# Methanol and Hydrogen from Biomass for Transportation, with Comparisons to Methanol and Hydrogen from Natural Gas and Coal

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*Center for Energy and Environmental Studies*
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# METHANOL AND HYDROGEN FROM BIOMASS FOR TRANSPORTATION, WITH COMPARISONS TO METHANOL AND HYDROGEN FROM NATURAL GAS AND COAL[1]

**Robert H. Williams**
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[1] This article is based on a paper of the same title presented at BioResources '94, Oct. 3-7, 1994, Bangalore, India. A shorter version appears in *Energy for Sustainable Development*, I(5), January 1995.

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# ABSTRACT

Methanol and hydrogen produced from biomass via indirectly heated gasifiers and their use in fuel cell vehicles would make it possible for biomass to be used for road transportation, with zero or near zero local air pollution and very low levels of lifecycle $CO_2$ emissions, if the biomass feedstock is grown sustainably. Moreover, because this approach to using biomass for transport fuels involves such an efficient use of land, it offers the potential for making major contributions in reducing dependence on insecure sources of oil in transportation.

Biomass-derived methanol and hydrogen would be roughly competitive with these fuels produced at much larger scale (to exploit scale economies) from coal, even with relatively high biomass feedstock prices. While biomass-derived methanol and hydrogen would not be able to compete with the production of these fuels from natural gas in the near term, natural gas prices are expected to rise substantially over the next decade or so. With natural prices expected by the year 2010, biomass would be nearly competitive with natural gas in the production of these fuels. A carbon tax that would increase the cost of owning and operating fuel cell vehicles on natural gas-derived fuels by less than 2% would be adequate to tip the economic balance in favor of biomass.

The production of methanol for export in developing regions could provide sustainable new income streams for rural areas in developing regions while bringing competition and fuel price stability to world markets in transport fuels.

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## Introduction

Deteriorating urban air quality, growing dependence on insecure energy sources, and global warming are major challenges forcing reexamination of the petroleum-fueled internal combustion engine vehicle (ICEV) as the basis for road transportation throughout the world.

Deteriorating urban air quality is the main driver forcing an examination of alternative options for meeting road transport needs today. To meet air quality goals in the face of continuing growth in transport fuel demand, tailpipe emission standards have been progressively tightened in many countries. But tailpipe emission control devices have been getting ever more complicated, expensive, and difficult to maintain. California regulators have decided that air quality goals for the state cannot be met simply by requiring further incremental reductions in tailpipe emissions and that more radical measures are needed; the state has mandated that 2% of new cars must be "zero-emission vehicles" by 1998, rising to 10% by 2003. Twelve eastern states collectively have asked the US Environmental Protection Agency to impose the California standards on them. In light of increasing air pollution in megacities like Mexico City and Sao Paulo (UNEP/WHO, 1992), such policies are likely to become widespread.

The spectre of energy insecurity that haunted energy planning in the 1970s is reemerging. Energy import dependence is rising (*e.g.* increasing in the US from 10 to 20% of total demand, 1985-1993), and the share of world oil production coming from the Middle East is growing (increasing from 18 to 28%, 1985-1993). While soft oil prices may persist for a few years, supplies will eventually tighten because of rapidly growing transport energy demand in developing countries and limited conventional oil resources outside the Middle East; the US Geological Survey projects that oil production will decline after the turn of the century in all regions except the Middle East (Masters, 1990).

For the long run, the most daunting challenge may be global warming. If society should decide to respond by trying to stabilize the atmosphere, it will be necessary to reduce carbon dioxide ($CO_2$) emissions from fossil fuel burning by 60% or more (IPCC, 1990).

Methanol (MeOH) and hydrogen ($H_2$) derived from biomass offer the potential for making major contributions to transport fuel requirements by addressing competitively all of these challenges, especially when used in fuel cell vehicles (FCVs).

## Fuel Cell Vehicles

In a fuel cell the chemical energy of fuel is converted directly into electricity without first burning the fuel to generate heat to run a heat engine. The fuel cell offers a quantum leap in energy efficiency and the virtual elimination of air pollution without the use of emission control devices. Dramatic technological advances for the proton exchange membrane (PEM) fuel cell in particular have focussed attention on this technology for motor vehicles. The FCV has the potential to compete with the petroleum-fueled ICEV in cost and performance (AGTD, 1994; Williams, 1993, 1994), while effectively addressing air quality, energy insecurity, and global warming concerns.

Ballard Power Systems, Inc., of Vancouver, Canada, introduced a prototype PEM fuel cell bus in 1993 and plans to introduce PEM fuel cell buses on a commercial basis beginning in 1998. In April 1994 Germany's Daimler Benz introduced a prototype PEM fuel cell light-duty vehicle (a van) and announced plans to develop the technology for commercial automotive applications. In the US, the FCV is a leading candidate technology for accelerated development under the "Partnership for a New Generation of Vehicles," a joint venture launched in September 1993 between the Clinton Administration and the US automobile industry; the partnership's goal is to develop in a decade production-ready prototypes of advanced, low-polluting, safe cars that could be run on secure energy sources, especially renewables, that would have up to three times the fuel economy of today's gasoline ICEVs of comparable performance, and that cost no more to own and operate.

Fuel cells used in vehicles would probably use as fuel either hydrogen[3] or a hydrogen carrier that is converted into $H_2$ onboard the vehicle. Hydrogen FCVs would be about three times as energy-efficient as comparable ICEVs (measured in km/liter of gasoline-equivalent). Hydrogen could be carried on the vehicle as liquefied $H_2$, in a metal hydride that releases $H_2$ when heated, or as compressed gas. The latter is the storage option favored at present.[4] Alternatively, an $H_2$ carrier could be delivered to and stored onboard the vehicle, to be converted to $H_2$ just before use. In liquid form such carriers (e.g. various alcohol and hydrocarbon fuels) are much easier to store and transport than gaseous $H_2$.

Methanol (MeOH)[5] is the preferred liquid carrier at present. Via steam reforming, MeOH can be converted to a mixture of $H_2$ and $CO_2$,[6] from which the fuel cell extracts the $H_2$ fuel. Methanol-

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[3] Hydrogen has a molecular weight of 2.016 kg/kmol, a higher heating value (HHV) of 141.77 MJ/kg, and a density of 0.0888 kg/m$^3$ at 0.101 MPa and 273 K.

[4] The storage challenge posed by gaseous $H_2$ for FCV applications is similar to that for ICEVs fueled with compressed natural gas: while $H_2$ has about 1/3 the heating value of natural gas, the fuel economy of FCVs (in km/liter of gasoline-equivalent) is about 3 times that for comparable ICEVs.

[5] Liquid methanol ($CH_3OH$) has a molecular weight of 32.04 kg/kmol, a higher heating value of 22.67 MJ/kg, and a density of 797 kg/m$^3$.

[6] In a steam reformer a mixture of MeOH and steam ($H_2O_{(g)}$) is heated over a catalyst and thereby converted into a mixture of $H_2$ and carbon monoxide (CO). Then, in "shift reactors," the fuel energy of the CO is "shifted" to $H_2$ via the "water-gas shift" reaction: $H_2O + CO \leftrightarrow H_2 + CO_2$. The resulting mixture of $H_2$ and $CO_2$ is delivered to the anode of the fuel cell. In passing the anode, the fuel cell extracts and consumes most of the $H_2$. The residual $H_2$ in the anode exhaust is burned to provide heat (through a heat exchanger) for the reformer.

fueled FCVs would be about 2 1/2 times as energy efficient as comparable ICEVs--somewhat less than for $H_2$-fueled FCVs because of energy losses in fuel reforming. Steam reforming conditions make this option for use onboard the vehicle more problematic for most other liquid fuels.[7] However, partial oxidation (POX) fuel converters could be used to reform a wide range of liquid hydrocarbons onboard vehicles. POX reforming entails some efficiency penalty,[8] but their use might facilitate the initial widespread introduction of FCVs by allowing them to operate on conventional hydrocarbon fuels (gasoline or diesel), with an eventual transition to more efficient FCVs operating on steam-reformed methanol or pure hydrogen.

For $H_2$-fueled FCVs local air pollutant emissions from the vehicle would be zero. For MeOH-fueled FCVs, these emissions would be non-zero but miniscule compared to ICEVs.

Methanol and $H_2$ are made commercially today in relatively small quantities from natural gas, mainly for chemical and petroleum refining markets. If large-scale fuel markets develop for MeOH and/or $H_2$, natural gas is likely to be the feedstock of choice initially. But when natural gas prices rise to appropriate levels, biomass- and coal-derived MeOH and $H_2$ will become competitive.

This paper explores the technology, economics, and environmental aspects of producing MeOH and $H_2$ from biomass for FCV applications, in relation to their production from natural gas and coal. But before getting into the details of this analysis, this biofuels strategy is discussed in relation to alternative biofuels strategies for transportation.

## Biomass-Derived Methanol and Hydrogen for FCVs vs. Alternative Biofuels Strategies

Methanol and $H_2$ derived from biomass can be used in ICEVs as well as FCVs, and such applications might well be pursued before FCVs become widely available. However, these biomass-derived fuels are far less attractive when used in ICEVs. One reason is that the high efficiency of the FCV makes it possible for biomass to contribute more than twice as much in displacing petroleum as with ICEVs, for a given amount of land (Table 1)--especially important in light of the fact that land

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[7] For example, to reform ethanol to $H_2$ requires temperatures in excess of 500 °C. The temperature mismatch between the on-board reformer and a PEM fuel cell, which operates near 100°C, gives rise to efficiency penalties and operational complexities. In contrast, reforming MeOH is relatively easy, because much lower temperatures are required (250 °C or less).

[8] In partial oxidation reforming, fuel is burned at relatively high temperatures in a limited supply of air, generating a gaseous mixture consisting mainly of $H_2$, CO, $CO_2$, and nitrogen ($N_2$). After partial oxidation, the CO is combined (in shift reactors) with $H_2O_{(g)}$ to form more $H_2$ and $CO_2$. The resulting gaseous stream is then delivered to the anode of the fuel cell where most of the $H_2$ is extracted for fuel.

With partial oxidation fuel converters, it is generally difficult to make use of the residual $H_2$ energy in the anode exhaust. As a result partial oxidation converters are less energy-efficient than steam reformers.

use constraints will ultimately limit the use of biomass for energy.[9] Another is that the use of these fuels in ICEVs offers limited air quality benefits. With MeOH-fueled ICEVs, CO emissions would be less and emissions of volatile organic compounds would be less problematic than for gasoline, but emissions of nitrogen oxides ($NO_x$) would probably not be reduced, and there would probably be some formaldehyde emissions. Moreover, it is now generally believed that MeOH, or any alcohol fuel, especially when blended with gasoline and used in flexible-fuel ICEVs, offers little or no air quality advantage relative to gasoline, other than reduced CO emissions (Calvert et al., 1993). Moreover, reformulated gasolines can meet or surpass reductions in the emission reactivity of alcohol-gasoline blends (BEST, 1991). Hydrogen would be a relatively clean fuel for ICEV applications--the only significant pollutant being $NO_x$ produced under lean combustion conditions (desirable to get high fuel economy); with only one pollutant to worry about, it would be possible, with appropriate control technology, to reduce its emissions to very low levels. However, because of the bulk of the storage system ($H_2$ has a relatively low volumetric energy density) and the relatively low efficiency of ICEVs, $H_2$-fueled ICEVs would have very limited range.

Very little attention has been given to MeOH or $H_2$ derived from biomass for vehicular applications. The focus of biofuels development has been largely on the production of ethanol via fermentation from grain, sugar beets, or sugar cane, or the production of rape methyl ester from rape seed oil. Such biofuels have attracted attention largely because the crops involved are familiar to farmers, and producing them for energy provides an alternative way for the farmer to make a living.

The largest commercial ethanol programs are in Brazil and the US. In 1989 Brazil produced from sugar cane 12 billion liters of fuel ethanol, which was used to support 4.2 million cars running on hydrated ethanol and the remaining 5 million cars on gasohol, a gasoline-ethanol blend (Goldemberg et al., 1993). In 1993 the US produced 4 billion liters of ethanol from maize, for gasohol applications. In both instances the technologies are presently uneconomic. However, substantial cost reductions are being made for cane-derived ethanol (Goldemberg et al., 1993), and there are good prospects for making it competitive at the present world oil price, if electricity is cogenerated from

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[9] If biomass is used to provide fuels for fuel cell vehicles, the time when land use constraints will begin limiting the further expansion of biomass production in providing fuels for transport can be pushed far into the future. To see this, note that in 1985 there were 389 million automobiles in the world, having an average fuel economy of 7.42 km/l (17.5 mpg), and consuming 27 EJ/year of petroleum (Lashoff and Tirpak, 1990). If this many cars had been methanol-fueled (hydrogen-fueled) FCVs with the fuel economies indicated in Table 1 (note a), primary biomass requirements would have been 12 EJ/year (9 EJ/year). Producing this much biomass would require about 40 (30) million hectares of plantations, assuming an average productivity of 15 dry tonnes/hectare/year. The amount of land worldwide that might be dedicated to biomass production for energy is of the order of 500 million hectares (Johansson et al., 1993).

cane residues with advanced technology along with ethanol from cane juice (Williams and Larson, 1993). But the prospects are poor for making ethanol economically from grain (Wyman et al., 1993).

The production of ethanol from grain or sugar beets and the production of rape methyl ester from rape seed oil also represent inefficient uses of land, with low yields of transport services (vehicle-km/hectare/year) compared to what is achievable with advanced technologies using woody biomass feedstocks (Table 1). These options also have marginal energy balances; for the Netherlands (energy output/fossil fuel input) ratios are 0.9 and 1.2 for ethanol from winter wheat and sugarbeets, respectively, and 1.8 for rape methyl ester derived from rape seed (Lysen et al., 1992). Greenhouse performance also tends to be marginal. For maize, the net fuel cycle emissions of $CO_2$ are approximately the same as from gasoline (Ho, 1989; Segal, 1989); total life cycle GHG emissions may even be somewhat higher than for reformulated gasoline (DeLuchi, 1991). In the case of rape methyl ester, life cycle $CO_2$ emissions are less than with fossil fuels (Lysen et al., 1992).

The problems posed by conventional "agro-fuels" have provided the motivation for an R&D effort aimed at developing enzymatic hydrolysis technology for the production of ethanol from ligno-cellulosic feedstocks, which can be grown at much lower cost than grain or sugar beets. Moreover, for a woody feedstock yield of 15 dry tonnes/hectare/year, which is generally believed to be achievable in large-scale production, the ethanol yield could be more than twice that from grain (Table 1). If the US Department of Energy's year-2000 goals for performance and cost are met, ethanol produced this way would be competitive with gasoline for oil prices less than $25/barrel; energy balances would also be favorable; life-cycle emissions of $CO_2$ for ethanol production and use in ICEVs (in g C/km) would be only about 2% of those from such vehicles operated on reformulated gasoline (Wyman et al., 1993). A shortcoming of the technology is that, when used in ICEVs, it may not lead to air quality improvements beyond what can be achieved with reformulated gasoline. (Ethanol offers less air quality benefits than methanol.) Also, while the yield (useful fuel/hectare/year) may prove to be comparable to that for MeOH derived from ligno-cellulosic feedstocks (Table 1), the lower efficiency of using ethanol in FCV applications (because fuel conversion would have to be by partial oxidation instead of steam reforming) restricts its attractiveness vis a vis MeOH.

In the remainder of this report, it is argued that the use of biomass-derived methanol or hydrogen in fuel cell vehicles offers is a way to use a renewable energy resource as a transportation fuel, with high efficiency, low emissions of local pollutants and greenhouse gases, and prospectively compteitive costs.

## Process Technology for Methanol and Hydrogen Production

The production of liquid MeOH or gaseous $H_2$ from biomass, coal or natural gas involves several similar basic processing steps. With biomass or coal, the feedstock is first gasified (by heating it to above 700 °C in the presence of little or no oxygen) into a synthesis gas (syngas) consisting of CO, $H_2$, $CO_2$, $H_2O_{(g)}$, and in some cases methane ($CH_4$) and small quantities of other hydrocarbons. The syngas exiting the gasifier is cooled and then quenched with a water spray to remove particulates and other contaminants. Additional cleanup of sulfur (S) compounds (especially important with coal) is needed to prevent poisoning of downstream catalysts. The syngas then undergoes a series of chemical reactions that lead to the desired end product. The equipment downstream of the gasifier for conversion to MeOH or $H_2$ is the same as that used to make MeOH or $H_2$ from natural gas.

All equipment for fuels production from coal is commercially available today. For biomass, suitable gasification technology is being developed. As a feedstock for gasification, biomass has some advantages relative to coal: it is more reactive and thus easier to gasify, and most biomass has a sufficiently low S content that costly S removal systems are not required. On the other hand, while coal facilities can benefit from economies of scale in capital costs, the higher cost of transporting biomass (due to its low volumetric energy density and the low efficiency of photosynthesis) will constrain biomass facilities to more modest sizes (Larson, 1993).

The processing of natural gas or syngas that contains hydrocarbons begins with reforming the hydrocarbons, i.e. converting them to CO and $H_2$, usually by reacting the syngas with $H_2O_{(g)}$ over a catalyst at close to 900 °C. [Reforming is not needed with some coal and biomass gasifier designs (discussed below), because no significant quantities of hydrocarbons are present in the syngas.] The CO:$H_2$ ratio of the gas is next adjusted to the level required for final processing using one or more shift reactors, wherein $H_2O_{(g)}$ reacts with CO over a catalyst to produce $H_2$ and $CO_2$.

For MeOH production (Fig. 1), a single shift reactor operating at 450 to 550 °C produces a molar $H_2$:CO ratio of about 2. (With natural gas as the feedstock, the high H:C ratio eliminates the need for the shift reactor.) Following the shift, the gas passes through a reactor containing a solvent that selectively dissolves $CO_2$ and $H_2O_{(g)}$ out of the gas. With $CO_2$ and $H_2O_{(g)}$ removed, the gas stream is then compressed to about 100 bar and fed to a MeOH synthesis reactor, wherein CO and $H_2$ combine over a catalyst at about 250 °C to form MeOH. Conversion of carbon as CO to carbon as MeOH is typically in excess of 98%. Water is removed from the MeOH in a final distillation step.

For $H_2$ production (Fig. 1), two shift reactors in series (the first operating at about 450 °C and the second at about 230 °C) convert to $H_2$ as much of the CO leaving the reformer as possible. The gas then enters a pressure swing adsorber (PSA), which separates gases by exploiting the ability of specially designed porous materials to selectively adsorb specific molecules at high pressure and desorb them at low pressure. The first PSA bed adsorbs $CO_2$ and $H_2O$, and the second desorbs all remaining components except $H_2$. Up to 97% of the $H_2$ fed to a PSA can be recovered as final product with greater than 99.999% purity. The $H_2$ is then compressed for storage or pipeline transmission.

## Biomass gasifiers

Since all process equipment needed to produce MeOH or $H_2$ from biomass is well established in commercial use, except for the gasifier, the discussion here emphasized the development status and projected performance of alternative gasifiers. Biomass gasifiers operate by direct or indirect heating. The former, which involves partial oxidation of the feedstock, is the basic principle used in coal gasification; the latter is not suitable for most coals due to their low reactivities compared to biomass.

*Directly-heated gasifiers.* Directly-heated gasifiers use air or oxygen to burn some of the feedstock *in situ*, thereby providing the heat needed to gasify the remaining feedstock. In the production of MeOH or $H_2$, oxygen is preferred, so as to minimize the gas volumes that must be treated downstream. (The presence of $N_2$ also reduces the efficiency of $H_2$ separation in the PSA step.) A disadvantage of $O_2$ use is increased costs. Because of the sensitivity to scale of capital costs for $O_2$ plants, there is a cost penalty with $O_2$ that is acute at the smaller scales that would characterize biomass facilities.

Fixed-bed, fluidized-bed, and entrained-bed gasifiers were originally developed for coal (Simbeck et al., 1993; Synthetic Fuels Associates, 1983). High-temperature entrained-bed gasification of biomass has not been the focus of any hardware development efforts, but there have been various efforts to adapt fixed and fluidized-bed designs to biomass. A number of atmospheric-pressure fluidized-bed units are commercially operating with biomass to supply gas for boilers or furnaces (Larson, 1993). For production of transport fuels from biomass, the fluidized-bed gasifier is likely to be the technology of choice among directly-heated gasifier designs, some reasons for which are discussed below.

In a fluidized-bed gasifier the feedstock, typically fed through the sidewall of the reactor, mixes into a bed of inert heat carrying material, such as sand, that is maintained in a fluidized state by injection of $O_2$ (and sometimes $H_2O_{(g)}$) from below. Fluidized-bed gasifiers have the flexibility to accept feedstocks with a wide range of sizes and bulk densities. Ash and unreacted char are carried out of the top of the reactor and separated from the product gas in a cyclone. In a bubbling fluidized-bed (Fig. 2a), the solids are removed via the cyclone for disposal. In a circulating fluidized-bed (Fig. 2b) the cyclone is an integral part of the gasifier, making it possible to recycle the solids for further reaction. In both designs, the product gas exits at 900-1000 °C--essentially the same as the temperature throughout the bed. The high average bed temperature facilitates cracking of high molecular weight hydrocarbons (tars and oils) that could otherwise complicate downstream processing.

Gasifiers operating at elevated pressures are preferred for MeOH and $H_2$ production to reduce the sizes and unit costs of downstream reactors and the cost of downstream gas compression. Several pressurized gasifiers are commercially available for coal but not for biomass. However, there are several ongoing efforts to demonstrate pressurized, directly-heated biomass gasifier technology, so that the technology is likely to be commercially ready before the end of the decade.

The directly-heated fluidized-bed gasification technology selected here for analysis is the pressurized bubbling fluidized-bed gasifier under development by the Institute of Gas Technology (IGT). At pilot scale, the IGT gasifier has had perhaps the most extensive operating experience with biomass of any pressurized bubbling-bed gasifier (Evans et al., 1988). Also, this technology has been the focus of several previous assessments of MeOH production from biomass (Wyman, et al, 1993; Stevens, 1991; OPPA, 1990). A scaled-up version of the IGT reactor has now been built in Hawaii, where it will be operated initially on sugarcane bagasse (100 tonnes/day capacity). Plans are to subsequently demonstrate *(i)* the production of electricity by burning the product gas in a gas turbine and *(ii)* the production of MeOH (Overend, et al., 1994). For the present analysis, projected performance characteristics of a commercial-scale IGT gasifier are considered (Table 2).

The present analysis also considers pressurized entrained-bed biomass gasification (specifically, with the Shell gasifier, which is commercially available for coal). The entrained flow gasifier, the predominant coal gasification technology, has not been seriously considered for biomass for two reasons. First, the feedstock must be crushed to a fine size (100-600 microns), which is relatively easily done with coal, but is energy and capital intensive with biomass. Second, because of its high reactivity, biomass does not require the same high peak gasification temperatures needed for coal; thus, the large amount of $O_2$ used to achieve high temperatures (in excess of 1300 °C for coal) and the attendant removal of ash as molten slag represent unnecessary costs for biomass. This gasifier is included in the present analysis to underscore the importance of choosing a gasifier (other than this one) that is well designed to take advantage of the unique properties of the biomass feedstock.

*Indirectly-heated gasifiers.* Indirectly-heated gasifiers obtain the heat needed to drive the gasification reactions from heat-exchange tubes or from an inert heat-carrying material like sand. Indirect heating gives rise to lower reactor temperatures than direct heating, but temperatures are sufficiently high (700-800 °C) for effective biomass gasification. The indirect heating makes possible the production of a gas undiluted by $N_2$, without the use of costly $O_2$.

To characterize MeOH and $H_2$ production based on indirectly-heated gasification, we consider projected performance characteristics (Table 2) for the Battelle-Columbus Laboratory (BCL) and the Manufacturing and Technology Conversion International (MTCI) designs. Both have been tested at pilot scale but need to be demonstrated at larger scale before they can be commercialized.

The BCL gasifier is an atmospheric-pressure twin, fast fluidized-bed unit that resembles fluid catalytic crackers commonly used in petrochemical refining (Fig. 2c). After drying, biomass is fed to the gasification bed, wherein pyrolysis (devolatilization) of the biomass occurs. The char (solid carbon) residue of this process is transferred to a separate combustion bed where it burns to heat sand that is circulated back to the gasification bed to provide the heat to drive the devolatilization.[10] The ash is removed from the combustor. A portion of the product gas is recirculated (after cleaning) to fluidize the gasification bed. The solids residence time during gasification is short, so that the specific throughput rate is relatively high (Table 2), which should reduce unit capital costs relative to other gasifiers with lower throughputs.

To date, the BCL technology has operated at bench-scale, at about 500 kg/hour capacity (Feldman, et al., 1988; Paisley and Litt, 1993). A 200 $kW_e$ gas turbine has recently been coupled to this unit to explore prospects for power generation with the technology. A BCL gasifier with a capacity of 200 tonne/day of wood chips will be built in the state of Vermont in the US (with 50% cost sharing of the installation cost by the US Dept. of Energy) for eventual use in power generation using a gas turbine (Paisley and Overend, 1994). The use of the BCL gasifier for MeOH production has also been evaluated by researchers at the National Renewable Energy Laboratory (Wyman et al, 1993). The performance characteristics assumed here for the BCL gasifier (Table 2) are based on a detailed pre-feasibility study for a commercial-scale BCL-gasifier/gas turbine facility (Breault and Morgan, 1992).

The MTCI gasifier (Fig. 2d) is an atmospheric-pressure fluidized-bed unit with in-bed heat exchanger tubes (MTCI, 1990; Durai-Swamy et al., 1991a). Some combustible gas taken from one or more downstream points in the process is burned in a pulse combustor that feeds the heat exchanger tubes, providing the heat for gasification. Pulse combustion creates a highly turbulent flow of hot combustion products in the heat exchanger tubes, leading to heat transfer rates about five times higher than those for a conventional heat exchanger (Parkinson, 1990). The combustion of gaseous fuel in the pulse combustor can be augmented by combustion of residual char from the gasifier and tars

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[10] To increase the gas production per unit of biomass input to the gasification bed, supplemental fuel (biomass or other fuel) can be added to the combustor. However, the best overall efficiency (counting all energy inputs to the gasification and combustion beds) is achieved when no supplemental fuel is used (Paisley, 1994a).

recovered from the quench step.[11] Reactor temperatures are typically low (about 700 °C) due to the indirect heating. In work to date, $H_2O_{(g)}$ is injected into the gasifier both to promote gasification reactions and to fluidize the bed. In future systems, recycled product gas may be substituted for some of the $H_2O_{(g)}$, because only a small fraction of the $H_2O_{(g)}$ presently used is actually consumed in the gasification reactions.[12] The MTCI design produces a gas with an unusually high $H_2$ and relatively low $CH_4$ content (Table 2).[13]

Most biomass-related development work with the MTCI technology has focussed on "unusual" biomass feedstocks: black liquor, the lignin-rich byproduct of cellulose extraction during chemical pulp production (Black, 1991); pulp-mill sludge (Durai-Swamy et al., 1991b); and municipal sewage and solid wastes (Durai-Swamy et al., 1991c). A pilot-scale MTCI black liquor gasifier of 48 tonnes/day capacity has operated since 1992 in Erode, India, and a similar unit has recently been installed in a pulp mill in New Bern, North Carolina (Durai-Swamy, 1994). Also, a pilot-scale pulp-mill sludge gasifier with a capacity of 24 tonnes/day has been test operated at a paper mill in Ontario, California.

### Process Modeling of Methanol and Hydrogen Production

Detailed thermodynamic process models of MeOH and $H_2$ production from biomass, from natural gas, and from coal were constructed using commercially-available process simulation software (ASPEN-PLUS). See Katofsky (1993) for details.

Table 2 gives the performance characteristics, based on experimental data, assumed for the IGT, MTCI, and BCL biomass gasifiers and for the Shell coal gasifier. With the exception of the Shell gasifier, the use of experimental data is required, because gas compositions cannot be predicted using ASPEN-PLUS. All chemical reactions within an ASPEN-PLUS model are assumed to reach equilibrium. In actual biomass gasifiers, the low temperatures and short residence times produce gas compositions that are far from chemical equilibrium. The Shell gasifier produces a high enough temperature that chemical equilibrium calculations give a reasonable match to empirical data.

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[11] The burning of char and tar in the pulse combustor has been proposed (Turdera and Zahradnik, 1992) but not demonstrated. Here it is assumed to account for 15.3% of the total energy released in the pulse combustor.

[12] In the present analysis it is assumed that only $H_2O_{(g)}$ is the fluidizing agent, since no data are available for performance with recycled product gas.

[13] This allows some significant simplifications in the downstream process design for MeOH or $H_2$ production. In particular, the low $CH_4$ fraction allows the reforming step to be avoided; $CH_4$ is carried through the process as an inert gas, is eventually isolated, and then directed to the pulse combustor. The high $H_2$ content permits the shift reactor to be eliminated (for $CH_3OH$ production) or significantly downsized (for $H_2$ production).

Downstream chemical reactors (reformer, shift, MeOH synthesis) were modeled assuming operating conditions commonly found in commercial processes today. Gas compositions leaving each reactor were calculated assuming chemical equilibrium. To account for the fact that equilibrium is never achieved in a real reactor, the temperature selected for the calculation (called the approach temperature) in each case differs by a specified amount from the temperature that would be observed in practical operation.[14] The equilibrium assumption coupled with the approach temperature concept is widely used in industrial practice (although is not suitable for biomass gasifier modeling, as discussed above), as it allows chemical reactors to be accurately modeled without resorting to much more complex and uncertain methods based on reaction kinetics. The performance levels assumed for downstream physical separation steps ($CO_2$ or $H_2$ removal) were based on vendors' performance estimates. Typical pressure losses through reactors and heat exchangers were included in the analysis, as were compressor efficiencies (Katofsky, 1993).

Because a significant amount of heating and cooling of process streams is required between reactors, how heat exchange is carried out between process streams that need heating and those that need cooling has a significant effect on overall plant efficiency. To ensure consistent and realistic modeling of process-to-process heat exchange from one case to another, "pinch" analysis was used (Katofsky, 1993). Pinch analysis was introduced about a decade ago as a tool to help analyze and optimize the energy performance of industrial processes (Linnhoff, 1993). Accounting for practical inefficiencies of heat transfer, pinch analysis identifies the amount of heat addition and heat rejection (cooling) needed from external sources, after first systematically matching process streams that need heating with those that need cooling so as to minimize wastage. In almost all cases considered here, all of the process heating needs can be met by heat exchange from other, hotter process streams. Waste heat that remains after this matching is assumed to be used to raise steam for expansion through a condensing steam turbine to produce electricity for onsite use. Shortfalls in on-site electricity supply, in heat, and in cooling needs are assumed to be provided from external sources, the energy requirements for which are accounted in the analysis.

In the case of natural gas feedstock, the methodology used here gives overall production efficiencies for MeOH and $H_2$ production that are consistent with estimates for actual commercial installations at relatively large scale (Moore, 1994).[15] This lends confidence in the results for

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[14] The approach temperature for endothermic (heat absorbing) reactions, such as steam reforming, is lower than the temperature that would be observed in a real reactor. For exothermic (heat releasing) reactions, such as MeOH synthesis, it would be higher than the actual observed temperature.

[15] In smaller facilities, the gains in efficiency that would come from the extensive levels of heat integration considered here are typically sacrificed to reduce capital costs.

biomass, for which there is no commercial experience, and for coal, for which there is very limited experience.

### Efficiencies of Fuel Production

Two measures are used here for the energy efficiency of MeOH or $H_2$ production: *(i)* the energy ratio (ER), defined as the energy content of the product fuel divided by the energy in the input feedstock (HHV basis), which takes no account of the energy required to provide electricity or heat from external sources, and *(ii)* the thermal efficiency (TE), defined as the energy content of the product fuel divided by the energy content of all energy inputs to the process, including the feedstock and additional amounts of feedstock used to generate the electricity and heat requirements not provided from byproduct process heat or purge gases (see notes *(g)* and *(h)* of Table 3). In effect, the TE gives the energy performance of a facility whose only energy input is the feedstock under consideration.

With biomass, regardless of the gasifier considered, the ERs and TEs for MeOH production (Table 3) are lower than for $H_2$ production (Table 4). This is also true when the feedstock is natural gas or coal. The primary reason is that the MeOH synthesis step is more exothermic (heat rejecting) than the analogous step (shift reaction plus PSA separation) in $H_2$ production (Katofsky, 1993). Also, all the ERs are higher than the TEs, because external energy inputs are required in all cases.

The efficiencies of MeOH or $H_2$ production from biomass are lower than with natural gas, as expected, since gasification consumes a non-negligible fraction of the feedstock energy, following which the syngas is processed in a manner similar to that for natural gas.[16] In contrast, the most efficient biomass-based MeOH production system is about as efficient as when coal is the feedstock.[17]

The best ER and TE for MeOH or $H_2$ production from biomass arise with the Shell gasifier, because this produces the largest volume of CO and $H_2$ per unit of biomass input (Table 2). Also, no energy-consuming reforming step is required because there are no hydrocarbons in the raw gas. Despite the high efficiencies with this gasifier, the high costs for this option with biomass (discussed below) make it unattractive relative to alternative options.

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[16] Our estimate of the TE for $CH_3OH$ production from natural gas is slightly higher than other recent estimates (OPPA, 1989; Wyman et al., 1993), most probably due to our use of pinch analysis to determine external energy input requirements. Our estimate of the TE for $H_2$ production from natural gas is consistent with present industrial practice (Moore, 1994).

[17] Another analysis (Wyman et al., 1993) of MeOH from coal found an energy ratio of 0.55. This is lower than the value estimated in the present analysis (0.65), primarily because the Wyman analysis assumed use of a Texaco coal gasifier, which has a much lower cold gas efficiency than the Shell gasifier (Synthetic Fuels Assoc., 1983), which is the gasifier assumed here.

The other directly-heated biomass gasifier, the bubbling fluidized-bed gasifier, is characterized by the lowest ERs and TEs of all the options considered, reflecting the relatively large amount of $CO_2$ in the product gas (Table 2). Unlike carbon that leaves the gasifier as CO or hydrocarbons, carbon as $CO_2$ is largely unconverted to final product through downstream processing.

The ERs and TEs for the indirectly-heated gasifiers are 5 to 15% higher than for the bubbling fluidized-bed cases, but less than for the entrained-bed gasifier. Indirectly heated gasifiers produce considerably less $CO_2$ per unit of biomass input than the IGT gasifier, but do produce some hydrocarbons due to the low operating temperatures (Table 2). With the MTCI gasifier, the hydrocarbon content is low enough that they can be carried through downstream processing without using a reformer, as discussed earlier, thereby avoiding reformer efficiency losses. However, the large steam requirement for fluidizing the gasifier necessitates a large input of fuel gas to the pulse combustor (to heat the fluidizing steam to the gasification temperature), which reduces the energy ratio. The ER for the MTCI case might be improved by using recycled product gas for fluidization instead of steam, as discussed earlier.

The efficiency for MeOH production calculated here with the IGT gasifier is consistent with results of other process studies (OPPA, 1990; Stevens, 1991; Wyman et al., 1993); the efficiency with the BCL gasifer is lower than estimated in earlier analyses by Katofsky (1993) and by Wyman et al. (1993), because improved estimates for the performance of the BCL gasifier have been used here. The authors are unaware of previous analyses of MeOH production using the MTCI gasifier or of hydrogen production from biomass using any gasifier.

## Costs of Fuel Production and Use

Fuel costs are examined here at the points of production and of sale to final consumers, measured in \$/GJ of fuel, and also in terms of the transportation service provided, measured in \$/vehicle-kilometer (v-km) of driving, for automotive applications. Moreover, two measures are provided of the cost per unit of service--the fuel cost/v-km, and the total cost/v-km, including the capital and other operating costs for the vehicle.

*Costs of producing MeOH.* Estimates of the levelized costs of MeOH produced from biomass, coal and natural gas are presented in Fig. 3.[18] (Table 5 gives details of the cost estimate.) For biomass,

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[18] All costs in this paper are expressed in constant 1991 US dollars using the US GNP deflator (Council of Economic Advisors, 1994). Cost estimates in Fig. 3 and Tables 5 and 6 assume a real (inflation-corrected) discount rate of 9.9%/year and a real capital charge rate of 15.1%/year (based on average financial parameters for major US corporations from 1984 to 1989).

the assumed plant capacity is 1650 dry tonnes input/day, corresponding to 3000 tonnes/day of green biomass (45% moisture).[19] For natural gas, the assumed feed rate is 1.64 million $Nm^3$/day--typical for a large modern natural-gas MeOH plant. For coal, the assumed feed rate is 5000 tonnes/day, so that the output capacity is about five times as large as for the biomass cases; the larger capacity is considered because transportation costs for coal are not as scale sensitive as for biomass (due to the higher volumetric energy density of coal), so that scale economies in the capital cost of the conversion plant can be exploited with coal.

Installed capital cost estimates ( 25%) for major unit operations were developed based on vendor quotes, other studies of MeOH production, and discussions with gasifier developers. Other cost elements (startup, working capital, maintenance, contingencies, overhead, etc.) are consistent with those used in other recent studies (Wyman et al., 1993; OPPA, 1990). For simplicity, external electricity and heat requirements are assumed to be purchased.

The costs presented in Fig. 3 are for "reference case" feedstock costs. The assumed cost of the biomass feedstock delivered to the conversion facility is \$2.0/GJ.[20] For natural gas and coal, prices are assumed to be \$4.1/GJ and \$1.45/GJ, respectively, which are likely to be typical lifecycle prices for industrial customers in the US for the period 2010-2035 (see note *(q)* of Table 5).[21]

The lowest costs for MeOH from biomass--\$11.2-11.5/GJ (see Table 5 and Fig. 3)--are for indirectly-heated gasifiers. For the Shell gasifier, high capital costs (especially for feedstock preparation, high-temperature gas cooling, and $O_2$ production), lead to the highest biomass-MeOH costs. The MeOH cost for the other directly-heated gasifier, the fluidized-bed IGT unit, is also relatively high due primarily to the capital costs for $O_2$ production and the lower energy ratio. The

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[19] For comparison, this biomass input capacity is of the same order as that for a modern pulp and paper making facility or a large cane sugar production facility.

[20] In a major US Department of Energy/Department of Agriculture study (Graham et al., 1995) estimates have been made of the costs of producing plantation biomass in the US at various levels of supply, with current and future biomass plantation technology as projected by bioenergy researchers at the Oak Ridge National Laboratory. A related study carried out by the US Environmental Protection Agency (Turnure et al., 1995) estimates biomass prices under the same technical assumptions about yields and costs but also takes into account the effects of land use competition, which leads to higher land rents and thus higher biomass prices. In the latter study it is estimated that with current biomass plantation technology (not yet commercial but as demonstrated in field trials) some 3 EJ/year of plantation biomass could be produced in the US at prices up to \$2/GJ, and that the potential supply at this price would increase to 8 EJ/year with plantation technology projected for the year 2020.
For developing countries of Africa, Latin America, and Asia, it has been estimated that about 70 EJ per year could be produced at this price in 2025, assuming that 10% of the area of these regions that is not forest, not wilderness and not needed for food production in 2025 is available for biomass production (Larson, Marrison, and Williams, 1995).

[21] For comparison, the average industrial prices for natural gas and coal in the US in 1990 were \$3.2/GJ and \$1.5/GJ, respectively.

cost of MeOH from coal is only 10% less than the lowest cost shown for MeOH from biomass, even though the assumed reference feedstock price is 28% less for coal than for biomass, and the capacity of the coal plant is five times as large as for biomass.

At the average 1990 natural gas price for US industry ($3.2/GJ) the production cost for MeOH from natural gas would be $8.2/GJ, 27% less than for the least costly biomass case shown in Fig. 3. However, natural gas prices are expected to rise in most parts of the world. At our reference natural gas price of $4.1/GJ for the period 2010-2035, MeOH from natural gas would be 16% less costly than from $2.0/GJ biomass.

In some regions (e.g. some developing countries) biomass-derived MeOH would probably be competitive with MeOH derived from natural gas. For example, it is estimated that 1.7 EJ/year of biomass (enough to provide 1.0 EJ/year of MeOH) could be produced on 4 million hectares of plantations in the northest of Brazil at a cost of $1.5/GJ or less.[22] Producing MeOH there from biomass at this price and shipping it to Rotterdam would cost the same as MeOH made in Europe from natural gas costing $5/GJ, 21% above the reference natural gas price.[23] Methanol imports of 1 EJ/year would be able to support 60 million fuel cell-powered cars in Europe (about half of the total number of private cars in the European Union at present).

Producing biomass-derived fuels in excess of domestic energy needs for export markets in such a fashion could provide sustainably a major source of income for rural areas of developing countries (Johansson et al., 1993), and competition from such biofuels in world energy markets would help keep the price of natural gas from rising too much in the long term.

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[22] Carpentieri et al. (1993) estimate that 12.6 EJ/year of biomass could be grown at an average productivity of 12.5 tonnes/hectare/year on 50.5 million hectares of plantations in the Brazilian northeast, at an average cost of $1.55/GJ. This total potential production includes 1.7 EJ/year that could be grown at an average productivity of 20.7 tonnes/hectare/year on the better available lands at a cost of $1.23/GJ. These costs include 85 km of transport to the conversion facility but do not include the cost of chipping. Chipping costs have been estimated by Perlack and Wright (1994) to be $5.13/dry tonne or $0.26/GJ of biomass. Thus the total cost of 1.7 EJ/year of biomass (including chipping) would be $1.5/GJ delivered to the conversion facility.

[23] For biomass feedstock costing $1.5/GJ, the plant-gate cost of MeOH would be $10.4/GJ. For delivery to Rotterdam, the cost of oceanic transport must be added to this production cost. The cost C (in $/GJ) of oceanic transport of MeOH via large (250,000 dead-weight tons) tankers fueled with fuel oil costing P ($/barrel) for a round-trip distance of RT nautical miles is given by (USDOE, 1989):

$$C = 2.03 \cdot RT \cdot 10^{-5} + 2.49 \cdot 10^{-2} + 1.55 \cdot P \cdot RT \cdot 10^{-7} + 2.59 \cdot P \cdot 10^{-5}.$$

For transport from the Brazilian northeast to Rotterdam, RT = 12,000 nautical miles. Moreover, the US DOE projects (EIA, 1995) that the cost of residual fuel oil for transport in 2010 will be $3.24/GJ or $21.5/barrel. Thus the transport cost would add $0.30/GJ to the cost of producing MeOH, bringing the total cost of MeOH produced from biomass in the Brazilian northeast and delivered to Rotterdam to $10.7/GJ--which would be competitive with MeOH produced in Europe from natural gas priced at $5/GJ.

*Costs of producing $H_2$.* Estimates of the levelized costs of $H_2$ produced from biomass, coal and natural gas, derived in the same manner as for MeOH, are also shown in Fig. 3. (Table 6 shows details.) The ranking of production costs for $H_2$ using various biomass gasifiers is the same as for MeOH, but $H_2$ production costs are as much as 20% less than for MeOH. This is due to significantly lower hardware costs and to higher efficiency. The difference in hardware costs is attributable to the simpler technology required: the PSA unit used in $H_2$ production costs about as much as the Selexol® unit used in MeOH production (to remove $CO_2$), but replaces both the Selexol® and MeOH synthesis units. The latter is one of the costliest process units at a MeOH production plant.

The estimated cost of producing $H_2$ from biomass is less, by a factor of two or more, than the cost of $H_2$ derived electrolytically from water using wind or photovoltaic power sources, assuming that the long-term cost reduction goals for these power sources can be met (Ogden et al., 1994; Takahashi, 1989).

The natural gas price at which $H_2$ derived from biomass costing \$2.0/GJ (reference feedstock price) becomes competitive with $H_2$ derived from natural gas is slightly higher than for MeOH (\$5.5 vs. \$5.4/GJ), whereas the biomass feedstock price at which $H_2$ derived from biomass can compete with $H_2$ derived from coal at its reference price is considerably higher than for MeOH (\$2.1 vs. \$1.3/GJ).

If global environmental costs associated with fuel production and use (e.g. net $CO_2$ release) were internalized, the economics of biomass-derived MeOH and $H_2$ production would improve relative to the fossil fuel options, as discussed below.

*Consumer costs.* The costs seen by the consumer differ from production costs because of the extra costs of delivering the fuel to the consumer and the characteristics of the vehicle that uses the fuel.

While the cost of producing $H_2$ from biomass or natural gas is about one-quarter less than that for producing MeOH, the cost of $H_2$ delivered to consumers is only about 6% less than for MeOH, largely because of the expense of compressing $H_2$ to high pressure at the refueling station (Fig. 4a and Table 7). But because $H_2$-fueled FCVs would be more energy-efficient than MeOH-fueled FCVs, the cost of $H_2$ fuel per v-km would be only four-fifths as much as for MeOH (Fig. 4b and Table 7). Going one step further and considering the total cost of owning and operating the vehicle (in cents/km), however, would tip the balance slightly in favor of MeOH, because $H_2$-fueled FCVs would probably have higher first costs (Fig. 5 and Table 9).

### Lifecycle $CO_2$ Emissions for Alternative Fuel/Vehicle Combinations

In addition to potential urban air pollution benefits, the use of MeOH or $H_2$ could also lead to reduced lifecycle $CO_2$ emissions (emissions over the entire cycle of fuel production and use) per km of vehicle travel, with especially large reductions for FCVs (Fig. 6, with additional details in Tables 8 and 9).

When natural gas is the feedstock, $CO_2$ emissions per km of vehicle travel would decrease relative to a gasoline-fueled ICEV: by 8% for a MeOH-fueled ICEV; by 16% for a $H_2$-fueled ICEV; by 56% for a MeOH-fueled FCV; and by 65% for a $H_2$-fueled FCV.

With coal as the feedstock, $CO_2$ emissions would increase relative to a gasoline-fueled ICEV 61% and 58% for MeOH and $H_2$ used in ICEVs, respectively, while emissions would decrease 24% and 34% for MeOH and $H_2$ used in FCVs, respectively.

In the case of biomass grown sustainably, the $CO_2$ released in fuel processing and use is offset by the $CO_2$ removed from the atmosphere during photosynthesis. However, since energy inputs (typically fossil fuel energy) are needed in the growing, harvesting and transport of the biomass, and in the delivery of the fuel to the end user, total lifecycle emissions, though small compared to fossil fuel systems, are non-zero. For fuels production based on the BCL gasifier, lifecycle $CO_2$ emissions for FCVs fueled with MeOH[24] and $H_2$ are 93% and 85% less, respectively, than for gasoline-fueled ICEVs, when the external electricity requirements for fuels production are provided by the average mix of electricity generation in the US. If instead this electricity were provided by biomass-fueled power systems (which is the more likely scenario and the one shown in Fig. 6), lifecycle emissions would be, respectively, 96% and 91% less than for gasoline-fueled ICEVs.[25]

A carbon tax on net $CO_2$ emissions associated with the production and use of automotive fuels could be used to internalize the costs of global warming. While it is beyond the scope of the present analysis to estimate the level of tax required to internalize these costs, a useful exercise is to ask what the tax would have to be to make the cost of fuels derived from biomas equal to the cost of these fuels derived from natural gas. The results of this exercise are presented in Fig. 7, which shows the cost of MeOH and $H_2$ production from natural gas, coal, and biomass as a function of the carbon tax on each of these fuels, assuming reference feedstock prices for natural gas and coal, and biomass feedstock

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[24] The lifecycle $CO_2$ emissions for MeOH produced using the BCL gasifier can also be expressed as 0.089 tonnes of C per tonne of MeOH produced. For comparison, Ellington et al. (1993) recently estimated that lifecycle $CO_2$ emissions associated with the production of MeOH from biomass would be 1.24 tonnes of carbon per tonne of MeOH produced from biomass and used in a vehicle. Ellington et al.'s estimate neglects the photosynthetic credit associated with biomass grown sustainably. Also, it assumes use of a biomass gasification technology that is considerably less efficient than the gasifiers we have considered here.

[25] Lifecycle $CO_2$ emissions could be eliminated completely if biofuels were used in place of fossil fuels in the production of biomass, if the use of artificial nitrogen fertilizers were eliminated (for example by growing nitrogen-fixing species in plantations), and if the downstream compressor work and transport energy requirements were provided by renewable energy sources.

prices of \$1.5/GJ and \$2.5/GJ as well as the reference feedstock price of \$2.0/GJ.[26] The carbon taxes needed to make MeOH and $H_2$ from biomass competitive with these fuels from natural gas are \$85/tC and \$95/tC, respectively, for the reference biomass feedstock cost. A carbon tax at such levels seems substantial, as it would increase the price of the natural gas feedstock by 21-29%. But to the consumer owning a FCV operated on natural gas-derived fuels, the tax would hardly be noticeable, as it would increase the cost of owning and operating the vehicle by less than 1.5%. The effect is small because fuel accounts for a small part of the cost of owning and operating a FCV (Fig. 5). And even with the tax in place for FCVs, the total cost of owning and operating the FCV would probably be less than the total cost of owning and operating a comparable gasoline ICEV with or without the carbon tax (Fig. 5).

## Conclusions

Biomass-derived methanol and hydrogen offer substantial improvements over conventional biofuels such as ethanol derived from maize, because *(i)* they can be produced more efficiently from a given amount of land and with fewer adverse environmental impacts, and *(ii)* they are well-suited for use in clean and efficient fuel cell vehicles. Biomass-derived methanol and hydrogen could make major contributions to energy requirements for road transportation when used in fuel cell vehicles. Such systems offer attractive economics in the post-2010 time-frame and the potential for very low emissions of both local air pollutants and low net $CO_2$ emissions if the biomass is grown sustainably.

The development of technologies for producing MeOH and $H_2$ from biomass should be coordinated with efforts to develop FCVs. The FCV represents an extrordinarily attractive market for biomass producers, and FCV developers should not hesitate to introduce their new products out of concern about the availability of fuels suitable for use in FCVs, in light of the prospect of large potential renewable supplies of cost-competitive biomass-derived MeOH and $H_2$.

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[26] The absolute lifecycle $CO_2$ emissions (indicated in relative terms in Fig. 6) are: 22.61 kg C/GJ for gasoline; 23.35 and 21.81 kg C/GJ for natural-gas derived MeOH and $H_2$, respectively; 40.79 and 41.20 kg C/GJ for coal-derived MeOH and $H_2$, respectively; 2.34 and 5.35 kg C/GJ for biomass-derived MeOH and $H_2$ respectively (assuming the BCL gasifier with electricity provided by biomass power plants).

Table 1. Energy Yield for Alternative Feedstock/Conversion Technologies

| Option | Feedstock Yield (dry tonnes/ha/year) | Transport Fuel Yield (GJ/ha/year) | Transport Services Yield[a] ($10^3$ v-km/ha/year) |
|---|---|---|---|
| Rape Methyl Ester (Netherlands, year 2000)[b] | 3.7 of Rapeseed | 47 | 21 (ICEV) |
| EthOH from Maize (US)[c] | 7.2 of Maize | 76 | 27 (ICEV) |
| EthOH from Wheat (Netherlands, year 2000)[d] | 6.6 of Wheat | 72 | 26 (ICEV) |
| EthOH from Sugar Beets (Netherlands, year 2000)[e] | 15.1 of Sugar Beets | 132 | 48 (ICEV) |
| EthOH from Sugar Cane (Brazil)[f] | 38.5 of Cane Stems | 111 | 40 (ICEV) |
| EthOH, Enzymatic Hydrolysis of Wood (present technology)[g] | 15 of Wood | 122 | 44 (ICEV) |
| EthOH, Enzymatic Hydrolysis of Wood (improved technology)[g] | 15 of Wood | 179 | 64 (ICEV) |
| MeOH, Thermochemical Gasification of Wood[h] | 15 of Wood | 177 | 64/133 (ICEV/FCV) |
| $H_2$, Thermochemical Gasification of Wood[h] | 15 of Wood | 213 | 84/189 (ICEV/FCV) |

(a) The fuel economy of the vehicles used (in liters of gasoline-equivalent per 100 km) are assumed to be: 6.30 for rape methyl ester (assumed to be the same as for diesel), 7.97 for ethanol, 7.90 for methanol, and 7.31 for hydrogen used in internal combustion engine vehicles (ICEVs); and 3.81 for methanol and 3.24 for hydrogen used in fuel cell vehicles (FCVs) (DeLuchi, 1991). Note that 1 liter of gasoline equivalent = 0.0348 GJ, HHV.

(b) Per tonne of seed: 370 liters of rape methyl ester plus (not listed) 1.4 tonnes of straw (Lysen *et al., 1992)*.

(c) For wet milling, assuming the US average maize yield, 1989-1992; per tonne of grain: 440 liters of ethanol plus (not listed) 0.35 tonne of stover (out of 1 tonne of total stover, assuming the rest must be left at the site for soil maintenance), 275 kg of corn gluten cattle feed, and 330 kg of $CO_2$ (Wyman *et al.*, 1993).

(d) Per tonne of seed: 455 liters of ethanol plus (not listed) 0.6 tonnes of straw (Lysen *et al.*, 1992).

(e) Per tonne of sugar beet: 364 liters of ethanol (Lysen *et al.*, 1992).

(f) For the average sugar cane yield in Brazil in 1987 (63.3 tonnes of harvested cane stems per hectare, wet weight); per tonne of wet cane stems: 73 liters of ethanol (Goldemberg *et al.*, 1993). In addition, (not listed) the dry weight of the attached tops and leaves amounts to 0.092 tonnes and that for the detached leaves amounts to 0.188 tonnes per tonne of wet stems--altogether some 18 dry tonnes per hectare per year (Alexander, 1985).

(g) Per tonne of feedstock: 338 liters of ethanol plus (not listed) 183 kWh (0.658 GJ) of electricity, present technology; 497 liters of ethanol plus (not listed) 101 kWh (0.365 GJ) of electricity, improved technology (Wyman *et al.*, 1993).

(h) For the indirectly-heated Battelle Columbus Laboratory biomass gasifier; per tonne of feedstock: 11.8 GJ of methanol or 14.2 GJ of hydrogen; per tonne of feedstock, external electricity requirements are 107 kWh (0.38 GJ) for methanol or 309 kWh (1.11 GJ) for hydrogen (see Tables 3 and 4).

Table 2. Characteristics of biomass gasifiers.

| Feedstock | Biomass | | | | Coal |
|---|---|---|---|---|---|
| **Gasifier design** | Bubbling fluid-bed (IGT)[a] | Indirectly-heated fluid bed (MTCI)[b] | Indirectly-heated fast fluidized-bed (BCL)[c] | Entrained-bed (Shell)[d] | Entrained-bed (Shell)[e] |
| **Feedstock characteristics** | | | | | |
| Dry, ash free composition | $CH_{1.52}O_{0.68}$ | $CH_{1.63}O_{0.66}$ | $CH_{1.54}O_{0.65}$ | $CH_{1.52}O_{0.68}$ | $CH_{0.91}O_{0.11}$ |
| HHV (GJ/dry tonne) | 19.28 | 19.40 | 19.46 | 19.28 | 29.69 |
| Initial moisture (%) | 45 | 45 | 45 | 45 | 5 |
| Moisture after drying (%) | 15 | 20 | 10 | 11 | 5 |
| **Additional inputs** | | | | | |
| Gasifier steam (kg/kg dry feed) | 0.3 | 1.37 | 0.019 | 0.03 | 0.03 |
| Gasifier oxygen (kg/kg dry feed) | 0.3 | 0 | 0 | 0.45 | 0.8 |
| Combustor air (kg/kg dry feed) | 0 | 2.52 | 2.06 | 0 | 0 |
| **Reactor characteristics** | | | | | |
| Exit temperature (°C) | 982 | 697 | 863 | 1085 | 1371 |
| Pressure (MPa) | 3.45 | 0.101 | 0.101 | 2.43 | 2.43 |
| Throughput (dry kg/m$^2$-s) | 1.9 | 0.07 | 4.09 | n.a. | n.a. |
| Solids residence time | minutes | minutes | ~1 second | ~1 second | ~1 second |
| **Product gas characteristics** | | | | | |
| Yield (kmol/tonne dry feed) | 82.0 | 146.8 | 45.8 | 79.3 | 92.4 |
| Molecular weight (kg/kmol) | 22.27 | 17.65 | 21.64 | 20.08 | 20.49 |
| HHV (MJ/Nm$^3$ raw gas) | 8.62 | 7.52 | 15.19 | 9.25 | 11.53 |
| HHV (MJ/kg raw gas) | 8.68 | 9.55 | 15.73 | 10.32 | 12.61 |
| Composition[f] (volume % wet) | | | | | |
| $H_2O$ | 31.8 | 49.5 | 19.9 | 18.4 | 2.1 |
| $H_2$ | 20.8 | 25.3 | 16.7 | 30.7 | 31.8 |
| CO | 15.0 | 11.2 | 37.1 | 39.0 | 64.3 |
| $CO_2$ | 23.9 | 9.9 | 8.90 | 11.8 | 1.7 |
| $CH_4$ | 8.2 | 4.0 | 12.6 | 0.1 | 0 |
| $C_2$+ | 0.3 | 0.2 | 4.8 | 0 | 0 |
| **Net carbon conversion[g] (%)** | 96.2 | 67.2 | 75.2 | 100 | >99 |
| **Cold gas efficiency[h] (%)** | 82.1 | 90.0 | 80.1 | 85.2 | 80.3 |

(a) Institute of Gas Technology (IGT) gasifier data from OPPA (1990). See also Overend, et al. (1994).

(b) Manufacturing and Technology Conversion International (MTCI) gasifier data from Durai-Swamy et al. (1991a).

(c) Battelle Columbus Laboratory (BCL) gasifer heat and mass balances from Breault and Morgan (1992). Some product gas is recycled as the fluidizing agent for the gasifier. An earlier analysis of the BCL gasifier by Katofsky (1993), since proven to be in error, considered the performance of the BCL gasifier at a higher gasification temperature. (The higher temperature was achieved by supplemental firing of the combustor with some fresh biomass to increase the heat production rate.) Paisley (1994) indicates that the most efficient operation (highest gas yield) for the BCL gasifier occurs with no supplemental fueling of the combustor. This is the case considered here.

(d) Entrained-bed biomass gasification is based on the Shell gasifier, which has never operated with biomass. However, because of the high operating temperature of entrained beds, this system can be modeled relatively simply assuming chemical equilibrium. Katofsky (1993) provides details of the performance estimate of this gasifier operating on biomass.

(e) From Synthetic Fuels Associates (1983).

(f) Small quantities of nitrogen- and sulfur-containing compounds and argon are not shown.

(g) This is the fraction of carbon input as biomass or coal that leaves as synthesis gas.

(h) Defined as: [the energy content of the product gas (HHV basis)]/(the energy content of all energy inputs to the gasifier).

Table 3. Heat and mass balance results for methanol production from natural gas, biomass, and coal.

| **Feedstock** | **Biomass** | | | | **N. Gas**[a] | **Coal** |
|---|---|---|---|---|---|---|
| **Process** | IGT gasifier | MTCI gasifier | BCL gasifier | Shell gasifier | steam reforming | Shell gasifier |
| **Energy Inputs** | | | | | | |
| Feedstock (GJ/GJ methanol product)[b] | 1.77 | 1.63 | 1.65 | 1.48 | 1.42 | 1.54 |
| Electricity (kWh/GJ methanol product) | | | | | | |
| Pumps | 1.12 | 0.01 | 0.03 | 0.30 | 0.08 | 0.10 |
| Compressors | 8.44 | 35.38 | 29.72 | 7.89 | 13.05 | 9.37 |
| Lockhopper | 1.53 | 0.00 | 0.00 | 1.03 | 0.00 | 0.69 |
| Oxygen[c] | 13.20 | 0.00 | 0.00 | 16.56 | 0.00 | 18.92 |
| Total | 24.29 | 35.39 | 29.74 | 25.77 | 13.13 | 29.08 |
| Steam (kg/kg dry feed)[d] | 1.02 | 1.37 | 0.38 | 0.92 | 3.23 | 1.91 |
| **Energy Ratio (ER)**[e] | 0.566 | 0.615 | 0.606 | 0.677 | 0.704 | 0.649 |
| **Fraction of Electricity Input From:** | | | | | | |
| Waste heat[f] | 0.615 | 0.609 | 0.696 | 0.309 | 0.446 | 0.416 |
| Purge gases[g] | 0.000 | 0.000 | 0.000 | 0.154 | 0.000 | 0.248 |
| External sources | 0.385 | 0.391 | 0.304 | 0.537 | 0.554 | 0.336 |
| **Thermal Efficiency**[h] | 0.539 | 0.568 | 0.576 | 0.610 | 0.674 | 0.613 |

(a) Natural gas is assumed to be available at the conversion site at 2.5 MPa and to have the following volumetric composition: 94.7% $CH_4$, 2.8% $C_2H_6$, 0.2% $CO_2$, and 2.3% ($N_2$+Ar).

(b) Excluding any feedstock burned for electricity production. The heating values of the input feedstocks are as follows. For biomass (see Table 2) with the IGT gasifier, 19.28 GJ/dry tonne; MTCI gasifier, 19.40 GJ/dry tonne; BCL gasifier, 19.46 GJ/dry tonne. For natural gas and coal, 39.51 MJ/Nm$^3$ and 29.70 GJ/dry tonne, respectively.

(c) Production of 99.5% purity $O_2$ is assumed to require 480 kWh/tonne, based on 442 kWh/tonne for 95% purity (Brown et al., 1987), plus an additional 9% for 99.5% purity (Klosek et al., 1986).

(d) This is the total amount of steam generated for the process, excluding steam that is used for electricity production.

(e) The energy ratio is defined as: [the energy content (HHV basis) of the product (methanol or hydrogen)]/(the energy content of the feedstock input to the process,excluding any additional feed used for electricity production).

(f) All waste heat that is available (after meeting as much of the process heating needs as possible) and that is of sufficient thermodynamic quality is assumed to be used to raise steam at 6.2 MPa and 400 °C. The steam is expanded in a condensing steam turbine operating with an exhaust pressure of 0.005 MPa and an isentropic efficiency of 75%. A 95% generator efficiency is assumed. Pinch analysis techniques were used to determine how much of process heating needs could be met with a highly-integrated process-to-process heat exchange system. (All heating needs could be thus met in the methanol production cases.) The pinch analysis also provided the magnitudes of waste heat remaining that was suitable for steam-electricity generation.

(g) Purge gases (from the methanol synthesis loop or the PSA loop) are assumed to be used for electricity production with efficiencies achievable in a gas turbine/steam turbine combined cycle (Katofsky, 1993).

(h) The thermal efficiency is defined as: [the energy content (HHV basis) of the product (methanol or hydrogen)]/(the sum of energy content of all primary-energy inputs to the process). The inputs include the feedstock plus additional feed used to produce the electricity and heat that must be provided from external sources. No external heat addition is required for any of the methanol cases. External cooling is required in the fossil fuel cases. Per GJ of methanol product, the cooling energy requirements are 0.260 GJ for natural gas and 0.216 GJ for coal (Katofsky, 1993). We assume "free" external cooling is available, such as river water, so cooling is not counted in the thermal efficiency. The efficiencies assumed for electricity production correspond to estimates for advanced gas-turbine based power systems using the same feedstock as the methanol plant (Katofsky, 1993). For natural gas (or purge gas), the efficiency (in %) is given by: 39.58 + 0.134x$MW_e$, where $MW_e$ is the required electrical capacity in megawatts. For biomass or coal, the efficiency is 37.00 + 0.047x$MW_e$.

Table 4. Heat and mass balance results for hydrogen production from natural gas, biomass, and coal.[a]

| **Feedstock** | **Biomass** | | | | **N. Gas**[b] | **Coal** |
|---|---|---|---|---|---|---|
| **Process** | IGT gasifier | MTCI gasifier | BCL gasifier | Shell gasifier | steam reforming | Shell gasifier |
| **Energy Inputs** | | | | | | |
| Feedstock (GJ/GJ hydrogen product)[b] | 1.50 | 1.32 | 1.37 | 1.27 | 1.11 | 1.29 |
| Electricity (kWh/GJ hydrogen product) | | | | | | |
| Pumps | 0.99 | 0.01 | 0.04 | 0.29 | 0.05 | 0.11 |
| Compressors | 7.77 | 26.21 | 22.84 | 6.21 | 7.69 | 8.28 |
| Lockhopper | 1.30 | 0.00 | 0.00 | 0.88 | 0.00 | 0.58 |
| Oxygen[b] | 11.17 | 0.00 | 0.00 | 14.22 | 0.00 | 15.87 |
| PSA[c] | 11.88 | 9.23 | 8.90 | 11.62 | 2.75 | 11.03 |
| Total | 33.11 | 35.45 | 31.79 | 33.23 | 10.49 | 35.87 |
| Steam (kg/kg dry feed)[d] | 1.30 | 1.37 | 0.95 | 1.65 | 2.66 | 2.99 |
| **Energy Ratio (ER)**[a] | 0.669 | 0.759 | 0.732 | 0.788 | 0.897 | 0.774 |
| **Fraction of Electricity Input From:** | | | | | | |
| Waste heat[b] | 0.109 | 0.033 | 0.317 | 0.032 | 0.219 | 0.086 |
| Purge gases[b] | 0.000 | 0.000 | 0.000 | 0.151 | 0.000 | 0.138 |
| External sources | 0.891 | 0.967 | 0.683 | 0.817 | 0.781 | 0.776 |
| **Thermal Efficiency**[e] | 0.564 | 0.611 | 0.636 | 0.645 | 0.844 | 0.640 |

(a) In all cases the production facility is designed to produce gaseous $H_2$ at a pressure of 75 bar.

(b) See note on the item in Table 3 that corresponds to this item.

(c) Electricity use is assumed to be 4.46 kWh/kmole of $CO_2$ (Solomon, 1991).

(d) This is the total amount of steam generated for the process, excluding steam that is used for electricity production.

(e) See note *(h)* to Table 3. External heating is required only for the coal case (0.031 GJ input/GJ $H_2$). For use in calculating the thermal efficiency, the efficiency of external heat production is assumed to be 80%. External cooling is required per GJ of product $H_2$ as follows: 0.137 GJ for the MTCI biomass case, 0.014 GJ for the Shell-biomass case, 0.040 GJ for the natural gas case, and 0.189 GJ for the coal case (Katofsky, 1993).

Table 5. Estimated production costs (in 1991$) for methanol from biomass, natural gas, and coal.

| FEEDSTOCK → | BIOMASS | | | | N. GAS | COAL |
|---|---|---|---|---|---|---|
| PROCESS → | IGT gasifier | MTCI gasifier | BCL gasifier | Shell gasifier | Steam reforming | Shell gasifier |
| **Feedstock input capacity** | | | | | | |
| Dry tonnes/day | 1650 | 1650 | 1650 | 1650 | 1224 | 5000 |
| GJ/hour | 1326 | 1334 | 1338 | 1326 | 2700 | 6188 |
| **Output production capacity**[a] | | | | | | |
| Tonnes/day | 794 | 868 | 858 | 950 | 2012 | 4252 |
| GJ/hour | 750 | 820 | 811 | 897 | 1901 | 4016 |
| **Annual feed and output** | | | | | | |
| Feed ($10^6$ GJ/year) | 10.45 | 10.52 | 10.55 | 10.45 | 21.29 | 48.79 |
| Product output ($10^6$ GJ/year) | 5.91 | 6.47 | 6.39 | 7.07 | 14.99 | 31.66 |
| **Installed Equipment Costs ($10^6$ $)** | | | | | | |
| Feed preparation, including drying[b] | 17.32 | 13.21 | 13.17 | 38.78 | 0.00 | 67.96 |
| Gasifier[c] | 29.74 | 33.72 | 12.72 | 29.74 | 0.00 | 120.06 |
| High temperature gas cooling[d] | 0.00 | 0.00 | 0.00 | 39.67 | 0.00 | 113.27 |
| Oxygen plant[e] | 21.55 | 0.00 | 0.00 | 28.77 | 0.00 | 95.42 |
| Sulfur removal[f] | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 36.25 |
| Reformer feed compressor[g] | 0.00 | 15.94 | 11.88 | 0.00 | 0.00 | 0.00 |
| Reformer[h] | 21.39 | 0.00 | 17.20 | 0.00 | 50.00 | 0.00 |
| Shift reactor[i] | 1.98 | 0.00 | 2.00 | 0.00 | 0.00 | 0.00 |
| $CO_2$ removal[j] | 20.20 | 15.38 | 14.34 | 22.05 | 0.00 | 59.50 |
| Methanol synthesis & purification[k] | 35.87 | 38.05 | 37.75 | 40.37 | 66.25 | 108.54 |
| Steam turbine cogeneration plant[l] | 17.18 | 22.85 | 22.12 | 16.70 | 17.11 | 57.67 |
| Utilities/auxiliaries[m] | 41.31 | 34.79 | 32.80 | 54.02 | 33.34 | 164.67 |
| Subtotal | 206.55 | 173.95 | 163.98 | 270.09 | 166.69 | 823.33 |
| Contingencies[n] | 41.31 | 34.79 | 32.80 | 54.02 | 33.34 | 164.67 |
| Owners costs, fees, profits[n] | 20.65 | 17.39 | 16.40 | 27.01 | 16.67 | 82.33 |
| Startup[o] | 10.33 | 8.70 | 8.20 | 13.50 | 8.33 | 41.17 |
| **Total Capital Requirement ($10^6$ $)** | 278.84 | 234.83 | 221.37 | 364.62 | 225.04 | 1111.49 |
| **Working Capital**[n] **($10^6$ $)** | 20.65 | 17.39 | 16.40 | 27.01 | 16.67 | 82.33 |
| **Land**[p] **($10^6$ $)** | 2.08 | 2.08 | 2.08 | 2.08 | 4.26 | 7.40 |
| **Variable Operating Costs ($10^6$ $/year)** | | | | | | |
| Feed[q] | 20.90 | 21.03 | 21.10 | 20.90 | 87.28 | 70.74 |
| Catalysts and chemicals[r] | 1.67 | 0.67 | 2.24 | 0.67 | 2.58 | 10.87 |
| Purchased energy[s] | 2.77 | 4.47 | 2.89 | 4.90 | 5.45 | 15.47 |
| Subtotal | 25.33 | 26.17 | 26.23 | 26.46 | 95.30 | 97.07 |
| **Fixed Operating Costs ($10^6$ $/year)** | | | | | | |
| Labor[t] | 1.08 | 1.08 | 1.08 | 1.08 | 1.00 | 3.14 |
| Maintenance[u] | 6.20 | 5.22 | 4.92 | 8.10 | 5.00 | 24.70 |
| General Overhead | 4.73 | 4.10 | 3.90 | 5.97 | 3.90 | 18.09 |
| Direct Overhead | 0.49 | 0.49 | 0.49 | 0.49 | 0.45 | 1.41 |
| Subtotal | 12.50 | 10.88 | 10.39 | 15.64 | 10.35 | 47.34 |
| **Total Operating Costs ($10^6$ $/year)** | 37.83 | 37.06 | 36.62 | 42.11 | 105.65 | 144.41 |
| PRODUCTION COST ($/GJ of $CH_3OH$) | | | | | | |
| Capital[v] | 7.50 | 5.78 | 5.52 | 8.19 | 2.41 | 5.58 |
| Labor & maintenance | 2.40 | 1.79 | 1.98 | 2.31 | 0.86 | 1.84 |
| Purchased energy | 0.47 | 0.69 | 0.45 | 0.69 | 0.36 | 0.49 |
| Feedstock | 3.53 | 3.25 | 3.30 | 2.95 | 5.82 | 2.23 |
| TOTAL PRODUCTION COST | 13.90 | 11.51 | 11.24 | 14.14 | 9.46 | 10.14 |
| Natural gas price ($/GJ) for same total cost[w] | 7.23 | 5.55 | 5.36 | 7.40 | 4.10 | 4.58 |
| Biomass price ($/GJ) for same total cost as coal[x] | -0.12 | 1.16 | 1.33 | -0.71 | n.a. | n.a. |

Notes to Table 5:

(a) Based on process energy ratios given in Table 3.

(b) Feed preparation costs include drying (and pulverizing for the Shell gasifier cases). Costs are scaled from other estimates according to feed capacity (dry tonne per day, dtpd) raised to the 0.7 power. The other estimates are as follows: for IGT, $18.5 million for 1814 dtpd (OPPA, 1990); for MTCI, $1.48 million for 72.5 dtpd (MTCI, 1994); for BCL, $7.51 million for 740 dtpd (Breault and Morgan, 1992); Shell-biomass, $41.44 million for 1814 dtpd (OPPA, 1990); and for Shell-coal, $94.3 million for 7982 dtpd (OPPA, 1989).

(c) Gasifier costs are scaled from other estimates according to feed capacity (dry tonne per day, dtpd) raised to the 0.7 power. The other estimates are as follows: for IGT, $31.78 million for 1814 dtpd (OPPA, 1990); for MTCI, $3.78 million for 72.5 dtpd (MTCI, 1994), which is for a pressurized reactor (no adjustment is made for the lower pressure unit considered here); for BCL, $7.25 million for 740 dtpd (Breault and Morgan, 1992); Shell-coal, $166.6 million for 7982 dtpd (OPPA, 1989). The Shell-gasifier with biomass is assumed to cost the same as the IGT gasifier, because both are pressurized and the lower cost associated with the higher throughput of an entrained-bed design is assumed to be offset by the higher cost associated with higher temperature operation.

(d) For the coal case, the cost of the high-temperature gas cooling system (plus the shift reactor) is scaled from an estimate of $157.1 million for a plant with a coal feed rate of 7982 dry tonnes per day (OPPA, 1989). A 0.7 power scaling factor is assumed. For the biomass case, the cost is scaled (using 0.7 factor) from that for the coal case according to the heat removal rate in the gas cooler: 215.6 MW in the coal case and 48.2 MW in the biomass case.

(e) The cost for oxygen plants (in million 1991$) is assumed to be $0.260 \times (tO_2pd)^{0.712}$, where $tO_2pd$ is the plant capacity in tonnes of 99.5% purity oxygen per day. This is based on estimates of the cost of 95% purity oxygen plants sold by the Air Products Company for use in integrated coal-gasifier/gas turbine facilities (Brown et al., 1987). The plants produce $O_2$ at 3.7 MPa and include 20 minutes of gaseous oxygen storage. For plant sizes of 1000 $tO_2pd$ or larger, the use of dual trains is assumed, each providing 50% of the capacity. Methanol or hydrogen production requires an $O_2$ purity of 99.5% (or higher). It is assumed that capital cost increases by 15% to produce $O_2$ of 99.5% purity instead of 95% purity (Klosek et al., 1986).

(f) $H_2S$ recovery is required with the coal system. This cost has been scaled using a 0.7 power factor according to the feed rate of dry coal from a baseline cost of $50.3 million for a coal feed rate of 7982 tonnes per day.

(g) Compressor cost is assumed to be $900 per kW of required capacity. There is no reformer in the MTCI case. The cost here refers to the cost of the compressor used to raise the pressure of the syngas before it enters the methanol synthesis loop.

(h) The reformer cost includes costs for boiler feedwater pumps, steam drum, induced-draft and forced-draft fans, all internal heat exchangers, including exchangers to cool the reformate to ambient temperature, desulfurizing vessels, local piping, controls, instrumentation, analyzers, initial catalyst charge and water treating equipment. For the natural gas case, the cost is based on an estimate of Moore (1994). For the other cases, the reformer cost is scaled (using 0.57 power) according to the total heat exchange duty of the reformer, including all preheating, after-cooling, and steam raising associated with the reformer. The total duty in the natural gas case is 767.8 MW.

(i) No shift reactor is required with natural gas or with the MTCI biomass gasifier. For the Shell gasifier cases with biomass and with coal, the shift reactor costs are included in the cost of high temperature gas cooling equipment. For the IGT and BCL gasifier cases, the shift reactor cost is scaled according to the volume flow of $H_2 + CO$, assuming a baseline cost of $9.02 million for a flow rate of 8819 kmol/hour and a scaling factor of 0.65 (Moore, 1994).

(j) For Union Carbide's SELEXOL process, leaving approximately 2% $CO_2$ in the exit gas. Costs are scaled according to volume of $CO_2$ removal raised to the 0.7 power. The baseline estimate is $14.3 million for 810 kmol/hour of $CO_2$ removal Epps (1991).

(k) Estimated cost with the ICI low-pressure methanol synthesis process, including the make-up compressor, recycle compressor, and synthesis loop equipment. A cost estimate of $66.25 million is assumed for a facility with production capacity of 2012 tonnes per day (Moore, 1994). The costs for other capacities have been scaled using a 0.66 power factor (Mansfield, 1991).

(l) The installed cost (in 1991 $/kW$_e$) for a condensing-extraction steam turbine cogeneration system, including a waste heat boiler, is assumed to be $3785 \cdot (MW_e)^{-0.374}$, where $MW_e$ is the electricity generating capacity of the facility. This is based on estimates of the cost for steam bottoming cycles in the range of 20 to 40 $MW_e$ for gas turbine/steam turbine combined cycles (Tittle et al., 1993).

(m) Assumed to be 25% of the sum of other installed hardware costs (Wyman et al., 1993; Moore, 1994).

(n) The following percent of installed equipment costs (given by Wyman et al., 1993 for methanol production) are adopted here: contingencies, 20%; owners costs, fees and profit, 10%; working capital, 10%.

(o) Startup costs are assumed to be 5% of installed hardware costs.

(p) The following costs for land, developed from estimates of Wyman et al. (1993), are assumed: for biomass and coal facilities, land cost (in million 1991$) is $423\text{x}(\text{tpd})^{1.147}$, where tpd is the dry feed capacity in tonnes per day; for the natural gas case the land cost (in 1991$) is assumed to be $0.18 per GJ/yr of natural gas feed capacity.

(q) Assuming levelized costs for delivered feedstocks of $2.0/GJ for biomass chips, $4.1/GJ for natural gas, and $1.45 for coal. It is expected that the price of plantation biomass will be no higher than this in many parts of the world by 2010. The assumed natural gas and coal prices are levelized prices for the period 2010-2035 based on prices projected for US industrial customers in 2010 (EIA, 1995) escalated for 25 years thereafter at rates of 1.0%/year for natural gas and 0.5%/year for coal.

(r) The costs for catalysts and chemicals for the IGT, BCL, and Shell-coal are those estimated by Wyman et al (1993) scaled linearly by production rate. The costs for the MTCI and Shell-biomass cases are scaled according to the cost estimated by Wyman et al. for a biomass case using the Koppers-Totzek entrained bed gasifier, which requires no reforming.

(s) Electricity is the only required external energy input. A cost of 5 cents per kWh is assumed. See Table 3 for quantities.

(t) Labor costs are based on Wyman et al. (1993). For natural gas, Wyman's estimate is used directly. For solid feedstocks, the following relationship for annual labor costs was derived from two BCL cases of different capacities considered by Wyman, et al: $10^6\text{S} = 889\text{x}(\text{dtpd})^{0.959}$, where dtpd is the plant feed rate in dry tonnes per day.

(u) Based on Wyman, et al. (1993), maintenance is assumed to be 3% of installed hardware costs, general overhead is 65% of labor and maintenance, and direct overhead is assumed to be 45% of labor.

(v) Annual capital charge rate of 15.1% is assumed, based on average financial parameters for major US corporations during the period 1984-1988 (9.91% real rate of return on equity, 6.2% real rate of return on debt, a 30% debt fraction, a 44% corporate income tax), a property and insurance rate of 1.5% per year, and a 25-year plant life. For land and working capital, the annual capital charge rate is taken to be 9.91% per year, the corporate discount rate.

(w) This is the price of natural gas at which the total levelized cost of methanol from natural gas would equal the total levelized cost of methanol from the alternative feedstock (biomass or coal).

(x) This is the price of biomass at which the total levelized cost of methanol from biomass would equal the total levelized cost of methanol from coal, with coal costing $1.45/GJ.

Table 6. Estimated production costs of hydrogen from natural gas, biomass, and coal (in 1991$).

| FEEDSTOCK → | BIOMASS | | | | N. GAS | COAL |
|---|---|---|---|---|---|---|
| PROCESS → | IGT gasifier | MTCI gasifier | BCL gasifier | Shell gasifier | Steam reforming | Shell gasifier |
| **Feedstock in input capacity** | | | | | | |
| Dry tonnes/day | 1650 | 1650 | 1650 | 1650 | 1224 | 5000 |
| GJ/hour | 1326 | 1334 | 1338 | 1326 | 2700 | 6188 |
| **Output production capacity**[a] | | | | | | |
| Million $Nm^3$/day | 1.69 | 1.93 | 1.87 | 1.99 | 4.62 | 9.13 |
| GJ/hour | 887 | 1012 | 979 | 1044 | 2422 | 4790 |
| **Annual feed and output** | | | | | | |
| Feed ($10^6$ GJ/year) | 10.45 | 10.52 | 10.55 | 10.45 | 21.29 | 48.79 |
| Product output ($10^6$ GJ/year) | 6.99 | 7.98 | 7.72 | 8.23 | 19.09 | 37.76 |
| **Installed Equipment Costs** ($10^6$ $) | | | | | | |
| Feed preparation[b] | 17.32 | 13.21 | 13.17 | 38.78 | 0.00 | 67.96 |
| Gasifier[b] | 29.74 | 33.72 | 12.72 | 29.74 | 0.00 | 120.06 |
| High temperature gas cooling[c] | 0.00 | 0.00 | 0.00 | 39.67 | 0.00 | 113.27 |
| Oxygen plant[b] | 21.55 | 0.00 | 0.00 | 28.77 | 0.00 | 95.42 |
| Sulfur removal[b] | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 36.25 |
| Reformer feed compressor[b] | 0.00 | 14.99 | 11.88 | 0.00 | 0.00 | 0.00 |
| Reformer[d] | 22.45 | 0.00 | 18.38 | 0.00 | 43.91 | 0.00 |
| Shift reactors[b] | 4.70 | 5.10 | 5.01 | 3.10 | 9.02 | 7.28 |
| PSA recycle compressor[e] | 3.66 | 4.82 | 2.43 | 1.61 | 7.11 | 16.43 |
| PSA (with $CO_2$ removal)[f] | 14.50 | 16.02 | 15.63 | 16.40 | 30.82 | 51.39 |
| Hydrogen compressor[d] | 2.53 | 5.31 | 5.82 | 4.23 | 9.64 | 19.22 |
| Steam turbine cogeneration plant | 7.84 | 4.21 | 15.87 | 12.04 | 11.08 | 37.19 |
| Utilities/auxiliaries[b] | 31.07 | 24.35 | 25.23 | 43.58 | 27.90 | 141.12 |
| Subtotal | 155.37 | 121.74 | 126.13 | 217.91 | 139.48 | 705.58 |
| Contingencies[b] | 31.07 | 24.35 | 25.23 | 43.58 | 27.90 | 141.12 |
| Owners costs, fees, profits[b] | 15.54 | 12.17 | 12.61 | 21.79 | 13.95 | 70.56 |
| Startup[b] | 7.77 | 6.09 | 6.31 | 10.90 | 6.97 | 35.28 |
| **Total Capital Requirement** ($10^6$ $) | 209.76 | 164.35 | 170.27 | 294.18 | 188.30 | 952.53 |
| **Working Capital**[b] ($10^6$ $) | 15.54 | 12.17 | 12.61 | 21.79 | 13.95 | 70.56 |
| **Land**[b] ($10^6$ $) | 2.08 | 2.08 | 2.08 | 2.08 | 4.26 | 7.40 |
| **Variable Operating Costs** ($10^6$ $/year) | | | | | | |
| Feed[b] | 20.90 | 21.03 | 21.10 | 20.90 | 87.28 | 70.74 |
| Catalysts and chemicals[b] | 1.67 | 0.67 | 2.24 | 0.67 | 2.58 | 10.87 |
| Purchased energy[g] | 10.31 | 13.68 | 8.38 | 11.18 | 7.82 | 52.55 |
| Subtotal | 32.88 | 35.38 | 31.71 | 32.74 | 97.68 | 138.80 |
| **Fixed Operating Costs**[b] ($10^6$ $/year) | | | | | | |
| Labor | 1.08 | 1.08 | 1.08 | 1.08 | 1.00 | 3.14 |
| Maintenance | 4.66 | 3.65 | 3.78 | 6.54 | 4.18 | 21.17 |
| General Overhead | 3.73 | 3.08 | 3.16 | 4.95 | 3.37 | 15.80 |
| Direct Overhead | 0.49 | 0.49 | 0.49 | 0.49 | 0.45 | 1.41 |
| Subtotal | 9.96 | 8.30 | 8.52 | 13.06 | 9.00 | 41.51 |
| **Total Operating Costs** ($10^6$ $/year) | 42.85 | 43.68 | 40.23 | 45.81 | 106.68 | 180.31 |
| PRODUCTION COSTS ($/ GJ of $H_2$) | | | | | | |
| Capital[b] | 4.78 | 3.29 | 3.52 | 5.68 | 1.58 | 4.01 |
| Labor & maintenance | 1.66 | 1.12 | 1.39 | 1.67 | 0.61 | 1.51 |
| Purchased energy | 1.48 | 1.71 | 1.08 | 1.36 | 0.41 | 1.51 |
| Feedstock | 2.99 | 2.64 | 2.73 | 2.54 | 4.57 | 1.87 |
| TOTAL PRODUCTION COST ($/GJ) | 10.91 | 8.76 | 8.73 | 11.24 | 7.17 | 8.91 |
| Natural gas price ($/GJ) for same total cost[h] | 7.45 | 5.52 | 5.50 | 7.75 | 4.10 | 5.66 |
| Biomass price ($/GJ) for same total cost as coal[i] | 0.66 | 2.12 | 2.13 | 0.16 | n.a. | n.a. |

Notes to Table 6:

(a) Based on the process energy ratios in Table 4.

(b) The basis for the cost estimate for this item is the same as for $CH_3OH$ production--see corresponding note to Table 5.

(c) For the Shell gasifier cases, this item includes the high temperature gas cooling and the shift reactors. The gas cooling costs are the same as for the corresponding methanol production cases, but additional shift reactor capacity is needed for hydrogen production. Thus, the costs estimated for the methanol case (Table 5) have been increased by an amount representing the added cost of the additional shift reactor capacity. The cost of the additional shift reactor capacity has been estimated by scaling according to incremental volume flow of $H_2$ + CO using a 0.65 power factor and assuming a baseline cost of $9.02 million for a flow of 8819 kmol/hour (Moore, 1994).

(d) The reformer cost includes costs for boiler feedwater pumps, steam drum, induced-draft and forced-draft fans, all internal heat exchangers, including exchangers to cool the reformate to ambient temperature, desulfurizing vessels, local piping, controls, instrumentation, analyzers, initial catalyst charge and water treating equipment. For the natural gas case, the cost is based on an estimate of Moore (1994). For the other cases, the reformer cost is scaled (using 0.57 power) according to the total heat exchange duty of the reformer, including all preheating, after-cooling, and steam raising associated with the reformer. The total duty in the natural gas case is 560.13 MW.

(e) Compressors are assumed to cost $900 per kW of capacity.

(f) Assuming use of the "Gemini-9" pressure swing adsorption system from Air Products, Inc., which removes $CO_2$ and $H_2O$ in a first bed and produces a fuel gas of 99.999% purity $H_2$ out of a second bed. The estimated cost for the natural gas case is $30.82 million for a hydrogen production rate of 8474 kmol/hour (Moore, 1994). For the other cases, cost are scaled according to the hydrogen production rate raised to the 0.7 power. The cost excludes the recycle compressor.

(g) For external electricity input, a cost of 5 cents per kWh is assumed. External heat input is charged at $4/GJ. Quantities are given in Table 4.

(h) This is the price of natural gas at which the total levelized cost of hydrogen from natural gas would equal the total levelized cost of hydrogen from the alternative feedstock (biomass or coal).

(i) This is the price of biomass at which the total levelized cost of hydrogen from biomass would equal the total levelized cost of hydrogen from coal, with coal costing $1.45/GJ.

Table 7. Summary of estimated delivered retail fuel prices[a] (in 1991 $).

| Energy carrier | Reformulated gasoline[b] | Methanol | | | Hydrogen | | |
|---|---|---|---|---|---|---|---|
| Feedstock | Crude | Biomass[c] | N. gas | Coal | Biomass[a] | N. gas | Coal |
| Feedstock cost ($/GJ) | $3.75<br>($23/bbl) | 2.0 | 4.1 | 1.45 | 2.0 | 4.1 | 1.45 |
| Cost components for delivered fuel ($/GJ) | | | | | | | |
| Production[d] | 6.76 | 11.24 | 9.46 | 10.14 | 8.73 | 7.17 | 8.91 |
| Transport to filling station[e] | 0.99 | 1.9 | 1.9 | 1.9 | 0.5 | 0.5 | 0.5 |
| Filling station cost[e] | 0.61 | 1.2 | 1.2 | 1.2 | 4.4 | 4.4 | 4.4 |
| Delivered retail price ($/GJ)<br>[$/liter gasoline equivalent[f]]<br>[$/gallon gasoline equivalent[f]] | 8.36<br>[0.29]<br>[1.10] | 14.34<br>[0.50]<br>[1.89] | 12.56<br>[0.44]<br>[1.66] | 13.24<br>[0.46]<br>[1.75] | 13.63<br>[0.48]<br>[1.80] | 12.07<br>[0.42]<br>[1.59] | 13.81<br>[0.48]<br>[1.82] |
| Vehicle type | ICEV | FCV | | | FCV | | |
| Fuel cost per unit of service[g] (cents per km) | 2.65 | 1.92 | 1.69 | 1.76 | 1.58 | 1.38 | 1.58 |

(a) Excluding retail fuel taxes.

(b) Based on a projected crude oil price in the US of $23/barrel in 2010 (EIA, 1995).

(c) Assumes use of the BCL gasifier for methanol and hydrogen production..

(d) Production costs for methanol and hydrogen derived from biomass, natural gas, and coal are from Tables 5 and 6. For reformulated gasoline, with 0.132 GJ/gallon, the production cost per gallon is 0.9·($/bbl crude)/42 + 0.25 + 0.15, where 0.9 is the fraction of a gallon of gasoline derived from crude oil, 42 is the number of gallons per barrel, $0.25/gallon is the estimated cost of refining standard formula gasoline and $0.15/gallon is the additional cost for refining reformulated gasoline (DeLuchi, 1992).

(e) See Ogden, et al. (1994).

(f) The fuel price in $ per unit volume of gasoline equivalent is calculated as the $/GJ price times the higher heating value of gasoline (0.035 GJ/liter or 0.132 GJ/gallon).

(g) This is the fuel price (in $ per liter) divided by the gasoline-equivalent fuel economy (in km per liter). The assumed ICEV is a year-2000 version of the Ford Taurus, having a fuel economy of 11.0 km/liter (25.8 miles per gallon) when operated on reformulated gasoline. The FCV version of this automobile would have a gasoline-equivalent fuel economy of 26.1 km/l (61.5 mpg) when operated on methanol and 30.4 km/l (71.6 mpg) when operated on compressed hydrogen (Ogden et al., 1994).

Table 8. Technical and cost characteristics of vehicles considered in the analysis.[a]

| Vehicle type ====> | ICEV[b] | FCV | |
|---|---|---|---|
| Fuel ====> | Ref. gasoline | Methanol | Hydrogen |
| Energy storage system | metal tank | metal tank | carbon/aluminum tank[c] |
| Driving range (km) | 640[d] | 560 | 400 |
| Max. power to wheels (kW)[e] | 101 | 76 | 76 |
| Vehicle life (km) | 193,080 | 241,350 | 241,350 |
| Annual vehicle use (km per year)[f] | 17,800 | 23,800 | 23,800 |
| Volume of fuel storage and fuel cell system (lit.)[g] | 63 | 301 | 310 |
| Weight of complete vehicle (kg)[h] | 1,415 | 1,328 | 1,311 |
| Coefficient of drag[i] | 0.28 | 0.23 | 0.23 |
| Fuel economy (gasoline-equivalent mpg, HHV)[j] | 25.8 | 61.5 | 71.6 |
| Fuel economy (gasoline-equivalent liters/100 km) | 9.1 | 3.8 | 3.3 |
| Full retail price of vehicle, incl. taxes (1991$) | 17,976 | 21,709 | 25,091 |
| Annual vehicle maintenance cost (1991$/yr) | 396 | 389 | 376 |

(a) From Ogden, Larson, and DeLuchi (1994), based on DeLuchi (1992) and unpublished updates thereof.

(b) The gasoline vehicle is a year-2000 version of the 1990 Ford Taurus.

(c) Carbon-wrapped aluminum ultra-high pressure vessel, storing $H_2$ at 550 bar. As storage pressure increases, the bulk of the storage system decreases but cost increases. This pressure represents a good balancing of these two opposing tendencies.

(d) The 1990 Ford Taurus has a 16.0-gallon gasoline tank. If in the year 2000 the tank were the same size and the vehicle got about 25 mpg in use, the drving range would be about 400 miles, or 640 km.

(e) For the FCVs, the peak power of the motor has been calculated given the peak power of the ICE, the desired high-end acceleration of the FCV relative to the high-end acceleration of the ICEV, and the mass, drag, and rolling resistance of the FCV and ICEV. The ratio of the maximum acceleration of the FCV at 60 mph to the maximum acceleration of the ICEV at 60 mph is assumed to be 0.80:1.00. Note, though, that the FCVs would perform better than the ICEVs at low speeds.

(f) It is assumed that the annual use of the FCV is higher than the ICEV because of the lower operating cost (DeLuchi, 1992).

(g) The volume of the fuel tank plus the fuel cell and methanol reformer.

(h) Including one passenger and fuel to 40% of vehicle capacity.

(i) The coefficient of drag is assumed to be lower for FCVs than for ICEVs because of the higher value of improving the efficiency of FCVs, due in turn to the higher cost of fuel storage.

(j) Gasoline-equivalent fuel economy in miles/gallon is calculated as the mile/million-Btu fuel economy of the alternative vehicle in combined city and highway driving in the year 2000, divided by 125,000 Btu/gallon-gasoline (34,830 kJ/liter). HHV indicates higher heating values for hydrogen and methanol. (On a lower heating value basis, the gasoline-equivalent fuel economy for hydrogen vehicles is 1.092 times higher than the higher-heating value basis. It is 1.054 times higher for methanol vehicles.)

The mpg fuel economy of the gasoline ICEV has been calculated from a detailed set of input parameters, including vehicle weight, powertrain efficiency, aerodynamic drag, the amount of city vs. highway driving, and other factors. The miles/million Btu fuel economy of the FCEVs were calculated from a detailed set of input parameters, including vehicle weight, powertrain efficiency, aerodynamic drag, the amount of city vs. highway driving, and other factors. An electric powertrain, consisting of the motor, controller, and transmission, is at least 6 times more efficient than an ICE powertrain, in combined city/highway driving (after accounting for regenerative braking). PEM fuel cells are about 45% efficient (after accounting for the energy consumption of auxiliaries); hence, the fuel cell/electric motor system would be almost 3 times as efficient as the ICE, before accounting for differences in vehicle weight and aerodynamic drag.

Table 9. Lifecycle cost of transportation (cents per km, 1991$) for ICEVs and FCVs.[a]

| | Lifecycle Transportation Cost (cents per km) | | | | | Break-even gasoline price ($/gal.)[b] |
|---|---|---|---|---|---|---|
| | Base vehicle, excluding fuel storage and fuel cell | Fuel storage and fuel cell | Non-fuel operating costs[c] | Fuel (excl. tax) | TOTAL | |
| ICEV fueled with reformulated gasoline | 11.60 | in base cost | 6.58 | 2.65 | 20.83 | n.a. |
| FCV fueled with | | | | | | |
| Methanol from | | | | | | |
| Biomass | 9.65 | 1.88 | 6.02 | 1.92 | 19.47 | 0.54 |
| Natural gas | 9.65 | 1.88 | 6.02 | 1.69 | 19.24 | 0.44 |
| Coal | 9.65 | 1.88 | 6.02 | 1.76 | 19.31 | 0.47 |
| Hydrogen from | | | | | | |
| Biomass | 9.88 | 2.53 | 6.01 | 1.58 | 20.00 | 0.76 |
| Natural gas | 9.88 | 2.53 | 6.01 | 1.38 | 19.80 | 0.67 |
| Coal | 9.88 | 2.53 | 6.01 | 1.58 | 20.00 | 0.76 |

(a) All cost components except fuel are from Ogden, et al (1994). Fuel costs are from Table 7.

(b) The breakeven gasoline price is the retail price of gasoline excluding taxes (which averaged $0.31/gallon or $0.082/liter in 1991 in the United States) at which the lifecycle consumer cost-per-km of the methanol or hydrogen vehicle would equal that of the gasoline vehicle. For comparison, the actual pre-tax gasoline price we assumed in our analysis is $1.10/gallon (Table 7).

(c) Includes regular maintenance and repairs, oil consumption, replacement tires, parking and tolls, registration fee, inspection and maintenance fee, accessories, and fuel taxes (assumed to be the same, in cents per km, for all cases, based on current US-average state-plus-Federal gasoline taxes).

Table 10. Lifecycle $CO_2$ emissions from the production of alternative energy carriers from fossil fuel feedstocks and their use in alternative automobiles (grams carbon per km of travel).

| Energy carrier | Reformulated gasoline | Methanol | | | | Hydrogen | | | |
|---|---|---|---|---|---|---|---|---|---|
| Feedstock | Crude Oil | Natural Gas | | Coal | | Natural Gas | | Coal | |
| Vehicle type[a] | ICEV | ICEV | FCV | ICEV | FCV | ICEV | FCV | ICEV | FCV |
| ACTIVITY | $CO_2$ EMISSIONS (GRAMS OF CARBON PER KM OF VEHICLE TRAVEL)[b] | | | | | | | | |
| Gas well $CO_2$[c] | - | 1.20 | 0.57 | - | - | 0.92 | 0.38 | - | - |
| Feedstock recovery[d] | 1.77 | 1.61 | 0.76 | 1.14 | 0.54 | 1.23 | 0.51 | 0.93 | 0.39 |
| Feed production[e] | - | 1.44 | 0.68 | - | - | 1.10 | 0.46 | - | - |
| Feedstock transport[f] | 0.81 | - | - | 0.61 | 0.29 | - | - | 0.50 | 0.21 |
| Fuel production[g] | | | | | | | | | |
| From feedstock | 8.70 | 9.00 | 4.26 | 60.32 | 28.54 | 42.43 | 17.72 | 87.21 | 36.42 |
| External electr. | 1.55 | 3.89 | 1.84 | 5.23 | 2.47 | 4.27 | 1.78 | 14.49 | 6.05 |
| Fuel transport to refueling station[h] | 0.57 | 2.18 | 1.04 | 1.25 | 0.59 | - | - | - | - |
| Compressors at refueling station[i] | - | - | - | - | - | 9.92 | 4.14 | 9.92 | 4.14 |
| End use | 58.22 | 46.55 | 22.02 | 46.55 | 22.02 | - | - | - | - |
| TOTAL EMISSIONS | | | | | | | | | |
| Grams C per km | 71.61 | 65.86 | 31.17 | 115.08 | 54.45 | 59.86 | 25.00 | 113.05 | 47.21 |
| Relative to gasoline ICEV | 1.00 | 0.92 | 0.44 | 1.61 | 0.76 | 0.84 | 0.35 | 1.58 | 0.66 |

(a) The gasoline ICEV is a year-2000 version of the 1990 Ford Taurus haveing a fuel economy of 11.0 km/l (25.8 mpg). The methanol and hydrogen vehicles would be comparable-duty vehicles. The ICEVs running on methanol and hydrogen would have fuel economies of 12.4 km/l (29.1 mpg) and 12.7 km/l (29.9 mpg), respectively. The FCVs cars running on methanol and hydrogen would have fuel economies of 26.1 km/l (61.5 mpg) and 30.4 km/l (71.6 mpg), respectively. For details, see Ogden et al. (1994).

(b) The following carbon emission rates are assumed in this analysis for fuels (in kg C/GJ): crude oil, 18.73; residual fuel, 19.42; gasoline, 18.31; diesel fuel, 18.65; coal, 24.60; natural gas, 13.87; and methanol, 16.41. Carbon emissions from electricity use are assumed to be 189.72 g C/kWh, which corresponds to emissions from primary energy sources representing the average mix of US electric power generating sources (56.34% coal, 9.43% natural gas, and 3.18% residual fuel), their respective average heat rates (10.86 MJ/kWh, 10.73 MJ/kWh, and 10.70 MJ/kWh), and transmission and distribution losses of 7.4%. Emissions from production and delivery of fuels to power plants are also included in the total per-kWh emissions.

(c) Based on estimated emissions of $CO_2$ from natural gas wells of 1,102 $gCO_2/GJ$ of gas (200 gC/GJ) [Table 7, DeLuchi (1991)].

(d) Estimated energy use during feedstock recovery is as follows, based on Tables 3 and 4 in DeLuchi (1991). Crude oil recovery: 0.0254 GJ/GJ of gasoline, 13% of which is consumed as crude oil, 14% as diesel fuel, 50% as natural gas, 17% as electricity, 4% as gasoline, and 10% as residual fuel. For natural gas recovery: 0.0279 GJ/GJ of natural gas, 1% of which is crude oil, 4% diesel fuel, 92% natural gas, 1% electricity, and 1% gasoline. For coal recovery: 0.0083 GJ/GJ of coal, 5% of which is crude oil, 48% diesel fuel, 1% natural gas, 37% electricity, 3% gasoline, and 6% coal.

(e) Estimated energy use during feedstock production is as follows, based on Tables 3 and 4 in DeLuchi (1991). Crude oil and coal: energy requirements are included in recovery. For natural gas: 0.0245 GJ/GJ of natural gas, 98% of which is natural gas and 2% is electricity.

(f) Energy requirements by fuel type for crude oil and coal transport are from Tables 3 and 4 in DeLuchi (1991). Estimated energy use during feedstock transportation is as follows, based on Tables 3 and 4 in DeLuchi (1991). Crude oil: 0.0116 GJ/GJ of gasoline, 13% of which is crude oil, 7.4% is electricity, and 91.3% is residual fuel. For natural gas, transportation energy use is zero because the fuel production facilities are assumed to be located at the wellhead. For coal: 0.0075 GJ/GJ of coal, 1.3% of which is crude oil, 74.2% of which is diesel fuel and 25.8% is residual fuel.

(g) Energy requirements for fuel production from feedstock are estimated to be as follows. For gasoline, 0.1847 GJ/GJ of gasoline, 77% of which is natural gas, 5% is electricity, 1% is residual fuel, and 16% is coal [Tables 3 and 4 in DeLuchi (1991)]. For natural gas and coal, the fraction of feedstock energy not converted to fuel is (1 - ER), where ER is the energy ratio given in Table 3 for methanol production and Table 4 for hydrogen production. Tables 3 and 4 also give electricity supplied from external sources. For methanol these are 7.274 kWh/GJ of methanol with natural gas feed and 9.771 kWh/GJ of methanol with coal. For hydrogen, these are 8.193 kWh/GJ of hydrogen with natural gas feed and 22.957 kWh/GJ of hydrogen.

(h) Energy requirements associated with gasoline and methanol delivery to the refueling station are based on Tables 3 and 4 in DeLuchi (1991). For gasoline, 0.0084 GJ/GJ of gasoline are needed, of which 6.9% is electricity, 70.5% is diesel fuel, and 22.6% is residual fuel. For methanol from natural gas, 0.0378 GJ/GJ of methanol are needed, of which 3% is electricity, 26% is diesel fuel, and 72% is residual fuel. For methanol from coal, 0.019 GJ/GJ are needed, or which 12% is electricity, 60% is diesel fuel, and 19% is residual fuel. Transport energy requirements for methanol from natural gas are higher than for methanol from coal because DeLuchi assumes that methanol is produced at remote natural gas sites, while methanol from coal is produced much closer to the point of use. Hydrogen is assumed to be sufficiently compressed at the production facility for pipeline delivery to the refueling station with no additional energy inputs.

(i) We calculate (using ASPEN PLUS process simulation software) that compression at the refueling station (from 50 to 8400 psia with 85% compressor efficiency) requires 19.06 kWh/GJ of hydrogen.

Table 11. Lifecycle $CO_2$ emissions from the production of methanol and hydrogen from biomass and their use in alternative automobiles (grams carbon per km of travel).

| Energy carrier | Methanol | | | | Hydrogen | | | |
|---|---|---|---|---|---|---|---|---|
| Gasifier type | IGT | | BCL | | IGT | | BCL | |
| Vehicle type[a] | ICEV | FCV | ICEV | FCV | ICEV | FCV | ICEV | FCV |
| ACTIVITY | $CO_2$ EMISSIONS (GRAMS OF CARBON PER KM OF VEHICLE TRAVEL)[b,c] | | | | | | | |
| Feedstock Production[d] | | | | | | | | |
| Establishment | 0.06 | 0.03 | 0.05 | 0.03 | 0.05 | 0.02 | 0.04 | 0.02 |
| Fertilizers | 1.08 | 0.51 | 1.00 | 0.48 | 0.89 | 0.37 | 0.81 | 0.34 |
| Herbicides | 0.16 | 0.08 | 0.15 | 0.07 | 0.13 | 0.06 | 0.12 | 0.05 |
| Equipment | 0.07 | 0.03 | 0.07 | 0.03 | 0.06 | 0.02 | 0.05 | 0.02 |
| Harvesting | 3.04 | 1.44 | 2.84 | 1.34 | 2.50 | 1.04 | 2.28 | 0.95 |
| Hauling | 1.00 | 0.47 | 0.93 | 0.44 | 0.82 | 0.34 | 0.75 | 0.31 |
| Subtotal, feed production | 5.39 | 2.55 | 5.04 | 2.38 | 4.44 | 1.86 | 4.06 | 1.70 |
| Photosynthetic credit[e] | - 122.03 | - 57.74 | - 113.98 | - 53.93 | - 100.48 | - 41.96 | - 91.83 | - 38.35 |
| Fuel Production[f] | | | | | | | | |
| From feedstock | 75.49 | 35.72 | 67.43 | 31.91 | 100.48 | 41.96 | 91.83 | 38.35 |
| External electricity from fossil fuels | 5.00 | 2.37 | 4.84 | 2.29 | 15.36 | 6.41 | 11.30 | 4.72 |
| Subtotal, fuel production | 80.49 | 38.09 | 72.27 | 34.20 | 115.84 | 48.37 | 103.14 | 43.07 |
| Fuel transport to refueling station[g] | 1.25 | 0.59 | 1.25 | 0.59 | - | - | - | - |
| Compressors at refueling station[h] | - | - | - | - | 9.92 | 4.14 | 9.92 | 4.14 |
| End use | 46.55 | 22.02 | 46.55 | 22.02 | - | - | - | - |
| TOTAL NET EMISSIONS | | | | | | | | |
| Grams C per km | 11.64 | 5.51 | 11.12 | 5.26 | 29.72 | 12.41 | 25.28 | 10.56 |
| Relative to gasoline ICEV | 0.16 | 0.077 | 0.16 | 0.073 | 0.42 | 0.17 | 0.35 | 0.15 |
| Total net emissions if external electricity for fuel production is generated from biomass[i] | | | | | | | | |
| Grams C per km | 6.95 | 3.29 | 6.59 | 3.12 | 15.29 | 6.39 | 14.67 | 6.13 |
| Relative to gasoline ICEV | 0.097 | 0.046 | 0.092 | 0.044 | 0.214 | 0.089 | 0.205 | 0.086 |

(a) For the fuel economies of the gasoline ICEV and FCVs considered here see note *(a)* of Table 10.

(b) The following carbon emission rates are assumed in this analysis for fuels (in kg C/GJ): crude oil, 18.73; residual fuel, 19.42; gasoline, 18.31; diesel fuel, 18.65; coal, 24.60; natural gas, 13.87; methanol, 16.41; and biomass, 24.50.

(c) Carbon emissions from electricity use are assumed to be 189.72 g C/kWh, which corresponds to emissions from primary energy sources representing the average mix of US electric power generating sources (56.34% coal, 9.43% natural gas, and 3.18% residual fuel), their respective average heat rates (10.86 MJ/kWh, 10.73 MJ/kWh, and 10.70 MJ/kWh), and T&D losses of 7.4%. Emissions associated with production and delivery of the fuels to power plants are also included in the total per-kWh emissions. We have also calculated the carbon emissions associated with electricity production from biomass (instead of fossil fuels). See note *(i)* below.

(d) Assumed biomass yield (11.3 dry tonnes per hectare per year, after counting harvesting and handling losses) and energy inputs for short-rotation intensive culture production of hybrid poplar are from Turhollow and Perlack (1991). Energy inputs are as follows. Plantation establishment requires 14 GJ/ha/yr of diesel fuel. Fertilizers require 0.24 GJ/ha/yr diesel fuel, 2.810 GJ/ha/yr natural gas, and 25.55 kWh/ha/yr electricity. Pesticides require 0.29 GJ/ha/yr diesel fuel, 0.10 GJ/ha/yr natural gas, and 1.825

kWh/ha/yr electricity. Equipment requires 0.17 GJ/ha/yr diesel fuel. Harvesting requires 7.31 GJ/ha/yr diesel fuel. Hauling requires 2.4 GJ/ha/yr diesel fuel.

(e) Assumes an uptake of 485.1 kg of carbon per dry tonne of biomass.

(f) The fraction of biomass feedstock that is not converted to fuel is (1 - ER), where ER are the energy ratios given in Table 3 for methanol production and Table 4 for hydrogen production. Tables 3 and 4 also give electricity supplied from external sources. For methanol this is 9.35 kWh/GJ of methanol with the IGT gasifier and 9.041 kWh/GJ of methanol with the BCL gasifier. For hydrogen, this is 29.50 kWh/GJ of hydrogen for the IGT gasifier and 21.71 kWh/GJ of hydrogen for the BCL gasifier. This electricity is assumed to be provided by the average US electric utility power mix--see note *(c)*.

(g) The energy requirements for methanol delivery to the refueling station are assumed to be the same as for methanol from coal. See note *(h)* in Table 10. Hydrogen is assumed to be sufficiently compressed at the production facility for pipeline delivery to the refueling station with no additional energy inputs.

(h) We calculate (using ASPEN PLUS process simulation software) that compression at the refueling station (from 50 to 8400 psia with 85% compressor efficiency) requires 19.06 kWh/GJ of hydrogen.

(i) The external electricity requirements for fuel production could be met by electricity produced from biomass, rather than from fossil fuels. The biomass consumption for electricity production is based on a heat rate corresponding to that estimated for a biomass-gasifier/steam-injected gas turbine power station. From Figure 5.5 in Katofsky (1993), this heat rate is (in GJ/kWh): $0.0036/(0.3239 + 0.00059*MW_e)$, where $MW_e$ is the required electricity production capacity. The external electricity supply requirements are given in Tables 3 and 4 for methanol and hydrogen, respectively.

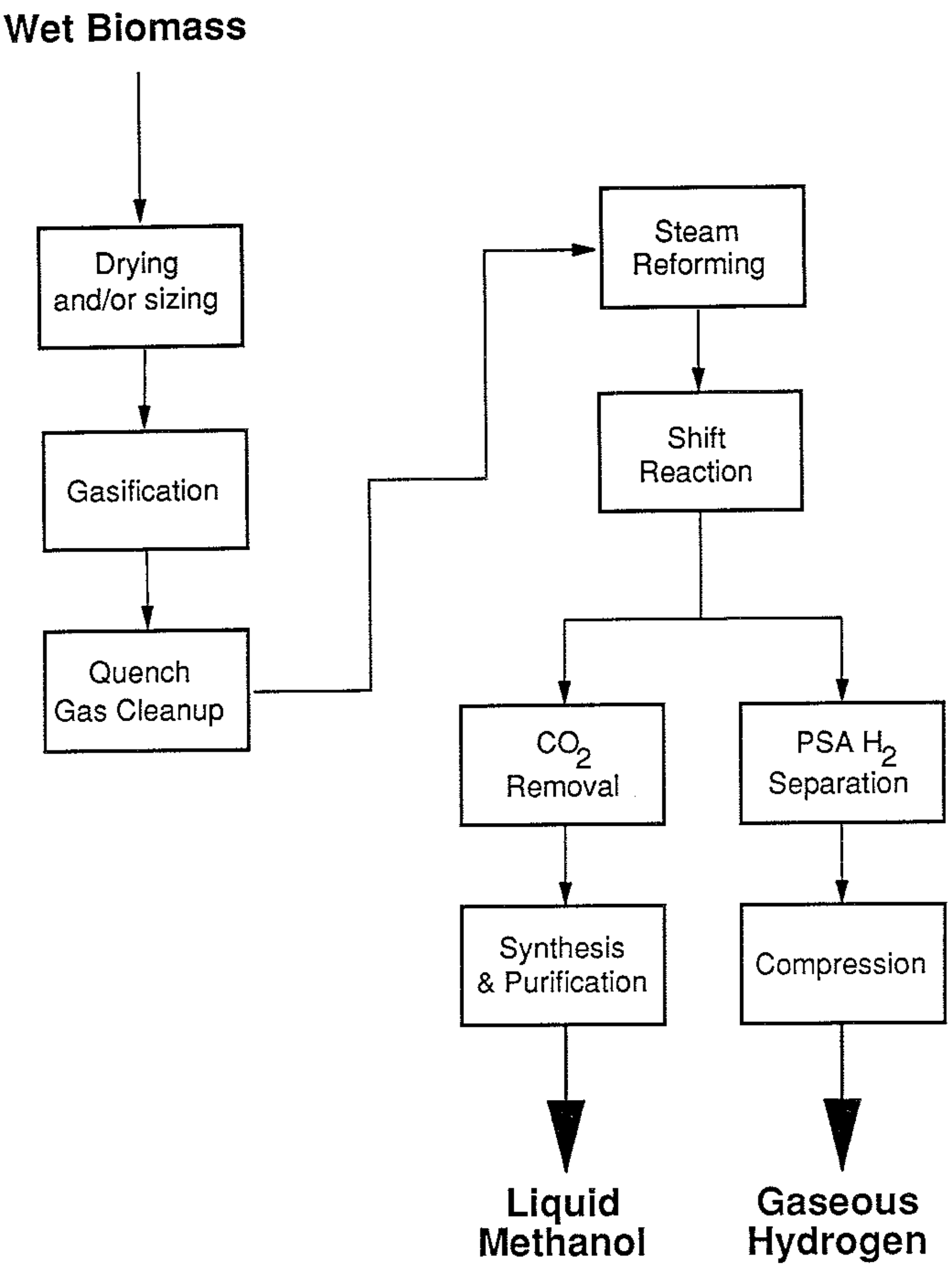
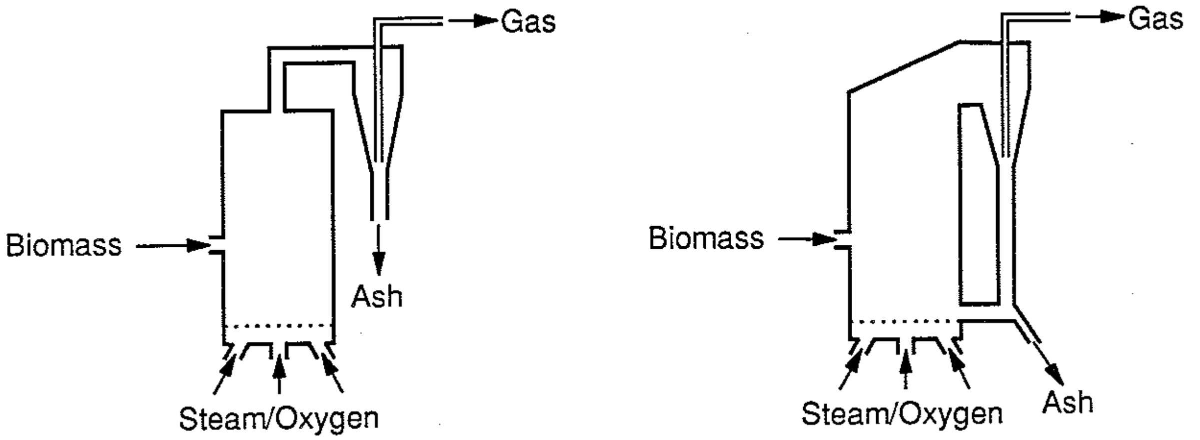
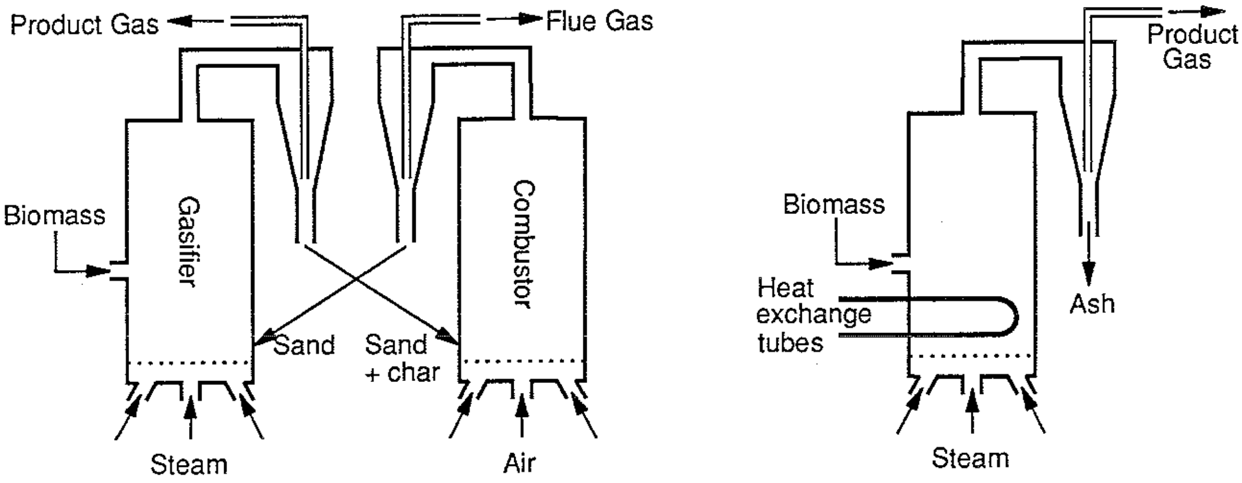


Fig. 1. Thermochemical processing steps in the production of methanol ($CH_3OH$) or hydrogen ($H_2$) from biomass. (PSA is pressure swing adsorption.)



(a) bubbling fluidized-bed gasifier (b) circulating fluidized-bed gasifier



(c) indirectly heated gasifier (BCL) (d) indirectly-heated gasifier (MTCI)

Fig. 2. Alternative designs for thermochemical gasification of biomass: two directly-heated gasifiers--a bubbling fluidized-bed gasifier *(a)* and a circulating fluidized bed gasifier *(b)*; and two indirectly-heated gasifiers that are being developed in the US. In the twin-fludized bed design *(c)*, pyrolytic gasification is driven by heat from circulating sand; the sand and char are separated from the product gas and sent to a char combustor where the sand is heated again. An alternative design *(d)* uses an in-bed heat exchanger to provide the heat for pyrolysis and char gasification. In both indirectly-heated gasifiers steam and (not shown) recirculated product gas, act as fluidizing agents.

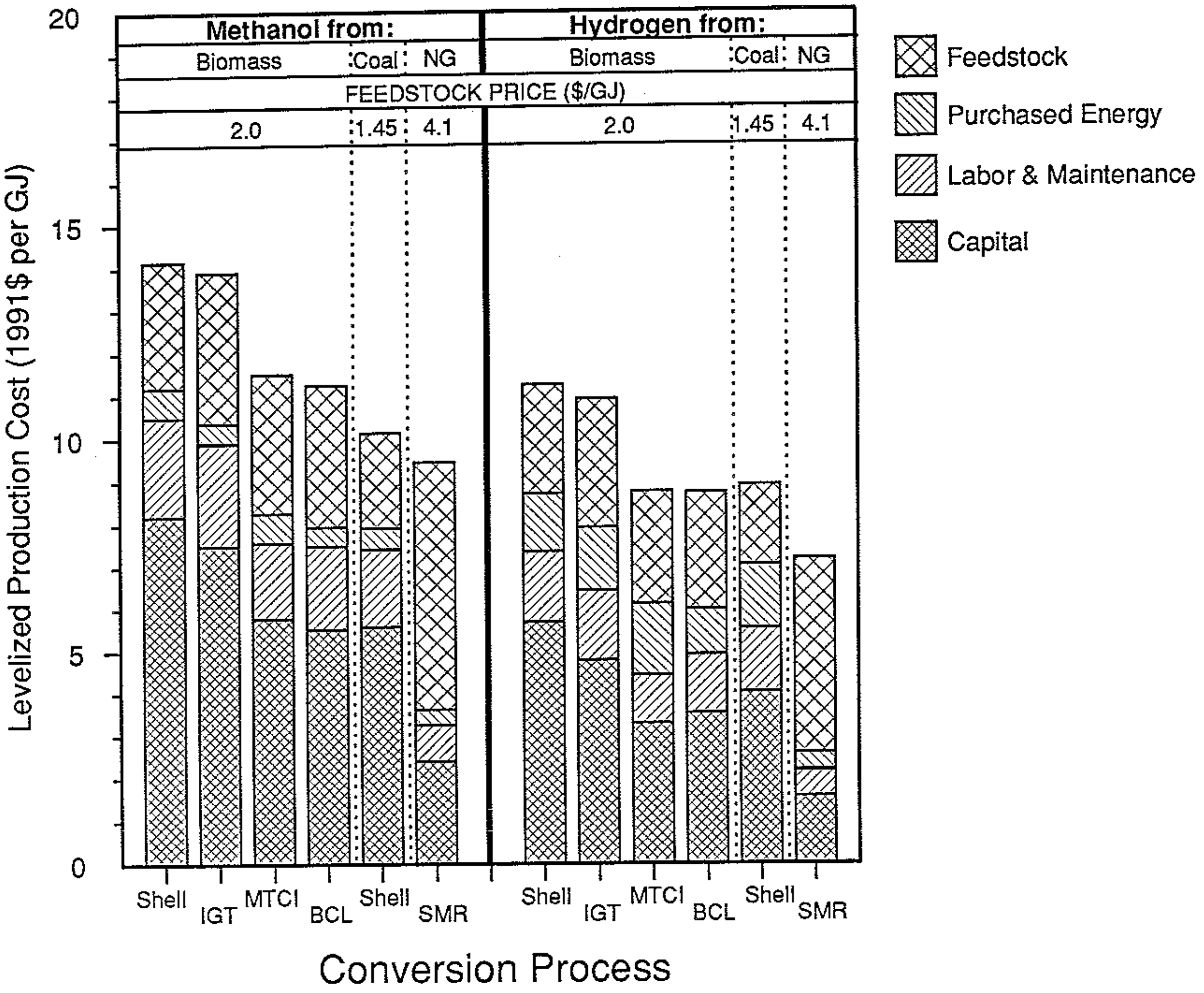
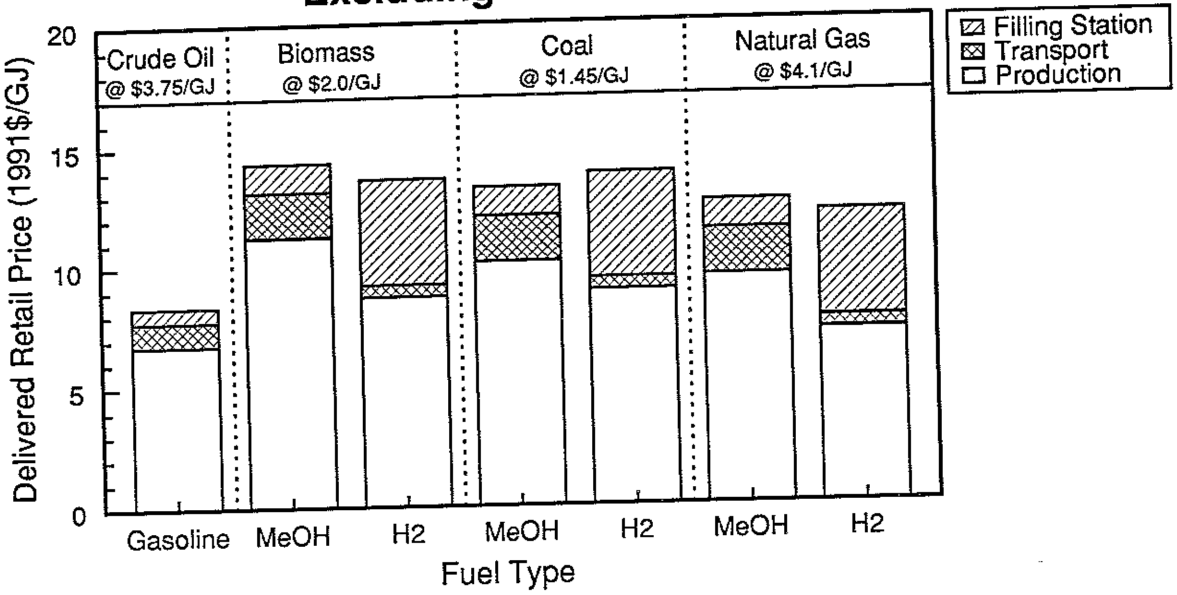


Fig. 3. Estimated total levelized costs (in 1991 US$ per GJ) of producing methanol and hydrogen from biomass (using alternative gasifiers), natural gas [via steam methane reforming (SMR)], and coal. See Tables 5 and 6 for details.

Estimated Delivered Retail Fuel Prices, Excluding Retail Taxes



Estimated Fuel Cost per Unit of Service

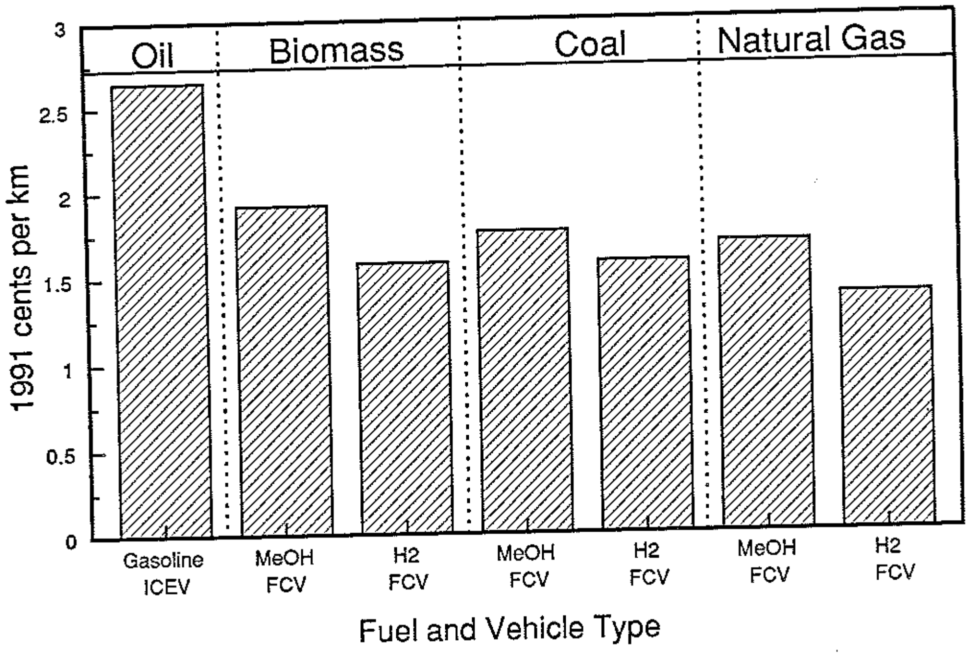


Fig. 4 Costs to consumer for alternative fuels. *(a)* Estimated retail price (in 1991 US$ per GJ), excluding retail taxes, of methanol and hydrogen produced from biomass, coal, and natural gas. *(b)* Estimated fuel cost to consumer per vehicle-km of transport service. See Table 7 for details.

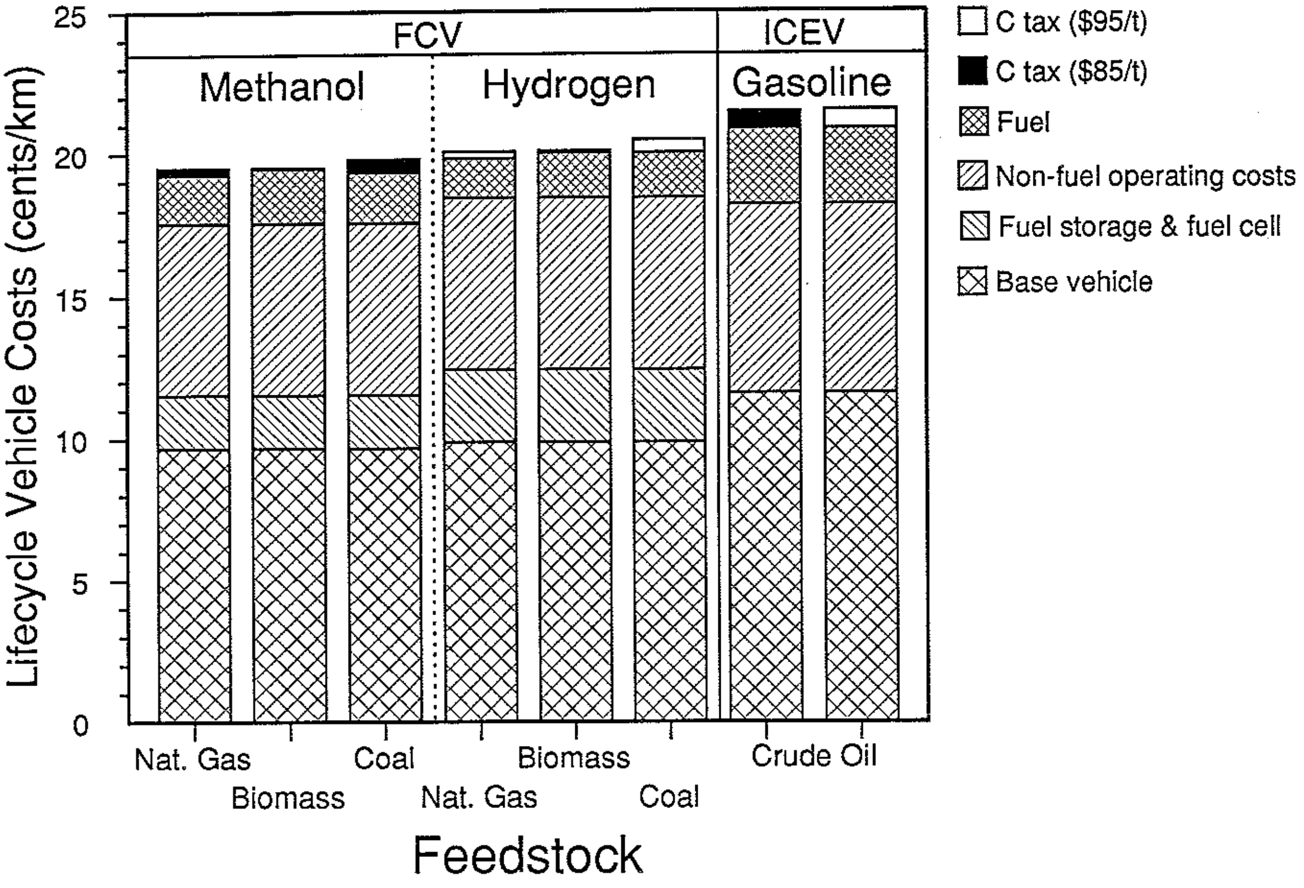


Fig. 5. Estimated lifecycle costs (in 1991 US cents per km) for FCVs fueled with methanol and hydrogen derived from alternative feedstocks, with a comparison to the lifecycle costs of an ICEV fueled by reformulated gasoline. See Table 9 for additional details.

Unlike the costs presented in Table 9, a carbon tax has sufficient to equate the lifecycle costs for operation with methanol or hydrogen derived from natural gas and biomass has been included. The tax needed is $85/tC for methanol and $95/tC for hydrogen. THis tax would increase the lifecycle costs for the natural gas FCV cases by less than 1.5%. The effect of the carbon tax for the gasoline ICEV is to raise the lifecycle cost about 3%.

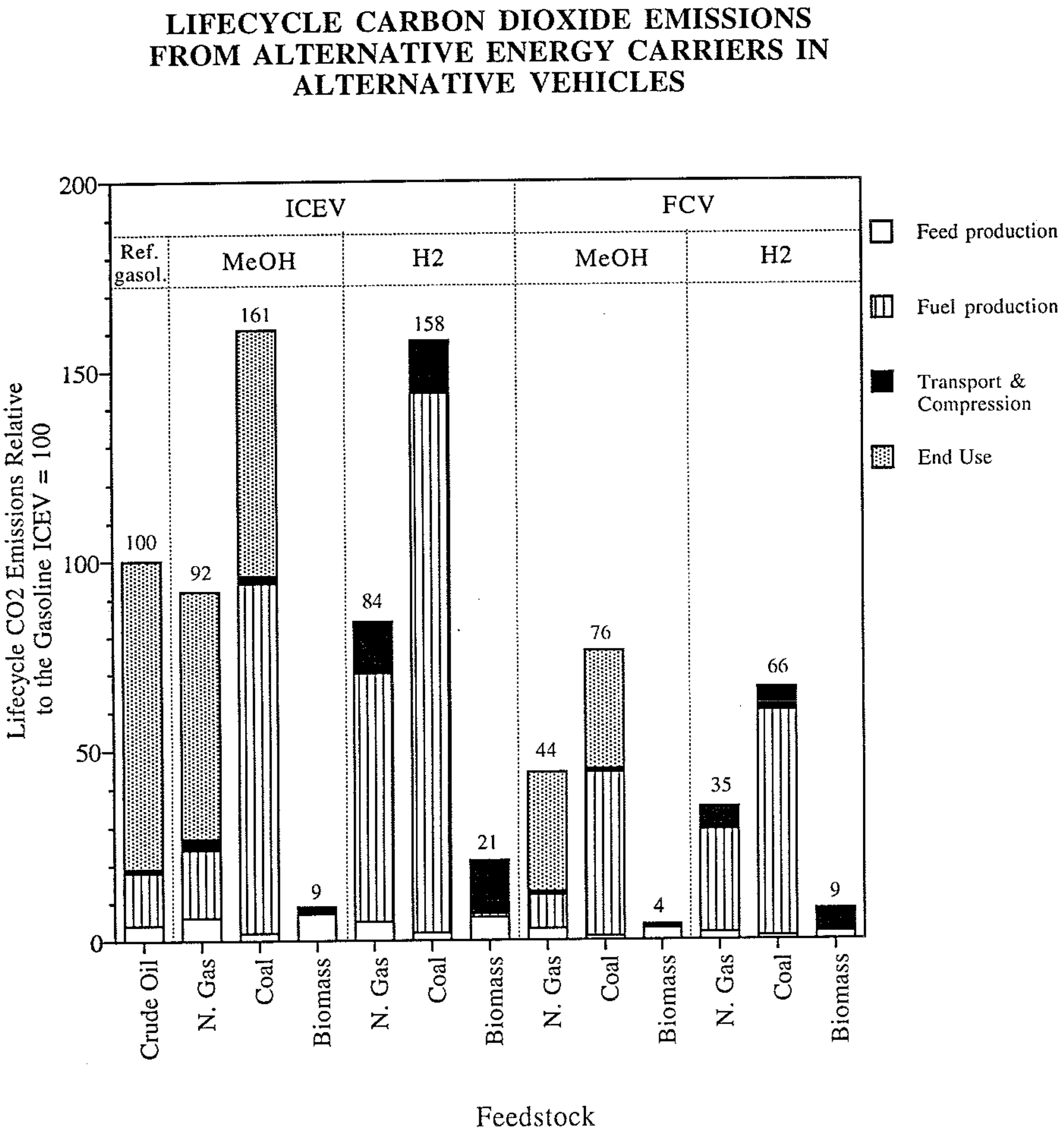
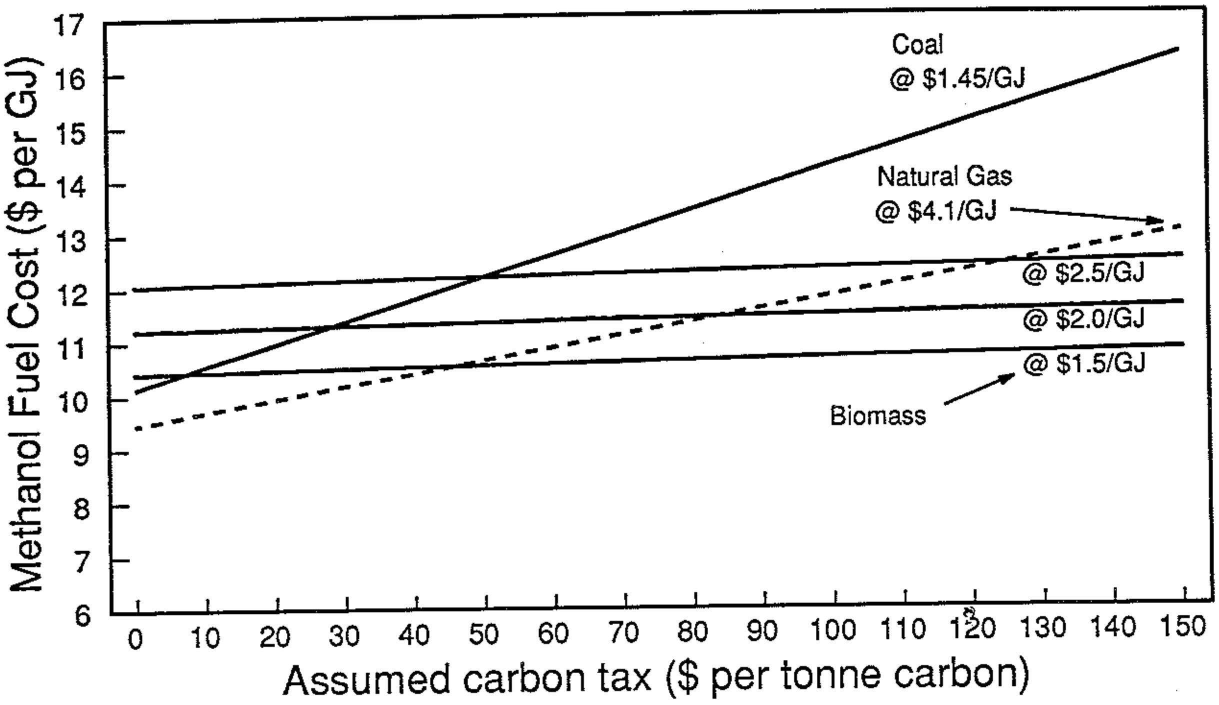


Fig. 6. Estimated lifecycle carbon dioxide emissions from methanol and hydrogen ICEVs and FCVs, with comparisons to emissions from a gasoline ICEV. See Tables 10 and 11 for details.



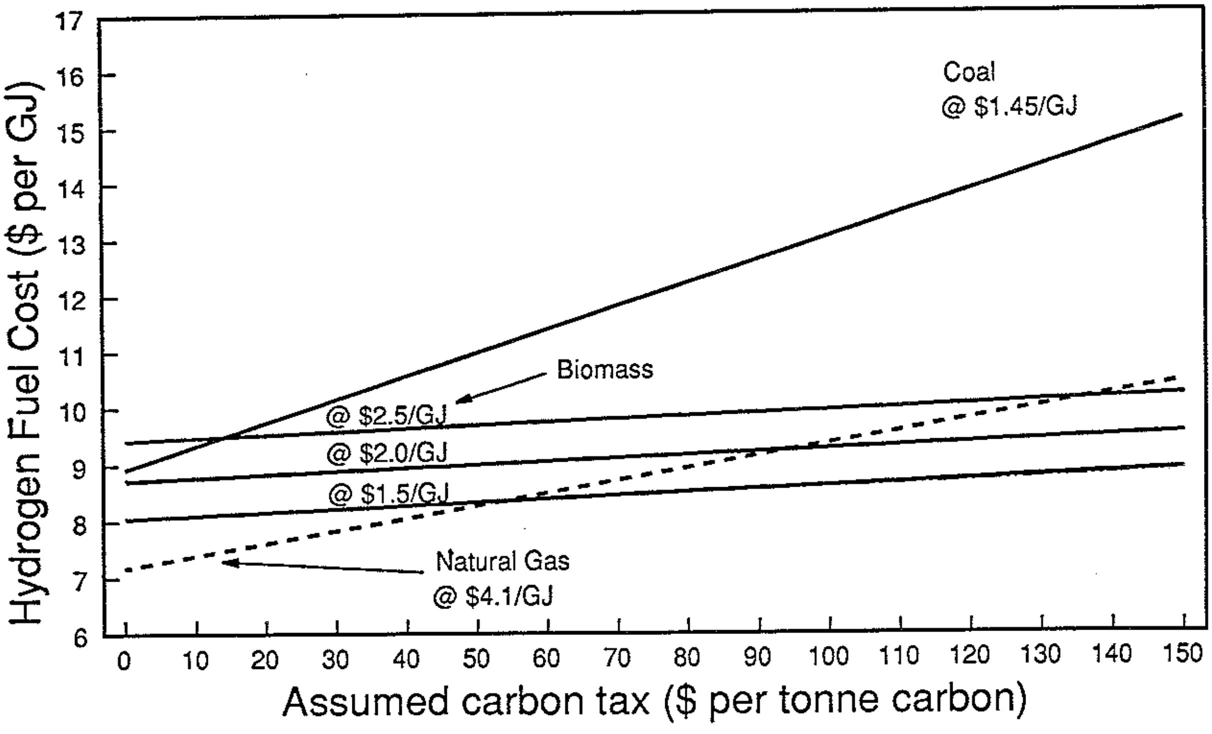


Fig. 7. The effect of a carbon tax on the cost of transport fuel production for alternative feedstocks. The costs on the ordinate axis (zero carbon tax) are fuel production costs from Tables 5 and 6 (BCL case for biomass) for the reference feedstock prices for natural gas, coal, and biomass, as well as alternative biomass prices of $1.5/GJ and $2.5/GJ. The prices of the alternative fuels with the carbon tax are calculated by adding to these prices carbon charges for the lifecycle $CO_2$ emission parameters indicated in Tables 10 and 11.

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