DESIGN OF AN OCEAN THERMAL ENERGY PLANT SHIP TO PRODUCE AMMONIA VIA HYDROGEN

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Abstract—The 43°F (24°C) temperature difference that exists between surface water and deep water at selected sites in tropical oceans can be used to drive a heat engine to produce electric power, electrolyze water, and produce ammonia from the resulting hydrogen plus nitrogen from the air. A baseline design has been developed for a 100-MWe Ocean Thermal Energy Conversion (OTEC) plant-ship that would produce 313 tons per day of ammonia. The cost estimates for this design have been extrapolated to 500-MWe plant-ships to produce ammonia (for fertilizers and chemicals) or liquid hydrogen for shipment to the U.S. It is judged that ammonia will be producible at competitive cost ($96/short ton in 1975 dollars) by the sixth and subsequent plant-ships in the mid-1980s. This production by OTEC/ammonia plants would conserve supplies of natural gas or other fossil fuels now used to produce ammonia on shore. For the longer term (1990s), liquid hydrogen from OTEC plants should become competitive as demands for this clean fuel and efficient ways for employing it in larger markets (fuel cells, transportation, etc.) come to maturity.

INTRODUCTION

THE Applied Physics Laboratory (APL) has conducted an analysis [1] of the maritime aspects of Ocean Thermal Energy Conversion (OTEC) plant-ships for deployment in tropical oceans to produce ammonia, aluminum, liquid hydrogen, or other energy-intensive products. This analysis was supported by the U.S. Maritime Administration (MARAD), Department of Commerce. Working with APL were the Sun Shipbuilding & Drydock Company (SS&DD), who provided substantial company-funded design and consulting efforts; Hydronautics, Inc., who conducted analysis on platform motions and other maritime aspects; the Woods Hole Oceanographic Institution for site selection and design criteria; Avondale Shipyards and Kaiser Aluminum and Chemical Corporation who contributed consulting efforts; the law firm of LeBoeuf, Lamb, Leiby, & MacRae, who addressed the international legal considerations; and many industrial organizations and consultants who provided information on the OTEC and product equipment.

The baseline design developed for a 100-MWe OTEC/ammonia plant-ship features the integration of a novel concept for low-cost OTEC heat exchangers with a simple, relatively low draft, rectangular concrete hull which can be built in and launched from existing U.S. shipyards. The concept for the heat exchangers, which employ large-diameter, multipass aluminum tubes with the ammonia working fluid inside the tube, was developed under APL in-house support and support from the U.S. Energy Research and Development Administration (ERDA) [2].

The present U.S. interest in ocean thermal energy was stimulated by the Andersons, who developed a closed Rankine cycle, OTEC plant concept in the early 1960s [3, 4]. There is still interest in the open-cycle process (vacuum flash vaporization of the warm water to drive a low-pressure steam turbine) that was initially demonstrated by CLAUDE in 1930 [5] and has been analyzed recently by BROWN and WECHSLER [6]. BECK [7] and ZENER and FETKOVICH [8] have conceived improvements on the open cycle that would use a "steam lift water pump" or a foam lift concept, respectively, to raise the warm water, which would then fall to drive hydroelectric turbine generators. However, most of the recent OTEC work in the U.S. has addressed closed Rankine cycle systems in which a working fluid (e.g. ammonia) is vaporized by heat exchange with warm, surface water. The vapor drives the power turbine and then is condensed by heat exchange with cold sea water pumped from 1100–4000 ft (340–1220 m) depth. Notable have been the plant concepts developed by the University of Massachusetts [9], Carnegie–Mellon University [10], and industrial teams headed by TRW Systems Group [11] and Lockheed Missiles and Space Co., Inc. [12].

There are two primary concepts for using OTEC power. One is to moor a plant near shore, e.g. in the Gulf Stream off the lower U.S. East Coast, and deliver to shore either electricity via cables [9, 13] or high pressure gaseous hydrogen. Development of a demonstration plant using this approach would offer greater visibility to the U.S. public than one in tropical oceans, and the

potential exists to deliver a significant amount of energy to the south-eastern states. However, plants in the Gulf Stream would have to be much more rugged to withstand hurricanes and large currents, and would have to operate with a much lower ocean temperature difference (ΔT, 30–34°F) than is available in oceans nearer the equator. These factors make such plants more expensive per unit of electric power developed on board.

The other concept is to base OTEC plants in selected tropical ocean sites where design winds and currents are much smaller and available ΔTs are not only much higher (40–43°F) but are less subject to seasonal variation. The available ΔT is a powerful factor; plant cost varies inversely with ΔT to a power of 2 to 2.5. The total energy potential from large siting areas in tropical oceans is much greater, too. The power generated by these OTEC plants can be used to make energy-intensive products (e.g. ammonia, aluminum, liquid hydrogen) on board which can be shipped to U.S. or foreign markets. This paper presents primarily the concepts and economic estimates for producing liquid ammonia (via gaseous hydrogen) and liquid hydrogen that have been developed [1]. Other studies of production of ammonia and hydrogen on OTEC plants [14] and of ocean industrial complexes based on OTEC plants [15] have been supported by ERDA. The writers have drawn on an earlier study by IGT about production of hydrogen from off-peak power at land-based plants reported at this Conference [16] for some of the figures related to costs of hydrogen liquefaction and storage systems.

THE BASELINE, 100-MWe PLANT CONCEPT

The baseline configuration for the 100 MWe(net), OTEC/Ammonia Demonstration Plant-Ship, Fig. 1, integrates the heat exchanger structure and the ship structure to form a barge-type platform. The platform has a beam of 196 ft for compatibility with existing shipyard capabilities and a draft when empty and without the cold water pipe which will permit the plant-ship to be built, launched and fitted out in 40-ft water depths available at shipyards and in waterways to the sea.

The platform arrangement is basically symmetric about a centrally located cold water pipe (CWP). The cold water is pumped from 2500-ft depth through this 60-ft-diam CWP to head ponds over the condensers, flows down by gravity, and is discharged from the condensers at the bottom of the platform. There are two condenser head ponds, one each fore and aft. Each has a sloped feed channel on each outboard side, deepest near the CWP, from which the water overflows transversely inward toward the pond centerline, thus providing a distributed flow over the length of the pond. Removable covers (not shown) may be placed over the ponds to mitigate free-surface dynamic effects and exclude sunlight to minimize growth of marine slimes (biofouling on the tubes). The warm water is pumped from the surface with 20 pumps supplying 20 separate evaporator head ponds. The smooth-sided, prestressed concrete panels that form the heat exchanger box compartment walls also serve as ship structure. The water from the evaporators and the condensers is discharged beneath the ship at a depth of 80 ft.

The arrangement allows clear vertical access over all the heat exchanger units and location of the turbines and associated equipment on decks above them with a minimum of piping. Areas at the ends of the platform are available for product manufacture, for repair and maintenance of heat exchangers, and for the conduct of several tests and programs simultaneously. A 200-ton gantry crane spans the beam and travels the length of the platform. The total "dry" operating weight is 63,800 long tons (LT). Provision is made for 10,000 LT of product and 39,700 LT of ballast. Total operating displacement is 139,300 LT. A propulsion capability for "grazing" at slightly less than 1 knot and for steering is provided by four 1200-SHP electric drive thrusters, one at each ship corner, so that they can perform efficiently in undisturbed water. The power requirement for normal stationkeeping against a 0.5-knot current is only 750 SHP total (0.6 MWe input). A total crew of 48 is required (31 on board and 17 ashore on leave).

The fundamental design decision for this OTEC plant-ship concept was the selection of submerged, modular heat exchangers with 4-in. diameter, multipass, aluminum tubes with the ammonia working fluid inside the tubes and the sea water outside. This approach was selected to avoid the need for heat exchanger shells, to reduce the number of tube joints, thus reducing fabrication and assembly costs, to facilitate maintenance and repairs and permit economical, *in situ* cleaning of the heat exchangers using a water jet system such as the Hydronautics Cavijet™ system [17] without interrupting operation, and to take maximum advantage of a modular
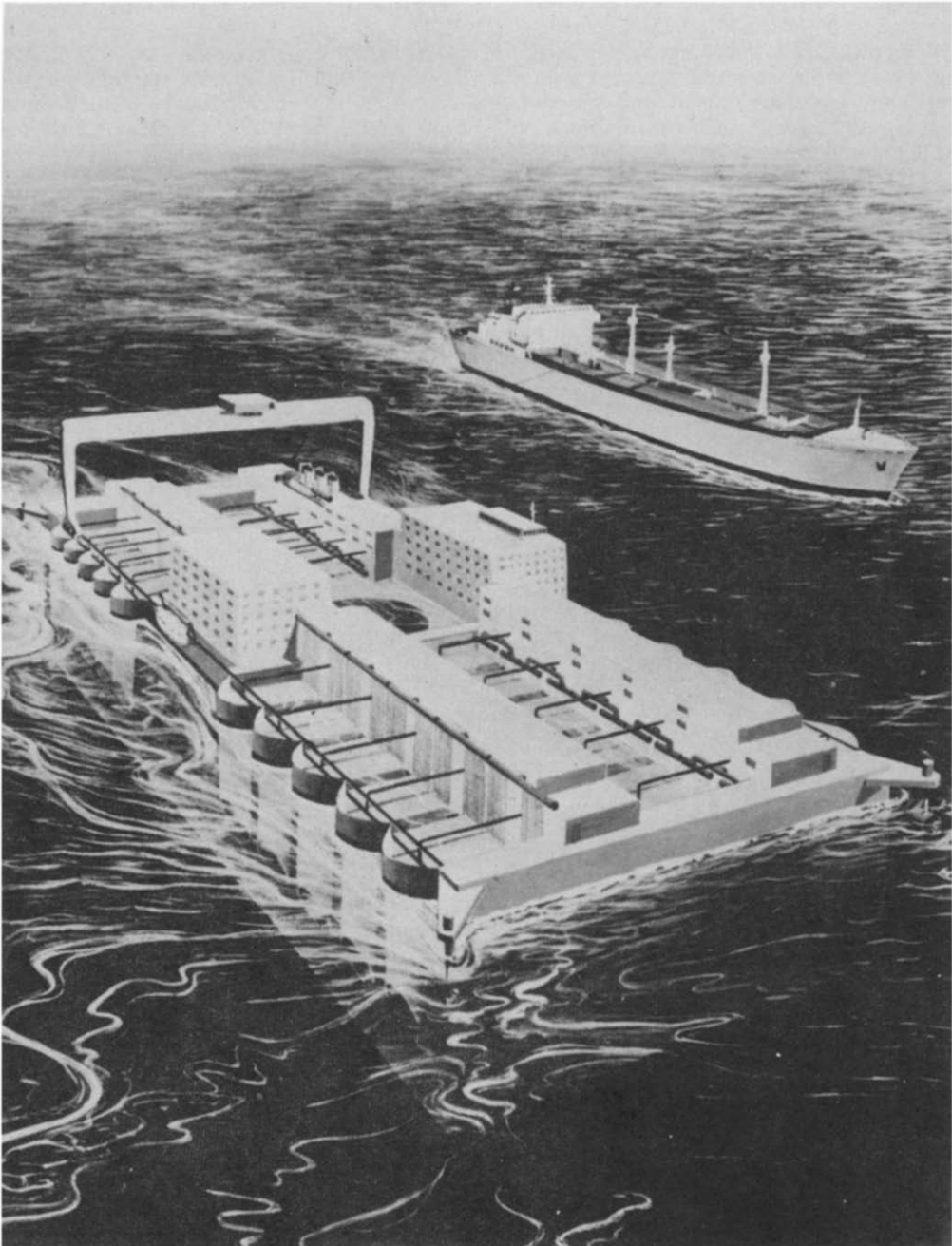
FIG. 1. The APL baseline design for a 100-MWe (net), demonstration size, Ocean Thermal Energy Conversion (OTEC)/ammonia plant-ship. The ammonia tanker in the background would pick up ammonia once a month.

approach. The use of 80 heat exchanger modules assembled into 20 power modules in a total plant will permit regularly scheduled maintenance operations, one module at a time, without a large reduction in the total power output.

The power module concept is illustrated in Fig. 2. In an evaporator element, each of the multiply-nested tubes is fed liquid ammonia via a downcomer pipe and a short riser, in which the liquid flows upward, to promote the onset of nucleate boiling. As the ensuing two-phase mixture flows upward through the multipass tube, 60% of the liquid is vaporized. The demister returns the liquid to the evaporator inlet, and the vapor drives the turbine and then goes to the top of the condenser. The ammonia is condensed completely as it flows down and is delivered through a manifold to the sump, from which it is pumped with the total pressure required for delivery to the evaporator. The condenser element is identical to the two-phase-flow portion of the evaporator element to simplify construction and inventory problems. Figure 2 shows only three nested tubes for simplicity, whereas six nested tubes will be used; it also shows square corners to make the return bends, but several alternatives that may offer lower cost are being evaluated.

The heat exchangers are fabricated from an aluminium alloy (possibly Alclad) to minimize weight and cost. Much experience with aluminum in marine environments [18] and in heat exchangers in water desalting plants [19] indicates that satisfactory life can be achieved; our design life is 20 years. Twenty-seven of the 6-tube elements are manifolded to form one
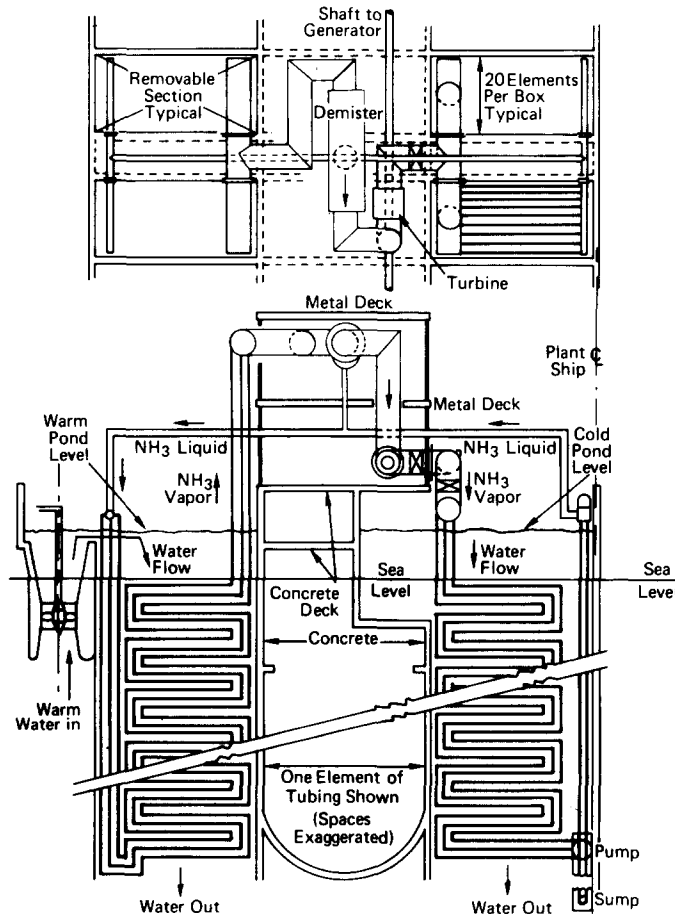

FIG. 2. Basic power unit comprising two each of evaporator and condenser modules powering one 6-MWt gross output turbine.

removable heat exchanger module, which, in conjunction with the mating condenser module, provides the heat transfer needed for 2.5 MWe net power. Two each of the evaporator and condenser modules drive one 6-MWe output turbine. Five turbines are connected on a shaft to drive each of four 29-MWe output, a.c. generator/transformer/rectifier sets. With 16 MWe of power output required for the pumps and propulsion, the total net output is 100 MWe, which is available for ammonia production. Table 1 presents the power summary for the OTEC plant and the ammonia synthesis process via water electrolysis to obtain H_2 and air liquefaction to obtain N_2 as discussed later.

TABLE 1. Power summary for 100 MWe (net) OTEC/ammonia demonstration plant-ship producing 313 STPD of liquid ammonia

Thermal energy input to OTEC power turbines		137.5 MWt
Turbine shaft power output ($\eta_t = 0.88$)[a]		121.0 MWt
Electric power output from generators ($\eta_g = 0.96$)		116.0 MWe
Less parasitic power (MWe) for:		
Ammonia pump input (40 pumps, $\eta_p = 0.8$, $\eta_m = 0.9$)	= 2.3	
Warm water pump input (20 pumps, $\eta_p = 0.8$, $\eta_m = 0.92$)	= 5.0	
Cold water pump input (19 pumps, $\eta_p = 0.8$, $\eta_m = 0.92$)	= 8.1	
Plant propulsion ($\frac{1}{2}$ kt speed)	= 0.6	
Total OTEC parasitic power		−16.0 MWe
Net OTEC power		100.0 MWe
Electrolysis system output, GH_2 ($\eta_e = 0.838$)		83.8 MWt
Ammonia plant output, LNH_3 ($\eta_{c_{theo}} = 0.893$; yield = 0.99)[b]		73.9 MWt
Overall OTEC/ammonia plant thermal efficiency as percentage		
of ideal Rankine cycle efficiency, 73.9/137.5		53.7%

[a] Thermal efficiency symbols: η_t = turbine; η_g = generator/rectifier; η_p = pump; η_m = pump motor; η_e = electrolysis cell; $\eta_{c_{theo}}$ = hydrogen to ammonia chemical conversion, theoretical.

[b] Assumes power requirement for air liquefaction and ammonia synthesis plant (approx. 6 MWe) is obtained from waste heat available from the electrolysis cells (~16 MWt) and the NH_3 synthesis converters (~8 MWt). The ammonia produced is: 73.9 MWt × 3.413 × 10^6 Btu/h-MWt × 24 h/day/(9668 Btu/lb × 2000 lb/short ton) = 313 short ton/day (STPD), where 9668 Btu/lb is the higher heating value of ammonia.

ENVIRONMENTAL OPERATING CONDITIONS AND PLATFORM MOTIONS

There are three siting areas each in the Atlantic and Pacific Oceans which A. R. Miller of the Woods Hole Oceanographic Institution found to have the following suitable characteristics, based on the available data:

(1) surface currents of approximately $\frac{1}{2}$ knot or less, and no substantial currents; deep currents of 0.2 knot or less;

(2) deep bathymetry which insures that there is no possibility of encountering an under-ocean peak extending up to $\frac{1}{2}$ mile below the surface;

(3) available water temperature differences between surface and the deep cold water exceeding 20°C (36°F), surface temperature of 25°C (77°F) or greater;

(4) normal winds of Beaufort force 3 to 4 (7–16 knots);

(5) normal sea states ≤5 (wave height ≤12 ft or 4 m).

For the preferred initial site, Atlantic 1, data are available based on observations back to 1888 and the nearby "Gate" experiment of 1974 (conducted by an international assemblage on site for 60 days with instruments continuously in the water). This 360,000 sq. nmi. area is located 300–900 miles off Brazil from 5° to 15° south latitude. Water depth exceeds 5000 ft (1500 m). There are no islands or other national land. Therefore, the preliminary opinion of the law firm LeBoeuf, Lamb, Leiby and MacRae is that an OTEC plant in this location will be a protected activity under the law of the high seas, similar to fishing.

The on-site temperature, wind and wave conditions have been characterized [1] by oceanographic data for four Marsden Squares covering and near site Atlantic 1. The predominant temperatures are 81–83°F (27–28°C) at the surface and 39–40°F (4–4.5°C) at 2500-ft depth, yielding a representative ΔT of 43°F (23.9°C), our design value. No Beaufort wind force greater than 6 was recorded. A 20-ft significant wave height is, therefore, the *extreme* on-site design condition. The most probable sea state is 3, but sea state 4 occurs with appreciable probability and therefore is the *normal* on-site design sea state, for which the significant wave height is 8 ft. For the latter conditions the calculated motions are:

Pitch = 0.02°; roll = 0.02°; yaw = 0.2°; heave = 2 ft; sway = 0.5 ft.

(For the worst expected sea condition (20-ft significant wave height) these values increase only to 0.1°, 0.1°, 0.7°, 12 ft, and 2.3 ft, respectively.) Such a stable platform should also be very useful for conducting oceanographic research.

HYDROGEN PRODUCTION BY ELECTROLYSIS

The background and basic tradeoffs in the electrolytic production of hydrogen were reviewed recently by KNOPKA and GREGORY [20]. A few points will be appropriate here as background for the information from two companies—Teledyne Energy Systems [21, 22] and General Electric [23–25]—upon which the present investigation is based.

The isothermal generation of hydrogen and oxygen by water electrolysis at 25°C (77°F) ideally requires that a quantity of energy be supplied equal to the heat of formation of liquid water, which is the same as the higher heating value (HHV) of hydrogen, 61,030 Btu/lb. This energy could all be provided as electrical energy at a cell potential, 1.47 V, in which case the quantity can also be expressed as 17.9 kWh/lb of H_2, or 94 kWh/1000 SCF.† In this isothermal case, no heat would be added from or wasted to the surroundings. However, a voltage as low as 1.23 V could still generate hydrogen (Fig. 3) at a very slow rate; this would correspond to an *electrical*
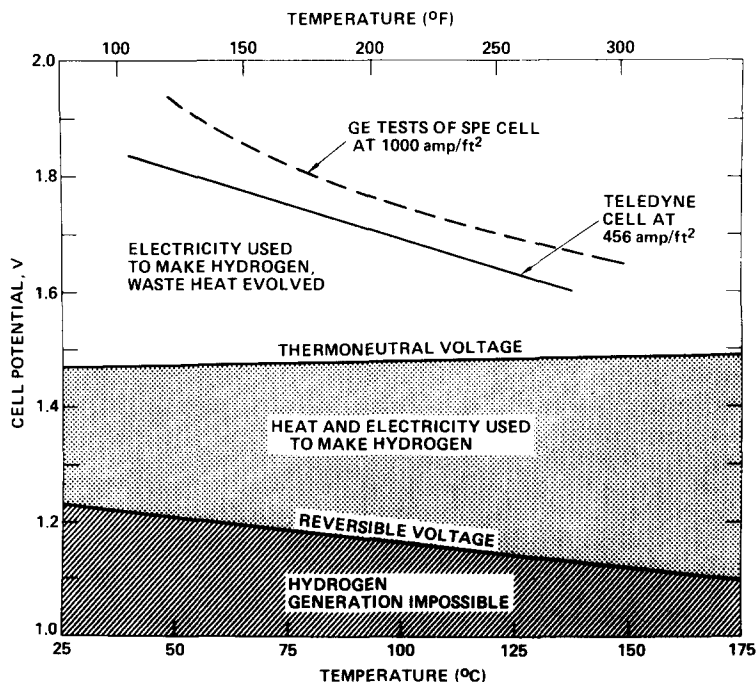

FIG. 3. Idealized operating conditions for electrolysis and temperature effect from [20] with added curves from Teledyne [22] and GE [25].

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†SCF = standard cubic feet measured at 68°F and 1 atm.

energy input of 79 kWh/1000 SCF, or 83.7% of the HHV of hydrogen; the other 16.3% would have to be supplied as heat. Since the "thermal efficiency" of electrolysis is defined as the thermoneutral voltage of 1.47 V divided by the actual voltage used, this means that an ultimate efficiency of 119.5% is theoretically possible if that heat is supplied by the surroundings.

At present most industrial electrolyzers do provide all of the energy as electrical energy and, in fact, operate at a voltage above the thermoneutral value due to inefficiencies inherent in a given design and to a desire to increase the rate of production of hydrogen. The rate of production increases as the applied current and hence the current density I/A (current per unit area of electrode) increases. However, the cell operates as a nonlinear resistor, so there are tradeoffs among E, I/A, and capital cost. Thus, many industrial units today operate at $E \approx 2.0$ V, or 23.8 kWh/lb, or 74% thermal efficiency. However, near-term advances promised by recent and planned developments mentioned hereinafter appear to assure improvements to the 1.5–1.8 V range, corresponding to 18–22 kWh/lb, or efficiencies in the 82–98% range.

Increasing the temperature in an electrolysis cell will decrease the cell's internal resistance and overvoltage because the diffusional and activation processes are accelerated. The solid curve in Fig. 3 was provided by Teledyne for a cell operating at 456 A/ft^2. For the GE solid polymer electrolyte (SPE) cell [23], the dashed curve was obtained in tests of a single cell at a current density of 1000 A/ft^2. The efficiency increased from 76% to 90% as the temperature was increased from 120°F to 300°F. The advantage of operating at high temperatures is, in some designs, offset by decreased life of the cell materials. If an attempt is made to achieve efficiencies very near (or ultimately in excess of) 100%, heat will have to be supplied from the surroundings, because the total enthalpy change of the reaction, as indicated by the thermoneutral line in Fig. 3, increases slightly with temperature.

The Teledyne electrolysis system

The Teledyne Energy Systems Company is currently manufacturing and marketing commercial systems ranging in sizes up to 430 SCFH of hydrogen and expects to have a breadboard model of a large-scale cell operating soon as described in [21].

In their proposed hydrogen plant system for the OTEC application [22], water is decomposed by the electrolysis of a 25% KOH (potassium hydroxide) solution in a filter-press type electrolysis module. The hydrogen is generated at 100 psig, separated from the electrolyte and piped directly to the heat exchangers and compressor of the ammonia synthesis loop. The installation for a 100-MWe demonstration plant would be made up of 16-MWe input unit plants, each containing five electrolysis modules run by auxiliaries. The hydrogen and oxygen gases generated are removed with the existing KOH solution. The KOH is separated from the gases in the primary separators, cooled from 180 to 165°F in heat exchangers and filtered before re-entering the five electrolysis modules from a main header. Teledyne has proposed that for the OTEC/ammonia plant application, the aforementioned heat exchangers would be in the ammonia synthesis plant, used to preheat the reactant gases.

Control equipment will monitor electrolysis module inlet temperature and flow rates; if a high temperature or a low flow is encountered, an alarm will be sounded, and if the condition persists, the plant will be shut down automatically and purged with nitrogen. The H_2 and O_2 product gas streams will be saturated with water vapor at 180°F; the heating values of these streams can be utilized in the ammonia plant. A deionized water system capable of producing pure water for all six plants required will be supplied. Input to this system will be potable water supplied from a fresh-water plant on the platform. (We have assumed throughout that a reverse osmosis desalinization system, which has a power requirement of only 0.2 MWe for a 100-MWe OTEC/ammonia plant-ship, will be used to provide fresh water [1].) The deionized make-up feed water is delivered to the hydrogen separators on demand from level switches in the separators.

Each electrolysis module may be run from 25% to 100% capacity by varying the voltage between 840 V and 860 V (1.68 to 1.72 V/cell) and the current between 950 and 3800 A. The cathode members of the cell are modified by a proprietary technique to achieve the current density required without an unacceptably high cell voltage. To date, this technique has been verified only by preliminary laboratory tests. Table 2 lists the performance characteristics per unit that would apply for a demonstration plant.

TABLE 2. Teledyne electrolysis plant performance and requirements (16.14-MW plant unit)

Hydrogen output	149,500 SCF/hr (776 lb/hr)
Cell voltage and power density	1.72 V, 791 W/ft^2
Voltage and current efficiencies	85.6%, 98%
Overall system efficiency	83.8%
Maximum operating pressure	100 psig
Gas purity after water removal	98 + %
Dew point	180°F; reduced to NH$_3$ synthesis requirement in NH$_3$ plant
Space requirement	25 ft × 25 ft × 15 ft high
Total weight	805,000 lb
Plant life	20 yr
Plant manning	Completely automatic with manual override (one man operation in manual mode).
Deionized water requirement	500,000 Ω/cm resistance (minimum)
Cooling	Must be available in NH$_3$ synthesis loop to cool KOH electrolyte to 165°F.
Power	3.2 MW per module; 440 V, 3∅ for pumps, remainder 220 V, 3∅.
Air	Oil-free air, 80 psig min. pressure.
Nitrogen	700 SCF per startup, 20,000 SCF for emergency purge use
Initial KOH charge	13,764 gal, 25% weight solution
Preventive maintenance	183 man-hr, first year; 124 man-hr 2nd through 20th years.

The General Electric SPE electrolysis system

The General Electric Company is developing a large-scale electrolysis plant and has received some funding from ERDA for this work. In their Solid Polymer Electrode (SPE) Water Electrolysis System [24], the 5-MWt-output (in hydrogen thermal equivalent) system depicted in Fig. 4 is their basic unit. The electrolysis module is 3 ft in diameter by 6 ft high, contains 580 cells of 2.9 ft^2, operates at 220°F, weighs 5000 lb, and generates approximately 53,800 SCFH of hydrogen. Distilled water is circulated through the cell anode cavities at a sufficient rate to remove internal waste heat as well as the amount needed for the electrolysis reaction. Part of this water is carried over to the cathode side charged with both the hydrogen and oxygen gases. The O$_2$–water mixture is passed through a heat exchanger to remove the waste heat from the system, and then to a water separator from which the water is recirculated to the pump inlet. The H$_2$–water mixture goes directly to a water separator. If required, gas driers can be added to the system downstream of the water separators. The system will be completely automated for unattended operation.

Although the 580-cell stack used in the electrolysis module had not yet been produced at this writing, continuing improvements in anode electrocatalysis are being demonstrated in the laboratory from which GE has projected that the cell voltage at 2000 A/ft^2 will be reduced to approximately 1.75 and 1.6 V by 1980 and 1985, respectively. In addition, two other areas of development are demonstrating significant potential for improved performance and the achievement of these goals: the aforementioned increase in operating temperature (Fig. 3), which reduces both the catalytic activation losses and the internal resistance of the electrolyte, and permits a voltage reduction of 0.25 V; and a reduction of the SPE thickness from 10 mils to 5 mils in thickness, permitting another 0.25 V reduction. With these two improvements alone, it will be possible to come within 0.05 V of their 1985 projected performance.

Life testing up to 4 years has been conducted on single units without maintenance and with *no* degradation of performance; testing is continuing. During the past 7 years that GE has been testing SPE cells (near 10^6 cell hours of operation), there have been only two cell failures, both
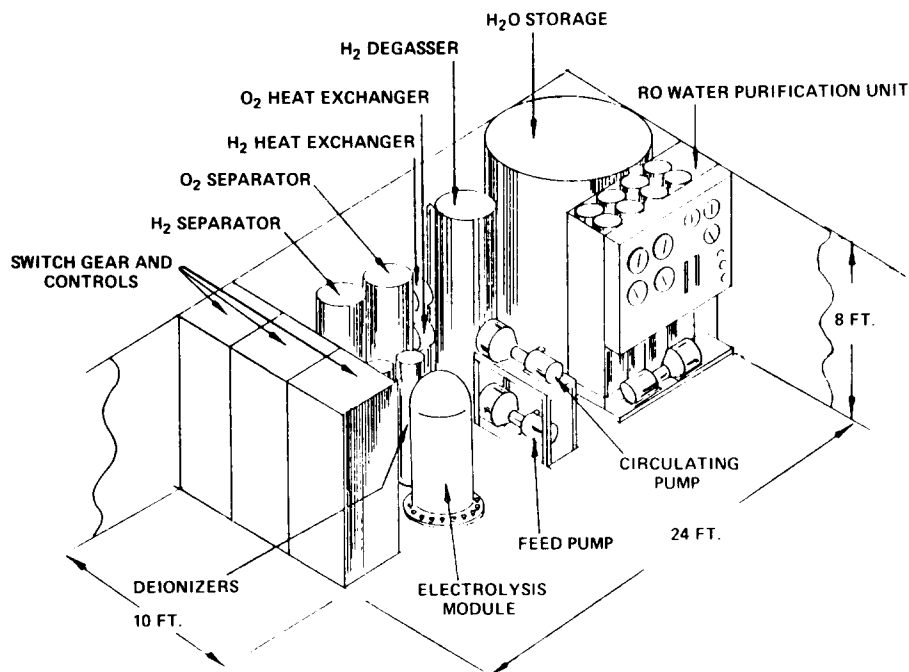
FIG. 4. Conceptual layout of General Electric's 5-MWt output SPE water electrolysis system
[25].

due to gasket failures [23]. With improved sealing techniques now available, GE confidently
projects a reliable, maintenance-free life potential of 10–20 years for the SPE electrolysis stack
in commercial service.

Cell efficiency is primarily a function of the cell voltage, since the coulombic losses at ambient
pressure are almost negligible. They estimate [25] that electrolysis efficiencies of 82% for the
pilot plant (1978), 86% for the demonstration plant (1980), and 90% for commercial plants
(1982–85) can be achieved if adequate government funding for development is provided. That
rate of development would correspond to the schedule advanced by APL for the development
and production of OTEC ammonia-producing plant-ships.

Summary of hydrogen output and cost data

The hydrogen output and cost information provided by Teledyne [22] and GE [25] is
summarized in Table 3. Teledyne's projected costs cover the electrolysis modules, auxiliary

TABLE 3. Summary of projected lead-time requirements, hydrogen outputs, and
electrolysis system costs in 1975 dollars from Teledyne [22] and General Electric
[25]

	100 MWe (1980)			500 MWe (1982-85)		
	Lead time, yr	1000 SCFH output	$/kWe input	Lead time, yr	1000 SCFH output	$/kWe input
Teledyne	2.5	926	142	3	4634[a]	128
General Electric[b]	2.5	932	198	3	4876	78–79

[a]Could exceed 5,000,000 SCF with R&D funding; see text.

[b]GE projections require adequate R&D funding; see text.

equipment, and installation; d.c. power is supplied directly at a suitable voltage from the OTEC generators. (Teledyne has advised that H_2 output can be increased by modifying the cathode members of the cell by a proprietary technique, by increasing cell operating temperature to 250°F and other developments. With R&D funding, Teledyne states the output for the 500-MWe plant could rise as much as 8–14% with no attendant capital cost increase.)

The GE plant cost is quoted for delivery to a U.S. shipyard, tested and installed. It does not include the power for the air dryers which would be supplied from an integrated total OTEC/ammonia plant. A funded pilot plant development program is presupposed. Hydrogen is generated at 14.7 psia. At 500 psia, the hydrogen output for the 500-MWe plant would be reduced to 4,605,000 SCFH, allowing a tradeoff between ammonia output and compressor investment.

The overall OTEC/ammonia plant investment and production cost estimates were based on use of the Teledyne system for the 100-MWe demonstration plant in 1980, since the cost projection is favorable and there appears to be less uncertainty in achieving their development goals with sufficient lead time to meet that schedule. The projections by GE are more favorable for the longer term and are used for the 500-MWe, commercial-size plants hereinafter.

AMMONIA PRODUCTION

The need for alternatives

Ammonia is considered the prime near-term candidate product for tropical OTEC plants for the following reasons. It is a vital commodity for use in fertilizers (75% of the U.S. ammonia consumption at present) and other chemicals. Furthermore, over 95% of U.S. ammonia production uses natural gas as the feedstock (H_2 source). In 1975, 16 million short tons were produced, which required 560 billion SCF of natural gas—nearly 3% of the total U.S. consumption of natural gas. Projections suggest that by 1980 ammonia production may require over 5% of the U.S. total, and by 1990, over 11%. These figures are approximately doubled when the projected petrochemical use of natural gas is added. The forthcoming additional pressure on prices and availability to the individual home-owner consumer is readily apparent. The attainable schedule which we have projected for building OTEC ammonia plant-ships would have 15 commercial-sized plants in operation by the end of 1985, saving 350 billion SCF of natural gas per year; the projected cumulative saving through the year 2000 is 12 trillion SCF.

The use of fertilizer in the U.S. in recent years (47 million tons in fiscal year 1974) accounted for approximately one-third of the U.S. food production [26]. A ton of fertilizer nutrients typically produces 8–10 tons of grain. Our ability to produce food over and above our basic needs contributed about $22 billion to our foreign trade in 1975. With the world population still expanding rapidly, the foreign trade food market will continue to expand. However, ammonia itself as a sale commodity no longer contributes to the U.S. trade balance. In the August 1974–May 1975 period, the U.S. exported only 335,000 tons of ammonia while importing 495,000 tons [27]. The total U.S. demand for ammonia is expected to increase by 10 million tons per year by 1985.

Thus, it is clear that the U.S. will need an alternative to ammonia production from natural gas. Ammonia can be made from petroleum (naptha) or coal, and for the longer term, coal would be the better of these two fossil fuel routes for the U.S. However, there will be many competing demands for coal, and many social and environmental problems will attend a substantial increase in coal mining and transportation capabilities, in addition to the environmental problems of using coal to make ammonia on shore. Table 4 indicates that our estimates for the cost of producing ammonia via hydrogen generated by OTEC plants and delivering it to the U.S. would be competitive with costs of making it on shore from coal, naptha or heavy fuel oil. It will be competitive with the cost from natural gas when that cost approaches an energy-equivalent cost compared to oil (see footnote a, Table 4). Successful development of OTEC/ammonia plant-ships could also lead to sale of that technology or the plant-ships themselves to other nations who do not have large supplies of fossil fuel.

The selected OTEC/ammonia process

Electrolytic hydrogen generated by equipment previously described could be used to synthesize ammonia by either of two approaches. The first approach studied [1, 28] was to use

TABLE 4. Comparison of estimated U.S. production costs in 1975 dollars for ammonia from various resources

Resource	NH$_3$ production cost ($/short ton)	Relative environmental impact
Natural gas	$ 58–107[a]	Intermediate
OTEC/ammonia plant-ship	88–96[b]	Least (some aspects favorable)
Naptha	98–111[a]	Intermediate
Heavy fuel oil	101–116[a]	Intermediate
Coal	98–136[a]	Greatest

[a]Estimates from TVA National Fertilizer Research Center data [33] based on $0.80–2.20/MSCF gas cost, $80–95/ton naptha cost, $11–14/bbl fuel oil cost and $22–42/ton coal cost. The $2.20/MSCF natural gas cost is equal to $14/bbl oil cost on an energy basis (900 Btu/SCF for gas, 5.7 × 10^6 Btu/bbl for oil).

[b]OTEC/ammonia plant-ship cost is for ammonia delivered FOB New Orleans, Louisiana.

approximately $\frac{1}{7}$ of the hydrogen in a burner to remove oxygen from the air, condense the resulting water to be used as feed to the electrolysis cells, and combine the remaining gas (mainly nitrogen but containing carbon oxides, argon, and other trace gases) with the other $\frac{6}{7}$ of the hydrogen to make the feed gas for the ammonia synthesizer. The other approach is to obtain the nitrogen by using an air liquefaction plant to remove the oxygen, water vapor, CO_2 and argon from the air, using a process similar to that used to produce liquid oxygen on Navy aircraft carriers today. The latter approach, used for our cost estimates herein, requires a slightly higher capital investment but, to the greater ammonia yield for a given OTEC power plant size, is presently judged to be more attractive. It is illustrated schematically in Fig. 5. We have not yet
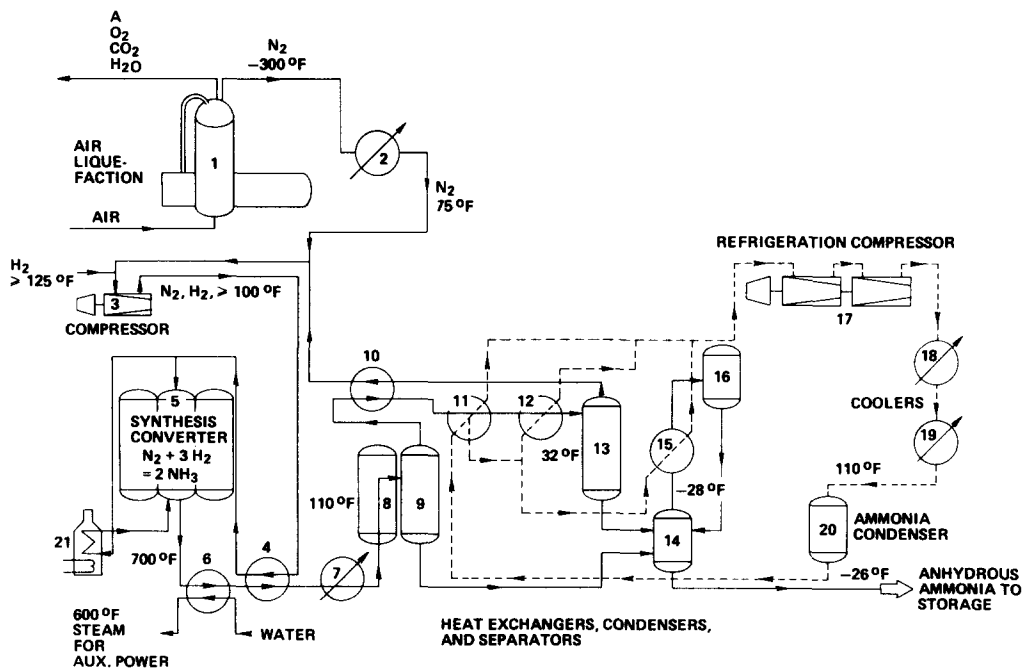

FIG. 5. Schematic diagram of the OTEC ammonia process using air liquefaction.

TABLE 5. Ammonia plant equipment for 100-MWe plant

Item	Deck area, ft^2	Wt. LT	Cost, $M (1975)
Air liquefaction	600	19	2.5
Gas compressor	150	16	1.2
Synth. converters (3) and catalyst	300	77	0.8
Waste heat boilers (2)	250	8	0.1
Various heat exchangers, coolers	830	106	0.7
Refrigeration compressor	150	16	1.3
Condensers and separators	1600	54	0.3
Liquid storage (10^4 LT)	(Hull)		0.8
Transfer pumps, hose	(Hull)	40	0.4
Installation and controls		1	2.3
Totals	3880	337	10.4

conducted a specific process engineering study or prepared a heat balance for this application. However, the assumed process efficiencies for the demonstration plant are indicated in Table 1. For purposes of our plant sizing and capital cost estimate we used the values listed in Table 5 for the items in Fig. 5. The demonstration-size plant-ship (Fig. 1) has ample space and buoyancy to accommodate the electrolysis and ammonia synthesis plant equipment.

Most of the costs in Table 5 are based on installed costs of equipment in the TVA National Fertilizer Research Center's experimental, 248 short ton per day (STPD) ammonia plant in 1970. For our 100-MWe OTEC/ammonia plant-ship, the equipment costs provided by TVA were escalated to 1975 dollars (1.57 ratio to 1970 costs per Chemical Engineering Plant Cost Index) and increased to the 313 STPD level (1.26 ratio). The estimate of $2.5 M for the air liquefaction equipment, based on information from Air Products, is in line with a $5 M estimate for a 1000-STPD plant [20].

The optimum commercial size for OTEC/ammonia plant-ships is expected to be in the 1000–1700 STPD range requiring approximately 300–500 MWe net.

An important factor is to exceed the 600 STPD level to gain the economies of using centrifugal (rather than reciprocating) compressors in the ammonia synthesis plant. A 500-MWe, 1697-STPD commercial size was chosen based on use of the GE electrolysis system at 90% cell efficiency and no advance in ammonia synthesis technology. With a 30% saving due to economies of scale, comparable to scale-up data in [33], the scale-up from 313 STPD (at $10.4 M) to 1697 STPD gives an ammonia synthesis plant cost of $39.4 M. It is assumed that the power requirement for the latter is met by waste heat recovery from the electrolysis and ammonia plants.

OTEC/ammonia plant-ship investment and production costs

The upper half of Table 6 shows estimated plant investments (1975 dollars) for the first, second, fourth, and sixth commercial-size plant ships. The annual production rate is 565,600 STPY based on 345-day/yr operation. All capital costs in Table 6 and subsequent tables include 12% for interest and escalation during plant construction. The "Subtotal OTEC Cost" covering the basic powerplant-ship was obtained by extrapolation from the estimated 100-MWe demonstration plant costs, which (including the aforementioned 12%) are: heat exchangers, $38.5 M; other OTEC cycle equipment, $33.8 M; platform; $32.5 M; cold water pipe (CWP), $7 M. The various components are scaled up by factors of two to five for economy. Appropriate "learning curve" improvements in cost for the lots of units on each ship (e.g. 100 identical evaporator modules) and for a multiple-ship program are assumed: 95% learning curve for the heat exchangers; 90% for the OTEC machinery, CWP, and ship outfit; and 80% for the hull including reuse of concrete forms. The resulting basic OTEC plant cost for the sixth and subsequent 500-MWe ships is $283 M or $566/kWe. This cost, with the 12.8% fixed charges listed in footnote d, Table 6, leads to an onboard power cost of 8.7 mills/kWh.

TABLE 6. Estimated capital and production costs for first, second, fourth, and sixth 500-MWe (net), 585,600-STPY commercial OTEC/ammonia plant-ships in 1975 dollars

	Ship 1	2	4	6
Capital costs, $M[a]				
Platform, $M	129	107	89	80
Heat exchangers	148	133	125	120
Water pumps	39	32	28	26
Turbines/gen./rect.	31	26	23	21
Cold water pipe	22	20	18	17
NH₃ pumps, controls, misc.	24	19	17	17
Deployment to site	2	2	2	2
Subtotal OTEC cost	395	339	303	283
Ammonia plant[b]	96	91	87	84
Total plant invest. (PI) $M	491	430	390	367
Working capital (120 Days), $M	16	14	13	13
Production cost, $/ST				
Chemicals, labor, overhead[c]	------ $5.13 --------			
Fixed charges (12.8% of PI)[d]	107	94	85	80
Production cost at sea	112	109	90	85
Interest on working capital at 8%	2	2	2	2
Shipment to U.S.	------ $9.00 --------			
Cost at U.S., $/ST	123	110	101	96
Gross profit ($180 Price)	32%	39%	44%	47%

[a]All capital costs include 12% for interest and escalation during construction.

[b]Includes desalting plant and GE electrolysis plant; see text.

[c]$0.60 for catalyst and chemicals; $4.53 for labor and overhead, 39-man crew plus 20-man standby crew at $45,000/yr cost per man.

[d]Comprises: maintenance materials, 1%; insurance on hull, indemnity, and construction, 2.8%; depreciation (20 years), 5%; and interest, 8% of ½ PI (PI = Total Plant Investment).

The "Ammonia Plant" costs in Table 6 include the GE electrolysis system ($89/kWe + 12% = $100/kWe or $50 M for the first ship), a reverse osmosis fresh-water system for the cells ($2 M for first ship), and the ammonia synthesis plant ($39.4 M + 12% = $44 m for first ship). A 95% learning curve brings this $96 M total down to $84 M for the sixth ship. Thus, the total OTEC/ammonia plant-ship investment is $367 M for the sixth and subsequent ships. The corresponding ammonia production cost in the lower half of Table 6, based on a costing method similar to that in [33], is $96/short ton including delivery to U.S. ports from distances of about 4000 miles. Compared to an f.o.b. price of $180/short ton, typical of quotations in *Chemical Marketing Reporter* in 1975, a gross profit margin of 47% of sales before taxes is estimated. The magnitude of the investment required for an OTEC/ammonia plant-ship remains a matter of concern. However, depending upon the income and tax situation for the investing company, the total cash requirement may be recovered in a relatively short time. An example was calculated [1] for a U.S. investor using double declining balance depreciation on a 20-yr life with no residual value after depreciation, an investment tax credit of 10% based upon the 1975 update of the Investment Credit and Accelerated Depreciation Restoration Act of 1967, a sales price of $180/ton in constant 1975 dollars, and an investing company in a 48% Federal Income Tax bracket with other income sufficient to allow use of the investment credit to at least the amount of the computed tax on ammonia profits. An average cash flow of $70 million per year would result, and the total plant investment of $367 million would be recovered in 5.3 years.

The "total plant investment" of $367 million (in 1975 dollars) for the sixth and subsequent commercial-size plant-ships in Table 6 is considered to represent state-of-the-art for the 1982–85 time period with the assumption that funding is provided for the development required on electrolysis systems to meet the GE projection of cost and performance, as well as the necessary program for development and demonstration of the OTEC equipment from the existing concepts and technology base. The analysis [1] included an estimate of reductions in plant cost that could be achieved by further R&D to refine those concepts and sharing the cost of the OTEC cold-water pipe and pumps with a mariculture operation. The resulting estimate of possible gross plant ship cost was $329 million. Thus, if all the estimated improvements were achieved, the resulting ammonia cost would be $88/ton, the lower end of the range shown for the OTEC/ammonia line in Table 4.

This OTEC plant-ship, which uses aluminum heat exchangers integrated with a low-cost, barge-type platform and is designed to take advantage of high ocean ΔT's at selected tropical sites where all environmental operating conditions are most favorable, is considerably lower in cost than the baseline designs that have been developed by the teams headed by TRW [11] and Lockheed [12]. The latter teams reported that costs also would be reduced by replacing their titanium-tubed heat exchangers by aluminum heat exchangers and making other design changes. However, they did not present complete cost breakouts for such changes or for commercial plants of larger size. For comparison purposes their baseline power-plant-ship concept costs were adjusted to 500-MWe size allowing economies of scale comparable to those estimated for the APL concept. Table 7 shows that the estimated plant investments for both the basic power-plant

TABLE 7. Comparison of 500-MWe (net), 586,000-STPY OTEC/ammonia plant-ship costs and ammonia production costs for APL and extrapolated TRW and Lockheed baseline designs in 1975 dollars

Design:	APL	TRW	Lockheed
ΔT (°F):	43	40	34
HX tube material:	Al	Ti	Ti
CWP depth (ft):	2500	4000	1500
Capital costs, $M[a]			
Heat exchangers (HX)	120	405	831
Other equipment	64	82	154
Platform	80	150	212
Cold water pipe (CWP)	17	47	24
Deployment to site	2	2	4
Subtotal OTEC cost	283	686	1225
Electrolysis & NH_3 plant	84	84	84
Total plant investment, $M	367	770	1309
Working capital (120 days, $M)	13	22	34
Production costs, $/ST:			
Catalyst, chemicals, labor and			
overhead, maintenance materials	11	11	11
Insurance (2.8% of PI)	18	37	63
Depreciation (for plant life, yr)	31 (20)	33 (40)	59 (38)
Interest (8% of ½ PI)	25	53	89
Production cost at sea	85	134	222
Interest on working capital (8%)	2	3	5
Shipment to U.S. port	9	9	9
Cost at U.S. port, $/ST	96	146	236
Cost relative to APL cost	1	1.52	2.46

[a]All capital costs include 12% for interest and escalation during construction.

ships and ammonia plant-ships are higher by factors near 2 for the TRW baseline concept and by factors greater than 3 for the Lockheed baseline concept. The resulting ammonia production costs, taking into account their longer plant life estimates (hence slower writeoffs) would be greater by factors of 1.5 and 2.5, respectively.

LIQUID HYDROGEN (LH₂) PRODUCTION

OTEC/LH₂ plant-ship investment and production costs

For OTEC/LH₂ plant ships, the General Electric Company's projection for electrolysis equipment was employed for a 500-MWe, 79,400 STPY commercial plant in 1985. For the estimates of hydrogen liquefaction and storage costs we drew upon [16] (which drew upon [29]). For liquefaction equipment the cost given was $0.5M/STPD of LH₂ produced plus 50% for installation for plant capacities from 10 to 250 STPD. It was stated that economies of scale could be expected near the upper end of this range and that technology improvements are expected. For the OTEC/LH₂ plant in 1985, a 30% reduction in cost was assumed, yielding a cost for liquefaction equipment of ($0.75 M/STPD) (0.7) (230 STPD) = $121 M. For LH₂ storage a cost of $1.8 M for 10^8 SCF was given. For 30-day storage on the plant-ship, the capacity required will be 26.4 times this large, and it was assumed that the scaled-up LH₂ storage and transfer system cost would increase with the $\frac{2}{3}$ power of the volume stored. Thus, a cost of $23 M was estimated. These costs were then increased by 12% for interest during construction and reduced by a 90% learning curve to the sixth plant-ship, as in Table 6. The resulting OTEC/LH₂ plant-ship cost is $438 M as shown in the first column of Table 8.

TABLE 8. Estimated capital and production costs for sixth and subsequent 500-MWe OTEC/LH₂ plant-ships in 1975 dollars

	1985	1990s
Capital costs, $M		
Basic powerplant-ship cost, deployed, $M	283	245[a]
Electrolysis, including fresh-water plant[b]	34	31
Hydrogen liquefaction	102	94
LH₂ storage (30 days) and transfer system	19	18
Total OTEC/LH₂ plant investment (PI), $M	438	388
Working capital (120 days), $M	14	13
Rated capacity (345 days/yr), ST/Y	79,000	92,100
Production costs, $/ST		
Labor and overhead[c]	42	43
Fixed charges (12.8% of PI)[d]	707	540
Production cost, $/ST	749	582
Interest on working capital (8%)	14	11
Shipment to U.S. port[e]	62	60
Cost at port per short ton, $/ST	825	653
Cost per 10^6 Btu (LHV, 19.4 lb)[f]	$8.00	$6.30
Gross profit margin, % (sales price, LHV)	68% ($25)	21% ($8)

[a]Reduced cost achievable with R&D program and sharing of cold-water sub-system cost with a mariculture operation.

[b]Hydrogen liquefaction efficiencies used (fractions of the 61,030-Btu/lb HHV of H₂ remaining after accounting for the liquefaction requirement) are 0.75 and 0.80 in 1985 and the 1990s, respectively. For the 1990s, 15% cost reductions in LH₂ plant components due to R&D and 98% electrolysis cell efficiency are assumed.

[c]Crew of 49 on board and 25 ashore, at $45,000/year each.

[d]See footnote d, Table 6.

[e]Estimated @ $54/ST plus 1% of total cost for boil-off loss.

[f]Lower heating value (LHV) = 51,570 Btu/lb; costs based on the HHV would be $6.80 and $5.40/$10^6$ Btu in 1985 and 1990s, respectively.

As previously mentioned, an estimate was made of numerous improvements in the basic OTEC power-plant-ship cost that might be achieved through R&D and sharing cold-water system cost with a mariculture operation. If these improvements are achieved by the 1990s, the estimated cost for the basic 500-MWe power-plant-ship will be reduced to $245 M. If 15% improvements in cost of each of the LH_2 plant components also are achieved through R&D, and hydrogen liquefaction efficiency as a percentage of the higher heating value of hydrogen is increased to 80%, the *possible* 1990 OTEC/LH_2 plant-ship investment (in 1975 dollars) becomes $388 M for a rated production capacity of 92,100 STPY as shown in the second column of Table 8. These plant investments lead to the estimated production and delivery costs shown in the lower half of Table 8. For the 1985 case, the delivered cost is $825/short ton or $8.00/$10^6$ Btu based on the lower heating value (LHV) of hydrogen. The possible cost for the 1990s is $653/ST or $6.30/$10^6$ Btu based on the LHV of H_2 (or $5.40/$10^6$ Btu based on the higher heating value).

The sales price of LH_2 depends strongly on the customer's volume requirement in the current market. For customers needing about one million SCF per day, the price is approximately $5/1000 SCF [16]. This price corresponds to $18.25/$10^6$ Btu (based on LHV of H_2). However, a NASA contract for LH_2 for the space shuttle during the period 1975–1985 [30] is calculated to correspond to a price near $26/$10^6$ Btu. Therefore, a price of $25/$10^6$ Btu is considered, for 1985, which would yield an estimated gross profit margin of 68% for the OTEC/LH_2 plant.

For hydrogen users in the 100 million SCF/day class, the forecast prices of LH_2 delivered from land-based, fossil-fuel-feedstock plants are considerably lower. For a plant using partial oxidation of coal, the cost estimated [16] for production of *gaseous* hydrogen, when adjusted to late 1975, to a higher coal price ($28.50/ton),† and to the lower heating value of H_2, is approximately $4/$10^6$ Btu. For production of LH_2, this would yield approximately $8/$10^6$ Btu. Thus, the essentially state-of-the-art OTEC/LH_2 plant in 1985 would have no chance of returning a substantial gross profit margin in competition with such land-based plants. However, there will be many other demands for fossil fuels that will make additional R&D on the OTEC/LH_2 resource very worthwhile to achieve the more attractive LH_2 cost of $6.30/$10^6$ Btu in the 1990s. Compared to a sale price of $8.00/$10^6$ Btu, this cost would yield a 21% gross profit margin.

This *cost* of LH_2 is still rather high for direct use as a fuel for, say, automobiles and commercial aircraft. However, it is not far greater than the cost of synthoil made from coal, which has been estimated to be $25/bbl or $4.31/$10^6$ Btu [31]. Since future automobile engines [32], aircraft engines, and other heat engines will operate at higher efficiency on LH_2 than on synthoil, operating costs could be nearly comparable, while atmospheric pollution could be reduced.

Cost of electricity from OTEC/LH_2 used in fuel cells on shore

An interesting possibility is that the LH_2 shipped from OTEC plants could be used in fuel cells on shore to generate electricity. The General Electric Company's electrolysis cells can be operated reversibly to serve as fuel cells. The efficiency of operation in this mode currently is 57%, but laboratory tests have shown 61% at 100 A/ft^2 in continuous operation. A reasonable projection for 1990 is judged to be 70% efficiency. Table 9 shows that the estimated 1990 cost (in constant 1975 dollars) of the electricity generated on shore by this approach is 29 mills/kWh, a figure that surely will be of interest. Still another possibility is to ship liquid ammonia to the desired site, there to decompose it to GH_2 for use in fuel cells [34].

CONCLUDING REMARKS

A concept has been developed for Ocean Thermal Energy Conversion plant-ships for use at selected sites in tropical oceans to produce energy-intensive products on board. It employs a novel concept for the OTEC heat exchangers, which are made of aluminum and are integrated structurally with the barge-like hull to minimize cost. These slowly grazing ($\frac{1}{2}$ knot) platforms will be very stable under normal operating conditions and could serve many other needs simultaneously (e.g. oceanographic research, and cold-water supply for mariculture).

The most attractive candidate product from such plants in the near term (early 1980s) is

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†In September 1974 the TVA negotiated a contract with Webster Country Coal Co. for 1.5 million tons/yr for 3 years at $28.50/ton, subject to price adjustment (*Energy Users Report*, Nov. 28, 1974).

TABLE 9. Possible 1990s cost for generating power with fuel cells on shore
using LH₂ from OTEC plant (1975 dollars)

Possible OTEC/LH₂ plant-ship investment at sea, $/kWe[a]	776
Efficiency for delivery of LH₂ to shore[b]	0.508
Effective OTEC/LH₂ plant-ship investment per kWe output from fuel cells, $/kWe	1528
Fuel cells installed on shore, $/kWe[c]	156
Storage system installed on shore, $/kWe	56
Subtotal cost per kWe output, $/kWe	1740
Recovery factor (partial recovery of energy of liquefaction)	1.06
Effective total investment, $/kWe	1642
Weighted fixed charge rate (13% on plant-ship, 15% onshore facility)	13.25%
Fixed charge, mills/kWe	24.8
Operating costs, ship + shore, mills/kWe	2.0
Fuel cost (transportation of LH₂ to shore), mills/kWe	2.1
Power cost at fuel cell plant, mills/kWe	28.9

[a]Plant-ship investment is $388 M for 500 MWe generated onboard.

[b]Product of the following: plant factor at sea (345 days/year), 0.945; electrolysis efficiency, 0.98; liquefaction efficiency fraction of HHV, 0.80; fuel cell efficiency, 0.70; storage efficiency (boil-off), 0.98.

[c]Based on 15% improvement on fuel cell cost due to R&D, 30% addition for installation, 0.70 fuel cell efficiency, and 0.90 plant factor on shore.

ammonia, which is presently produced mainly from natural gas; both are expected to be in increasingly short supply in the U.S. Estimated costs for producing ammonia on OTEC plant ships via electrolytically-produced hydrogen are attractive. This means for obtaining ammonia for U.S. agricultural and chemical needs would appear to be more attractive economically and environmentally than producing it from coal, the most abundant fossil fuel in the U.S. It could be very attractive to any nation not having an abundant and assured supply of natural gas or oil.

Tropical OTEC plants producing liquid hydrogen for delivery to the U.S. probably would have limited market access and be considered higher risk for investment in the near term (1985). However, an aggressive R&D program on OTEC and the technologies for water-electrolysis and hydrogen liquefaction, storage, and shipping could lead to a competitive position for LH₂ from OTEC plants in the 1990s. If fuel cell technology also advances as expected, this LH₂ could serve for production of electricity on shore as well as for transportation, heating, and chemical and industrial uses.

The key sister technology to OTEC for both ammonia production and LH₂ production is water electrolysis. It is strongly recommended that at least two paths for development of electrolysis systems be given priority support along with OTEC.

A strong reason for giving high priority to the development of OTEC technology *now* can be appreciated by considering the situation if the sixth OTEC/ammonia plant existed *now*. The ammonia production cost would be only $96 per ton. If this plant-ship proved to be still fully usable with only routine maintenance beyond the 20 years we have projected, the forecast production cost for 1996 and beyond (depreciation no longer included) would be only $131/ton, increasing at, say, only 5% per year thereafter, while the production cost using coal feedstock would have exceeded $200/ton by 1996 and would increase at 7%/yr.

We believe it would be possible to develop and produce enough OTEC/ammonia plants by the mid-1980s to meet the projected additional U.S. requirement for ammonia by that time, estimated at 10 million tons per year. After satisfying this need, many additional OTEC plants could be used to produce liquid hydrogen or aluminum or other energy-intensive products. The attractive potential for related mariculture operations also warrants further study.

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