a Tandem Mirror Reactor (TMR) as an energy source to drive the GA Sulfur-Iodine Cycle and thus to produce hydrogen. In parallel studies, the Brookhaven National Laboratory is investigating the prospects for hydrogen production by high temperature steam electrolysis driven by a Tokamak Reactor.

In this report we summarize (1) the progress to date on developing conceptual TMR blanket designs as a heat source, (2) adapting the GA Cycle to the TMR heat source, and (3) giving a preliminary assessment of hydrogen production costs. This summary draws on the efforts of a number of individual contributors (Werner and others, 1981; Werner and others, 1982). In particular, we acknowledge the contributions of Richard Werner, Terry Galloway, and Gary Johnson at LLNL; Fred Ribe and Gene Woodruff at UW; Gottfried Besenbruch and John Norman at GA; and of the consultants Myron Hoffman at the University of California at Davis, and Don Rowe of Rowe and Associates in Bellevue, Washington.

THE FUSION ENERGY SOURCE

Thermochemical cycles for hydrogen production can be readily adapted to the use of the LLNL Tandem Mirror Reactor design with its linear topology and with its output of both thermal energy and high voltage DC electrical energy. The TMR design is based on the D-T reaction which produces an energetic neutron and an alpha particle, i.e.,



A schematic illustration of the TMR is shown in Fig. 1. Plasma confinement within the central cell, which is 100 m long, is provided by high-field magnets that function as end plugs. The central cell is where the main D-T reaction occurs, and a blanket is provided for two functions: (1) to breed T as fuel for the reactor by utilizing neutron reactions with lithium and (2) to provide process heat for the thermochemical cycle by utilizing neutron moderation and nuclear reactions in the blanket. Direct converters located outside the end-plug magnets are used to trap alpha particles and to very efficiently convert the ionic charge of the alpha particles to high voltage DC electricity and to use the alpha particle impact energy as additional process heat. About 20% of the total reactor output

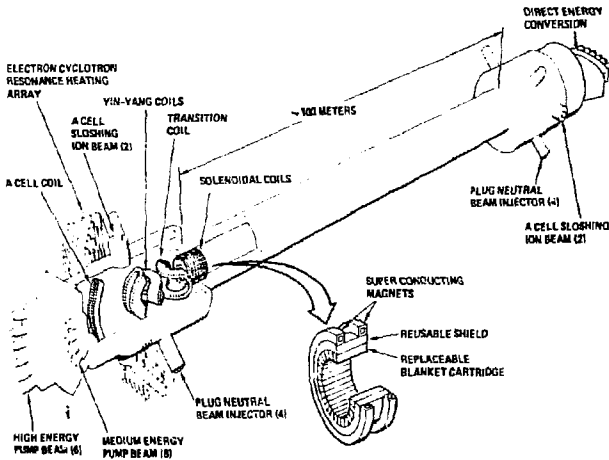


Fig. 1. Schematic view of the tandem mirror reactor.

energy comes from the direct converters, and about 60% of this is in the form of DC electricity. In more advanced TMR designs the proportionate amount of direct converter power to blanket power may be even higher.

Adapting thermochemical hydrogen cycles to the TMR is favorable for several reasons:

- The relatively open structure of the TMR central cell lends itself well to relatively simple modular blanket designs and to good accessibility for process heat utilization.
- Process heat temperatures and amounts that are required for thermochemical cycles are within the performance available from TMR blanket designs.
- TMR designs provide high conversion efficiency electricity from direct converters. A portion of this electricity is available for process use in thermochemical cycles.

In designing the fusion energy source our emphasis thus far has been on blanket designs and on the blanket/process interface. The blanket design studies will be described next.

BLANKET DESIGN

Conceptual designs of fusion blankets as heat sources for thermochemical cycles have been under study for three years at both LLNL and UW. Concepts studied at LLNL are:

- Li-Na Cauldron Blanket
- Heat Pipe Blanket
- Fixed Bed Li_2O Blanket

Concepts studied at UW are:

- Flowing Li_2O Microsphere Blanket
- Two-Temperature-Zone Blanket

Brief descriptions of each of these blanket concepts are as follows:

Li-Na Cauldron Blanket

In this design (Werner and others, 1981) a Li-Na (50-50 a/o) liquid pool surrounds the plasma and is maintained at ~ 1200 K by allowing the Na to boil at a pressure of about 1 atm. The heat for boiling is provided by moderation of neutrons in the Li-Na mixture with the main heat generation occurring near the plasma first wall. The Na vapor is condensed onto heat exchanger tubes located in a dome region above the pool, and the latent heat of condensation is transferred to a high pressure (~ 35 atm) helium coolant stream for use in the thermochemical hydrogen plant. Na is recycled back to the pool for reheating. Tritium generated by neutrons reacting with the Li is removed through a Nb diffusion membrane located in the dome region above the pool. Modular clamshell type of construction is used for the blanket design, with modules about 2 m wide in order to fit within the solenoid coil structure of the TMR central cell. Although the simplicity of operation makes this blanket design very attractive, it does pose two main concerns, which are: (1) a large vapor volume fraction ($\sim 60\%$) is established near the first wall because of the relatively high heat transport that occurs in that region, and (2) there are safety concerns associated with an accident involving large amounts of hot alkali metals.

Heat Pipe Blanket

This design (Werner and others, 1982) which is also modular, evolved from the cauldron concept and represents solutions to the main concerns over the cauldron. Here liquid $\text{Li}_17\text{Pb}_{83}$ is used as the pool fluid to avoid the safety concerns associated with the reactivity problems of liquid Li-Na. Heat transport out of the pool is accomplished using heat pipes (see Fig. 2) in place of the boiling approach used in the cauldron. We find, however, that a relatively large number of heat pipes are required, and an individual heat pipe is not a simple tube with a wick fitted against the inner wall. Rather, the heat pipe structure is quite complicated having a flattened tube structure with a sandwich wall where a layer of SiC is used as an insulator insert in the wall structure to prevent eddy current effects on the heat pipe fluid from the TMR magnetic field surrounding the blanket. Heat removed from the pool by heat pipes is transferred to high pressure helium for use in the thermochemical cycle. The heat pipes also serve the function of removing the tritium by utilizing Nb diffusion membranes in the liquid $\text{Li}_17\text{Pb}_{83}$ region to introduce tritium into the heat pipe and a Nb diffusion membrane at the far end of the heat pipe for exit of the tritium. We use the heat pipe blanket as the heat source for one of the two blanket options selected in this paper to illustrate the use of magnetic fusion for thermochemical hydrogen production.

Fixed Bed Li_2O Blanket

The conceptual design of a fixed bed Li_2O blanket is currently underway at LLNL (Werner, 1982). As presently envisioned, the design consists of an array of metallic tubes filled with Li_2O cartridges and cooled on the shell side with helium at about 40-50 atm pressure. The blanket structure is again modular.

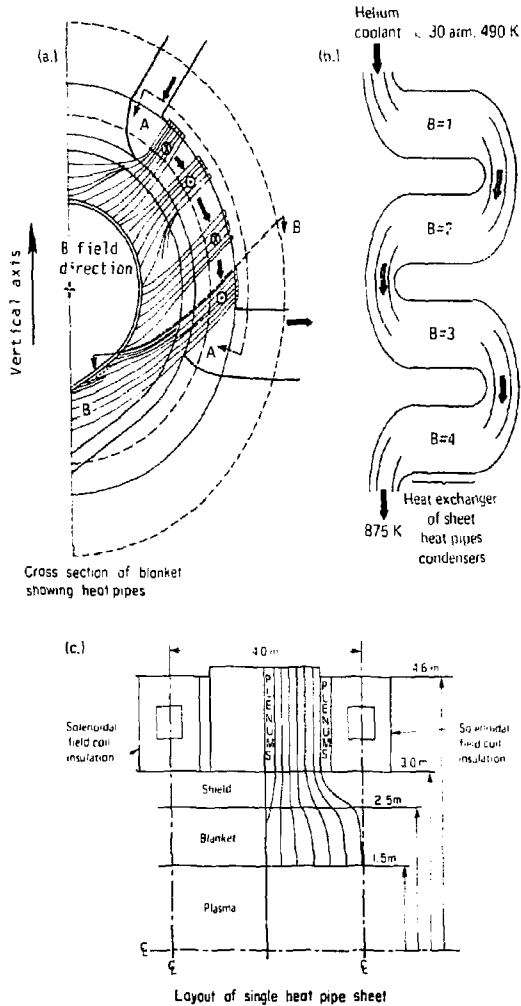


Fig. 2. The heat pipe blanket module: (a) end view of a TMR heat pipe blanket module showing the four bundles of heat pipes; (b) side view of the reactor showing the arrangement of the four-pass, heat-pipe-condenser heat exchanger (Section A-A); (c) Section B-B of the heat pipe blanket showing a typical sheet of flat heat pipes tapered to fit around the solenoidal magnet coil.

Tritium is removed from the Li_2O cartridges by using a separate stream of slowly flowing helium as a purge gas within the Li_2O containment tube region. Heat from the main helium stream is transferred to steam through an intermediate heat exchanger and the steam carries the process heat to the thermochemical plant. We believe that this fixed bed helium-cooled design (although still in its early stages) should provide the best blanket alternative in terms of simplicity of design, reliability, safety, and cost.

Flowing Li_2O Microsphere Blanket

The flowing Li_2O microsphere blanket was initially designed (Werner and others, 1981) at UW to provide high temperature (~ 2000 K) heat to the thermochemical process. Subsequently, it was used as one alternative for the low temperature zone (~ 900 K) of the two-temperature-zone blanket, which is described below (Werner and others, 1982). The original flowing Li_2O microsphere design uses 0.4 mm diameter Li_2O microspheres for the flowing bed. The microspheres flow by gravity through 10 channels on each side of the plasma region. The channels are placed at increasing radial distances from the plasma center and are separated by metal baffles. Flow rates through the channels (~ 20 cm/s) are carefully controlled at exit orifices in each channel to give a balanced exit temperature, i.e., higher flow rates are used closer in to the plasma because of higher heat deposition. Heat from exiting microspheres is transferred to liquid sodium which carries it to the thermochemical plant. Heat transfer is with a shell and tube heat exchanger with sodium on the tube side. Pressurized helium is an alternative to the sodium for heat transport to the plant. After heat transfer, the Li_2O microspheres are returned to the blanket using an Archimedes screw. Tritium is removed from the Li_2O microsphere region using a helium purge gas. The flowing Li_2O microsphere blanket presents a significant advantage in providing a relatively safe and simple design by not requiring containment of a high pressure gas as the coolant for the blanket. The high use temperature of 1200 K, however, presents significant problems of sintering of the Li_2O microspheres and of interaction of the Li_2O with the metal baffles.

Two-Temperature-Zone Blanket

In the two-temperature-zone blanket, (Werner and others, 1982) a lower temperature zone (~ 900 K) close in to the plasma provides about 70-80% of the process heat for the thermochemical cycle, and all of the tritium breeding. The high temperature zone is located further out radially from the plasma. It provides a higher temperature (~ 2000 K) and about 20-30% of the process heat for the thermochemical cycle. The design of the low temperature zone is similar to the flowing Li_2O microsphere design given above, except that the temperatures are significantly lower. Thus, the sintering problem is avoided and the baffle reaction problem is reduced. In this latest design, the heat is transferred to high pressure helium for use in the process and tritium is removed with a helium purge gas as before. Two alternative designs based on liquid $\text{Li}_{17}\text{Pb}_{83}$ were also investigated in place of the Li_2O for the low temperature zone. In one, liquid $\text{Li}_{17}\text{Pb}_{83}$ is pumped through flow tubes surrounding the plasma. In the other, high pressure helium is pumped through tubes in a stagnant $\text{Li}_{17}\text{Pb}_{83}$ pool. In these latter cases, heat is transferred through an intermediate heat exchanger to high pressure helium for the thermochemical process. Heating is provided in the high temperature zone by moderating and absorbing neutrons in Ta rod bundles that are placed inside of tubes that carry the high pressure helium coolant. Two versions of the two-temperature-zone blanket are illustrated in Fig. 3. The two-temperature-zone blanket is the second of the two blanket design options that we have selected in this paper to illustrate the use of magnetic fusion for thermochemical hydrogen production.

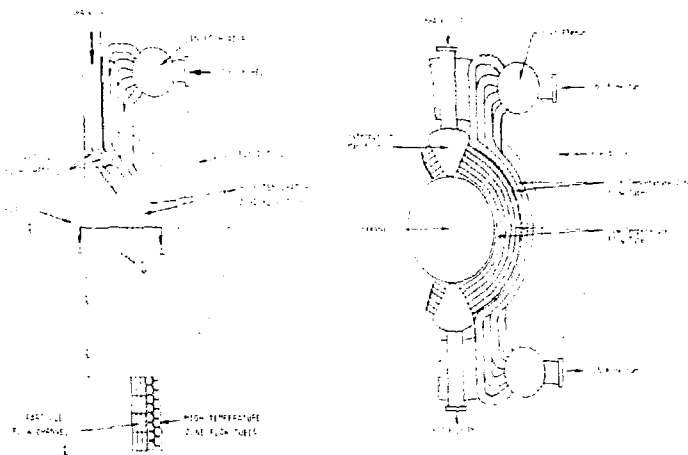


Fig. 3. Two versions of the two-temperature-zone blanket are illustrated. The low-temperature zone in the diagram on the left utilizes flowing $\text{Li}_{17}\text{Pb}_{83}$ microspheres for heat removal, while in the diagram on the right, flowing liquid $\text{Li}_{17}\text{Pb}_{83}$ is used. Helium is passed over a Ta moderator in the high-temperature zone in both cases.

TMR/ SO_3 DECOMPOSER INTERFACE

For the two blanket options selected for illustration (i.e., heat pipe blanket and two-temperature-zone blanket), the main difference in coupling the TMR to the GA Cycle lies in the manner in which the SO_3 decomposer is designed. For the heat pipe blanket we use a Joule-boosted decomposer and for the two-temperature-zone blanket, a fluidized bed decomposer. These approaches are described below.

Joule-Boosted Decomposer Concept

One of the most significant recent advances at LLNL for interfacing the TMR to the high temperature step of H_2SO_4 -based thermochemical cycles has been the introduction of the Joule-boosted decomposer concept and its use with the heat pipe blanket (Werner and others, 1982). By Joule-boosting, we refer especially to the utilization of electrically heated commercial SiC furnace elements in place of using a heat exchanger to transfer process heat to the highest temperature step of a thermochemical hydrogen cycle, i.e., in our case the SO_3 decomposition step in the GA Sulfur-Iodine Cycle.

The use of Joule-boosting can significantly lower temperature and materials requirements for the blanket, and in addition improve the operation and reliability of the high temperature process units in the chemical plant. Tandem Mirror Reactors have a distinct advantage in coupling with a Joule-boosted decomposer design in that surplus electricity in the direct converter can be used

to provide a significant fraction of the electrical requirements for the decomposer. In future D-D cycle fusion reactors, sufficient net electricity (~25% of the total output power) could be available to provide all of the electrical requirements for the Joule-booster decomposer. We therefore view the Joule-booster decomposer concept as an important step forward toward achieving the Fusion/Synfuels tie.

The most important advantage of the Joule-booster approach from the TMR blanket standpoint is that the temperature requirements for the blanket are reduced by about 200 K. This greatly reduces the severity of a number of materials problems for the blanket, especially in regard to materials corrosion and materials creep strengths. Corrosion and creep strength are design limiting factors for many blanket designs, and they are materials characteristics that degrade exponentially with increasing temperature. Hence, the reduction in blanket temperatures by 200 K is a major change, and puts us into a new regime where more reliable and lower cost materials can be utilized in blanket designs.

The Joule-booster decomposer (see Fig. 4) is used in the chemical plant to decompose SO_3 into SO_2 and O_2 . The decomposer design is relatively simple. Electrical power is supplied to commercially available SiC heating elements arranged in an array with the decomposing gases passing through the array.

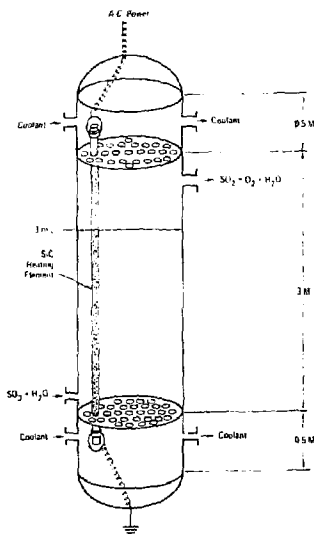


Fig. 4. Schematic illustration of the Joule-booster SO_3 decomposer design.

Electrodes are cooled in isolated end chambers, and the vessel consists of a silica brick-lined Teflon-coated steel. The heating elements are operated at a temperature of ~ 1250 K, which is a very conservative temperature for SiC, and the gas which consists of SO_3 , SO_2 , O_2 , and H_2O , is at a total pressure of 7 atm.

In considering further the advantages of the Joule-boosted decomposer in the chemical plant, we find the following:

- A SiC furnace decomposer unit is simpler, more reliable, and much lower-cost than a heat exchanger unit for providing heat for SO_3 decomposition. SiC furnaces have proven to be reliable low-cost designs through many years of use, and recent work at Oak Ridge National Laboratory has shown excellent compatibility of SiC with the gases present in the SO_3 decomposer (Tiegs, 1981).
- The Joule-boosted decomposer gives higher decomposition yields for SO_3 than with a heat exchanger design (40% vs 60%). This means less recycle of SO_3 , smaller and less costly equipment for a number of process units, and reduced pumping power for circulating of chemicals.
- There is no need for a catalyst in the Joule-boosted decomposer, since it operates at a sufficiently high temperature (~ 1250 K) that SO_3 decomposition kinetics are rapid. This not only reduces costs but improves reliability.
- The Joule-boosted decomposer provides the best possible safety isolation of the chemical plant from the blanket, and vice versa, i.e., the chemical plant is isolated from tritium in the blanket, and the blanket is isolated from O_2 and corrosive sulfur-containing gases in the decomposer.

A disadvantage of the Joule-boosted decomposer concept is that the efficiency for hydrogen production is reduced compared to a heat exchanger approach because of the need to provide additional electrical power. However, since only about 20% of the total energy for the CA Cycle is required for SO_3 decomposition, the use of additional power does not seriously reduce the overall cycle efficiency. Furthermore, the reduced requirement for pumping power for recycle of chemicals as compared with a heat exchanger approach, partially offsets the electrical needs. In the final analysis, we believe that the gain in reliability and safety of the Joule-boosted case more than offsets the decrease in efficiency.

Fluidized Bed Decomposer

By coupling a fluidized bed decomposer to the high temperature zone of the two-temperature-zone blanket, we find that we can transfer heat to the decomposing SO_3 gas with a relatively small degradation in temperature, as shown in Fig. 5 (Werner and others 1981; Werner and others 1982). Thus, using a helium source temperature of 1100 K in the tube side of a shell and tube heat exchanger we expect to attain a temperature of ~ 1100 K in the gaseous decomposition products exiting from the bed. However, at this temperature decomposition rates are sufficiently slow that a catalyst is required. We calculate a 55% decomposition yield of SO_3 at 1100 K by using either a CuO catalyst, or (more expensively) a Pt catalyst on a titania support. The catalyst is provided in the form of 0.5 mm diameter spheres which also serve as the fluidized particle medium in the bed.

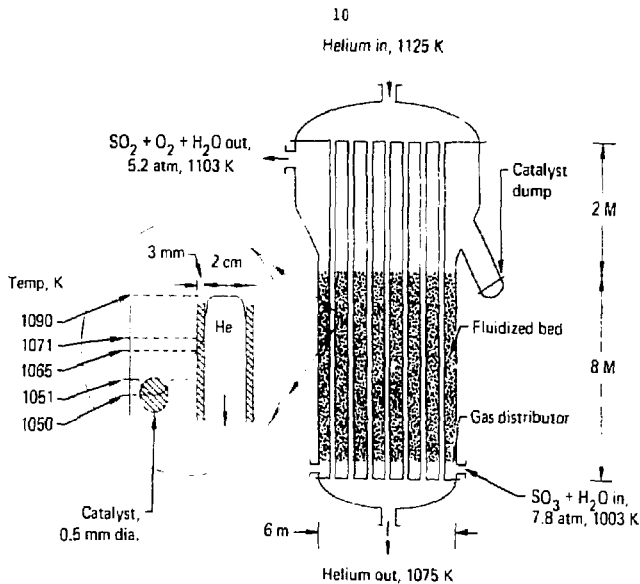
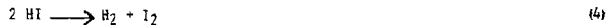
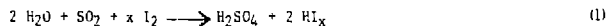


Fig. 5. Schematic illustration of the fluidized bed SO_3 decomposer.

GA WATER-SPLITTING PROCESS

The GA water-splitting cycle (Besenbruch and others, 1980; Norman and others, 1982) may be described by the four equations:



These equations represent three chemical reactions and one separations process. In these equations the species HI_x represents an aqueous complex of HI and I_2 formed by reaction 1. The H_2SO_4 product of reaction 1 is obtained as a moderately concentrated aqueous solution which is immiscible with HI_x .

The overall process is divided into five sections for design purposes. Sections I through IV roughly correspond to the four equations and section V represents the heat and energy transmission equipment required to match the fusion heat source to the chemical process.

As indicated in the simplified schematic flow diagram (Fig. 6), there are a small number of streams which pass between sections thus permitting the individual sections to be designed, to a large part, separately. For this effort (Werner and

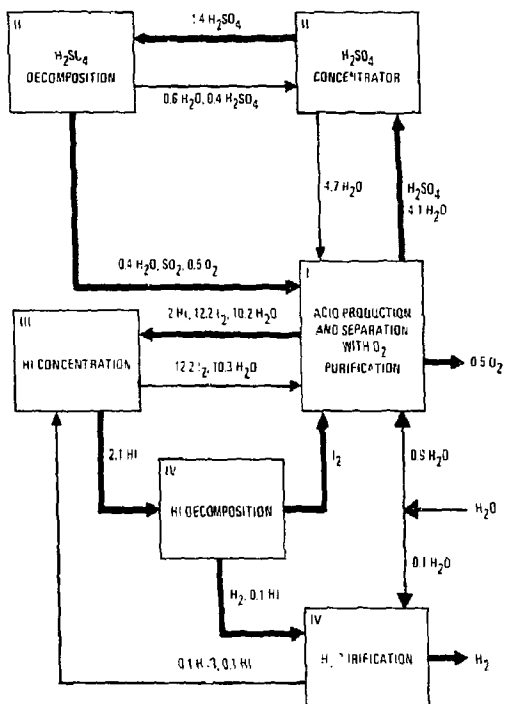


Fig. 6. Simplified schematic flow diagram of the 1981 version of the sulfur-iodine cycle.

others, 1982), the designs for sections I, III, and IV were provided by GA and the designs for sections II and V for both the Joule-booster and Fluidized bed decomposer concepts were provided by LLNL.

Design Philosophy

A stipulation, from the beginning, was that the process design be such that one could rationally expect to design, build, and operate the water-splitting plant using technology available today. The final design met this criterion. Secondly, the process design was to aim for minimum hydrogen product cost rather than simply maximum thermal efficiency. Where the net effect of process decisions on hydrogen cost was uncertain, the decision was made in favor of high efficiency, but high thermal efficiency was never knowingly selected over low hydrogen cost.

Section I - Acid Production

The one cost consideration which dominates the design of section I is the high cost of heat transfer equipment suitable for use with the corrosive H_2SO_4 solution. The least expensive materials alternative, niobium, costs in excess of \$200/kg for

fabricated tubing. Hence, we used the following design techniques to minimize usage of niobium:

- Cool the recycle streams entering section I thus reducing the amount of exothermic heat of reaction removed from section I to maintain the desired outlet stream temperature.
- Operate reaction 1 isothermally using excess iodine instead of low temperature to shift the equilibrium towards the products.
- Use direct contact between immiscible liquids or between liquids and gases.
- Apply enhanced heat transfer techniques such as two-phase flow and spiral-fluted tubing to increase heat transfer rates.

Using the above techniques, a flow diagram was developed as illustrated in Fig. 7. The majority of chemical reaction 1 takes place in the heat exchanger reactor (R101). Here the mixed SO_2/O_2 stream formed as a result of reaction 2, is reacted during cocurrent flow with internal recycle streams containing I_2 , H_2O , H_2SO_4 , and HI . The exothermic heat of reaction is transferred to the waste heat recovery system while the reaction takes place in spiral-fluted tubes.

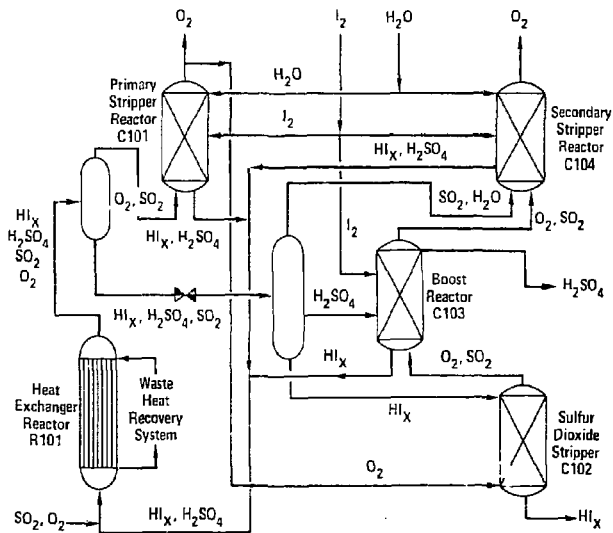


Fig. 7. Schematic of section I, reaction of SO_2 with iodine and water to give H_2SO_4 and HI_x as immiscible liquid products.

Part of the unreacted SO_2 is dissolved in the liquid product of R101 and the remainder must be scrubbed from the product oxygen in the primary stripper reactor (C101). The primary function of the stripper reactor is to remove the SO_2 from the O_2 by titrating with I_2 according to reaction 1. The second function is to act as a direct contact heat exchanger, cooling the O_2 product and pre-heating the water entering section I.

Other equipment in section I include the stripper (C102) where SO_2 is removed from the HI_x phase, the boost reactor (C103) where the concentration of the H_2SO_4 phase is increased from 57% to 57% through contact with fresh I_2 and the secondary stripper reactor (C104) where the recycle strip oxygen is purified prior to discharge. All of these vessels are fluorocarbon-lined mild steel.

Section II - H_2SO_4 Processing

The Joule-boosted and fluidized-bed decomposers have already been described, but before the decomposition can take place, the sulfuric acid must be concentrated from 57% to 98% and the 98% azeotrope vaporized. Fig. 8 shows a simplified version of the flowsheet for the Joule-boosted version of section II. The fluidized bed version is similar except the temperatures are lower.

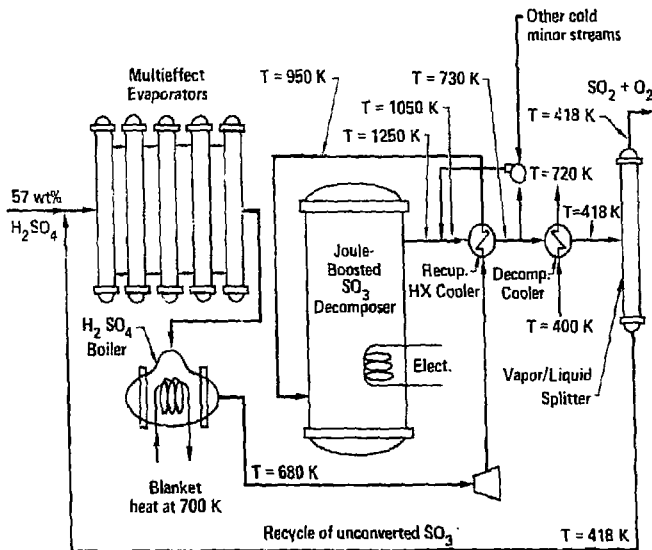


Fig. 8. Schematic of section II, the H_2SO_4 processing step.

The sulfuric acid concentration takes place in a five stage multi-effect evaporator. Heat input to the evaporator is via silicon-carbide U-tube heat exchangers. Flash drums are constructed from fluorocarbon-lined mild steel. The fluorocarbon provides corrosion protection and is itself protected from temperature effects by a thermally insulating internal liner made up of silica bricks. Similar materials of construction are employed in the H_2SO_4 vaporizer.

The exit gases from the decomposer are first quenched with recycle acid in the Joule-boosted case to drop the temperature to 1050 K and thus prevent recombination of SO_2 and O_2 . The gases are then heat exchanged with the vaporized azeotrope in an Incoloy-800H recuperator. This metallic heat exchanger is satisfactory in the gaseous environment and temperatures involved. At lower temperatures, where liquid H_2SO_4 is present, and at much higher temperatures, metallic heat exchangers would be unsuitable.

Final cooling of the decomposer product gases takes place in silicon-carbide heat exchangers in the multi-effect evaporators.

Section III - HI Concentration

Both water and iodine must be separated from the HI prior to its decomposition. H_3PO_4 is the agent which aids both separations. By using countercurrent contact of HI_x with concentrated H_3PO_4 (C302, Fig. 9), water and HI are extracted into

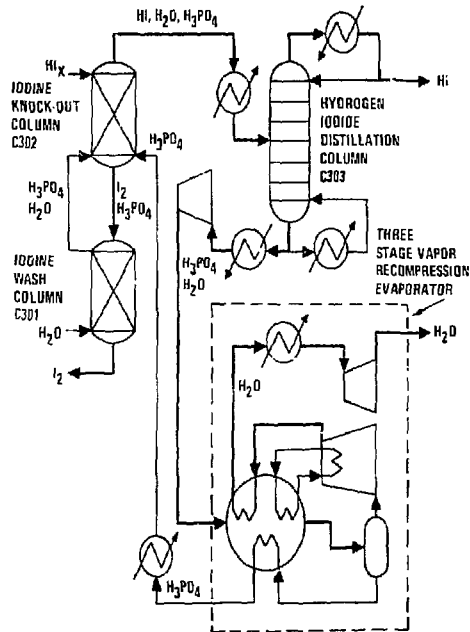


Fig. 9. Schematic of section III, separation of aqueous HI_x into HI , I_2 , and H_2O .

the H_3PO_4 -containing phase and iodine is separated out. Iodine is then water washed and recycled to section I. H_3PO_4 lowers the water activity, thus allowing the $\text{HI}-\text{H}_2\text{O}$ azeotrope to be broken, and HI to be distilled away from the $\text{H}_3\text{PO}_4-\text{H}_2\text{O}$. This distillation is carried out at 9 atm so that the HI product can be taken off as a liquid.

Phosphoric acid must be concentrated for recycle and this is the most costly single step in the CA Cycle, as it is presently constituted. The water is removed in three stages of vapor-recompression-driven flash-evaporation. Both the large Hastelloy-C heat exchangers and the compressors that are required for this process contribute significantly to the overall cost.

Section IV - HI Decomposition

The HI decomposition (reaction IV) takes place at 80 atm in the liquid phase. Unreacted HI is separated from the iodine product via distillation in a Hastelloy-B distillation column (C401, Fig. 10). HI and a trace of H_2S are removed from the hydrogen product before it is delivered at pipeline pressures of 50 atm.

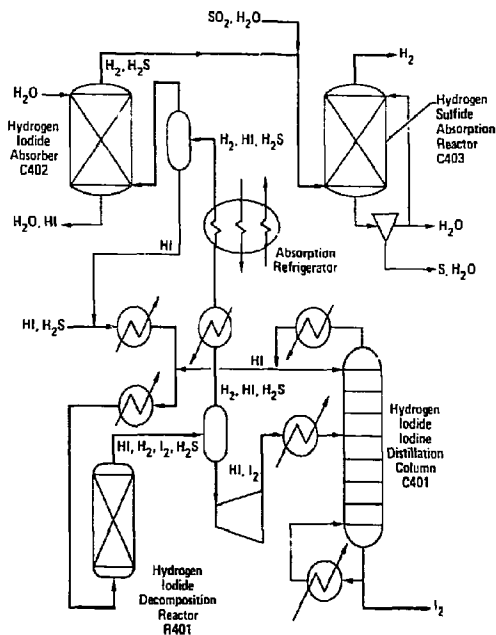


Fig. 10. Schematic of section IV, the HI decomposition step.

ESTIMATED HYDROGEN COST

We have estimated the cost of hydrogen for a 4900 mole/s thermochemical hydrogen plant driven by a 2930 MW TMR (Werner and others, 1982), as described below.

Costing Procedures

The capital cost for the fusion plant was obtained by updating a previous cost study (Henning, 1981) for an electricity producing TMR. The original study gave a range of estimated capital costs to account for uncertainties in the level of technology required for the confinement physics designs. We have chosen the high cost figure for the presentation here.

The cost of the thermochemical plant was based upon preliminary equipment sizing calculations. Standard chemical engineering costing techniques (Peters and Timmerhaus, 1980; Guthrie, 1969) are available for predicting total capital costs of chemical plants based upon FOB costs of the major pieces of equipment. Major piping runs are treated as additional equipment items. The only piping of significance was found to be the helium piping and its cost was included in the Section V costs. An overall facility plot plan was also prepared based on the equipment sizing calculations (Fig. 11).

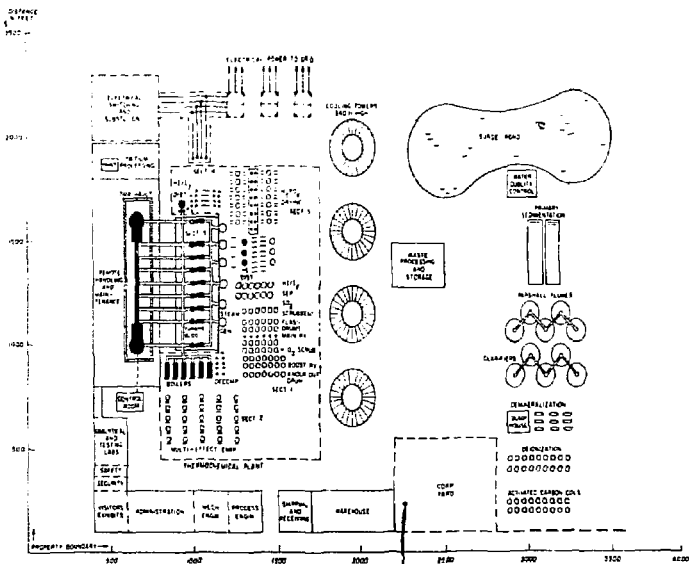


Fig. 11. TMR-synfuels thermochemical hydrogen production plant plot plan.

Correlations are available (Peters and Timmerhaus, 1980; Guthrie, 1969) for predicting FOB equipment costs for most of the thermochemical plant. Where the standard works could not provide capital costs, as in most of Section II and the power recovery systems, vendor estimates or other special sources of FOB cost data were used. When only pre-1980 costs were available, the industry standard, Marshall and Swift (M&S), equipment cost index was used to estimate 1980 costs.

It is recognized that some of the best costing techniques are maintained as proprietary by architect and engineer (A&E), chemical and oil companies. A proprietary costing method, available at GA was used to spot check costs for a representative number of items. No overall bias could be observed, although variation between different types of equipment were noted.

Capital costs for the TMR and the chemical plant are summarized in Table 1. We believe that these cost estimates are within a goal that we set beforehand of $\pm 30\%$ accuracy.

Economic Basis

A standard costing manual, prepared by Battelle (Schulte and others, 1978) for use by the fusion community, was used as a basis for predicting financing costs.

The time frame for the cost estimate is after fusion-based thermochemical hydrogen production has been commercially established, circa 2030-2050. The constant dollar financing method (Schulte and others, 1978), with a base of July 1980, was used to facilitate comparison with today's costs. For this financing method the interest rate is assumed to be 5% above the underlying rate of inflation with all results then expressed in terms of the base year. The net result is an estimate of the hydrogen production costs which would have been in effect if the plant had begun commercial operation in the base year.

Total capital costs for each section of the thermochemical plant and for the TMR are presented in Table 1 for the two different blanket/decomposer combinations previously discussed. Total plant investment, which includes financing during construction, is given for two different financing methods. The first method assumes that the facility is completely debt financed by a publicly-owned utility. The second method assumes that an investor-owned utility finances half the project through equity and half through debt. Construction time is assumed to be three years for the chemical plant and eight years for the TMR. Investment is assumed to be linear during the construction period.

Hydrogen Production Cost

The cost of hydrogen is given in Table 2 for both blanket/decomposer combinations and both types of utility financing methods. Operating life of the TMR was taken as 30 years as recommended by the Battelle report (Schulte and others, 1978), whereas the design life of the chemical plant was taken to be 20 years. Operating costs for the TMR were also taken from the Battelle report. Operating costs for the thermochemical plant were based on historical data from the chemical industry (Guthrie, 1970), using the cost of an equivalent chemical plant constructed from carbon steel as a basis.

The 10% return on investment is applicable to the equity portion of the private utility finance cases. Note that this 10% is in addition to the underlying inflation rate.

The final hydrogen cost takes into account a calculated 74.4% availability of the combined TMR-thermochemical plant. A 77% availability for the TMR comes from the Battelle report (Schulte and others, 1978). A 90% stream availability for the chemical plant takes into account 5% unscheduled and 5% scheduled downtimes.

These quantities are reasonable considering the amount of redundancy that we have designed into the chemical plant, and the number of parallel trains (usually five or six) used to carry out each of the major processes in the plant.

We have used the energy equivalence of the hydrogen product, taken as the higher heating value of hydrogen (285.77kJ/mole), plus an added energy of RT lnP (9.7kJ/mole) associated with the 50 atm pressure of the hydrogen product to arrive at values for hydrogen production costs of about \$12-14/GJ.

TABLE 1 Summary of Capital Costs - M\$ July 1980

	Heat Pipe Blanket with Joule-Boosted Decomposer	Two Zone Blanket with Fluidized Bed Decomposer
<u>Total Capital Cost</u>		
Section I	92.5	92.5
Section II	97.2	136.6
Section III	593.0	593.0
Section IV	58.3	58.3
Section V	368.9	243.7
Tandem Mirror Reactor	1336.0	1336.0
	<u>2545.9</u>	<u>2460.0</u>
<u>Total Plant Investment</u>		
Public Utility (100% debt)	2841.1	2750.4
Private Utility (50% debt, 50% equity)	2693.6	2605.3

TABLE 2 Hydrogen Production Costs - July 1980

	Heat Pipe Blanket with Joule-Boosted Decomposer		Two Zone Blanket with Fluidized Bed Decomposer	
	Public Utility	Private Utility	Public Utility	Private Utility
Total Investment - M\$	2841.1	2693.6	2570.4	2605.3
Annual Finance Charges - M\$	201.7	100.8	194.5	97.3
10% Return of Equity	0	127.3	0	123.0
Annual Operating Cost	204.7	204.7	197.5	197.5
Total Annual Cost	406.4	490.6	392.0	472.3
Hydrogen Production Cost \$/GJ	11.95	14.43	11.53	13.89

SUMMARY AND CONCLUSIONS

We have found that it is feasible both from an engineering and economic standpoint to couple a TMR fusion driver to the GA Sulfur-Iodine Cycle to produce hydrogen. By carrying out our current studies on a completely integrated system, we have been able to delineate some of the key areas where future work is required to improve system reliability and to lower costs. In particular, further work is needed on TMR blanket designs and on the HI_x purification step in the GA Cycle. We believe that our introduction of the Joule-boosted decomposer for SO_2 decomposition is an important contribution in that it will lead to more reliable and lower cost blanket designs in the future because of the lowered requirement on blanket temperatures. The cost of hydrogen production (\$12-14/GJ) that we obtain here is expected to be cost competitive with fossil fuels at the time we expect fusion to become a commercial reality (in 30-40y).

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