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HYDROGEN ENERGY SYSTEMS STUDIES

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TABLE OF CONTENTS

LIST OF FIGURES

FIGURES - CHAPTER 1

FIGURES - CHAPTER 2

LIST OF TABLES

TABLES - CHAPTER 1

TABLES - CHAPTER 2

HYDROGEN ENERGY SYSTEMS STUDIES

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SUMMARY

Hydrogen is being considered as a low polluting energy carrier, especially for use in zero emission vehicles. If hydrogen is derived from renewable sources (biomass, solar, wind), it would be possible in principle to produce and use energy on a large scale with greatly reduced greenhouse gas emissions and very little local pollution.

The results of our previous studies suggest that use of hydrogen from natural gas might be an important first step toward a hydrogen economy based on renewables. Because of infrastructure considerations (the difficulty and cost of storing, transmitting and distributing hydrogen), hydrogen produced at the end-user's site could be a key feature in the early development of hydrogen energy systems. Technologies for producing hydrogen from natural gas at small scale could be critical to getting started with hydrogen as an energy carrier.

Several options exist for producing hydrogen from natural gas at small scale. (Processes for producing hydrogen from methane are known as "reforming".) Catalytic steam methane reforming is a well known, commercially available technology. Small scale partial oxidation systems are now being commercialized. Autothermal reformers are also under development. In the first chapter of this report, we assess the technical and economic prospects for small scale reformer technologies, addressing the following questions:

> 1) What are the performance, cost and emissions of small scale steam reformer technology now on the market? How does this compare to partial oxidation and autothermal systems?

2) How do the performance and cost of reformer technologies depend on scale? What critical technologies limit cost and performance of small scale hydrogen production systems? What are the prospects for potential cost reductions and performance improvements as these technologies advance?

3) How would reductions in the reformer capital cost impact the delivered cost of hydrogen transportation fuel?

We found that there appear to be real opportunities to reduce the capital cost of small scale steam reformers. Our results can be summarized as follows:

* Two types of small scale steam methane reformers exist: "conventional" steam methane reformers, which are well established, commercially available technology and are widely used in chemical applications; and "fuel cell type" steam reformers, which are being developed as part of fuel cell cogeneration packages, and have just become commercially available as part of a system for stand-alone hydrogen production.

* Conventional steam methane reformers produce hydrogen at relatively high pressure (so that downstream purification equipment such as PSAs will work well), and temperature (so that methane conversion to hydrogen will be high), and high purity (99.999% hydrogen). Heat transfer is accomplished via radiative heating of long catalyst filled tubes. They are custom designed.

* The installed capital cost for a conventional steam methane reformer hydrogen plant exhibits strong scale economies in the size range of interest, and is given by:

$$\text{Installed cost} = \$3 \times 10^6 \times (\text{H2 plant capacity in million scf/day})^{0.3}$$

The cost of conventional SMR plants might be reduced if the design were standardized.

* The main capital cost drivers for small scale conventional SMR hydrogen plants are the reformer vessel (which accunted for about 43% of the capital cost in a hydrogen plant producing 1.6 million scf H2/day), and the PSA (39%).

Compressors account for 11% and shift reactors 7%. Engineering costs are significant for small, one of a kind conventional SMR plants.

* The majority of the cost for a conventional reformer vessel is for the reformer tubes, which are made of high alloy steels to withstand the pressures and temperatures in the reformer.

* Fuel cell steam methane reformers produce hydrogen at relatively low pressure (and temperature) and low purity (only CO is removed), as required by fuel cells. Heat transfer is accomplished in a compact design via convective heat transfer. They are a standardized design.

* Fuel cell steam methane reformers offer several design features which could reduce the capital costs as compared to "conventional" small scale steam reforming: use of convective rather than radiative heat transfer which reduces heat exchanger area; lower pressure operation, which reduces the cost of reformer steels (stainless steel can be used instead of high alloy steels; and standardized design, which reduces engineering costs.

* The approximate cost of fuel cell reformer vessel plus shift reactors is about

$$\$0.44 \text{ million} \times (\text{capacity in million scf } H_2/\text{day})^{0.7}$$

assuming mass production of several hundred units/year. This does not include the cost of downstream compression and purification in a PSA.

* The costs of downstream purification are significant, but even when compression and PSA are added to a fuel cell reformer, adapting fuel cell steam methane reformers for stationary hydrogen production looks promising. The capital cost of small scale steam reformer systems might be reduced by 25-70%. as compared to conventional small scale SMR systems. Some of the savings would be achieved via standardized design, and some via lower materials costs in the reformer vessel because of lower pressure.

* Operating pressure of the steam reformer is an important issue. At higher pressure (15-25 atm), the cost of the reformer materials is higher (alloys rather than stainless steel must be used for tubes), but there is no need for costly dowstream compression prior to the PSA. There are system trade-offs between lower pressure, less costly reformers and more costly downstream clean-up and compression equipment. Even among fairly low pressure reformers, reforming at 5 atm rather than 1 atm appeared to offer advantages in terms of reduced compression power requirements and costs.

* By using lower cost fuel cell reformers, the delivered cost of hydrogen transportation fuel produced via onsite reforming might be significantly reduced at small station size (less than a million scf H2/day). For a hydrogen refueling station dispensing 0.1 million scf H2/day, the delivered cost of hydrogen transportation fuel would be reduced by about 40% from $40/GJ (for a conventional SMR) to $25/GJ (for a fuel cell type SMR). For a hydrogen refueling station dispensing 1 million scf H2/day, the delivered cost of hydrogen would be reduced by about 20% from $14.2/GJ (for a conventional SMR) to $11.5/GJ (for a fuel cell type SMR). These costs are competitive with pipeline delivered hydrogen from centralized production via SMR or biomass gasification. With low cost fuel cell reformers, the decentralized production of hydrogen from natural gas becomes approximately competitive with centralized production plus distribution.

* Hydrogen purification technologies at small scale are an important part of the cost of a small scale reformer system (especially for POX). According to our preliminary estimates the cost of the PSA may account for a large majority of the system capital cost for a POX system, and about 1/3-1/2 of that for a small scale steam reformer.

* We estimate the capital cost for a small scale POX plant producing hydrogen from methane would be comparable to that of a fuel cell steam reformer plant. The POX reactor vessel would be quite inexpensive compared to a steam reformer vessel, but the downstream clean-up and need to process nitrogen for an air fired POX would add considerably to the cost. The energy conversion efficiency [=hydrogen out

(HHV)/methane in (HHV)] for a POX-based hydrogen production system would probably be about 65-75% slightly less than that for a small steam methane reformer which might have an efficiency of 80-85%.

* Key issues for reducing the cost of small scale hydrogen production systems are:
1) lower pressure reformer operation, which gives a lower reformer vessel capital cost because lower cost steels can be used, and also appears to give a lower overall system cost
2) standardization of the reformer design (to reduce engineering costs)
3) use of compact, convective heat transfer reformer vessel design, as in fuel cell reformers to reduce the heat exchanger area needed.
4) Reduction of the cost of hydrogen purification systems (PSA). Some of this could be achieved through mass production.
5) Pursuit of innovative concepts such as sorbent enhanced reforming which might get around some of the trade-offs between the cost of the reformer vessel and the cost of the downstream purification.
6) For POX-based small scale hydrogen production systems, the POX vessel alone will be almost negligible in cost. But the relatively low hydrogen content and high nitrogen content of the syngas will increase the cost of downstream processing and purification equipment. Bringing these costs down is crucial. Enhancing the oxygen content into the POX may help, if this can be done inexpensively.

In Chapter 2, we discuss potential markets for hydrogen transportation fuel in the Southern California area. The results of this study are as follows:

* If fuel cell vehicles capture a significant fraction of the emerging ZEV market, several hundred thousand fuel cell cars could be on the road in the Los Angeles area by 2010.

* If hydrogen fuel cell cars captured half the ZEV market, about 50 million scf of hydrogen per day would be required, a capacity similar to that in a good sized oil refinery.

* The type of refueling network would depend on the geographical concentration of the demand, as well as the relative economics of various supply options.

* A more detailed case study of hydrogen infrastructure in the Southern California area will be presented in a forthcoming report.

1.0 ASSESSMENT OF TECHNOLOGIES FOR PRODUCING HYDROGEN FROM NATURAL GAS AT SMALL SCALE (TASK 1)

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1.1 INTRODUCTION

One of the major challenges facing the development of hydrogen as an energy carrier is the current lack of infrastructure. If hydrogen could be produced near the point of use (onsite) from widely available energy carriers such as electricity or natural gas, the difficulty and cost of building a hydrogen transmission and distribution system might be postponed, at least in the near term.

Our previous studies suggest that small scale reforming of natural gas could be an important technology in the early stages of a hydrogen economy (Ogden et.al. 1995). There are several reasons why onsite production of hydrogen from natural gas might be attractive for applications such as hydrogen vehicle refueling stations or fuel cell cogeneration systems:

1) Natural gas is widely distributed today and would be available in most urban areas of the US.

2) The technology for producing hydrogen from natural gas is well known. Steam reforming is commercially available and is the most widely used method of hydrogen production. Fuel cell manufacturers are developing lower cost steam methane reformers. Partial oxidation systems are now entering the market. Another option is autothermal reforming, which now under development.

3) Steam reforming of natural gas is currently the least expensive and most widely used method of hydrogen production over a wide range of plant sizes (e.g. for hydrogen demands above about 0.1-0.5 million scf/day -- See Box 1).

At present, steam reforming of natural gas is the dominant method of industrial hydrogen production. Most steam reforming units built to date have much larger capacities than the 0.1-2.0 million scf/day needed for hydrogen refueling stations or fuel cell cogeneration applications. (Typical applications are for refineries or ammonia production, where 25-100 million scf/day might be needed.) However, a number of smaller reformers have been built for chemical processing needs. Small scale steam reformers are commercially available today from several manufacturers including Haldor Topsoe, Howe-Baker Engineers, Hydrochem, KTI and Engelhard.

Steam methane reformers in this size range have also been developed as part of fuel cell cogeneration packages (Table 2). Several fuel cell manufacturers (International Fuel Cells, Ballard Power Systems, Chiyoda) are developing fuel cell reformer designs which are more compact and offer higher efficiencies. The first hydrogen production system using a fuel cell type reformer has recently been commercialized by Praxair using an International Fuel Cells reformer (Farris 1996).

Small scale partial oxidation systems have recently been commercialized and autothermal systems are also under development.

Our previous studies (Ogden et.al. 1995) indicate that with commercially available small scale steam reformer technology (in the range 0.1-2.0 million scf/day), it would be possible to produce and deliver hydrogen transportation fuel at a cost roughly competitive with other sources (truck delivered liquid hydrogen or electrolytic hydrogen from off-peak power). The reformer capital cost is a major contributor to the total delivered cost of hydrogen transportation fuel, especially at small plant size (e.g. less than 0.5 million scf/day), where it can account for 1/3 to 1/2 of the delivered hydrogen cost. Clearly, if the cost of the reformer could be reduced, this could lower the cost of hydrogen transportation fuel, making onsite small scale hydrogen production more economically attractive.

The goal of this study is to better understand the technical options for low cost, small scale production of hydrogen from natural gas, considering:

a) steam reforming,
b) partial oxidation
c) autothermal reforming.

Our focus is on stationary hydrogen production systems (rather than onboard vehicle reformers), in the range 0.1-2.0 million scf H2/day. This is about the size range needed for hydrogen vehicle refueling stations or cogeneration systems in commercial buildings. [For example, a refueling station serving a fleet of 200 Ballard type PEM fuel cell buses might require about 1.4 million scf H2/day. A 200 kW, natural gas fueled, fuel cell cogeneration system might require about 140,000 scf H2/day.]

We have addressed the following questions:

* What are the performance and cost of small scale steam methane reformers now on the market? How does this compare to partial oxidation and autothermal reformer systems?

* How do the performance and cost of reformer technologies depend on scale? What critical technologies limit cost and performance of small scale hydrogen production systems? What are the prospects for potential cost reductions and performance improvements as these technologies advance?

* How would reductions in the reformer capital cost impact the delivered cost of hydrogen transportation fuel?

1.2. APPROACH

First, we gathered cost and performance data on commercially available reformer technologies and those now under development, including: small scale steam reformer systems for industrial and fuel cell applications, partial oxidation systems and autothermal reforming systems. (Table 1 lists various small scale hydrogen production systems.)

To better understand the fundamentals of small scale reforming, we carried out computer modelling of:

a) equilibrium and kinetics of reformer reactions,
b) conversion of methane in a single steam reformer tube, and
c) hydrogen plants based on steam reforming, including upstream and downstream processing and heat recovery.

Throughout the study, we held discussions with reformer manufacturers and developers (Howe-Baker, KTI/Air Products and Chemicals, Inc., Haldor-Topsoe), fuel cell developers (International Fuel Cells, Ballard Power Systems, H-Power), developers of partial oxidation (POX) systems (Arthur D.Little, Hydrogen Burner Technology) and industrial gas companies (Air Products and Chemicals Inc., Praxair) focussing on cost and design issues for small scale reformers.

To gain further insight on small scale reformer design, we surveyed the patent literature on small scale reformer systems.

1.3. SURVEY OF TECHNOLOGIES FOR SMALL SCALE PRODUCTION OF HYDROGEN FROM NATURAL GAS

1.3.1. CATALYTIC STEAM REFORMING

Catalytic steam reforming of natural gas is a well known, commercially available process for hydrogen production (Rostrup-Nielsen 1984, Twigg 1989). Hydrogen production is accomplished in several steps: steam reforming, water gas shift reaction, and hydrogen purification. (Figure 1 shows material flows for a typical hydrogen production plant based on steam reforming of natural gas.)

The steam reforming reaction

$$CH_4 + H_2O <-> CO + 3\ H_2 \qquad \Delta h = +206.16\ \text{kJ/mol}\ CH_4 \qquad (1)$$

is endothermic and is favored at higher temperatures, and lower pressures. (This is illustrated in Figure 2, which shows how the equilibrium conversion of methane varies with temperature and pressure.) Typical reformers operate at anywhere from 3 atm, 700^oC to 15-25 atm, 850^oC. External heat needed to drive the reaction is often provided by the combustion of 20% of the incoming natural gas feedstock (exhaust from fuel cell anode or purge gas from the hydrogen purification system are often used in addition). Heat transfer to the reactants is accomplished indirectly through a heat exchanger. Methane and steam react in catalyst filled tubes (typically the "steam-to-carbon" ratio is about 3 or more to avoid coking).

After reforming, the resulting syngas is sent to one or more shift reactors, where the hydrogen output is increased via the water-gas shift reaction:

$$CO + H_2O \leftrightarrow CO_2 + H_2 \qquad \Delta h = -41.15 \text{ kJ/mol CO} \quad (2)$$

This reaction is favored at temperatures of less than about 600^oC, and can take place as low as 200^oC, with sufficiently active catalysts. The gas exiting the shift reactor contains mostly H2 (70-80%) plus CO_2, CH_4, and small quantities of H_2O and CO.

Hydrogen is then purified. The degree of purification depends on the application. For industrial hydrogen, pressure swing absorption (PSA) systems or palladium membranes are used to produce hydrogen at up to 99.999% purity. For PEM or phosphoric acid fuel cells closely coupled to reformers, diluents such as CO_2 and CH_4 are tolerable. However, CO must be reduced to less than about 10 ppm for PEM fuel cells, so a CO removal system such as preferential oxidation must be used.

In a preferential oxidation system, the gas is passed over a catalyst bed, with added air. At certain temperature and stoichiometry conditions, the reaction

$$CO + 1/2\ O_2 \rightarrow CO_2 \qquad (3)$$

is strongly favored over hydrogen oxidation, so that CO is removed to the level of several ppm.

1.3.1.1 COMPARISON OF SMALL SCALE STEAM METHANE REFORMERS: "CONVENTIONAL " SMALL SCALE STEAM METHANE REFORMERS VS. FUEL CELL STEAM REFORMERS

In surveying the design of small methane steam reformers, we found that there were two general types:

* "conventional" steam methane reformers, which are commercially available today and are used in chemical applications

* fuel cell steam methane reformers, which are now under development and have just been commercialized for stand-alone hydrogen production.

DESIGN CONSIDERATIONS FOR "CONVENTIONAL" SMALL SCALE REFORMERS

Conventional designs are used to make high purity hydrogen for chemical applications (see Rostrup-Nielsen 1984 and Twigg 1989 for excellent descriptions of conventional steam methane reformer technology). They are essentially small versions of the systems used in refineries. Generally, the methane feed plus steam are introduced into high alloy nickel chromium tubes containing nickel based catalysts. In most designs, heat transfer is largely radiative, with tubes heated via radiation from a flame. Because the heat flux (and, thus, the reaction kinetics) are limited by the need for indirect heating, heat exchangers in steam reformers have relatively large surface areas, and can be quite bulky. (Typical tube lengths for commercially available reformers are 12 meters. See Figure 3.)

The most important design considerations for chemical hydrogen plants are long tube lifetime, long catalyst lifetime, reliability, and production at a particular set of pressure, temperature and purity requirements. Compactness is not generally an issue. These systems are custom designed, and, not surprisingly, at small size (less than 1 million scf H2/day), engineering costs are quite important. Because downstream hydrogen purification systems such as pressure swing adsorption (PSA) allow a higher recovery of hydrogen at high inlet pressure (several hundred psi), high pressure (15-25 atm) operation of the reformer is common, which means higher cost reformer tube materials such as high alloy steels must be used. Because of the thermodynamics of the steam reforming reaction, acheiving good conversion of methane to hydrogen at higher pressure requires higher temperature operation (c. 800-850^{o}C) (Figure 2). The tube lifetime depends sensitively on the temperature of operation, and the pressure (Rostrup-Nielsen 1984). Tube lifetime is adversely effected by tube temperature gradients ("hot spots"), and uniform tube heating is desirable.

COST OF "CONVENTIONAL" STEAM REFORMER SYSTEMS

Data on commercially available "conventional" steam methane reforming plants were gathered from several vendors (Howe-Baker, Hydrochem, Haldor-Topsoe, KTI/Air Products and Chemicals, Inc.). In discussions with vendors, we found that the minimum commercially available size is about 100,000 scf H_2/day. Below this size, plant capital costs decrease very little, and other sources of hydrogen (such as truck delivery or electrolysis) are usually less expensive (King 1993).

Manufacturers' installed capital cost data for steam reforming plants in the range 0.1 to 2.0 million scf H_2/day are shown in Figure 5 (costs include the reformer, shift reactor, PSA system, and NOx control). The capital cost of commercially available small scale "conventional" steam reformers ranges from about $1,500,000 for a plant producing 100,000 scf H_2/day to $3,000,000 for a plant producing 1 million scf/day. There are strong scale economies in this size range, with installed capital costs varying as

$$\text{Installed cost} = \$3 \times 10^6 \times (\text{plant capacity in million scf/day})^{0.3} \quad (4)$$

One of the main capital cost drivers for commercially available steam reformers is the reformer tubes. Because of the high temperatures and pressures and long lifetime required for chemical plant applications, tube materials are made of expensive alloy steels.

The costs of individual components in a conventional plant (e.g.the reformer vessel, the shift reactors, the PSA, compressors) have been estimated by Moore 1992 for systems producing 1.6, 16 and 160 million scf H2/day (Moore 1992).

To get a rough idea of the relative costs of various components in a small (1.6 million scf/day) conventional steam reformer plant, we show the contributions of the reformer vessel (43%), the shift reactors (7%) the PSA (39%), and compressors (11%) (Figure 10). The reformer vessel and the PSA are the dominant costs.

The energy efficiency (=hydrogen energy out/methane in) of small scale reformers is quite high, typically about 80-85%.

To achieve long equipment lifetime, good performance and low emissions, manufacturers indicated that the reformer should be run continuously, preferably at constant output. The reformer start-up time is 4 to 6 hours, and the response time to changes in load is relatively slow, so that following a rapidly varying hydrogen demand would be difficult. Small scale reformers are designed for automated operation, with occasional monitoring.

The only significant pollutants emitted from a steam reforming system would be nitrogen oxides (NOx). NOx can be controlled through a low NOx burner design or with selective catalytic reduction (SCR) to levels that meet stringent California air quality standards.

Performance and capital and O&M costs for a typical small scale conventional steam methane reformer are shown in Table 3.

Manufacturers of commercially available, "conventional" small scale reformers commented that these systems had not yet been optimized for cost or compactness. They agreed that it was likely that the cost might be brought down by standardizing the design. This has not occurred since conventional reformers are customed designed for particular chemical plants.

DESIGN CONSIDERATIONS FOR FUEL CELL REFORMERS

A number of groups are developing small scale natural gas reformers for fuel cell cogeneration systems in the 200-250 kWe range (see Table 2). (This corresponds to a reformer output of about 120,000 - 150,000 scf of hydrogen/day.)

Fuel cell reformers are designed for low cost and compactness. Acheiving these goals is largely a problem of designing the reformer vessel heat exchange system. A number of low cost designs have been proposed for small scale steam methane reformers, where the heat flow path is made more compact (Sederquist 1978, Stahl et.al. 1989, Ohsaki 1993, Buswell 1994, Sederquist 1995, Buswell 1996, Haldor-Topsoe 1988, Haldor-Topsoe 1993) Instead of radiative heat transfer, which requires large heat exchanger areas, fuel cell reformers rely on convective heat exchange. The hot gases travel through several catalyst beds to assure efficient heat transfer (Figure 4). The annular reformer consists of a series of concentric annuli rather than a set of parallel, long catalyst filled tubes. The design is standardized to reduce engineering costs.

A key point influencing the design of fuel cell reformers is that fuel cells [even proton exchange membrane (PEM) fuel cells] require lower purity hydrogen than many chemical applications. Fuel cell feedstream diluents such as CO_2, CH_4, H_2O, and N_2, which would be removed by the PSA in a conventional steam methane reformer, are acceptable input for the anode of a PEM fuel cell (although the fuel cell performance may suffer somewhat because of diluents). For use in PEM fuel cells, CO must be reduced to less than about 10 ppm. This can be accomplished in a preferential oxidation system. The downstream hydrogen purification equipment required is less for fuel cell reformers; only CO must be removed, after the last shift reactor stage.

Also, the operating pressure of a PEM or PAFC fuel cell is at most a few atmospheres. So low pressure reformers (1-5 atm) with a shift stage and minimal purification are acceptable for coupling to a fuel cell. Lower pressure operation means less expensive materials can be used for the reformer (such as stainless steel instead of alloy steels). The reformer can also operate at lower temperature, which enhances material lifetimes.

Some important performance and design features of conventional and fuel cell reformers are contrasted in Table 4.

In summary:

> * Conventional reformers produce hydrogen at relatively high pressure (so that downstream purification equipment such as PSAs will work well), and temperature (so that methane conversion to hydrogen will be high), and high purity. They are custom designed.
>
> * Fuel cell reformers produce hydrogen at relatively low pressure (and temperature) and low purity (only CO is removed), as required by fuel cells. They are a standardized design.

COST OF FUEL CELL REFORMERS

Current prices for phosphoric acid fuel cell cogeneration packages in the 200 kWe size range are about $3000/kWe. Fuel cell reformers contribute about one third of the total capital cost of a fuel cell cogeneration system (Prater 1994, Wheeler 1995). (The fuel cell stack and power conditioning equipment/electronics each contribute another third). So a fuel cell reformer today costs about $1000/kWe. In the near term, with modest increases in production volume (from a few per year today to hundreds per year), costs for PAFC or PEM fuel cell cogeneration systems in the 200 kWe range are projected to reach \$1500/kWe or \$500/kWe for the reformer.

A steam reformer for a 200 kWe fuel cell produces about 120,000 scf H2/day and costs about \$200,000 assuming \$1000/kW for the reformer contribution (or \$100,000 if the refomer costs \$500/kW). Using a scaling factor of 0.7 over the range 0.1-2.0 million scf H2/day, the capital cost for a fuel cell reformer is given by

$$\$0.88 \text{ million} \times \text{(capacity in million scf H2/day)}^{0.7}, \quad (5a)$$

for today's technology (\$1000/kWe for the reformer)

$$\$0.44 \text{ million} \times \text{(capacity in million scf H2/day)}^{0.7}, \quad (5b)$$

assuming production of hundreds of units per year (\$500/kWe for the reformer),

Comparing these costs to those for conventional reformers (Eq. 4, Fig 6), we see that fuel cell reformers appear to be considerably less costly. However, as discussed above, the pressure, temperature and purity requirements for conventional reformers are quite different than for fuel cell reformers. This is illustrated in Figures 7a and 7b, where a conventional small scale steam reformer and a fuel cell reformer are compared. The conventional reformer produces 99.999% pure hydrogen, whereas the fuel cell reformer produces a gas which is only about 70% hydrogen (or 77% if the water is removed).

Figure 6 is somewhat misleading, as the hydrogen purity is lower from the fuel cell reformer. To make the two systems comparable, the output of the fuel cell reformer would have to be purified to 99.999%. This could be accomplished with a PSA unit or membranes. To make either of these alternatives work well, an additional compression stage would be needed (Figure 7c) after the shift reactor. This is the approach taken by Praxair and IFC, which recently announced a new small scale hydrogen production system based on the reformer from IFC's 200 kW fuel cell coupled with a PSA purification system (Farris 1996).

The operation and maintenance costs of fuel cell reformers have not been quantified. However, a realistic goal for the O&M costs for an entire 200 kWe fuel cell cogeneration system has been estimated to be about 1 cent/kWh (Appleby 1993). If we assume that O&M costs for a fuel cell reformer hydrogen plant would be similar to those for a fuel cell cogeneration system, 1 cent/kWh would correspond to about $18,000/year for a hydrogen plant producing 100,000 scf H2/day. We assume that O&M costs scale linearly with system capacity.

We now investigate whether it might be possible to reduce the cost of small scale hydrogen systems by adopting some of the design features of fuel cell reformers. This involves a trade-off between lower cost, low pressure fuel cell reformers with more expensive PSA purification vs. higher pressure higher cost conventional reformers with less expensive purification.

1.3.1.2. POTENTIAL FOR LOW COST, SMALL SCALE HYDROGEN PRODUCTION USING FUEL CELL REFORMERS

In this section we make an approximate estimate of the capital cost of a small scale hydrogen plant with a fuel cell reformer plus compression and a PSA clean-up system.

DESIGN OF A SMALL SCALE HYDROGEN PLANT BASED ON A FUEL CELL REFORMER WITH COMPRESSION AND PSA

A simplified process diagram for a hydrogen plant based on a fuel cell reformer is shown in Figure 8. Corresponding material flows and conditions for the various streams are given in Table 5a. (For clarity, the various heat exchangers used to cool and heat streams are not fully drawn. Also, compression of natural gas from line pressure to 100 psig and sulfur removal are not shown.)

The fuel cell reformer design is based on a recent patent by Ballard Power Systems, for a steam methane reformer coupled to a 200 kWe PEM fuel cell (Buswell 1994). Material flows and temperature and pressure conditions for the reformer, shift reactor and heat recovery boiler are from Ballard's patent, but instead of integrating it with a PEM fuel cell, we have added a single stage compressor and a PSA unit.

We found that integrating the Ballard reformer with a PSA instead a fuel cell required only minor re-routing of process streams. The main difference is the fuel supply for the reformer burner. In the original Ballard patent, all the fuel for the reformer burner was supplied from the fuel cell anode exhaust. In our design, the PSA purge gas is used for reformer burner fuel. The PSA recovery is chosen to be 72%, so that the reformer purge gas has enough energy to supply all the fuel requirements for the reformer burner.

For comparison, a process flow diagram for a conventional small scale reformer is shown in Figure 9 and Table 5b. The material flows and conditions are based on simulations by Katofsky (Katofsky1993).

Comparing Tables 5a and 5b, we see that the conventional reformer operates at higher pressure and temperature than the fuel cell reformer, and does not require a compressor prior to the PSA. Also, the conventional system uses some of the natural gas feed as well as the PSA purge gas for reformer burner fuel.

We now estimate the cost of the individual components making up the hydrogen plant: the fuel cell reformer vessel, the shift reactors, the compressor and the PSA.

CAPITAL COST OF A FUEL CELL STEAM REFORMER VESSEL

As discussed above the reformer cost depends on the type of heat transfer, and the operating pressure and temperature. We found a number of reasons why fuel cell reformers are less expensive than today's conventional small scale steam reformers.

> *Convective heat transfer (rather than radiative) => less heat exchanger area
>
> * Lower pressure operation => lower cost steels for reformer tubes (stainless steel can be used instead of alloy steels)
>
> * Standardized design for fuel cell reformers vs. specialized conditions for chemical applications => lower engineering costs. (It is possible that standardized design might bring down the cost of conventional reformers, as well, if a large enough market existed for a particular small scale model.)

The capital cost of the reformer plus shift reactors is taken to be

$$\$0.44 \text{ million} \times (\text{capacity in million scf H2/day})^{0.7} \quad (6)$$

This assumes that fuel cell reformers are produced in quantities of hundreds per year and have costs of \$500/kWe. This estimate covers the cost of the reformer, shift reactors, heat exchangers, water treatment and heat recovery boiler shown in Figure 8, but not the compressor or PSA.

CAPITAL COST OF SMALL SCALE PSA UNITS

Pressure swing adsorption is used to purify hydrogen from the fuel cell refomer. The cost for small scale PSA units is shown in Figure 11 as a function of plant size, based on conversations with UOP, a supplier of PSA systems (Ernst Ecker UOP 1996). Comparing the cost of the PSA to the total plant cost, we see that the capital cost of the PSA is roughly 30-40% of the total capital cost of a conventional hydrogen plant in the range 0.1-4 million scf H2/day. The capital cost for a first of a kind PSA system capable of removing CO, CO_2, CH_4, and H_2O would be about

\$500,000,
for PSA inlet total gas flow rates of 1 million scf/day or less

$\$500,000 \times (\text{gas flow rate in million scf/day})^{0.7}$,
for PSA inlet gas flow rates > 1 million scf/day (7)

The PSA inlet gas flow rate is related to the hydrogen flow rate based on the composition of the gas leaving the shift reactor, and the hydrogen recovery in the PSA system. For example, for a gas having 77% hydrogen by volume at the PSA inlet, and a PSA hydrogen recovery of 72%,

the PSA inlet gas flow rate = H2 flow rate out of PSA/(0.77 x 0.72)

The capital cost of the PSA would be about

\$500,000,
for hydrogen production rates < 0.55 million scf/day

and

$\$500,000 / (0.55)^{0.7} \times (\text{hydrogen flow rate in million scf/day})^{0.7}$,
for hydrogen production rates > 0.55 million scf/day (7)

The PSA capital cost might be reduced by 50%, if the engineering costs could be spread over a number of units (say 10 or more).

Although this is not an issue for steam methane reformers, if nitrogen removal were required (as in some partial oxidation systems), the cost of the PSA would be approximately doubled, because more expensive adsorbent materials are needed.

CAPITAL COST OF SMALL SCALE HYDROGEN COMPRESSORS

After the last shift reactor stage, the hydrogen-rich gas must be compressed prior to the PSA. Because the gas is about 77% hydrogen, we have assumed that a hydrogen compressor is used. The cost of small scale hydrogen compressors has been estimated by various authors. We have used a figure of $2000/kW (where kW refers to the electrical input to the compressor) for multi-stage compressors and $1000/kW for single stage compressors (see Ogden et.al 1995 and Dennis 1994). It is assumed that there is no scale economy for compressors in this size range. The cost of compression can be estimated based on the power input requirement. This is

$$P_{cm} = Q_{MMscfd} \text{ x } 114/n_C \text{ x } N \text{ x } [(P_2/P_1)^{0.291/N} - 1]/(n_{PSA} \text{ x } f_{H2}) \quad (9)$$

where:

P_{cm} = compressor power requirement (kW)
Q_{MMscfd} = hydrogen flow rate in million scf/day
P_1=inlet pressure
P_2=outlet pressure
N = number of compressor stages
n_C = compressor efficiency = 55%
n_{PSA} = H2 recovery in the PSA = 72%
f_{H2} = fraction H2 by volume at the PSA inlet = 77%

The compressor capital cost is then (assuming that each stage has a compression ratio of 4.):

Capital cost for a single stage compressor =
$1000/kW x Pcm = $184,000 x (hydrogen plant capacity in million scf/day) (10)

Capital cost for a multi-stage compressor =

$$\$2000/\text{kW} \times \text{Pcm} = \$368{,}000 \times N \times (\text{hydrogen plant capacity in million scf/day}) \qquad (11)$$

Fuel cell reformers operate at anywhere from 1 atm to 5 atm pressure. To get a high recovery of hydrogen (80%) in the PSA, we assume that the PSA inlet pressure is 250 psia (about 17 atm). Pre-PSA compressor power requirements and costs are quite sensitive to the reformer operating pressure. If the reformer operates at 5 atm (as shown in some fuel cell reformer designs e.g. Stahl 1993, Buswell 1994) it may be possible to use a single stage compressor. If the reformer operates at atmospheric pressure, 2 stages will be needed. With two stages, the compressor power requirement is about doubled, and the capital cost is quadrupled.

It is important to note that hydrogen compressors are not mass-produced items at present, and that the costs might be reduced (Thomas 1996).

APPROXIMATE ESTIMATE OF THE CAPITAL COST OF A SMALL SCALE HYDROGEN PLANT WITH A FUEL CELL STEAM REFORMER AND A PSA

Combining the estimates from the preceding sections (Eq. 6,7,10,11), we calculate that the total capital cost of a fuel cell reformer with 1-stage compression and a PSA would be about:

$$\$1 \text{ million} \times \left[\underbrace{(0.44+0.375) \times (\text{million scf H2/day})^{0.7}}_{\text{reformer + PSA}} + \underbrace{0.184 \times (\text{million scf H2/day})}_{\text{+ 1-stage compressor}}\right], \qquad (12)$$

for a hydrogen flow rate of more than 0.55 million scf/day

and

$$\$1 \text{ million} \times [(\underbrace{0.44 \times (\text{million scf H2/day})^{0.7}}_{\text{reformer}} \underbrace{+ 0.25}_{\text{+ PSA}} \underbrace{+ 0.184 \times (\text{million scf H2/day})}_{\text{+ 1-stage compressor}}],$$

for a hydrogen flow rate of less than 0.55 million scf/day

For a plant producing 1 million scf H2/day, the capital cost would be about $1 million, one third that of a conventional SMR plant, which would cost about $3 million. [This assumes that several tens to hundreds of units are built so that the cost of fuel cell reformer is

\$500/kWe and the cost of the PSA is \$0.25 million x (gas flow rate in million scf/day)$^{0.7}$, for gas flow rates > 1 million scf/day]

If a 2 stage compressor were needed, the total capital cost would be

$$\$1\text{ million} \times \Big[\underbrace{(0.44+0.375)\times(\text{million scf H2/day})^{0.7}}_{\text{reformer+PSA}} + \underbrace{0.736\times(\text{million scf H2/day})}_{\text{+ 2-stage compressor}}\Big], \quad (13)$$

for a hydrogen flow rate of more than 0.55 million scf/day

and

$$\$1\text{ million} \times \Big[\underbrace{(0.44\times(\text{million scf H2/day})^{0.7}}_{\text{reformer}} \underbrace{+\ 0.25\ +}_{\text{+ PSA +}} \underbrace{0.736\times(\text{million scf H2/day})}_{\text{2-stage compressor}}\Big],$$

for a hydrogen flow rate of less than 0.55 million scf/day

For a plant producing 1 million scf H2/day, the capital cost would be about \$1.6 million, about half that of a conventional SMR plant.

Figure 12 compares the cost of fuel cell reformer with 1-stage compression and PSA cleanup versus a conventional SMR system vs. hydrogen production capacity. The capital cost for a fuel cell reformer based system is about 30-75% that of a comparable conventional SMR system over the range 0.1-2.0 million scf H2/day.

1.3.1.3. EFFECT OF SMALL SCALE STEAM REFORMER COST REDUCTIONS ON THE DELIVERED COST OF HYDROGEN TRANSPORTATION FUEL

One possible application is onsite production of hydrogen transportation fuel. A hydrogen refueling station based on onsite reforming of natural gas is shown in Figure 13a. Figure 13b shows the potential impact of reducing the small scale reformer capital cost on the delivered cost of hydrogen transportation fuel. Assuming that the reformer capital cost followed the curve shown in Figure 12, the delivered cost of hydrogen transportation fuel could be significantly reduced, especially at small station size. For example, the delivered cost of hydrogen would be reduced from about \$40/GJ to about \$25/GJ for a hydrogen refueling station dispensing 100,000 scf/day, (enough for perhaps 8 Ballard PEMFC buses/day or 80 fuel cell automobiles/day). The capital and operating costs and delivered cost of hydrogen transportation fuel are given in Tables 6 and 7 for hydrogen refueling stations using fuel cell and conventional steam methane reformers.

1.3.2. PARTIAL OXIDATION

Small scale partial oxidation systems have recently become commercially available, but are still undergoing intensive R&D (Moard 1995, Loftus 1994, Mitchell et.al. 1995). In partial oxidation reforming, methane is oxidized to produce carbon monoxide and hydrogen according to

$$CH_4 + 1/2\ O_2 \ -> \ CO + 2\ H_2\ , \qquad \Delta h^o = -36 \text{ MJ/kmol } CH_4$$

The reaction is exothermic and no indirect heat exchanger is needed. No catalyst is required because of the high temperature. However, the hydrogen yield per mole of methane input (and the system efficiency) can be significantly enhanced by use of catalysts (Loftus 1994). The reactor is more compact than a steam reformer, where heat must be added indirectly via a heat exchanger. The efficiency of the partial oxidation unit is relatively high (70-80%). However, partial oxidation systems are typically less energy efficient than steam reforming because of the higher temperatures involved (which exacerbates heat losses) and the problem of heat recovery. (In a steam methane reforming plant, heat can be recovered from the flue gas to raise steam for the reaction and the PSA purge gas can be used as a reformer burner fuel to help provide heat for the endothermic steam reforming reaction. In a POX reactor where the reaction is exothermic, the energy in the PSA purge gas cannot be as fully recovered.) A hydrogen plant based on partial oxidation would include a shift reactor and hydrogen purification equipment. Partial oxidation systems have a fast response time, making them attractive for following rapidly varying loads, and can handle a variety of fuels.

Partial oxidation systems are under development by Arthur D. Little (ADL 1994, Loftus 1994, Mitchell et.al. 1995) and by Hydrogen Burner Technology (Moard 1995).

Because they are more compact, and do not require indirect heat exchange (as in steam reforming), it has been suggested that partial oxidation systems could be lower cost than steam reformers.

Hydrogen Burner Technology has estimated capital costs for a range of stationary hydrogen production systems, which include a POX, shift reactor and PSA (Moard 1995) The costs for POX hydrogen plants are plotted vs. plant size, and contrasted with the costs of a hydrogen production plant based on fuel cell steam reformers and conventional small scale steam reformers (Figure 14). The capital costs for POX-based hydrogen plants in the 0.1-0.5 million scf/day range are less than those for a conventional SMR system, but similar to those for a fuel cell SMR system. At larger plant capacities, the POX capital costs appear to be slightly lower.

Another cost estimate of small scale POX systems has been carried out by Mitchell (Mitchell 1996) for a small catalytic POX unit designed for coupling to a 50 kWe fuel cell. (Assuming this POX unit was coupled to a PSA with an 80% recovery, it would produce perhaps 25,000 scf H2/day.) The POX plus shift reactors are estimated to cost \$15-25/kWe or \$750-1250 in mass production. This does not include a PROX unit or the PSA clean-up, but does include a steam generator. Assuming a cost scale factor of 0.7 and extrapolating from 25,000 to 1,000,000 scf H2/day, the cost of the POX vessel plus shift reactors and steam generator would be about

$$\$15{,}500 \times (\text{H2 production capacity in million scf H2/day})^{0.7} \quad (14)$$

This is much lower than the projected costs for a steam methane reformer. The POX reactor vessel is only a minor contributor to the total cost of the hydrogen plant.

Although the POX reactor by itself is quite inexpensive, the gas must still be cleaned up to 99.999% purity. It turns out that the cost of the PSA unit is an important issue for POX systems. In a POX system, where the reaction takes place in air, nitrogen is a major diluent. In Hydrogen Burner Technology's POX system, N_2 is perhaps 47% of the input to the PSA. This drives up PSA costs in two ways. First, the adsorbent materials needed to separate nitrogen from hydrogen are about twice as expensive as those used to separate CO_2, CO, CH_4, and H_2O. Second, since hydrogen is only about 45% of the PSA inlet gas (as compared to 77% for a SMR system), the PSA must be larger to accomodate the larger total gas flow. One would expect the PSA for a POX based hydrogen production system to cost several times more than a PSA for a SMR based system (Figure 15). Our very rough estimate for the cost of a mass produced PSA needed for a POX is comparable to the total system cost estimated by Hydrogen Burner Technology in Figure 14. This is consistent with the ADL estimate, where the POX and shift reactors account for only a few percent of the total capital cost of the system.

Developing low cost purification technologies is key if POX systems are to be used for stationary hydrogen production. Another approach is using pure oxygen feed to the POX, which incurs high capital costs for small scale oxygen production, but eliminates the need to deal with nitrogen downstream. Oxygen enrichment of incoming air is another way of reducing, but not eliminating the amount of nitrogen.

1.3.3. AUTOTHERMAL REFORMING

In autothermal reformers, some fuel undergoes partial oxidation, which supplies heat to drive a catalytic steam reforming reaction. The autothermal reformer requires no external heat source and no indirect heat exchangers. This system has a higher efficiency than partial oxidation, and is only slightly more bulky. Autothermal reforming systems are not commercially available at present, but are now being developed by Rolls Royce, (Shoesmith 1992) International Fuel Cells, and Chiyoda. Because they are more compact and less costly, this approach may eventually offer lower cost than steam reforming.

1.3.4. OTHER APPROACHES TO SMALL SCALE REFORMING

1.3.4.1. SORBENT ENHANCED STEAM REFORMING

Recently several authors have investigated the possibility of sorbent enhanced steam methane reforming (Lyon 1996, Sircar 1996, Han and Harrison 1994). Here an absorbent (such as calcium oxide) is mixed with the steam reforming catalyst, removing the CO and CO_2 as the steam reforming reaction progresses. The resulting syngas has a substantially higher fraction of hydrogen than that produced in a catalytic steam reforming system. This reduces the need for downstream processing and purification, which can be expensive in a small scale steam reformer. Moreover, the reaction can take place at lower temperature and pressure, reducing heat losses and material costs. These systems are still at the laboratory stage, and show promise for low cost. Issues include catalyst and sorbent lifetime and system design.

1.3.4.2. THERMOCATALYTIC CRACKING

In this approach, methane is broken down into carbon and hydrogen at high temperature (Muradov 1996). The primary issues here are low energy efficiency of conversion and coking.

1.4. CRITICAL FACTORS LIMITING THE COST AND PERFORMANCE OF SMALL SCALE HYDROGEN PRODUCTION SYSTEMS

What factors currently limit the cost and performance of small scale systems producing hydrogen from methane?

1.4.1. FACTORS LIMITING THE COST

The two most costly components of a "conventional" small steam methane reformer plant are the reformer vessel and the PSA purification system (Figure 10).

Factors Limiting the Cost of the Steam Reformer Vessel

One of the main cost contributors for a conventional steam reformer vessel is the cost of the reformer tubes, which dominates the cost of the catalyst. Several strategies were identified for reducing the cost of the reformer tubes: operation at lower pressure and temperature (allowing the use of less expensive reformer tube materials) and use of convective rather than radiative heat transfer (reducing the heat exchanger area required). Standardization of design could also reduce the cost of the reformer vessel by reducing high engineering costs for one of a kind projects.

With lower pressure operation, better heat exchanger design and standardization, it appears that the capital cost of the reformer vessel could be reduced by a factor of 3 to 7, depending on the reformer size.

System Trade-offs: The Effect of Steam Reformer Pressure on System Cost

It is clear that lower pressure operation can reduce the cost of the steam reformer vessel. However, when the entire hydrogen production system is considered, we found that lower pressure reformer vessel operation also meant increasing the cost of downstream processing and purification equipment.

A "conventional" steam reformer vessel operates at high pressure and temperature (requiring expensive alloy steels), and produces syngas at pressure, which means that no downstream compression is needed prior to purification in a PSA (Figure 7a).

A fuel cell type steam methane reformer operates at lower pressure and temperature and can use less expensive materials (stainless steels), saving on reformer costs. However, the syngas requires compression prior to the PSA (Figure 7c).

Finding the lowest cost system involves a trade-off between using a lower pressure, lower cost reformer vessel with compression and PSA vs. a higher cost, higher pressure reformer vessel with no compression prior to the PSA. Our results suggest that it may be worth paying the price in extra compression in order to reduce reformer vessel costs.

Comparison of POX and Steam Reforming

Unlike a steam reformer, a partial oxidation system does not require indirect heat transfer. So the POX reaction vessel is much simpler and less expensive. Instead of catalyst filled tubes, the system, consists of a reactor, possibly with a catalytic section (Loftus 1994). Moreover, a POX system can be operated at pressures suitable for use with a PSA (Moard 1995). The capital cost of the POX reactor will be only a small fraction of that for a steam reformer vessel.

However, the fraction of hydrogen in the syngas is much lower for air-fired POX reactors than for steam reformers. Moreover, significant fractions of nitrogen are present in the syngas. So the clean-up system for a POX is more expensive, both because a larger volume of gas must be processed in the PSA and because nitrogen is more costly to separate from hydrogen than CO_2 and H_2O. It appears that the overall system cost for stationary hydrogen production may be comparable for mass produced POX and fuel cell steam reformers.

Importance of Small Scale Purification Technologies

Unless the purification and reforming can somehow be done in one vessel (as proposed in sorbent-enhanced reforming), it may be difficult to get away from a trade-off between reformer vessel cost and purification system cost. Clearly, development of low cost ways of purifying hydrogen at small scale could reduce the cost of hydrogen, especially for POX.

Manufacturers of PSA systems indicated that standardized designs for PSA systems could reduce the cost by roughly a factor of two (Ecker 1996).

Alternatives to PSA exist, but most are suitable for lower purity hydrogen production. For hydrogen of 99.999% purity, only palladium membranes can compete. Like PSA systems, membranes generally work better with a pressure drop of several hundred psi. Moreover, the cost of palladium membranes is still quite high. In the size range 0.1-2.0 million scf H2/day, PSA would be preferable.

Standardization of Design

Another important cost issue is standardization of the design. Fuel cell manufacturers have addressed this question, in estimating the cost of small scale steam methane reformers in mass production. It appears that the cost of the reformer part of a 200 kWe fuel cell cogeneration system could be brought down by about a factor of two: from $1000/kWe today to $500/kWe if tens to hundreds were manufactured each year. Materials costs for a PAFC fuel cell cogeneration system are estimated to be about $600/kWe (Appleby). Assuming that the reformer accounts for about 1/3 of this total, the materials cost for the reformer might be about $200/kWe or $40,000 for a system producing 120,000 scf H2/day.

Manufacturers of small scale "conventional" steam reformers have not addressed this issue, since they have not been focussed on potential markets requiring standardization, but rather on individually engineered chemical applications. Even in mass production, it seems likely that the cost of "conventional" reformer vessels would be higher than those for fuel cell type reformers, because of the more expensive tube materials needed, and the greater heat exchanger areas required.

Similarly, PSA systems and hydrogen compressors are not currently mass produced items. Costs might be reduced for these components as well.

1.4.2. FACTORS LIMITING THE PERFORMANCE OF SMALL SCALE REFORMERS

For small scale stationary hydrogen production, concerns about response time and transients are not so important as for vehicles. So we concentrate on conversion efficiency of methane to hydrogen as a measure of performance.

Limits on Conversion Efficiency of Methane to Hydrogen

Both conventional and fuel cell type steam reformer systems offer high energy conversion efficiency, on the order of 80-85% [where energy conversion efficiency = H2 out (HHV)/CH4 in (HHV)]. This is close to the limits possible, according to pinch theory (Katofsky 1993). These efficiencies are realized via extensive heat integration (Katofsky 1993, Buswell et.al. 1994). In particular the energy in the PSA purge gas can be recovered for use as reformer burner fuel, and reformer flue gases can provide heat to raise steam needed for the reaction.

In a POX system, effective heat integration would appear to be more difficult, since the reaction is exothermic, and there is not as much demand for heat from flue gases or energy from the PSA purge gases (which have lower heating value because of large amounts of nitrogen present). However, modelling by researchers at Arthur D. Little (ADL 1994) indicated that the total system efficiency for POX based systems coupled to a fuel cell were only a few percentage points lower than those for steam reforming. In each case the fuel cell anode exhaust gas (which is roughly analogous to the PSA purge gas in a stationary hydrogen production system) was utilized (for reformer furnace fuel in the steam reformer system, for fuel preheating in the POX system and for driving an air compressor in the both systems). In a stationary POX system it would probably be possible to utilize the PSA purge gas for fuel preheating and air compression.

Our simulations suggest that it should be possible to convert about 70-80% of the energy in methane to hydrogen plus CO in the POX reactor, without running into sooting limits. Assuming that all the CO is converted to hydrogen in shift reactors, the question is how much hydrogen can be recovered in the PSA. For small PSA units, this is about 80%. The purge gas of the PSA contains the remaining 20% of the hydrogen, plus CO2, CH4 and N2. Some of the energy in the purge gas can be recovered by burning it and using the combustion products for preheating methane and running an air compressor.

Actual energy conversion efficiencies for Hydrogen Burner's stationary POX system coupled to a PSA are estimated at about 55% (on a HHV basis). (These estimates do not include recovery of purge gas energy). Adapting ADL's POX system designs (ADL 1994) to stationary hydrogen production might give an energy conversion efficiency of up to 65-75%, depending on the PSA hydrogen recovery, and how much of PSA purge gas energy could be recouped for preheating and parasitic loads such as air compression.

The main factor limiting the energy conversion efficiency of small scale reformer systems is the efficacy of heat integration at small scale, especially for recovering energy in PSA purge gases.

The compactness of small scale reformers, will help limit heat losses. These may actually be less important in a compact, annular small steam reformer with convective heat transfer than for a conventional large scale plant with radiative heating of long tubes.

1.5. CONCLUSIONS

We surveyed existing small scale reformer technologies suitable for stand-alone hydrogen production of 0.1-2.0 million scf H2/day.

> * Two types of small scale steam methane reformers exist: "conventional" steam methane reformers, which are well established, commercially available technology and are widely used in chemical applications; and "fuel cell type" steam reformers, which are being developed as part of fuel cell cogeneration packages, and have just become commercially available as part of a system for stand-alone hydrogen production.
>
> * Conventional steam methane reformers produce hydrogen at relatively high pressure (so that downstream purification equipment such as PSAs will work well), and temperature (so that methane conversion to hydrogen will be high), and high

purity (99.999% hydrogen). Heat transfer is accomplished via radiative heating of long catalyst filled tubes. They are custom designed.

* The installed capital cost for a conventional steam methane reformer hydrogen plant exhibits strong scale economies in the size range of interest, and is given by:

Installed cost = $\$3 \times 10^6 \times (\text{H2 plant capacity in million scf/day})^{0.3}$

The cost of conventional SMR plants might be reduced if the design were standardized.

* The main capital cost drivers for small scale conventional SMR hydrogen plants are the reformer vessel (43%), and the PSA (39%). Compressors account for 11% and shift reactors 7%. Engineering costs are significant for small, one of a kind conventional SMR plants.

* The majority of the cost for a conventional reformer vessel is for the reformer tubes, which are made of high alloy steels to withstand the pressures and temperatures in the reformer.

* Fuel cell steam methane reformers produce hydrogen at relatively low pressure (and temperature) and low purity (only CO is removed), as required by fuel cells. Heat transfer is accomplished in a compact annular design via convective heat transfer. They are a standardized design.

* Fuel cell steam methane reformers offer several design features which could reduce the capital costs as compared to "conventional" small scale steam reforming: use of convective rather than radiative heat transfer which reduces heat exchanger area; lower pressure operation, which reduces the cost of reformer steels (stainless steel can be used instead of high alloy steels; and standardized design, which reduces engineering costs.

* The approximate cost of fuel cell reformer vessel plus shift reactors is about

$\$0.44 \text{ million} \times (\text{capacity in million scf H2/day})^{0.7}$

assuming mass production of several hundred units/year. This does not include the cost of downstream compression and purification in a PSA.

* The costs of downstream purification are significant, but even when compression and PSA are added to a fuel cell reformer, adapting fuel cell steam methane reformers for stationary hydrogen production looks promising. The capital cost of small scale steam reformer systems might be reduced by 25-70%. as compared to conventional small scale SMR systems. Some of the savings would be achieved via standardized design, and some via lower materials costs in the reformer vessel because of lower pressure.

* Operating pressure of the steam reformer is an important issue. At higher pressure (15-25 atm), the cost of the reformer materials is higher (alloys rather than stainless steel must be used for tubes), but there is no need for costly dowstream compression prior to the PSA. There are system trade-offs between lower pressure, less costly reformers and more costly downstream clean-up and compression equipment. Even among fairly low pressure reformers, reforming at 5 atm rather than 1 atm appeared to offer advantages in terms of reduced compression power requirements and costs.

* By using lower cost fuel cell reformers, the delivered cost of hydrogen transportation fuel produced via onsite reforming might be significantly reduced at small station size (less than a million scf H2/day). For a hydrogen refueling station dispensing 0.1 million scf H2/day, the delivered cost of hydrogen transportation fuel would be reduced by about 40% from $40/GJ (for a conventional SMR) to $25/GJ (for a fuel cell type SMR). For a hydrogen refueling station dispensing 1 million scf H2/day, the delivered cost of hydrogen would be reduced by about 20% from $14.2/GJ (for a conventional SMR) to $11.5/GJ (for a fuel cell type SMR). These costs are competitive with pipeline delivered hydrogen from centralized production via SMR or biomass gasification. With low cost fuel cell reformers, the decentralized production of hydrogen from natural gas becomes approximately competitive with centralized production plus distribution.

* Hydrogen purification technologies at small scale are an important part of the cost of a small scale reformer system (especially for POX). According to our preliminary estimates the cost of the PSA may account for a large majority of the system capital cost for a POX system, and about 1/3-1/2 of that for a small scale steam reformer.

* We estimate the capital cost for a small scale POX plant producing hydrogen from methane would be comparable to that of a fuel cell steam reformer plant. The POX reactor vessel would be quite inexpensive compared to a steam reformer vessel, but the downstream clean-up and need to process nitrogen for an air fired POX would add considerably to the cost. The energy conversion efficiency [=hydrogen out (HHV)/methane in (HHV)] for a POX-based hydrogen production system would probably be about 65-75% somewhat less than that for a small steam methane reformer which might have an efficiency of 80-85%..

* Key issues for reducing the cost of small scale hydrogen production systems are:
1) lower pressure reformer operation, which gives a lower reformer vessel capital cost because lower cost steels can be used, and also appears to give a lower overall system cost
2) standardization of the reformer design (to reduce engineering costs)
3) use of compact, convective heat transfer reformer vessel design, as in fuel cell reformers to reduce the heat exchanger area needed.
4) Reduction of the cost of hydrogen purification systems (PSA). Some of this could be achieved through mass production.
5) Pursuit of innovative concepts such as sorbent enhanced reforming which might get around some of the trade-offs between the cost of the reformer vessel and the cost of the downstream purification.
6) For POX-based small scale hydrogen production systems, the POX vessel alone will be almost negligible in cost. But the relatively low hydrogen content and high nitrogen content of the syngas will increase the cost of downstream processing and purification equipment. Bringing these costs down is crucial. Enhancing the oxygen content into the POX may help, iif this can be done inexpensively.

1.6. RECOMMENDATIONS FOR FUTURE WORK

We recommend that R&D on small scale hydrogen production systems based on fuel cell reformers be undertaken in parallel with R&D on fuel cell vehicles.

The leading existing candidates for small scale hydrogen production (steam reforming and POX) exhibit system trade-offs between the cost of the "front end" (e.g. the reformer vessel or POX reactor) and the cost of the downstream processing and clean-up equipment. Unfortunately, less expensive reactors seem to require more expensive clean-up systems. The net change in total system cost (reactor + clean-up) is unclear. Further conceptual designs of small reformer plants might illuminate the lowest cost configurations.

Approaches such as sorbent-enhanced reforming may get around the trade-off problem altogether by accomplishing the reforming and purification in one step. We recommend further R&D on this idea.

Small scale purification systems are an important cost driver for small scale reformer systems. These systems deserve further study. Innovative concepts for less costly hydrogen purification should be pursued.

Heat integration for small reformer systems for stationary hydrogen production should be looked at in more detail. It would be interesting to examine how reformer pressure effects hydrogen production system cost and design .

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APPENDIX 1.A.
REVIEW OF SELECTED PATENTS RELATED TO SMALL SCALE STEAM METHANE REFORMER DESIGN

Steam methane reforming is a well known, commercial process. A number of reviews of the design of large scale steam reformers have been written (e.g. Rostrup-Nielsen 1984, Twigg 1989). However, there is a scarcity of material published on the design of small scale steam methane reformers. Some chemical engineering handbooks contained estimates for reformer cost and performance, but the data were for reformers larger than the 0.1-2.0 million scf/day range of interest for fuel cell cogeneration systems and vehicle refueling stations.

Although manufacturers and developers of small reformers were helpful in discussions of system performance and cost, and in providing general guidance about the design, it proved difficult to obtain technical details about what was inside the "black box". We found patents to be an excellent resource for probing more deeply into the intricacies of small scale reformer design.

Using LEXIS-NEXIS, we searched the patent literature, over the past few decades. Most patents on small scale reformers are held by a few companies (International Fuel Cells and its parent company United Technologies Corporation, Ballard Power Systems, Haldor-Topsoe). There are several Japanese companies, which have patented small scale reformer designs, as well. Recent patented research on small scale reformers tends to be related to fuel cell applications rather than conventional chemical hydrogen production.

We found the following patents to be particularly useful:

U.S. Patent No. 4,071,330 (Jan. 31, 1978)
STEAM REFORMING PROCESS AND APPARATUS THEREFOR
Inventor: Richard Sederquist

U.S. Patent No. 4,098,588 (Jul. 4, 1978)
MULTI-TUBE CATALYTIC REACTION APPARATUS
Inventors: Richard Buswell, Richard Sederquist, Daniel Snopkowski

U.S. Patent No. 4,830,834 (May 16, 1989)

REACTOR FOR THE CATALYTIC REFORMING OF HYDROCARBONS
Inventors: Henrik Stahl, Jens R. Rostrup-Nielsen

U.S. Patent No. 5,360,679 (Nov. 1, 1994)
HYDROCARBON FUELED SOLID POLYMER FUEL CELL ELECTRIC POWER SYSTEM
Inventors: Richard Buswell, Joseph Clausi, Ronald Cohen, Craig Louie, David Watkins

U.S. Patent No. 5,470,360 (Nov. 28, 1995)
FUEL CELL POWER PLANT REFORMER BURNER GAS FLOW CONTROL SYSTEM
Inventor: Richard Sederquist

U.S. Patent No. 5,484,577 (Jan. 16, 1996)
CATALYTIC HYDROCARBON REFORMER WITH ENHANCED INTERNAL HEAT TRANSFER MECHANISM
Inventors: Richard Buswell, Ronald Cohen, Joseph Clausi, Stanley Leavitt, David Watkins

These are briefly reviewed below.

U.S. Patent No. 4,071,330
Jan. 31, 1978

STEAM REFORMING PROCESS AND APPARATUS THEREFOR

Inventor: Richard Sederquist
Assignee: United Technologies Corp.
(parent company of International Fuel Cells)

This patent was referenced in most other reformer patents as "typical" of small scale tubular reformers.

Background of invention

In steam reforming a mixture of steam and hydrocarbon fuel is passed through a vessel containing a steam reforming catalyst. The vessel is a filled tube inside a furnace. The furnace provides the heat for the endothermic reaction. It is possible to convert all of the hydrocarbons to H2, CO and CO2. Disadvantages are that to acheive very high conversion rates you need 1) high furnace temperatures. 2) larger or more reactors, 3) burning additional fuel in the furnace.

Summary of Invention

This invention is an improved, compact, efficient steam reformer capable on converting all the hydrocarbons to CH4, CO, CO2, and H2.

A tubular catalytic reactor is contained inside a furnace, which is divided into an enhanced heat transfer section and a burner cavity. Annular first bed of catalyst outside a coaxial annual second bed of catalyst. The product gases from the first bed, without adding heat are passed through a second bed essentially adiabatically, resulting in a reduction in the amount of unreacted higher hydrocarbons. The catalysts are nickel based, the tubes are nickel steel. Single or multiple tubes are used. The product gases are used to impart heat to the first bed.

U.S. Patent No. 4,098,588
Jul. 4, 1978

MULTI-TUBE CATALYTIC REACTION APPARATUS

Inventors: Richard Buswell, Richard Sederquist, Daniel Snopkowski
Assignee: United Technologies Corp.
(parent company of International Fuel Cells)

This patent is a variation on the one above (Sederquist 1978). The object of the invention is to have high reactor efficiency, compact apparatus, and capability to operate at high heating rates.

Background

Typical commercial reformers use radiant heating from furnace walls, and are designed for high product yield, rather than compactness. Typical tubes are 20-40 feet long and the primary mode of heat transfer is radiant heating, This necessitates the tubes have wide spacing and that they be placed close to the wall to assure a uniform heating from the walls. The rate of heat transfer is very high 20,000-25,000 BTU/h/ft^2 of reaction tube surface area. But reactor thermal efficiency is only 40-60%. Large amounts of waste heat were lost, and the apparatus was large and bulky.

Summary of invention

The purpose of this invention is to make steam reformers more compact and efficient. A number of tubes are evenly spaced around in the reactor. There is a burner cavity and an enhanced heat recovery zone where heat transfer material is located around the tubes. This helps assure a uniform heating of all the tubes, which is. normally a problem for multiple tubes in a reactor. The tubes closer to the wall will receive more radiantly heating. In this invention, about 2/3 of the furnace volume is filled with a heat transfer packing material, which keeps the tubes at uniform temperature.. The packing material can be anything which can withstand the temperatures of the furnace environment (ceramics or metals). This allows the tubes to be more closely spaced. There is one catalyst bed, which is heated by furnace and by the heat transfer material and by using heat from the product gases, which is carried back by the catalyst bed.

U.S. Patent No. 4,830,834
May 16, 1989

REACTOR FOR THE CATALYTIC REFORMING OF HYDROCARBONS

Inventors: Henrik Stahl, Jens R. Rostrup-Nielsen
Assignee: Electric Power Research Institute

This patent covers work by engineers at Haldor-Topsoe in developing a compact steam methane reformer for coupling to a phosphoric acid fuel cell cogeneration system. This project was supported by EPRI in the late 1980s.

Background

In some previous reformer designs, heat was recovered from the product stream. This invention extends this concept.

Summary of invention

The purpose of this invention is to utilize heat from the product stream of the reformed gas to supply part of the heat needed for the endothermic steam methane reforming reaction. The reformer is arranged in a series of concentric tubes. The heat transfer mechanism is primarily convective rather than radiative, which reduces the surface area needed for the tubes. The catalyst is arranged in two beds, which allows the system to be more compact.

U.S. Patent No. 5,360,679
Nov. 1, 1994

HYDROCARBON FUELED SOLID POLYMER FUEL CELL ELECTRIC POWER SYSTEM

Inventors: Richard Buswell, Joseph Clausi, Ronald Cohen, Craig Louie, David Watkins
Assignee: Ballard Power Systems Inc.

This patent gives a detailed design for a PEM fuel cell cogeneration system with a steam methane reformer, with several material and energy flow diagrams. The flow diagrams contain a number of ingenious ideas for heat integration in fuel cell reformers.

U.S. Patent No. 5,470,360
Nov. 28, 1995

FUEL CELL POWER PLANT REFORMER BURNER GAS FLOW CONTROL SYSTEM

Inventor: Richard Sederquist
Assignee: International Fuel Cells Corporation

This patent discusses how the burner flue gases are directed uniformly toward the tubes to promote more even heating of the catalyst tubes.

U.S. Patent No. 5,484,577
Jan. 16, 1996

CATALYTIC HYDROCARBON REFORMER WITH ENHANCED INTERNAL HEAT TRANSFER MECHANISM

Inventors: Richard Buswell, Ronald Cohen, Joseph Clausi, Stanley Leavitt, David Watkins
Assignee: Ballard Power Systems Inc.

Background

For use in fuel cells, conventional reformers have disadvantages:

1) because they operate at very high pressure and temperature differentials, they must be made of thick walled, rugged portions of expensive materials.

2) they tend to be quite large, which increases cost.

Object of invention to:

1) Reduce the operating temperature of the reformer, because of enhanced heat transfer between the chamber containing the burner and the reaction chamber

2) reduce overall volume of assembly

3) minimize the use of costly high temp materials for components

4) Use lower pressure differential between burner combustion gases and process fuel reaction gases.

5) Provide a catalytic reformer with reduced thermal gardients to extend life of catalysts and internal components

Multiple tubes can be used. Instead of a heat exchange medium to assure uniform heating of the tubes, a thin annuar gap with turbulent flow is used.

APPENDIX 1.B.
MEETINGS WITH MANUFACTURERS OF SMALL SCALE REFORMER EQUIPMENT

HALDOR-TOPSOE (December 1994)

At the 1994 Fuel Cell Seminar Joan Ogden met with representatives from Haldor Topsoe, a company which manufactures steam methane reformers. She discussed the economics of small scale steam methane reforming, and the potential for reducing the capital cost of small scale steam reformers.

BALLARD POWER SYSTEMS (December 1994)

At the 1994 Fuel Cell Seminar Joan Ogden met with representatives from Ballard Power Systems, a fuel cell developer. She discussed the potential for low cost steam methane reforming, and the design differences between fuel cell and conventional steam methane reformers.

H-POWER (February 1995).

Joan Ogden, Bob Williams and Sivan Kartha met with engineers from H-Power to discuss their experience with developing small scale steam reformers for use with fuel cells.

INTERNATIONAL FUEL CELLS (April 1995).

Joan Ogden, Sivan Kartha and Margaret Steinbugler of Princeton CEES visited with engineers at International Fuel Cells, South Windsor, CT. We presented results from our studies of hydrogen infrastructure, and discussed the present status and projections for small scale reformer technology.

HYDROGEN BURNER TECHNOLOGY, INC. (June 1995)

Joan Ogden visited Hydrogen Burner Technology in Costa Mesa, CA, where small scale partial oxidation systems are under development. She discussed the potential economics of small scale POX units with engineers there, and toured their labs, where several small POX units were being constructed.

ARTHUR D. LITTLE (February 1996)

Joan Ogden and Tom Kreutz of CEES met with researchers at Arthur D. Little, Cambridge, MA, where small scale partial oxidation systems are being developed. We presented preliminary results of our assessment of the economics of small scale steam reforming, and also some modelling results for POX systems. We also visited their laboratories, where an experimental POX system was being tested.

AIR PRODUCTS AND CHEMICALS, INC. (March 1996)

Joan Ogden and Tom Kreutz from Princeton CEES met with engineers from Air Products and Chemicals at Allentown, PA. We presented preliminary results of our assessment of small scale reformer technologies, and learned about a joint venture APCI and KTI have to develop a small scale reformer for chemical applications. We discussed reformer design considerations, especially the trade-offs involved in systems with low pressure reformers.

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Table 0. Conversion Factors and Economic Assumptions

1 GJ (Gigajoule) = 10^9 Joules = 0.95 Million BTU
1 EJ (Exajoule) = 10^{18} Joules = 0.95 Quadrillion (10^{15}) BTUs

1 million standard cubic feet (scf) = 28,300 Normal cubic meters (m_N^3) = 362 GJ (HHV)
1 million scf/day = 2.80 tons/day = 4.19 MW H_2 (based on the HHV of hydrogen)

1 scf H_2 = 362 kJ (HHV) = 344 BTU (HHV)
1 lb H_2 = 64.4 MJ (HHV) = 61.4 kBTU (HHV) = 178.5 scf

1 m_N^3 = 12.8 MJ (HHV)
1 kg H_2 =141.9 MJ (HHV) = 393 scf

1 gallon gasoline = 130.8 MJ (HHV)
$1/gallon gasoline = $7.67/GJ (HHV)

All costs are given in constant $1993.

Capital recovery factor for hydrogen production systems, distribution systems and refueling stations = 15%

Passenger car owners' real discount rate is assumed to be 3.8%
The real discount rate is taken to be 10% for urban buses.

Table 1. Characteristics Of Small-Scale Methane Reformers

Company	Capacity H2 out (scf/day)	Feed Pressure	Feed Purification	Reformer Temp	Shift reactor	Hydrogen Purification	Efficiency	Dynamic response	Physical Size and Cost
CATALYTIC STEAM REFORMERS									
Halder-Topsoe	67,000-3.3 million	Supply pressure	Zinc-oxide	1525°F	high and low temp.	99.999% w/PSA	80%	Cold start in 3 hours, 25-100% load in 30 seconds	$500/kW, 30' high, 8' diameter
Howe-Baker Engineers	1-5 million	300 psig	Zinc-oxide	1550-1600°F	High temp (675-800°F)	99.999% w/PSA	80%	Cold start in several hours	$3 million for 1 million scf/day
Hydro-chem	120,000-5 million	275 psig	Zinc-oxide		High temp	99.99% pure w/PSA		Cold start in 2-4 hours, 100% to 30% load in 1 hr	40' high, 40x80' area, $1 million for 120,000 scf/day
KTI/Air Products and Chemicals, Inc.	0.1-3 million	Supply pressure				99.99% pure w/PSA	80%		$3 million for 1 million scf/day
International Fuel Cells/ONSI	48,000-140,000	Supply pressure		Steam Reforming Fuel Processor Unit in Fuel Cell Cogeneration Package	High temp.	Less than 75% pure H2, product contains H2, CO, CO2	84%	Cold start in less than 3 hours, Idle to 100% load in 1 second	10' high, 4' diameter, $500/kWe for reformerBeing developed as part of a PAFC cogen package
Ballard Power Systems	6000 120,000	Supply pressure				99.99% pure (for use with PEM fuel cell)	80% eff.		Being devel. as part of a fuel cell package est. $500/kWe
Chiyoda (JAPAN)				Stea, Reformer used with 100-200 kW Phosphoric Acid Fuel Cellls					
PARTIAL OXIDATION SYSTEMS									
Arthur D. Little	12,500		Zinc oxide	1600-1800°C (catalytic POX)	high and low temp.	Pref. oxidation	80% eff	Fast	R&D only, not commercial yet
Hydrogen Burner Technology	12,000 96,000			2500°F (POX)	High and low temp	Palladium membrane 99.999% pure	75-80%	Fast	less than $200,000 for 96,000 scf/day unit with production of a few systems
AUTOTHERMAL REFORMERS									
Rolls-Royce/ Johnson-Mathey	7-25 kW H_2 (1700-6200 scf H_2/day)	ambient pressure		Exeprimental Autothermal Reformer Unit			>90%	Fast	60 liters
International Fuel Cells/ONSI	1.5-20 kW			Autothermal Reformer under development				Fast	

Table 2.
Fuel Cell Steam Methane Reformers

Haldor-Topsoe:	Heat Exchange Reformer for PAFC applications. 0.85 million scf/day unit built (corresp. to 1.25 MWe FC)
IFC:	Reformers for Phosphoric Acid Fuel Cell cogeneration systems PC25 series for 200 kWe FC Joint venture w/ Praxair for H2 production system based on PC-25 fuel cell reformer
BALLARD :	10 kWe FC reformer for PEM Size: 52" x 48" x 86" 250 kWe FC multi-tube reformer for PEM Size: 27,600 lb, 1200 cu.ft., 8' x 8' x 18' operational mid-1996; 50 units by 1998
OSAKA GAS:	Built Haldor-Topsoe type multi-cylinder single tube reformer for FC. Has plans to build multi-tube, more compact unit [10.5 m x 8 m x 3.6 m (ht)]

TABLE 3. OPERATING CONDITIONS AND COSTS FOR A SMALL SCALE HYDROGEN PLANT WITH STEAM REFORMING OF NATURAL GAS

Operating Temperature	845-870°C
Operating pressure	10 atmospheres
Hydrogen production capacity	0.366 million scf/day
Hydrogen purification	Pressure Swing Adsoprtion (PSA)
Outlet pressure from PSA	200 psig
Hydrogen purity	99.999%
Installed capital cost	$2.22 million
Natural gas consumption	0.400 MBTU/1000 scf H_2
Energy conversion efficiency	86% (HHV)
Electricity consumption	0.6 kWh/1000 scf H_2 = 220 kwh/day
Cooling water	200 gallons/1000 scf H_2 =73,200 gallons/day
Boiler water consumption	40 lb./1000 scf H_2 =760 gallons/day
Catalysts	$0.65/1000 scf H_2 =$8683/yr
Misc. O&M	3.5% x capital cost =$77,000/yr
Equipment lifetime	20 years
Land area	80 ft x 30 ft
Height	40 ft
Labor requirements	Automated with occasional monitoring
Emissions (NOx) control	Selective catalytic reduction

Source: D. King, Howe-Baker Engineering, private communications, 1993

Table 4.
Comparison Of Small Scale (0.1-2 Million scf H_2/Day) Steam Methane Reformers For Chemical And Fuel Cell Applications

	SMR for Chemical Applications[a]	Fuel Cell SMR for PEM system
Reformer Operating temp °C	845-870°C	745°C [b]
Reformer Operating Pressure (psig)	225-375	70-85 [b]
Reformer tube material	Nickel chrome alloy steels	Stainless steel
Heat Transfer	Radiative	Convective
Tube length (ft)	30	6-8
Energy conversion efficiency $CH_4 \rightarrow H_2$	85%	81.4% (reformer) [b] 70.7% (fuel proc. system incl. aux)
Purification Technology	PSA	Selective catalytic oxidation to convert $CO \rightarrow CO_2$
Outlet Gas Composition (% volume)		
H_2	99.999%	68.8% [b]
CH_4		3.9%
N_2		0.7%
CO_2		17.2%
H_2O		9.3%
Outlet Gas Pressure (psig)	200 (from PSA)	65 (from PROX)
Capital Cost	\$3x10^6 x (capacity in million scf/day)$^{0.3}$	\$500/kWe (mass produced)
120,000 scf H_2/day	\$1.6 x 10^6	\$100,000
Lifetime (yr)	20	>15

a. From Howe-Baker
b. From Ballard Power Systems U.S. Patent No. 5,360,679 Nov. 1, 1994

Table 5a. Material Flows for a Small Scale Hydrogen Production Plant (100,000 scf H2/day) Using a Fuel Cell Type Reformer, 1-stage Compressor and PSA (see Figure 9)

No.	Name	Temp °C	Press. (barg)	Flow volume kmol/h	Components (% by volume)						
					H20	H2	CO	CO2	CH4	O2	N2
1	NG Feed	272	6.8	2.22	0	0	0	0	100	0	0
2	Refomer Feed	343	6.3	8.87	75%	0	0	0	25%	0	0
3	Reformer Bed	743	5.4	12.5	32%	50%	8.0%	6.4%	3.3%	0	0
3a	Reformer Exit	454	5.3	12.5	32%	50%	8.0%	6.4%	3.3%	0	0
4	High Temp Shift Inlet	193	5.2	12.5	32%	50%	8.0%	6.4%	3.3%	0	0
5	Hight Temp Shift Outlet	281	5.1	12.5	25%	57%	0.6%	13.9%	3.3%	0	0
6	Low Temp Shift Inlet	193	5.0	12.5	25%	57%	0.6%	13.9%	3.3%	0	0
7	Low Temp Shift Outlet	199	4.9	12.5	25%	58%	0.15%	14.3%	3.3%	0	0
8	Compressor Inlet	40	4.8	9.38	0	77%	0.2%	19%	4.4%	0	0
9	Compressor Outlet	110	17.0	9.38	0	77%	0.2%	19%	4.4%	0	0
10	PSA Inlet	40	17.0	9.38	0	77%	0.2%	19%	4.4%	0	0
11	H2 Product	40	17.0	5.20	0	100%	0	0	0	0	0
11a	PSA Purge Gas = Reformer Burner Fuel	40	1.3	4.17	0	40%	0.5%	49%	11.4%	0	0
12	Reformer Burner Air	343	4.1	21.9	8.8%	0	0	0	0	9.5%	81.7%
12a	Reformer Burner	1166	4.1	25.5	19.7%	0	0	8.7%	0	1.3%	70.3%
13	Reformer Exhaust Gas	713	4.0	25.5	19.7%	0	0	8.7%	0	1.3%	70.3%
14	Reformer Exhaust Gas at Evap. Inlet	630	4.0	25.5	19.7%	0	0	8.7%	0	1.3%	70.3%
15	Reformer Exhaust Gas at Evap. Outlet	346	3.9	25.5	19.7%	0	0	8.7%	0	1.3%	70.3%
16	Evaporator Feedwater	155	10.2	6.65	100%	0	0	0	0	0	0
17	Steam in Reformer Feed	343	6.3	6.65	100%	0	0	0	0	0	0

PSA Recovery	72%	1 stage recip. compressor	150 scfm
Conversion of CH4 -> H2 inReformer	81.4%	Compressor inlet pressure (barg)	4.8
H2 Plant Energy Conv.Eff. = H2 out(HHV)/CH4 in (HHV)	77%	Compressor outlet pressure (barg)	17.0
		Compressor Power req.	17.5 kWe

Table 5b. Material Flows for a Small Scale Hydrogen Production Plant (1 million scf H2/day) Using a Conventional Reformer w/PSA (see Figure 8)

No.	Name	Temp °C	Press. (barg)	Flow volume kmol/h	Components (% by volume)						
					H20	H2	CO	CO2	CH4	O2	N2
0	NG Feed	20	25.0	19.0	0	0	0	0.2%	97.8%	0	2%
1	NG to Reformer Feed	227	24.5	15.7	0	0	0	0.2%	97.8%	0	2%
2	Refomer Feed	577	24.0	62.8	75%	0	0	0.1%	24.3%	0	0.6%
3	Reformer Exit	900	23.5	90.6	32%	50%	10.3%	5.0%	2.0%	0	0
4	High Temp Shift Inlet	350	23.0	90.6	32%	50%	10.3%	5.0%	2.0%	0	0
5	Hight Temp Shift Outlet	427	22.5	90.6	25%	57%	3.4%	11.9%	2.0%	0	0
6	Low Temp Shift Inlet	227	22.0	90.6	25%	57%	3.4%	11.9%	2.0%	0	0
7	Low Temp Shift Outlet	227	21.5	90.6	22%	60	0.5%	14.9%	2.0%	0	0
8	PSA Inlet	40	21.0	71.0	0.3%	77%	0.6%	19%	2.6%	0	0
9	NG to Reformer Furnace	20	25.0	3.3	0	0	0	0.2%	97.8%	0	2%
10	H2 Product	40	20.3	53.0	0	100%	0	0	0	0	0
11	PSA Purge Gas = Reformer Burner Fuel	40	1.3	4.3	0	40%	9.9%	0%	42%	0	0
12	Reformer Burner Air	727	1.0	62.3	0	0	0	0	0	21%	79%
13	Reformer Exhaust Gas at Ref. Outlet	920	1.0	69.0	17%	0	0	8.1%	0	2.5%	72%
14	Reformer Exhaust Gas at Stack	120	1.0	69.0	17%	0	0	8.1%	0	2.5%	72%
16	Steam in Reformer Feed	273	24.5	48.5	100%	0	0	0	0	0	0
17	Condensate	40	21.0	19.6	100%	0	0	0	0	0	0

PSA Recovery	90%
Conversion of CH4 -> H2 in Reformer	88%
H2 Plant Energy Conv.Eff. = H2 out(HHV)/CH4 in (HHV)	85%

Table 6. Hydrogen Refueling Station with Onsite Reforming of Natural Gas: Fuel Cell Reformer w/PSA System: Capital and O&M Costs and Delivered Hydrogen Cost

	REFORMER CAPACITY (SCF/D)		
REFUELING STATION	**100,000**	**366,000**	**1,000,000**
INSTALLED CAPITAL COST($)			
HYDROGEN PRODUCTION			
Reformer+Shift Reactors	88,000	218,000	440,000
1-stage Compressor	18,400	67,000	184,000
PSA	250,000	250,000	375,000
Sub-Total Hydrogen Production	356,400	535,000	999,000
Storage cylinders	157,500	534,240	1,575,000
Storage compressor	51,900	190,000	519,000
Priority Panel and Sequencer	10,500	10,500	10,500
Dispenser	50,000	100,000	275,000
TOTAL	626,300	1,369,740	3,378,500
OPERATING COSTS ($/YR)			
Natural gas (@ $4/MBTU)	64,468	235,951	644,680
Reformer O&M	18,000	65,880	180,000
Storage cylinder O&M	1500	5200	15,000
Compressor electricity- 6 c/kwh	14217	52,034	142,170
Compressor O&M	3000	3000	3000
Labor	131,400	131,400	131,400
TOTAL	232,585	493,465	1,028,465
DELIVERED H2 COST ($/GJ)			
Hydrogen plant capital	4.05	1.66	1.13
Natural gas feedstock	4.88	4.88	4.88
Reformer O&M	1.36	1.36	1.36
Storage cylinder capital	1.66	1.66	1.66
Compressor capital	0.59	0.59	0.59
Compressor O&M	0.23	0.062	0.022
Compressor electricity	1.08	1.08	1.08
Dispenser+controls	0.69	0.34	0.32
Labor	9.94	2.72	1.00
TOTAL	24.5	14.3	11.5
FUEL CELL CARS FUELED/DAY	82	300	820

Table 7. Hydrogen Refueling Station w/ Conventional Steam Methane Reformer w/PSA: System Capital and O&M Costs and Delivered Hydrogen Cost

	REFORMER CAPACITY (SCF/D)		
REFUELING STATION	**100,000**	**366,000**	**1,000,000**
INSTALLED CAPITAL COST($)			
Reformer Plant	1,500,000	2,220,000	3,000,000
Storage cylinders	157,500	534,240	1,575,000
Storage compressor	51,900	190,000	519,000
Priority Panel and Sequencer	10,500	10,500	10,500
Dispenser	50,000	100,000	275,000
TOTAL	1,769,900	3,054,740	5,379,500
OPERATING COSTS ($/YR)			
Natural gas (@ $4/MBTU)	58,400	213,744	584,000
Reformer O&M	63,680	118,560	216,800
Storage cylinder O&M	1500	5200	15,000
Compressor electricity- 6 c/kwh	14217	52,034	142,170
Compressor O&M	3000	3000	3000
Labor	131,400	131,400	131,400
TOTAL	272,197	525,938	1,092,370
DELIVERED H2 COST ($/GJ)			
Refomer plant capital	17.03	6.89	3.41
Natural gas feedstock	4.42	4.42	4.42
Reformer O&M	4.82	2.45	1.64
Storage cylinder capital	1.66	1.66	1.66
Compressor capital	0.59	0.59	0.59
Compressor O&M	0.23	0.062	0.022
Compressor electricity	1.08	1.08	1.08
Dispenser+controls	0.69	0.34	0.32
Labor	9.94	2.72	1.00
TOTAL	40.4	20.2	14.2
FUEL CELL CARS FUELED/DAY	82	300	820

TABLE 8. COMPARISON OF SMALL SCALE TECHNOLOGIES FOR PRODUCING HYDROGEN FROM METHANE

		REFORMER TYPE			
	CONV. SMR	MASS PRODUCED FUEL CELL SMR W/ 1-STAGE COMPRESSOR + PSA			MASS PROD. POX W/ PSA
Hydrogen Plant Energy Conv. Eff. = H2 out /CH4 in (HHV/HHV)	80-85%	80%			65-75%
Plant Capacity (million scf H2/day)	Hydrogen Plant Capital Cost (million $)				
0.1	1.5	0.36			0.39
0.366	2.2	0.54			0.85
1.0	3.0	1.0			1.2
2.0	3.7	1.7			1.6
O&M Costs ($/yr)	5% of total capital cost	$180,000 per year per million scf/day of H2 production capacity (equiv. to 1 cent/kWhe for fuel cell cogen. system)			N.A.
Capital Cost Contributions of Hydrogen Plant Components	For plant capacity of 1.6 million scf/day	For a range of H2 plant capacities in million scf/day			
		0.1	0.366	1.0	
Reformer Vessel+ Shift Reactors	50%	25%	41%	44%	<10%
Compressors	11%	5%	13%	18%	
PSA	39%	70%	47%	38%	>90%
Critical Factors Limiting the Capital Cost of the Hydrogen Plant	High pressure op. => High Temp, high cost reformer tube steels Radiative Heat Transfer => Large Tube Areas Not standard design => high enginering cost	Low pressure reformer => extra compression stage must be used prior to PSA O&M Costs not known yet			Presence of nitrogen in syngas => higher cost down stream process and purif. equip., High cost of PSA to remove N2

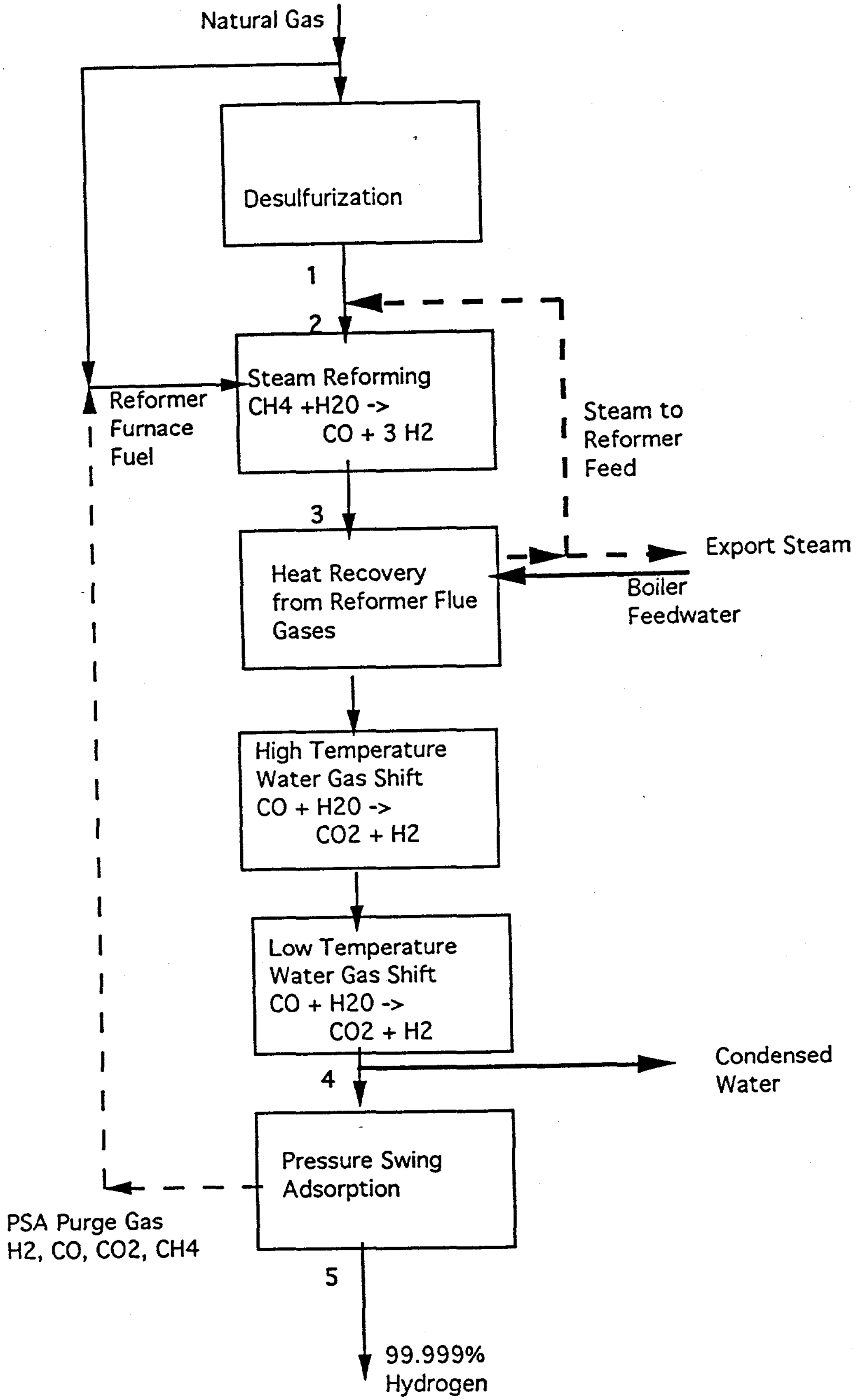


Figure 1. Process steps for production of hydrogen via steam reforming of natural gas, using PSA purification

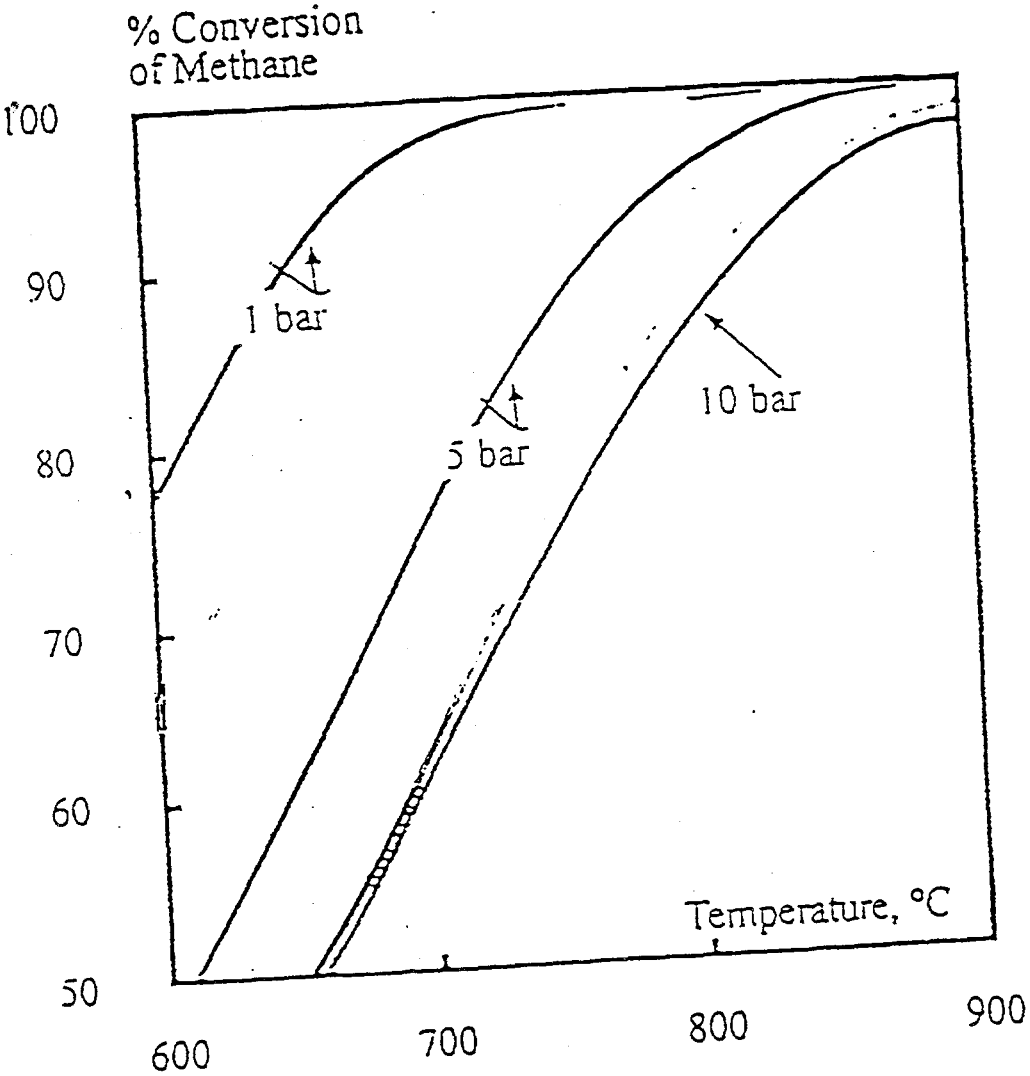
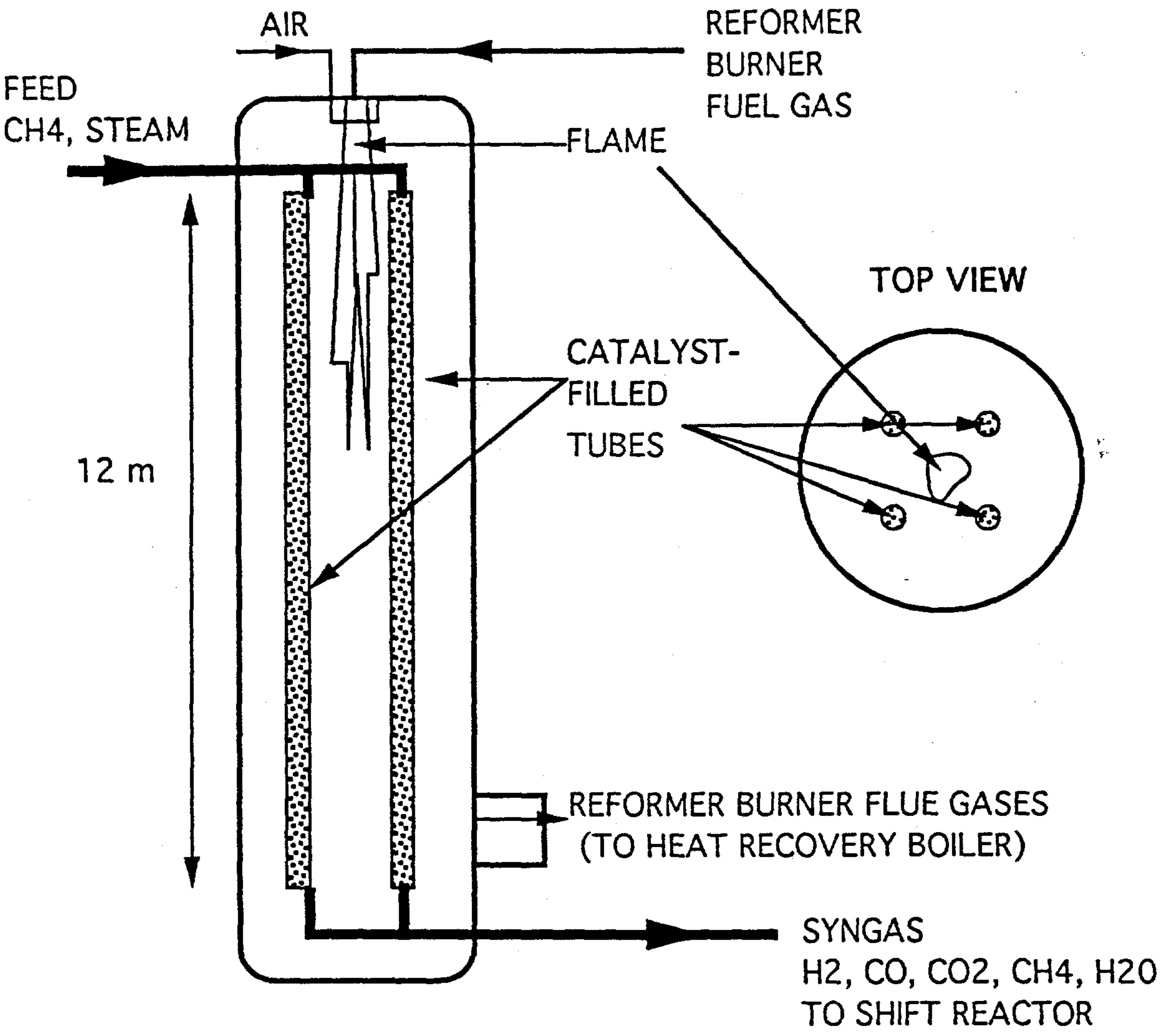
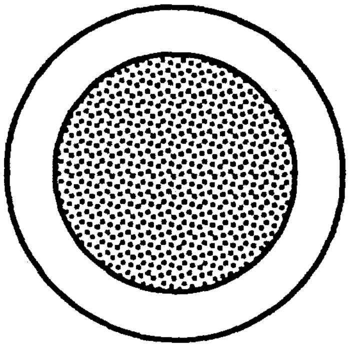


Figure ~~3~~ 2. Equilibrium methane conversion in steam reforming as a function of reformer operating pressure and temperature. Acheiving the same conversion of methane to hydrogen at higher pressure requires higher temperature.

FIGURE 3. CONVENTIONAL SMALL SCALE STEAM METHANE REFORMER DESIGN



SINGLE REFORMER TUBE



TYPICAL
CATALYST
MATERIALS:
NICKEL ON
ALUMINA
SUPPORT

TYPICAL TUBE WALL
MATERIALS:
HIGH ALLOY STEELS:
HK-40, IN 519

TYPICAL DIMENSIONS:
WALL THICKNESS: 10-20 mm
TUBE DIAMETER: 70-160 mm
TUBE LENGTH: 12 m

FIGURE 4. COMPACT, TUBULAR, SMALL SCALE STEAM METHANE REFORMER DESIGNED FOR FUEL CELL APPLICATIONS, WITH CONVECTIVE HEAT TRANSFER (Based on Haldor-Topsoe "Heat Exchange Reformer")

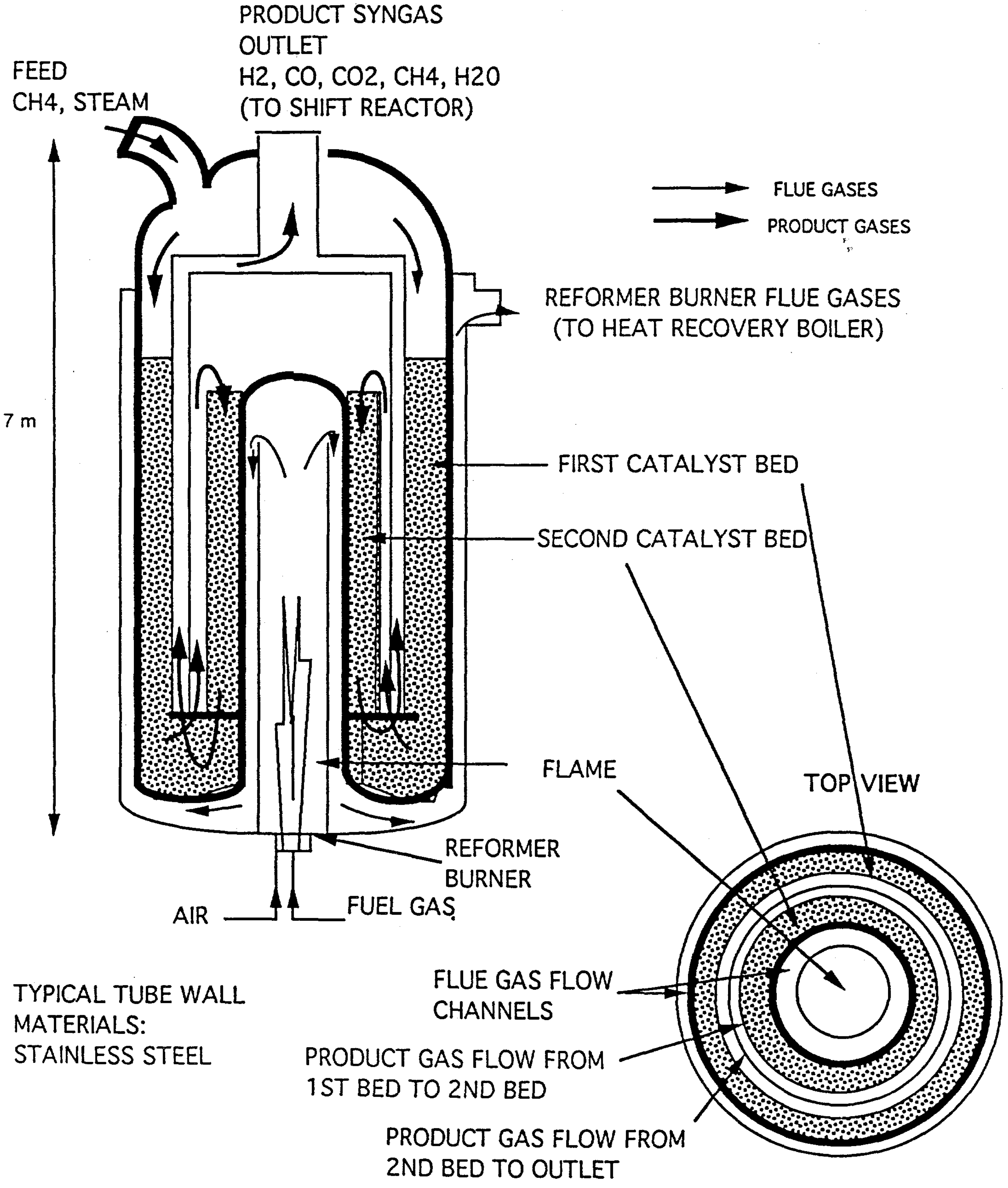


Figure 5.
Capital Cost of Conventional Small Scale Steam Methane Reformers

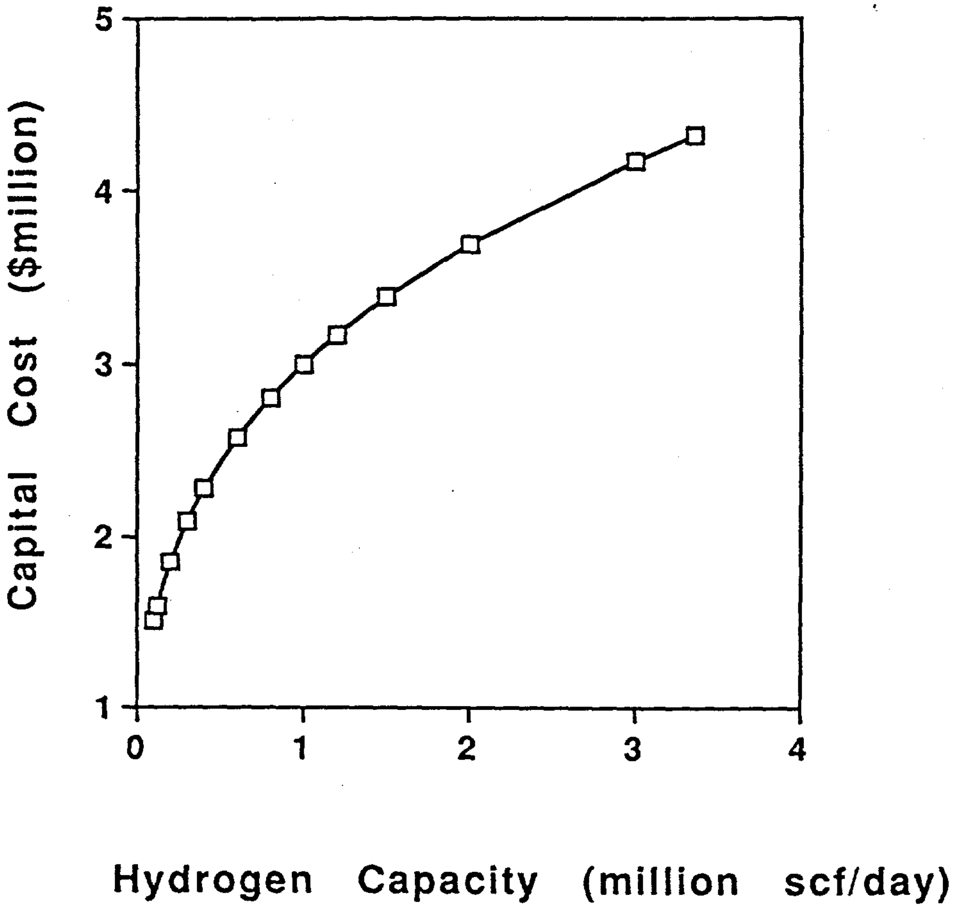
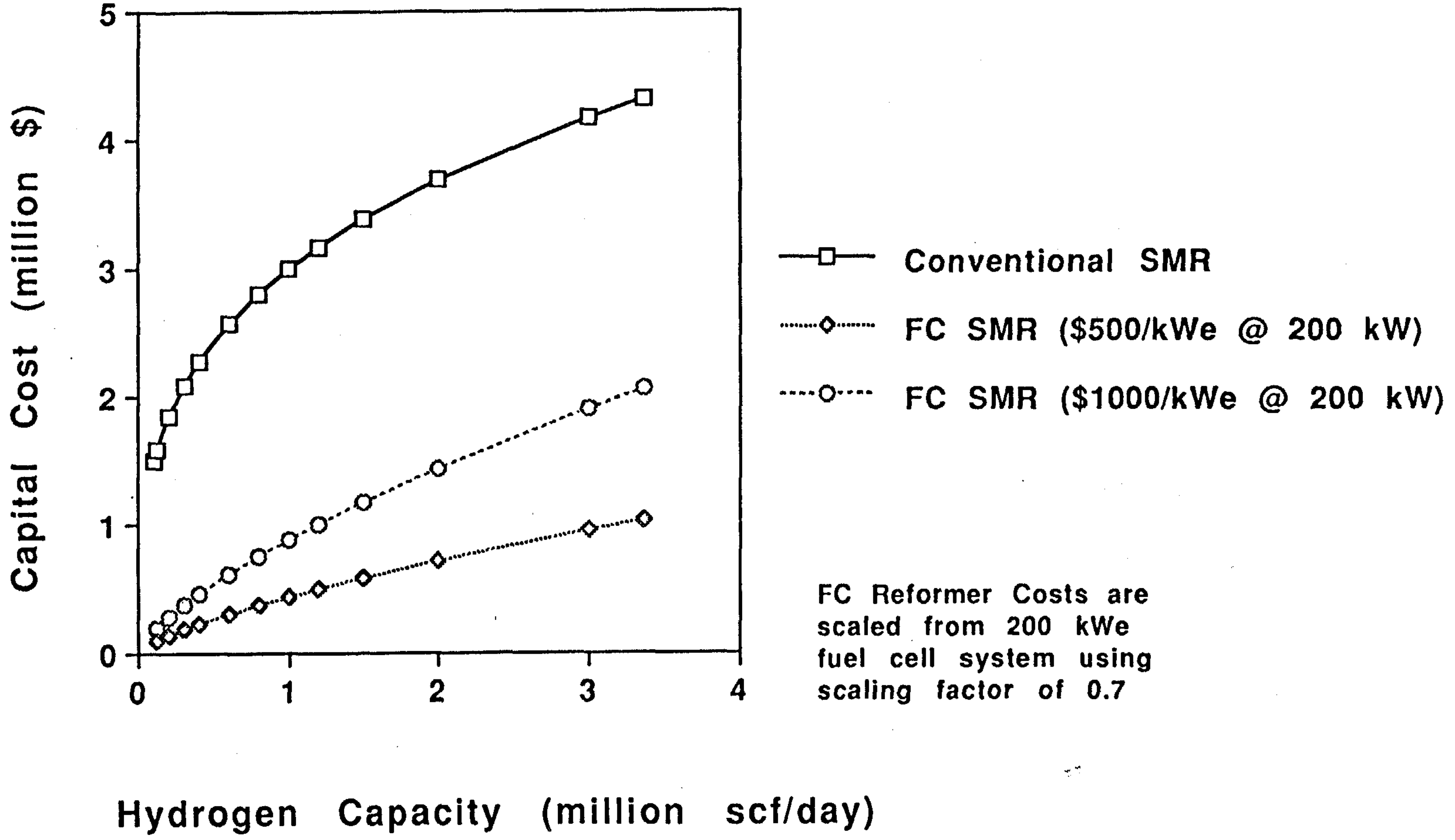


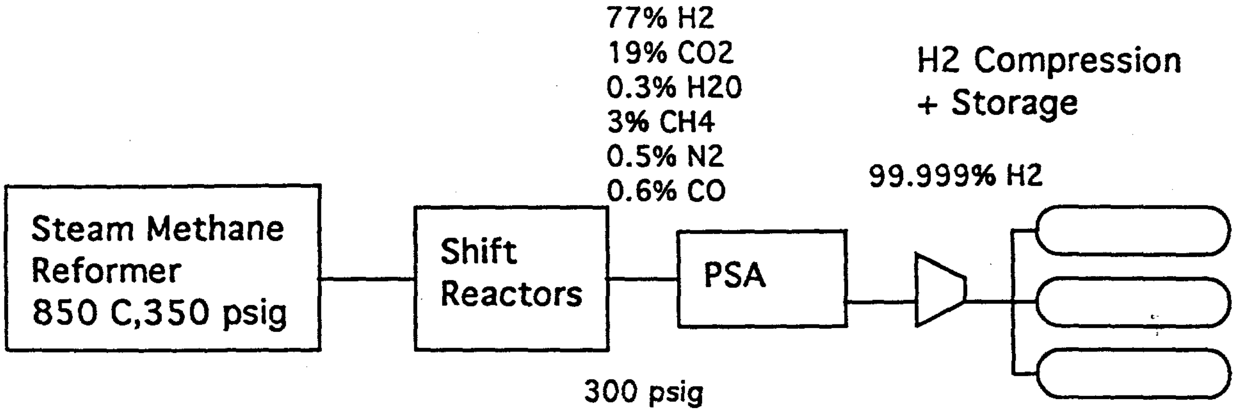
Figure 6.
Capital Cost of Small Scale Steam Methane Reformers



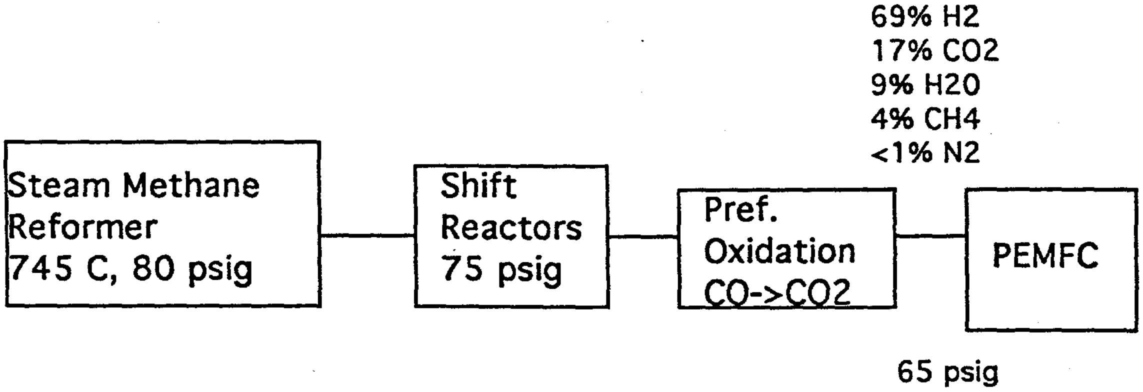
FC Reformer Costs are scaled from 200 kWe fuel cell system using scaling factor of 0.7

Figure 7.
STATIONARY H2 PRODUCTION FROM METHANE
WITH CONVENTIONAL AND FUEL CELL SMR

a) Conventional Small SMR



b) Fuel Cell SMR



c) Fuel cell SMR w/PSA

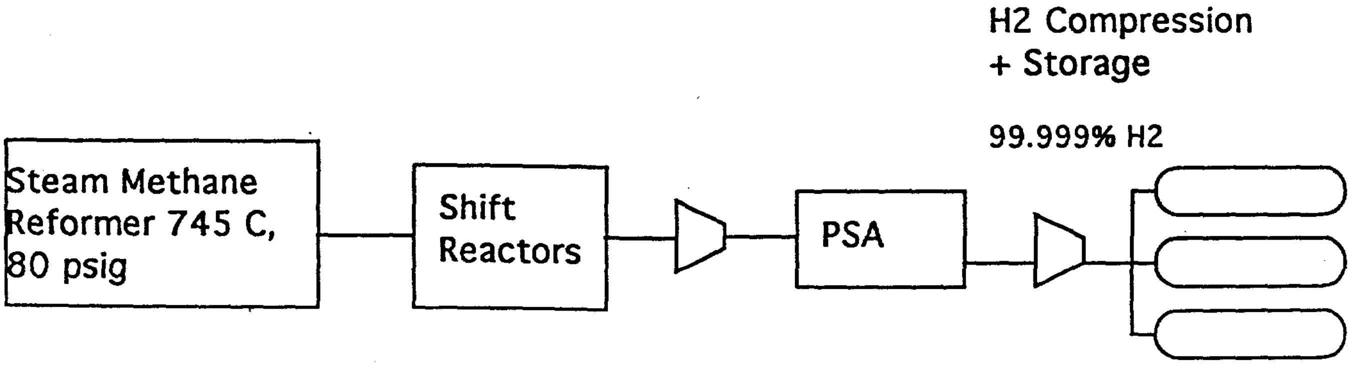


FIGURE 8. FLOWS IN A SMALL SCALE HYDROGEN PRODUCTION PLANT: CONVENTIONAL STEAM REFORMER w/PSA

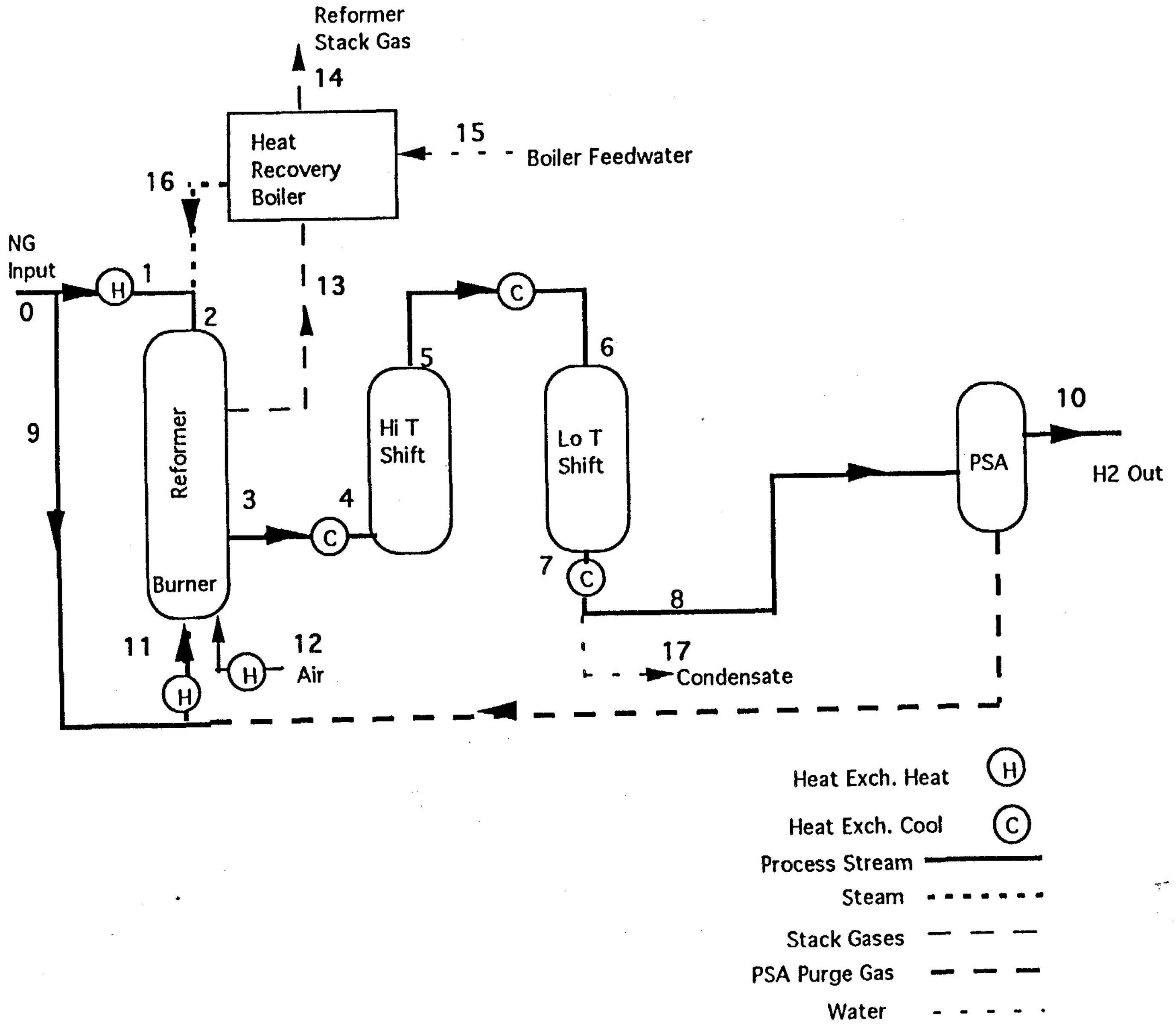


FIGURE 9. FLOWS IN A SMALL SCALE HYDROGEN PRODUCTION PLANT:
FUEL CELL STEAM REFORMER + 1-STAGE COMPRESSOR w/ PSA

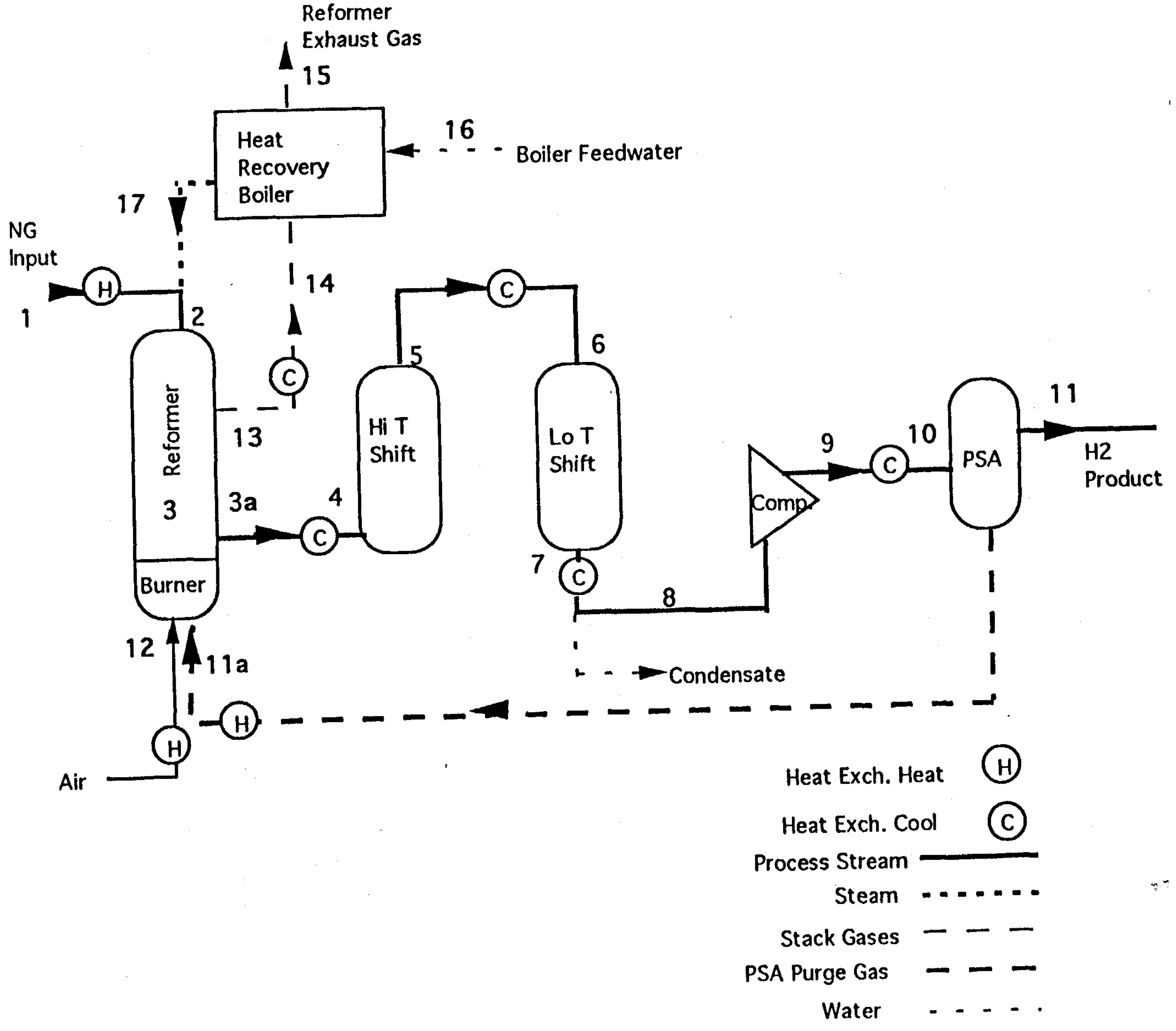
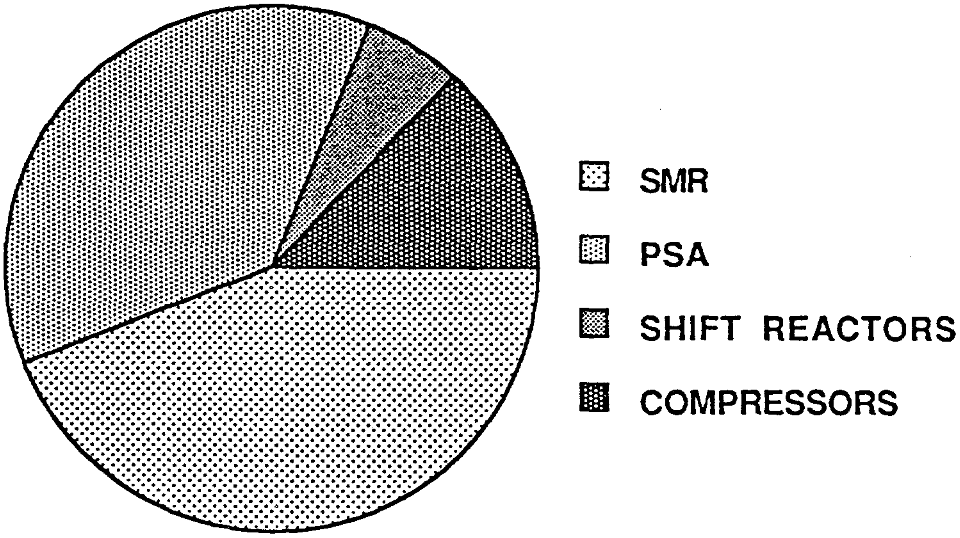


FIGURE 10.
CAPITAL COST COMPONENTS OF CONVENTIONAL SMALL SCALE STEAM METHANE REFORMER



Based on costs for a
1.6 million scf/day
Reformer

Figure 11. Capital Cost of Small Scale PSA Units as Compared Total Hydrogen Plant Cost (Conventional SMR)

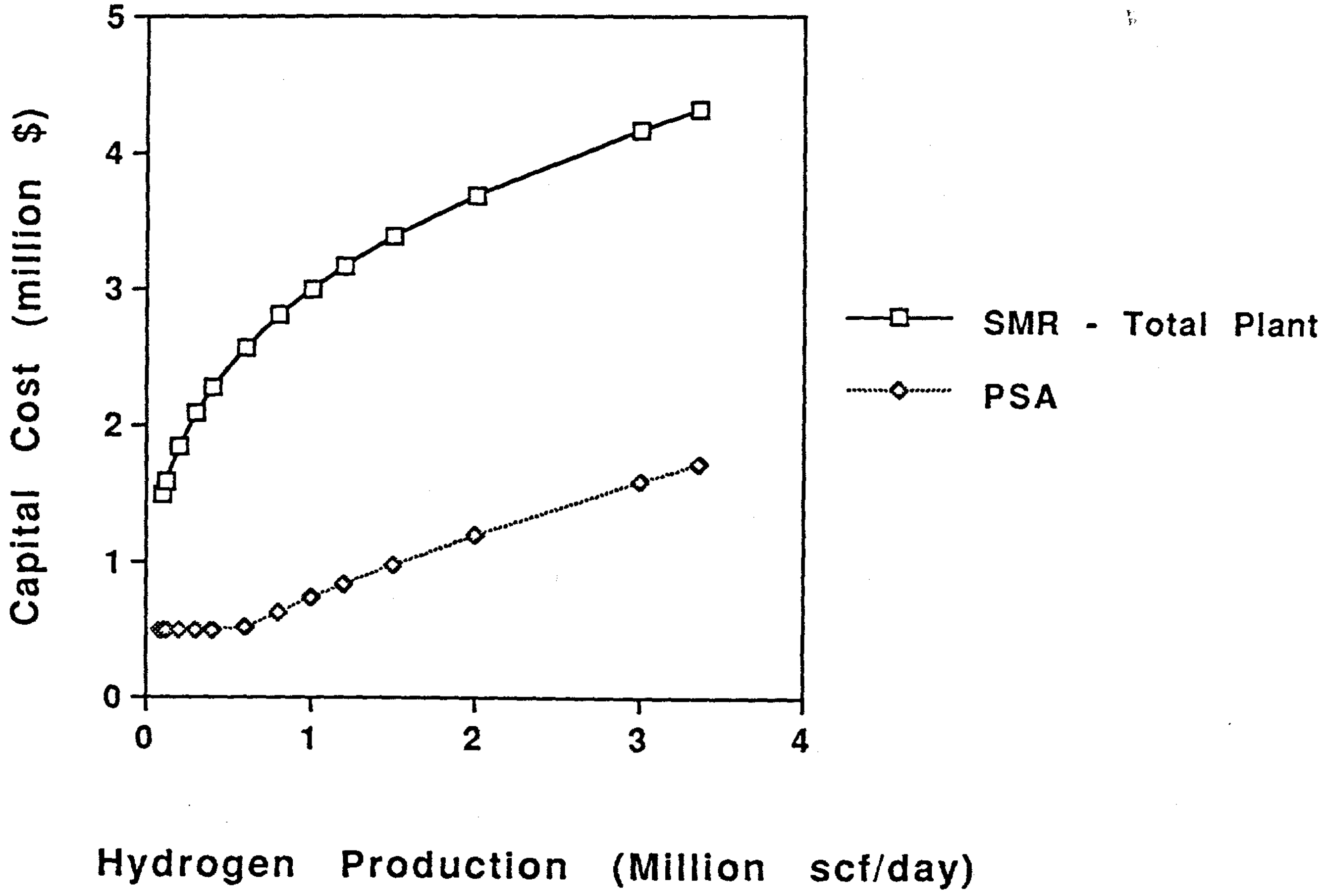
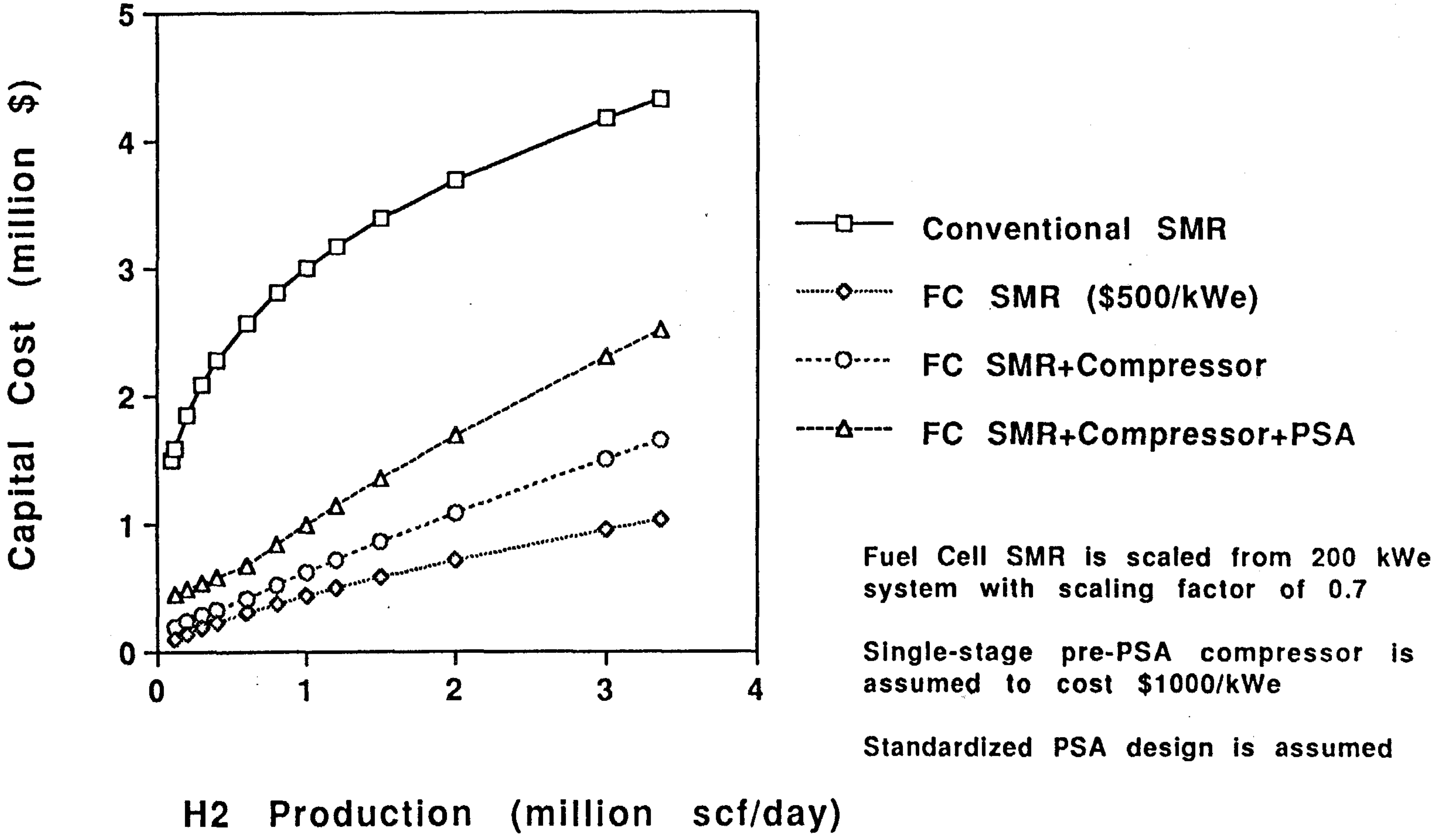


Figure 12.

Capital Cost of Small Scale Hydrogen Plants Based on Conventional and Fuel Cell Steam Reformers



Fuel Cell SMR is scaled from 200 kWe system with scaling factor of 0.7

Single-stage pre-PSA compressor is assumed to cost $1000/kWe

Standardized PSA design is assumed

FIGURE 13a. REFUELING STATION WITH ONSITE REFORMING OF NATURAL GAS

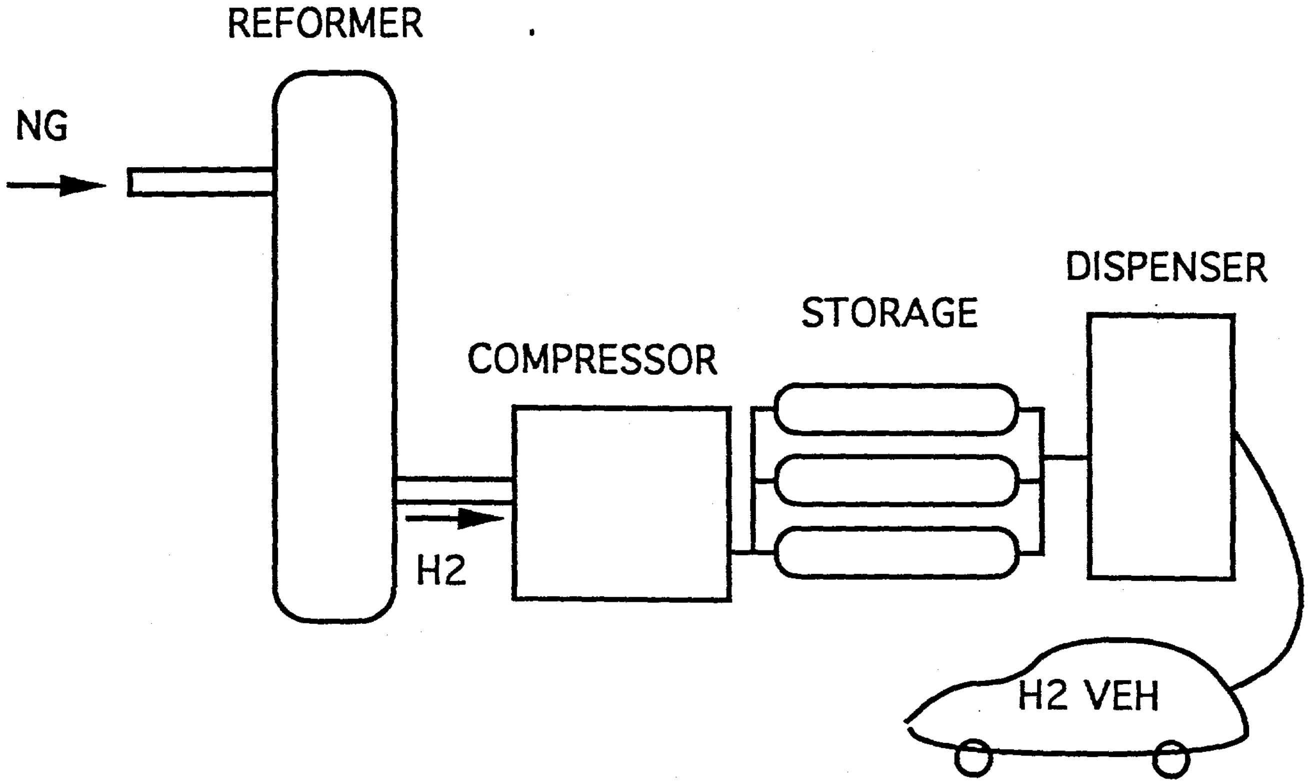


Figure 13b. Delivered Cost of Hydrogen Transportation Fuel: Onsite Reforming of Natural Gas

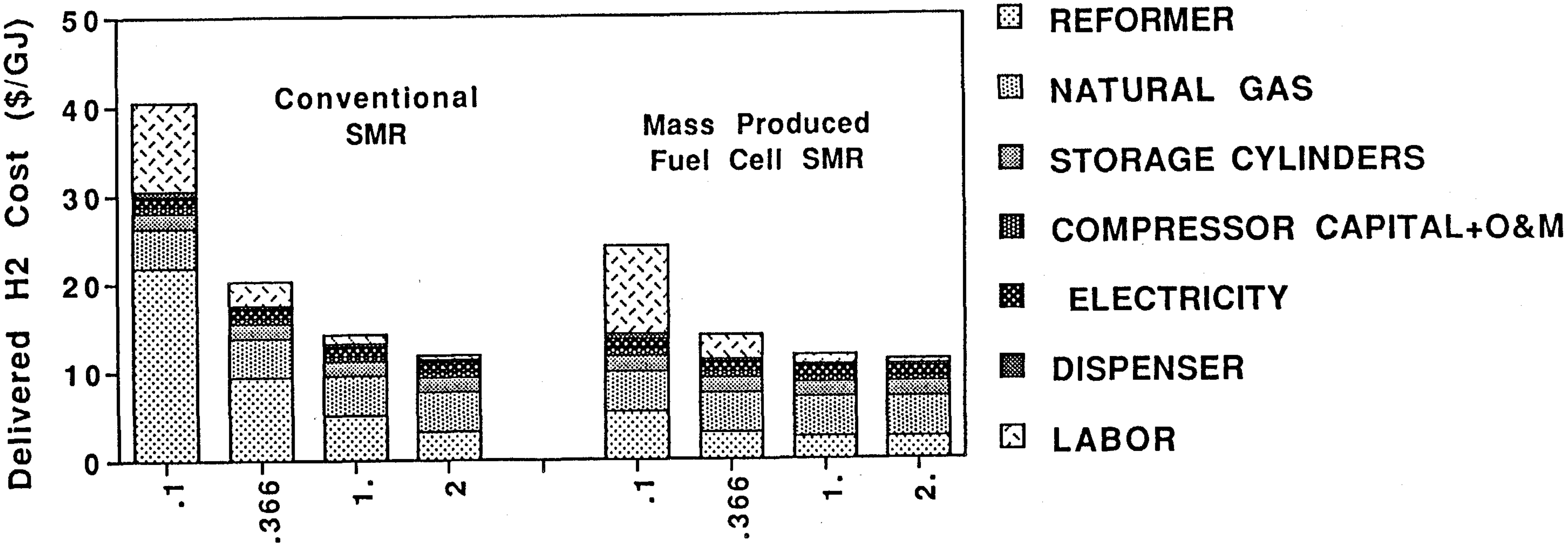


Figure 13c. Delivered Cost of Hydrogen Transportation Fuel: Onsite vs. Centralized Reforming of Natural Gas

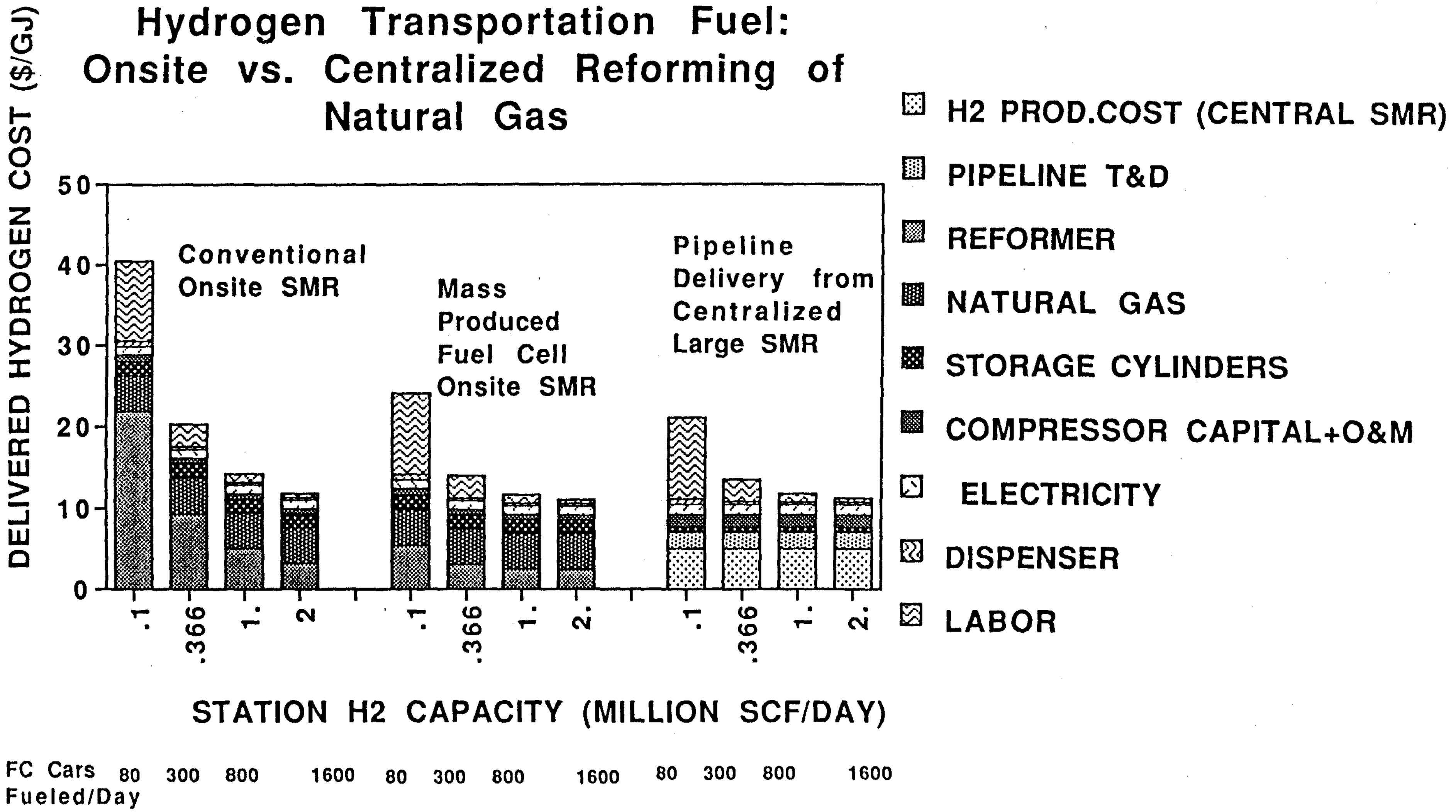


Figure 14.
Capital Cost of Small Scale Reformers

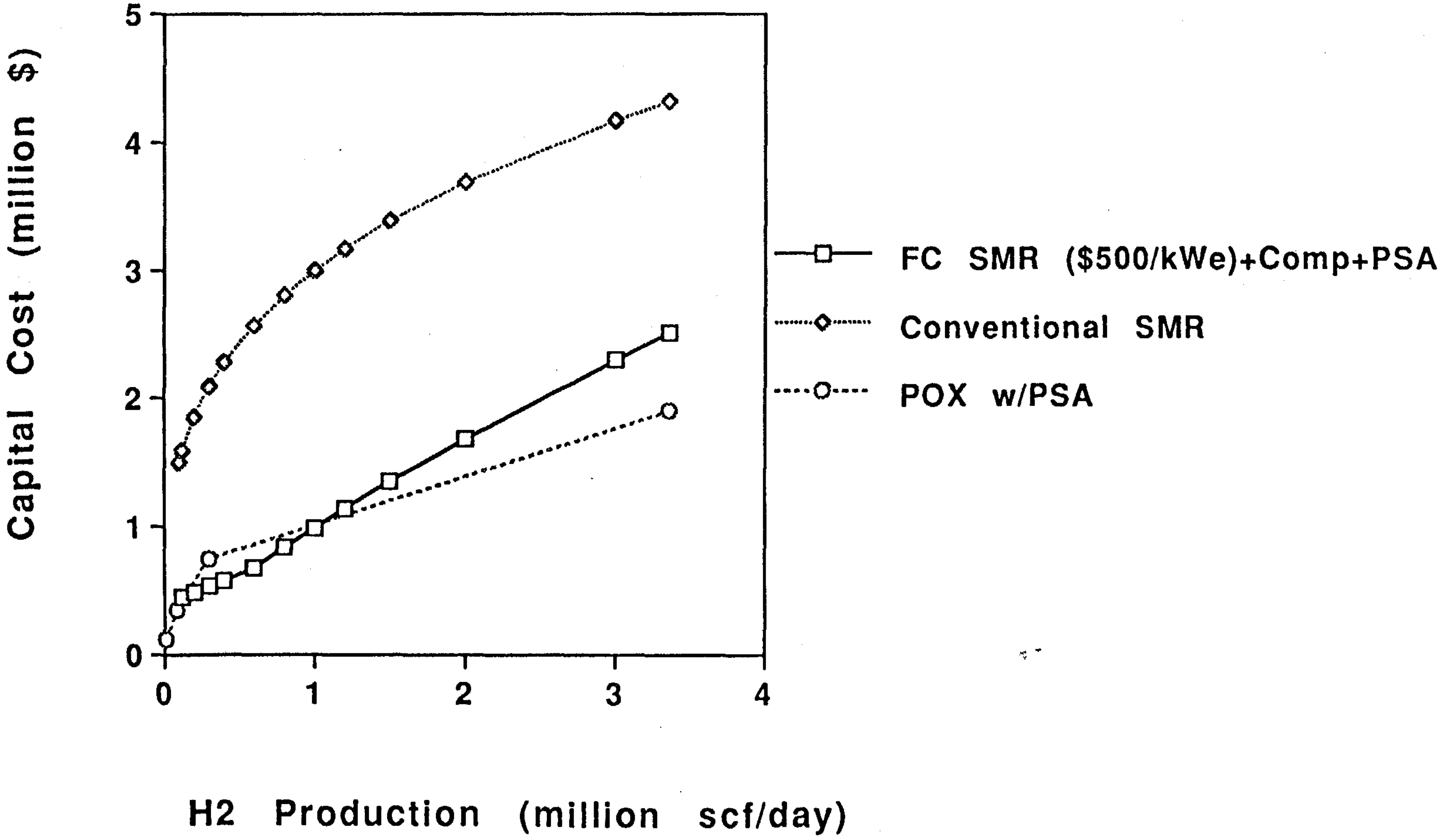
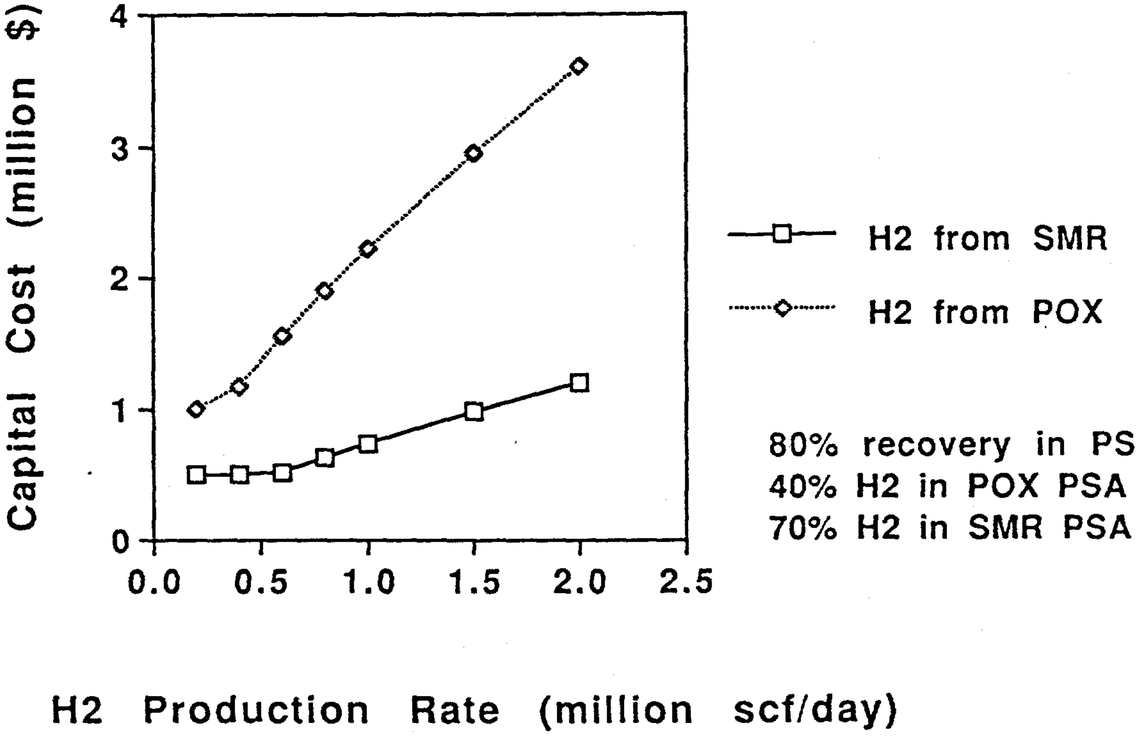


Figure 15.
Capital Cost of Small Scale PSA Units for Use with SMR and POX



80% recovery in PSA
40% H2 in POX PSA Feed
70% H2 in SMR PSA Feed

For commercial one of a kind projects. Costs might be reduced 50% in mass production

2.0. CASE STUDY: DEVELOPMENT OF HYDROGEN TRANSPORTATION FUEL MARKETS IN SOUTHERN CALIFORNIA (TASK 2)

In this Chapter, we examine the potential demand for hydrogen as a transportation fuel in Southern California. We consider how the mandated ZEV market might be served by hydrogen fuel cell vehicles, as well as potential markets for hydrogen buses.

2.1. INTRODUCTION

In 1990 the California Air Resources Board (CARB 1990) mandated that low polluting vehicles must be phased in starting in 1994 (Table 2.1a). The original mandate required that zero emission vehicles (ZEVs) must be offered in starting in 1998. The start date for ZEVs has recently been revised to 2003. (Table 2.1b lists the orginal and revised mandated schedules for introduction of zero emission passenger cars.)

The only near term technologies which could strictly meet the requirement for zero tailpipe emissions are electric battery cars and hydrogen fuel cell cars. (Fuel cell cars with onboard hydrogen production from methanol would be near zero emissions. Hydrogen internal combustion engine cars would also have near zero emissions.) As compared to electric battery vehicles, hydrogen fuel cell vehicles offer the advantages of longer range (a range of 250 miles or 400 km should be possible, even with compressed gas storage), and faster refueling time (several minutes for compressed gas hydrogen refueling as compared to several hours to recharge batteries) (Ogden, Larson and Delucchi 1994, DeLuchi 1992).

Electric battery vehicles are already commercially available, while fuel cell vehicles are still in the demonstration stage. However, there are reasons to believe that fuel cell vehicles could be commercialized over the next few years. Ballard has demonstrated a proton exchange membrane (PEM) fuel cell bus, will be testing small fleets of PEM fuel cell buses starting this year, and plans to commercialize a full sized transit bus based on proton exchange membrane (PEM) fuel cells in 1998. Fuel cell automobiles are also undergoing rapid development in the US, Europe and Japan. The Big Three automakers in the US are assessing fuel cell vehicles as part of the PNGV program. Daimler-Benz is in a joint venture with Ballard has recently demonstrated PEMFC vans (the NECAR-I and NECAR-II), and is considering developing fuel cell automobiles. Developers of fuel cell vehicles have projected that prototype PEMFC cars could be available around 2000 and mass produced fuel cell automobiles might be available as early as 2005-2010.

Given the potential advantages of fuel cell cars as compared to battery cars, it is possible that fleets of fuel cell vehicles could be introduced in California in the early part of the next century, in response to the zero emissions vehicle (ZEV) mandate. In this task we estimate the size of potential hydrogen transportation fuel markets in the Southern California area. We address the following questions:

1) What are projections for ZEV populations in Southern California in the 1996-2010 time frame?

2) What would the projected hydrogen demand be assuming that some fraction of the projected ZEV market is met by hydrogen vehicles?

3) How many hydrogen refueling stations would be needed to meet this projection?

2.2. PROJECTIONS FOR VEHICLE POPULATIONS AND ENERGY USE IN THE LOS ANGELES BASIN

2.2.1. Projections of Zero Emission Vehicle Populations

Data were obtained from the South Coast Air Quality Management District for current and projected numbers of automobiles, vehicle miles traveled, and gasoline consumed in each county (Los Angeles, San Bernadino, Orange and Riverside) in the South Coast Air Basin (Ranji George, private communications 1994, 1995). These are shown in Table 2.2. We see that by 2010, over 9 million passenger cars will be operating in the Los Angeles Basin. (If light trucks are considered, a category which includes the increasingly popular "sport-utility" vehicles, the total number of light duty vehicles is close to 11 million.)

From Tables 2.1 and 2.2, the ZEV population can be estimated, assuming that mandated levels of ZEV passenger cars are introduced on the time scale shown in Table 2.1. We further assume that 1) the projected vehicle population grows linearly, 2) ZEVs have a lifetime of 10 years and 3) the number of ZEVs/year is held fixed at 10% of new vehicles sold each year between 2003 and 2010. Table 2.3 estimates the ZEVs sold per year and the cumulative number of ZEVs as a function of year from 1995-2010 under the original and revised ZEV mandates. Under the original mandate, by 2010 over 800,000 zero emission vehicles would be on the road in the LA Basin. Even though the mandate has been altered to delay introduction of ZEV cars until 2003, the cumulative number of ZEVs is still projected to be over 700,000 in 2010 (see Figure 2.1)

2.3. PROJECTED HYDROGEN DEMANDS

2.3.1. Hydrogen Demand for a Single Fuel Cell Car or Bus

The hydrogen demand for a PEM fuel cell mid-size passenger car is given in Table 2.4. The performance of the PEMFC car is based on estimates by Delucchi (Ogden, Larson and Delucchi 1994, Delucchi 1992). The annual mileage and projected annual energy use is based on average driving patterns in the Los Angeles Basin (see Table 2.2).

The hydrogen demand for a PEMFC bus is estimated in Table 2.4, based on Ballard performance estimates for a PEMFC bus (Larson et.al. 1996), and Los Angeles bus annual mileage (Chaiboonma 1996).

2.3.2. Scenarios for Commercialization of Fuel Cell Vehicles

Projected numbers of fuel cell vehicles can be estimated assuming that some fraction of the potential ZEV market is served. Three possible scenarios for introducing hydrogen fuel cell vehicles are shown in Table 2.5. In the "base case" half the ZEV market is captured by hydrogen fuel cell cars starting in 2003, and 10% of all new buses are fuel cell buses starting in 1998 (the year Ballard has planned to commercialize PEMFC buses). For comparison, earlier and later commercialization scenarios are given (see Table 2.5 and Figure 2.2).

2.3.3. Total Hydrogen Demand for ZEVs in the LA Basin

The total hydrogen demand for the three scenarios above can be estimated using the information in Tables 2.3-2.5. The total hydrogen demand for fuel cell cars and buses is shown for each scenario in Table 2.6. The hydrogen demand for the base case is plotted in Figure 2.3.

By 2010 a hydrogen demand of about 55 million scf/day would have developed, for the assumptions in our base case. This is about as much hydrogen as would be produced at a good sized refinery today. Almost all the hydrogen would be for passenger cars. If 10% of the new bus market goes to PEMFCs starting in 1998, this would amount to about 300 buses by 2010 (about as many buses as would be served by a large bus depot). This would require about 2 million scf/day. If all the buses in the LA Basin were fuel cell buses, this would require about 20 million scf/day.

2.4. A POSSIBLE HYDROGEN REFUELING NETWORK

How many refueling stations would be needed to meet the demand for a growing number of ZEVs?

The answer is fairly straightforward for buses, which are garaged and fueled at night in centralized depots serving 60-200 buses. Here refueling sites would probably be located at or near the bus depots.

For passenger cars, the number of public refueling stations would depend on the geographical concentration of the demand. Our previous studies (Ogden et.al 1995) have indicated that the delivered hydrogen cost is lower when at least 800 cars per day are refueled (or 1 million scf H2/day is dispensed). For Los Angeles driving patterns, this means a total fleet of about 6500 cars would have to be located within acceptable driving distance (perhaps a few miles) of a refueling station If 350,000 fuel cell cars are present (as postulated for 2010 in Scenario I), this would require about 53 stations in the LA area, each serving 800 cars/day. The LA area is about 25 x 50 miles, so this would mean about 1 station for every 25 square miles. The average driver would be about 2-3 miles from a hydrogen station.

Another approach is to make smaller scale hydrogen refueling stations more economically attractive. The idea of garage refueling for hydrogen cars has been proposed by Directed Technologies, Inc., using small electrolyzers. Improved small scale reformer technology may also offer lower costs (see Chapter 1).

2.5 CONCLUSIONS AND RECOMMENDATIONS FOR FUTURE WORK

The results can be summarized as follows:

* If fuel cell vehicles capture a significant fraction of the emerging ZEV market, several hundred thousand fuel cell cars could be on the road in the Los Angeles area by 2010.

* If hydrogen fuel cell cars captured half the ZEV market, about 50 million scf of hydrogen per day would be required, a capacity similar to that in a good sized oil refinery.

* The type of refueling network would depend on the geographical concentration of the demand, as well as the relative economics of various supply options.

A detailed study of hydrogen supply and demand in Southern California is needed to fully address the implications of the ZEV market for hydrogen. This will be reported in a separate report which is forthcoming (Ogden et.al 1996).

2.6. ACKNOWLEDGMENTS

For useful conversations, the authors would like to thank Archana Agrawal (California Air Resources Board), Eck Chaiboonma (Los Angeles Metropolitan Transit Authority), Mark Delucchi (UC Davis), Ranji George (South Coast Air Quality Management District), Paul Lancaster (Ballard Power Systems), Eric Larson (Princeton University), Jon Leonard (SCAQMD), Alan Lloyd (Desert Research Institute), Margaret Steinbugler (Princeton University), Sandy Thomas (Directed Technologies, Inc.), and Robert Williams (Princeton University).

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Ogden, J.M., T. Kreutz, M. Steinbugler, A. Cox and J. White, "Case Study of Developing a Hydrogen Vehicle Refueling Infrastructure in Southern California," annual report for USDOE Contract No. DE-FG36-95G010061, forthcoming 1996.

TABLE 2.1. ORIGINAL MANDATED (1990) IMPLEMENTATION RATES FOR LOW POLLUTING PASSENGER CARS IN THE SOUTH COAST AIR BASIN

MODEL YEAR	NOx =0.39 g/mi	NOx = 0.25 g/mi	TLEV	LEV	ULEV	ZEV
1994	10%	80%	10%			
1995		85%	15%			
1996		80%	20%			
1997		73%		25%	2%	
1998		48%		48%	2%	2%
1999		23%		73%	2%	2%
2000				96%	2%	2%
2001				90%	5%	5%
2002				85%	10%	5%
2003				75%	15%	10%

SOURCE: A.Lloyd, South Coast Air Quality Management District, 1991.

TABLE 2.1.b. ORIGINAL (1990) AND REVISED (1996) IMPLEMENTATION RATES FOR ZEVS IN CALIFORNIA

Year	Original ZEV Mandate	1996 Revised ZEV Mandate
1998	2%	0%
1999	2%	0%
2000	2%	0%
2001	5%	0%
2002	5%	0%
2003	10%	10%

TABLE 2.2.
DATA AND PROJECTIONS FOR VEHICLE POPULATIONS, FUEL ECONOMY, ANNUAL MILEAGE AND ENERGY USE FOR PASSENGER CARS, LIGHT TRUCKS AND URBAN BUSES IN THE SOUTH COAST AIR BASIN

Passenger Cars					
Year	# Vehicles	Average Fuel Economy (mpg)	Average Miles/Yr/ Vehcle	Average Energy Use/Yr/ Vehicle (GJ/yr)	Energy Use All Passenger Cars (EJ/yr)
1995	7,419,502	23.6	11,311	62.7	0.46
2000	8,141,691	25.8	11,379	57.7	0.47
2005	8,753,995	27.6	11,035	52.3	0.46
2010	9,365,800	29.0	10,724	48.4	0.45

Light Trucks					
Year	# Vehicles	Average Fuel Economy (mpg)	Average Miles/Yr/ Vehicle	Average Energy Use/Yr/ Vehicle (GJ/yr)	Energy Use (EJ/yr)
1995	1,368,212	17.1	11,854	90.7	0.12
2000	1,513,177	18.0	11,960	86.9	0.13
2005	1,639,484	18.3	11,633	83.1	0.14
2010	1,765,701	18.4	11,213	79.7	0.14

Urban Buses					
Year	# Vehicles	Average Fuel Economy (mpg)	Average Miles/Yr/ Vehcle	Average Energy Use/Yr/ Vehicle (GJ/yr)	Energy Use (EJ/yr)
1995	2926	3.5	50,646	1892	0.0034
2000	3076	3.5	50,668	1894	0.0034
2005	3188	3.5	50,720	1895	0.0033
2010	3300	3.5	50,658	1893	0.0033

Source: Ranji George, SCAQMD, private communications 1995. Passenger cars and light trucks under 3750 lb. are subject to the ZEV mandate.

TABLE 2.3. PROJECTED NUMBERS OF ZEV PASSENGER CARS IN THE SOUTH COAST AIR BASIN 1995-2010

WITH ORIGINAL ZEV MANDATE

YEAR	ZEVS AS % OF NEW CARS SOLD EACH YEAR	ZEV PASSENGER CARS/YR	CUMULATIVE # ZEV PASSENGER CARS
1995	0	0	0
1996	0	0	0
1997	0	0	0
1998	2%	16,137	16,137
1999	2%	16,386	32,524
2000	2%	16,635	49,158
2001	5%	42,148	91,306
2002	5%	42,709	134,016
2003	10%	86,541	220,557
2004	10%	87,663	308,220
2005	10%	88,785	397,005
2006	10%	89,908	486,913
2007	10%	91,030	577,943
2008	10%	92,152	670,095
2009	10%	93,275	763,370
2010	10%	94,397	857,767

WITH REVISED ZEV MANDATE

YEAR	ZEVS AS % OF NEW CARS SOLD EACH YEAR	ZEV PASSENGER CARS/YR	CUMULATIVE # ZEV PASSENGER CARS
1995	0	0	0
1996	0	0	0
1997	0	0	0
1998	0	0	0
1999	0	0	0
2000	0	0	0
2001	0	0	0
2002	0	0	0
2003	10%	86,541	86,541
2004	10%	87,663	174,204
2005	10%	88,785	262,989
2006	10%	89,908	352,897
2007	10%	91,030	443,927
2008	10%	92,152	536,080
2009	10%	93,275	629,354
2010	10%	94,397	723,751

TABLE 2.4. ASSUMED CHARACTERISTICS OF FUEL CELL VEHICLES

	PEM FC Bus	PEM FC Car
Fuel economy	52 scf H2/mile = 6.9 mpg Diesel equivalent [a]	71.4 mpg gasoline equiv.[b]
Miles/yr	50,000[c]	11,140[d]
Fuel Storage	H2 gas @3600 psi	H2 gas @5000 psi
Hydrogen stored onboard (scf)	13,000[a]	1200
Range (mi)	250[a]	250
Energy use per year (GJ/yr)[e]	976	20
Hydrogen use per year (million scf/yr)[f]	2.60	0.056

a. Based on the efficiency of the Ballard Phase II PEMFC bus (Larson, Worrell, and Chen 1996). The mile per gallon gasoline equivalent efficiency for a fuel cell vehicle is estimated assuming that 1 gallon of gasoline contains 0.1308 GJ (HHV) and that 1 scf of hydrogen contains 362 kJ (HHV).

b. Based on estimates by Delucchi for a PEMFC automobile (Ogden, Larson and Delucchi 1994).

c. Typical annual mileage for a bus in the LA Basin (E. Chaiboonma, LA Metropolitan Transit Authority, private communications 1995, 1996).

d. Typical annual mileage for a passenger car in the LA Basin. (R. George, SCAQMD, private communications 1995, 1996).

e. Energy use was estimated assuming that the HHV of gasoline is 0.1308 GJ/gallon.

f. Hydrogen use was estimated based on the HHV of hydrogen, 362 GJ = 1 million scf

TABLE 2.5. POSSIBLE SCENARIOS FOR INTRODUCTION OF FUEL CELL VEHICLES (FCVS) IN THE LOS ANGELES BASIN

Scenario I (base case)	FCVs = 50% of ZEV Market = 5% of all new passenger cars, starting in 2003 10% of new Buses = FCVs, starting in 1998
Scenario II (earlier introduction of FCVs)	FCVs = 100% of ZEV Market = 10% of all new passenger cars, starting in 2003 100% of new Buses = FCVs, starting in 1998
Scenario III (later introduction of FCVs)	FCVs = 50% of ZEV Market = 5% of all new passenger cars, starting in 2005 10% of new Buses = FCVs, starting in 2000

Table 2.6.

NUMBERS OF ZEVS AND PROJECTED HYDROGEN DEMAND (IN MILLION SCF H2/DAY) IN THE LA BASIN FOR THREE DEMAND SCENARIOS (SEE TABLE 2.5)

Year	Scenario I				Scenario II				Scenario III			
	Cum. # Cars 1000s	H2 for Cars	Cum. # Buses	H2 for Buses	Cum. # Cars 1000s	H2 for Cars	Cum. # Buses	H2 for Buses	Cum. # Cars 1000s	H2 for Cars	Cum. # Buses	H2 for Buses
1998	0	0	30	0.21	0	0	300	2.1	0	0	0	0
1999	0	0	60	0.43	0	0	600	4.3	0	0	0	0
2000	0	0	90	0.64	0	0	900	6.4	0	0	30	0.21
2001	0	0	120	0.85	0	0	1200	8.5	0	0	60	0.43
2002	0	0	150	1.07	0	0	1500	10.7	0	0	90	0.64
2003	42.5	6.6	180	1.28	85.1	13.2	1800	12.8	0	0	120	0.85
2004	85.7	13.2	210	1.50	171	26.4	2100	15.0	0	0	150	1.07
2005	129	19.8	240	1.71	259	39.6	2400	17.1	43.7	6.7	180	1.28
2006	174	26.4	270	1.92	348	52.9	2700	19.2	88.1	13.4	210	1.50
2007	219	33.1	300	2.13	438	66.2	3000	21.3	133	20.1	240	1.71
2008	264	39.8	300	2.13	529	79.5	3000	21.3	179	26.9	270	1.92
2009	311	46.5	300	2.13	621	92.9	3000	21.3	225	33.6	300	2.13
2010	358	53.2	300	2.13	713	106.3	3000	21.3	272	40.4	300	2.13

Figure 2.1.
Cumulative Number of Zero Emission
Passenger Cars in the LA Basin

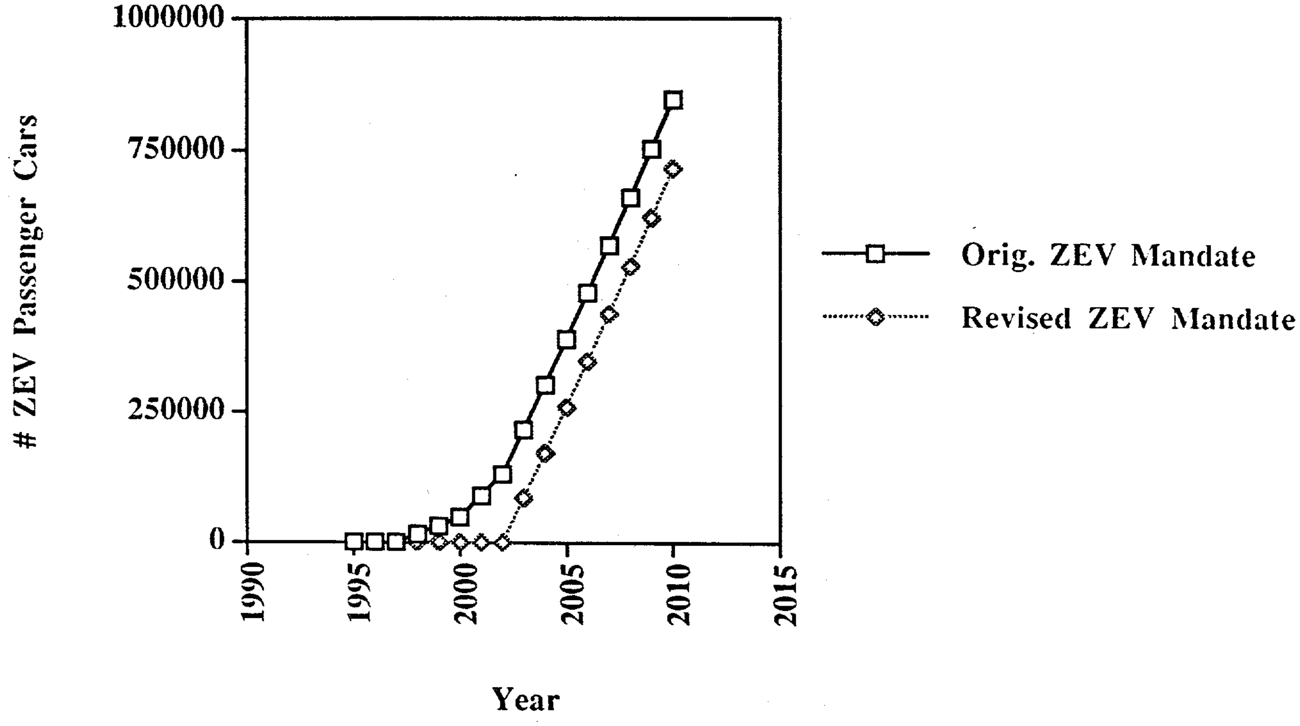


Figure 2.2
Projections for Fuel Cell Passenger Cars in the LA Basin for Various Commercialization Scenarios

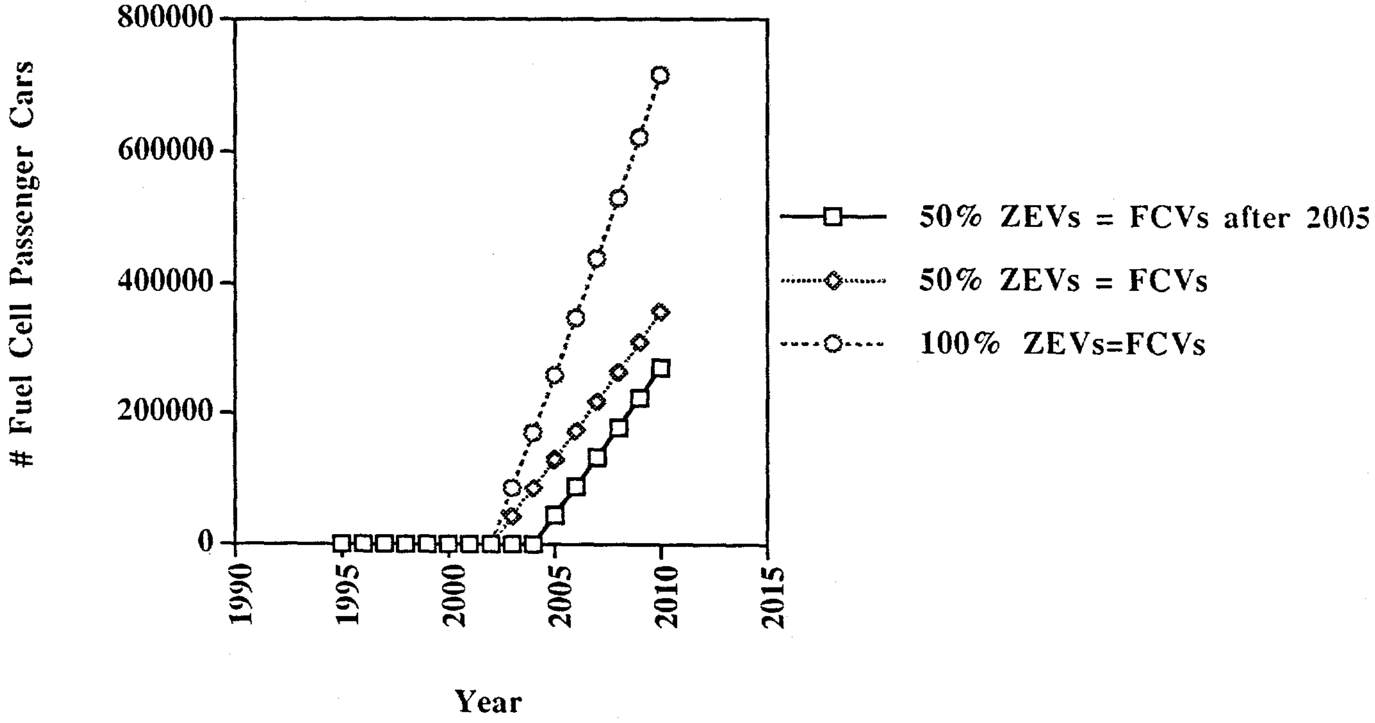
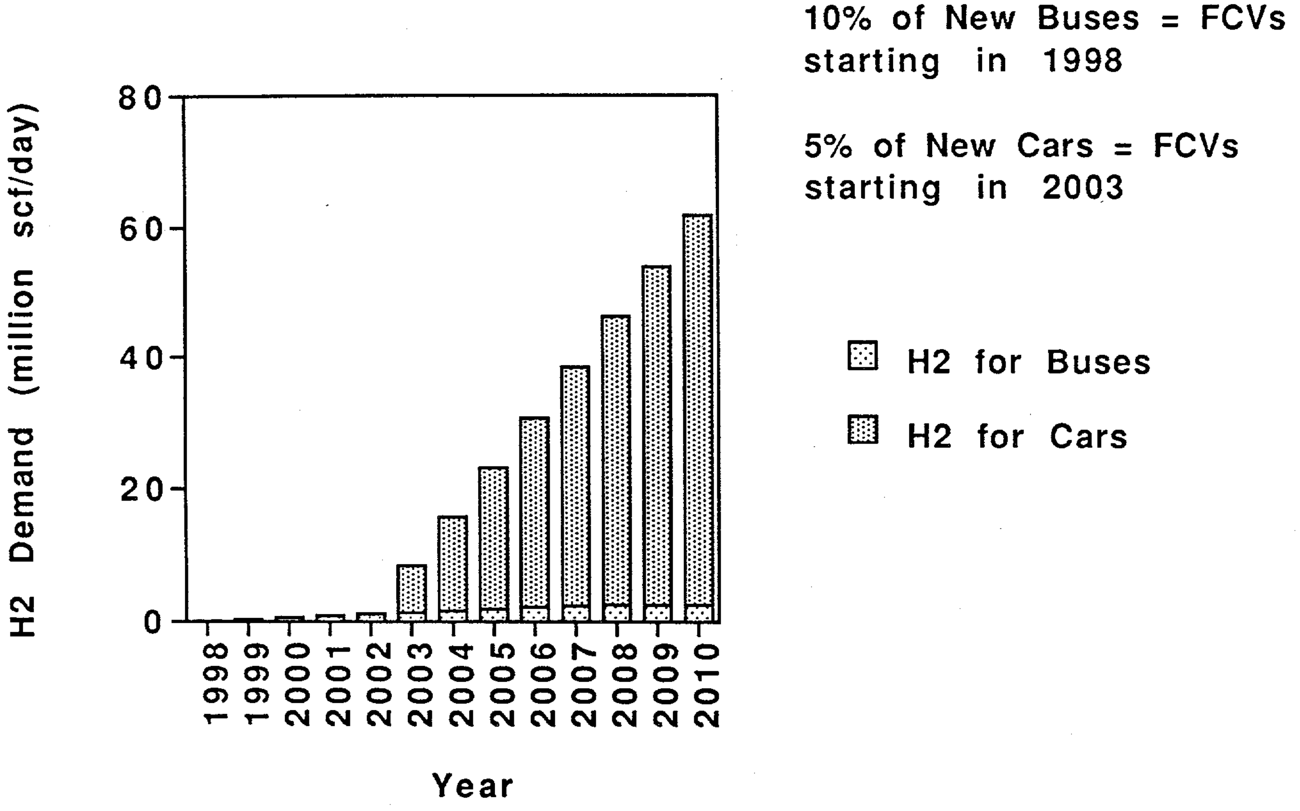


Figure 2.3.
A PossibleScenario for
Introducing Fuel Cell Vehicles
in the LA Basin



10% of New Buses = FCVs
starting in 1998

5% of New Cars = FCVs
starting in 2003

HYDROGEN ENERGY SYSTEMS STUDIES

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ABSTRACT

Hydrogen is being considered as a low polluting energy carrier, especially for use in zero emission vehicles. If hydrogen is derived from renewable sources (biomass, solar, wind), it would be possible in principle to produce and use energy on a large scale with greatly reduced greenhouse gas emissions and very little local pollution.

The results of our previous studies suggest that use of hydrogen from natural gas might be an important first step toward a hydrogen economy based on renewables. Because of infrastructure considerations (the difficulty and cost of storing, transmitting and distributing hydrogen), hydrogen produced from natural gas at the end-user's site could be a key feature in the early development of hydrogen energy systems. In the first chapter of this report, we assess the technical and economic prospects for small scale technologies for producing hydrogen from natural gas (steam reformers, autothermal reformers and partial oxidation systems), addressing the following questions:

1) What are the performance, cost and emissions of small scale steam reformer technology now on the market? How does this compare to partial oxidation and autothermal systems?

2) How do the performance and cost of reformer technologies depend on scale? What critical technologies limit cost and performance of small scale hydrogen production systems? What are the prospects for potential cost reductions and performance improvements as these technologies advance?

3) How would reductions in the reformer capital cost impact the delivered cost of hydrogen transportation fuel?

In the second chapter of this report we estimate the potential demand for hydrogen transportation fuel in Southern California.