Solar thermochemical production of ammonia from water, air and sunlight: thermodynamic and economic analyses

Ronald Michalsky, Bryon J. Parman, Vincent Amanor-Boadu, Peter H. Pfromm

How to cite this manuscript

If you make reference to this version of the manuscript, use the following information:


Published Version Information


Copyright: © 2012 Elsevier Ltd.

Digital Object Identifier (DOI): doi:10.1016/j.energy.2012.03.062

Publisher’s Link: http://www.sciencedirect.com/science/article/pii/S0360544212002654

This item was retrieved from the K-State Research Exchange (K-REx), the institutional repository of Kansas State University. K-REx is available at http://krex.ksu.edu
Solar thermochemical production of ammonia from water, air and sunlight: thermodynamic and economic analyses

Ronald Michalsky¹,², Bryon J. Parman³,⁴, Vincent Amanor-Boadu⁴, Peter H. Pfromm²

¹ NSF IGERT associate in biorefining, ² Department of Chemical Engineering, ³ NSF IGERT trainee in biorefining, ⁴ Department of Agricultural Economics, Kansas State University, Manhattan, Kansas, 66506, U.S.A.

Abstract

Ammonia is an important input into agriculture and is used widely as base chemical for the chemical industry. It has recently been proposed as a sustainable transportation fuel and convenient one-way hydrogen carrier. Employing typical meteorological data for Palmdale, CA, solar energy is considered here as an inexpensive and renewable energy alternative in the synthesis of NH₃ at ambient pressure and without natural gas. Thermodynamic process analysis shows that a molybdenum-based solar thermochemical NH₃ production cycle, conducted at or below 1500 K, combined with solar thermochemical H₂ production from water may operate at a net-efficiency ranging from 23-30% (lower heating value of NH₃ relative to the total energy input). Net present value optimization indicates ecologically and economically sustainable NH₃ synthesis at above about 160 tons NH₃ per day, dependent primarily on heliostat costs (varied between 90 and 164 dollars/m²), NH₃ yields (ranging from 13.9 mol% to stoichiometric conversion of fixed and reduced nitrogen to NH₃), and the NH₃ sales price. Economically feasible production at an optimum plant capacity near 900 tons NH₃ per day is shown at relative conservative technical assumptions and at a reasonable NH₃ sales price of about 534 ± 28 dollars per ton NH₃.

Corresponding Author: e-mail: pfromm@ksu.edu; phone: +1 785 532 4312; fax: 785-532-7372, postal address: 1036 Durland Hall, Kansas State University, Manhattan, KS 66506

Keywords: Hydrogen; Methane; Fertilizer; Solar; Electricity; Economics; Infrastructure.
1. Introduction

The U.S. Census Bureau estimates that between 1900 and 2000, the world’s population grew from 1.6 billion to 6.0 billion, and is projected to reach 9.0 billion by 2050 [1]. Global human population growth is projected to increase the strain on current natural resources, such as land, fossil hydrocarbons, and fresh water, if technological advances are not made in the production of products and services using these resources.

While technological advances in agriculture in the 20th century - chemical fertilizers, mechanization, breeding, genetic improvement, chemical pest control, processing and storage systems - have contributed to vastly increase the productivity of land globally, the interface between energy and fertilizer production, especially ammonia, promises to challenge the agricultural productivity in the future. At present, over 100 million metric tons of NH₃ [2] are produced annually, driven by increasing food demand and the need for higher crop yields [3]. NH₃ is the single-most important synthetic fertilizer, accounting for 58 wt% of all fertilizer consumed for example in the USA in 2007 [4]. Its role in the production of bio-energy feedstock and its potential use in solar-derived H₂ storage [5-7] or as a liquid fuel [8, 9] augment its criticality and importance in the global economy.

NH₃ easily reaches the U.S. Department of Energy 2015 hydrogen storage target for H₂-based transportation fuels [8, 10] or it can be blended into diesel for direct combustion in modified diesel engines releasing mainly H₂O and N₂ as combustion-products [11]. If these competitive uses and the duty of the agricultural industry to feed a growing global population at reasonable prices are to be realized new and innovative NH₃ synthesis technology will likely be required.

Industrially, the Haber-Bosch process synthesizes NH₃ by shifting the reaction equilibrium of a N₂/H₂ gas mixture at high pressure (about 30 MPa) towards formation of ideally 22.7 mol% NH₃ (relative to stoichiometric conversion) at 673-873 K and in presence of a catalyst [12]. The energy-intensive process [13], including natural gas/steam reforming for H₂ production (accounting for approximately 84% of the total energy required), consumes 28-40 GJ/t NH₃ in form of natural gas [12, 14] (about 1-2% of the world’s annual energy production [15]). Approximately 2.3 t of fossil-derived CO₂ are generated per t NH₃ synthesized [14]. Employing steam-reforming of coal increases the energy required for NH₃ production even further (about 47.6-165.9 GJ/t NH₃) and increases the associated generation of fossil CO₂ (16.7 t
Economies of scale have dictated current Haber-Bosch facilities producing above about 1,500 t NH₃ per day, consuming significant quantities of natural gas and influencing that commodity’s price trend. This in turn has a direct impact on NH₃ prices and their volatility.

Various alternatives proposed for nitrogen fixation from the atmosphere via synthesis of NH₃, including catalytic formation of NH₃ near ambient temperature and pressure in the liquid phase [16] and electrochemical NH₃ synthesis [17] have not yet reached maturity. Solar thermochemical NH₃ synthesis at ambient pressure is a proposed remedy to some of the difficulties associated with the Haber-Bosch process [5, 18-21]. Reactive NH₃ synthesis via a two-step solar thermochemical cycle of metal oxide nitridation and metal nitride hydrolysis [19, 22] has been demonstrated to form significant quantities of NH₃ from air and water at near 0.1 MPa [19]. The process neither requires a catalyst nor a fossil hydrogen source. The energy required for the generation of H₂ via H₂O splitting and for the reductive cleavage of N₂ is supplied in form of solar energy [18, 20]. Concentrated solar radiation, absorbed at elevated temperature in an endothermic metal oxide reduction, creates a metal nitride in the presence of N₂. The fixed nitrogen is, thereafter, released from the solid metal nitride as NH₃ in an exothermic steam hydrolysis reaction. Given the abundance of solar radiation in many areas of the world, this approach has the potential to produce NH₃ sustainably and facilitates simultaneously storage of intermittently available solar energy [23, 24].

Solid carbon (biomass or charcoal) has been suggested as reducing agent of the metal oxide in the process discussed above [18, 20]. However, carbon may not be available in the right quantities and proximity to the manufacturing plant, requiring transportation or production, using up arable land and requiring expensive and energy-intensive processing [25, 26]. On the other hand, reactants forming metal oxides which can be reduced with H₂ unfortunately tend to not fix 0.1 MPa N₂ in form of metal nitrides and show low NH₃ yields when reacting their nitrides with steam [21]. Molybdenum considered here represents a trade-off [21]: The oxide, MoO₂, that is formed during nitride (Mo₂N) hydrolysis at above 800 K can be reduced [27] and nitridated with moderate yields in H₂/N₂ gas mixtures in the range of 800 to 1500 K [28, 29]. Given the relative high ionicity of the nitride [30, 31], significant quantities of NH₃ are liberated during the hydrolysis of Mo₂N at atmospheric pressure.

The work presented here conceptually assesses the technical and economic attractiveness of Mo-based solar thermochemical NH₃ synthesis in the absence of any carbonaceous material or
natural gas as feedstock or for energy (Fig.1). Experimentation towards the technical feasibility of the reaction cycle is described elsewhere [20].

A thermodynamic analysis for synthesizing NH$_3$ in a two-step solar thermochemical reaction cycle from H$_2$ and N$_2$ with a Mo-based reactant at 0.1 MPa (Fig.1) is presented in section 2.1. Section 2.2 determines plant capacity and energy efficiency and CO$_2$ emissions of the Haber-Bosch process implemented with natural gas as a benchmark for a process analysis of the proposed reaction cycle (Fig.2) in section 2.3. The analysis simulates an implementation of the reaction cycle with H$_2$ generated via a well-studied solar thermochemical H$_2$O splitting cycle using zinc [24, 32, 33]. In practice, the proposed NH$_3$ synthesis may be implemented with other solar-to-hydrogen technologies [33]. Section 3.1 estimates investment costs for unit operations and chemical commodities used in section 3.2 that develops an economical optimization model for scalable solar thermochemical NH$_3$ synthesis. The model is evaluated and discussed employing net present value calculations in section 3.3. Section 3.4 shows that the proposed concept is economically attractive under fairly conservative assumptions.

2. Thermochemical NH$_3$ synthesis cycle

The Gibbs free energy of formation for a metal nitride is relatively small compared to the corresponding oxide. Thus, only a few metals such as Mo allow simultaneously for oxide reduction with H$_2$ and reductive cleavage of dinitrogen at 0.1 MPa [21, 28, 29]. Mo is shown here to be a promising reactant for solar thermochemical NH$_3$ synthesis.

To estimate the equilibrium reaction yield achievable in a system closed to mass transfer, the free energy of reaction, $\Delta_{\text{rxn}}^\circ$, was computed based on the literature [27]. The absolute error of energy of formation data was estimated previously with $\pm 3$ kJ [34] and was taken as 2% of the value in kJ/mol. The computed formation of Mo$_2$N was extrapolated at $>$ 800 K using a linear fit ($R^2 > 0.999$). With the free energy computations in hand, the equilibrium constants, $K_{eq}$, were determined at atmospheric pressure taking the total number of chemical species in the system for simplicity as the arithmetic mean of the number of reactants and the number of products at complete conversion [35]. This allows solving the elemental mol balances of the given reaction system symbolically (“live” Symbolics, Mathcad 13) as a function of $K_{eq}$, that is yielding the equilibrium composition of the reaction system as a function of temperature, $T$, at 0.1 MPa.
2.1 Thermodynamic analysis

Conversion of solar energy is accomplished by thermochemical reduction of Mo(IV) oxide with H₂ to Mo metal (Eq.1):

\[
2\text{MoO}_2^{(s)} + 4\text{H}_2 \leftrightarrow 2\text{Mo}^{(s)} + 4\text{H}_2\text{O} \\
\Delta_{\text{rxn}}h^{1500K} \approx 130 \pm 25 \text{ kJ/mol N}
\]

\(\Delta_{\text{rxn}}g\) of Eq.1 indicates that the reaction equilibrium is favored thermodynamically at above ca. 1428 K (Fig.3A). The chemical energy stored in the endothermic metal formation (the enthalpy of reaction, \(\Delta_{\text{rxn}}h\) [27]) allows subsequently for slightly exothermic N₂ fixation in form of metal nitridation (Eq.2) favorably at lower temperatures (Fig.3B):

\[
2\text{Mo}^{(s)} + \frac{1}{2}\text{N}_2 \leftrightarrow \text{Mo}_2\text{N}^{(s)} \\
\Delta_{\text{rxn}}h^{800K} \approx -63.2 \pm 0.8 \text{ kJ/mol N}
\]

The yield \(Y = \frac{\text{mol solid reaction product formed}}{\text{mol solid product at stoichiometric conversion}}\) for reaction 1 or 2 is below stoichiometric conversion at thermodynamic equilibrium (Eq.1 at ca. 1428 K, Eq. 2 at ca. 1115 K, assuming 0.1 MPa) (Fig.3). However, given the non-equilibrium situation (mass exchange) in an actual flow-through reactor, and assuming a high effective reactant surface, stoichiometric conversion for both reactions is assumed below. The over-stoichiometric supply of H₂ (section 2.3), may account for excess MoO₂ formed during nitride hydrolysis from Mo that may not have converted to Mo₂N during the nitridation.

The assumptions below (see 2.3) are based on experimental results (manuscript in preparation) for hydrolysis of 25.5 ± 0.5 wt% Mo₂N powder (balance Mo, 17 ± 8 μm average particle diameter, 429 ± 3 m²/kg BET surface area) at 773 K with steam supplied for 1 h at a rate of 0.91
± 0.02 ml(STP)/min. The simulation assumes either \( Y_3 = 100 \text{ mol\%}, \ Y_4 = 0 \) (stoichiometric conversion), or \( Y_3 = 13.9 \text{ mol\%}, \ Y_4 = 9.1 \text{ mol\%}. \) Decreasing the hydrolysis temperature (Fig.4) or increasing the availability of effective reactive surface sites may allow increasing the yield of \( \text{NH}_3 \) in practice. Endothermic decomposition of \( \text{Mo}_2\text{N} \) (Eq.2, when favoring the species on the left side) was neglected at this point due to a positive \( \Delta_{\text{rxn}} \) at 800 K. Oxidation of \( \text{MoO}_2 \) to \( \text{MoO}_3 \) [28] is not favored at equilibrium (105.2 kJ/mol oxide \( \Delta_{\text{rxn}} \) at 800 K [27]) but driven when the equilibrium is not established due to \( \text{MoO}_3 \) vapor formation (\( \text{MoO}_3 \) boils at about 1428 K). This can be reduced or avoided by low hydrolysis temperatures, low steam flow rates and short reaction times.

The overall reaction represents an alternative for realizing the Haber-Bosch reaction (\( \frac{1}{2}\text{N}_2 + \frac{3}{2}\text{H}_2 \leftrightarrow \text{NH}_3 \)) near 0.1 MPa. The enthalpy required for breaking the \( \text{N}_2 \) triple bond is supplied indirectly in form of concentrated solar radiation providing the heat (at 1500 K) for the endothermic reduction of \( \text{Mo(IV)} \) oxide to \( \text{Mo metal} \) (Eq.1). The metal is utilized to cleave and electrochemically reduce dinitrogen (Eq.2) increasing the metallic oxidation state formally to \( \text{Mo}^{3+} \) in \( \text{Mo}_2\text{N} \) [30, 31]. \( \text{Mo}_2\text{N} \) is further oxidized to \( \text{Mo(IV)} \) when reacted with \( \text{H}_2\text{O} \) to liberate \( \text{NH}_3 \) (Eq.3). The heat released from Eq. 2-4 is partly integrated [36] (see 2.3). The significant amount of energy required to form \( \text{H}_2 \) from \( \text{H}_2\text{O} \) (Fig.2) is supplied as solar radiation at 2000 K employing a two-step solar thermochemical cycle of endothermic \( \text{ZnO} \) dissociation (about 679.2 kJ per 3/2 mol \( \text{H}_2 \) \( \Delta_{\text{rxn}} \) at 2000 K), quenching of the \( \text{Zn}/\text{O}_2 \) vapor leaving the reactor, and exothermic oxidation of the condensed \( \text{Zn} \) with \( \text{H}_2\text{O} \) at 400 K (about -156.6 kJ per 3/2 mol \( \text{H}_2 \) \( \Delta_{\text{rxn}} \) at 400 K) recycling \( \text{ZnO} \) and producing \( \text{H}_2 \) (both computed at 0.1 MPa and assumed with stoichiometric conversion). This well-studied cycle has been discussed elsewhere [24, 32].

2.2 The scale of industrial \( \text{NH}_3 \) synthesis

As a benchmark, the Aspen Plus (V7.0) Ammonia Model [37] was used to simulate the industrial \( \text{NH}_3 \) production using natural gas as a feedstock. The model comprises a reforming unit converting a desulfurized hydrocarbon feed with steam (primary reformer, 3.1-3.3 MPa, 775-1064 K) and air (secondary reformer, 2.9-3.1 MPa, 1251-1530 K) into \( \text{H}_2 \) and carbon oxides. Subsequently, \( \text{CO} \) is converted catalytically (2.7-2.9 MPa, 483-721 K) to \( \text{CO}_2 \) that is removed with \( \text{NH}_3 \) forming an ammonium hydrogen carbonate byproduct. The synthesis gas obtained is freed from traces of \( \text{CO} \) and \( \text{CO}_2 \) employing a nickel catalyst to form \( \text{CH}_4 \)
(methanizer). Thereafter, NH\textsubscript{3} is synthesized at 28.4-29.2 MPa and 686-799 K over a promoted iron catalyst. The major fraction of the 23.9 mol\% NH\textsubscript{3} in the synthesis loop (about 33.9 mol\% nitrogen-to-NH\textsubscript{3} conversion) is liquefied via refrigeration (27.5 MPa, 288-304 K) and stored at 3 MPa. The model estimates the thermodynamic properties of gases at high temperature and pressure using a modified Redlich-Kwong equation of state (RKS-BM). Liquid and vapor properties in the CO\textsubscript{2} scrubbing unit are modeled with an electrolyte NRTL or a Redlich-Kwong equation of state model respectively. A detailed description of the Haber-Bosch modeling is provided elsewhere [37].

Overall, the process converts approximately 35.9 t/h natural gas (80.0 mol\% CH\textsubscript{4}, 17.7 mol\% C\textsubscript{2}H\textsubscript{6}, balance hydrocarbons and air, at 3.8 MPa and 303 K) with 308.5 t/h air at 302 K and 69.0 t/h water at 293 K (both at 0.1 MPa) to 27.5 t/h liquid anhydrous ammonia (99.6 mol\% NH\textsubscript{3}, 2 MPa, 306 K), 126.9 t/h ammonium bicarbonate salt (98.7 mol\% NH\textsubscript{4}HCO\textsubscript{3}, 0.1 MPa, 293 K), and 259.0 t/h flue gas (7.0 mol\% CO\textsubscript{2}, 16.1 mol\% H\textsubscript{2}O, balance N\textsubscript{2}, O\textsubscript{2}, and Ar, 0.1 MPa, 333 K). To compare this to the discontinuous operation of the solar thermochemical NH\textsubscript{3} synthesis in section 2.3, this equates production of about 1,324 t NH\textsubscript{3} per day (as anhydrous ammonia or ammonium salt, assuming 24 h/d operation).

The enthalpy balance of the process indicates a net heat duty of about 289 MW, mainly due to the heat required in the reforming unit and the CO\textsubscript{2} stripper and the electricity consumed for synthesis gas compression. Taking the lower heating value (LHV) of natural gas at 31.89 GJ/t [38] results in further consumption of about 32.6 t/h natural gas and 549.7 t/h air generating 582.3 t/h flue gas (9.9 mol\% CO\textsubscript{2}, 18.3 mol\% H\textsubscript{2}O, balance N\textsubscript{2} and Ar). This yields total CO\textsubscript{2} emissions of the process - flue gas - at about 2.17 t CO\textsubscript{2} per t NH\textsubscript{3}. The energy efficiency can be estimated with 46.9\% (the LHV of 1,324 t/d NH\textsubscript{3} relative to the LHV of 1,644 t/d natural gas).

2.3 Numerical process analysis

Given its conceptual state a conceivable solar thermochemical process that is converting air, desalinated water, and sunlight into liquid ammonia and compressed oxygen (Fig.5) was analyzed similar to other thermochemical processes reported in the literature [23, 36, 39]. To estimate the plant layout mass and energy balances were solved iteratively (Generalized Reduced Gradient nonlinear optimization code, 10\textsuperscript{2} iteration steps, 10\textsuperscript{-4} minimum sensitivity, Excel 2003) at steady-state and as a function of a variable NH\textsubscript{3} capacity. Two scenarios were computed: First,
assuming \( Y_3 = 100 \text{ mol}\% , \ Y_4 = 0 \) (see 2.1), and the ratio of gaseous reactant required at minimum to the amount of gaseous reactant supplied to any reaction, \( r_{\text{gas}} \), of 90 mol\%, “ideal operation”. The second, rather “conservative operation”, assumes \( Y_3 = 13.9 \text{ mol}\% , \ Y_4 = 9.1 \text{ mol}\% , \) and \( r_{\text{gas}} = 67 \text{ mol}\% . \)

To analyze the performance of the envisioned plant located in a suitable geographic region typical meteorological data from the updated National Solar Radiation Data Base (NSRDB) was used [40, 41]. The hourly direct normal irradiance values provided by the database were averaged over a typical meteorological year to identify several regions in the southwestern U.S. with an averaged normal irradiance in the range of 7.01-7.81 kWh/m\(^2\)/d. The analysis presented here assumes a yearly-averaged direct normal irradiance of 7.48 kWh/m\(^2\)/d (that is 1 kW/m\(^2\) for annualized 7.48 h/d), based on data for 1997-2005, Palmdale Airport, CA, USA (approximately 80 km linear distance to the Pacific).

Solar radiation is concentrated via heliostats and absorbed at 1500 or 2000 K respectively (Fig.5) with an efficiency of solar radiation converted to chemically stored energy (dimensionless mean flux concentration ratio at \( 10^4 \), for a detailed description see [24]) assumed at 0.78 (Fig.6). The radiation is received by a series of reactors at 100 m above ground (similar in appearance to the Solar Two power tower, Mojave Desert, CA, USA; or the PS10 and PS20 plants near Seville, Spain) splitting H\(_2\)O into H\(_2\) and O\(_2\) (2.5 x 2.5 x 5.0 m reactor volume, \( V_R \), Mo-alloy), or converting H\(_2\) and N\(_2\) to NH\(_3\) (15 x 15 x 25 m \( V_R \), ceramic lining). \( V_R \) was estimated assuming 1 min residence time of the gaseous species at a given temperature in the reactor (using a molar ratio of N\(_2\) sweep gas to ZnO of 0.1 [32]). In practice \( V_R \) will be determined empirically by reaction kinetics for a reactant with optimized composition and by the heliostat area required per reactor for providing heat at a desired temperature.

Sensible, \( \Delta_{\text{sens}}h \), and latent, \( \Delta_{\text{lat}}h \), heat and the \( \Delta_{\text{rxn}}h \) of exothermic reactions [27] are recovered and integrated at a ratio of 0.6 (that is heat losses at about 40\%) [36] estimating heat exchange areas of 1.44 x 10\(^4\) m\(^2\) for temperatures up to 2000 K (lined with molybdenum disilicide) or 8.47 x 10\(^4\) m\(^2\) < 1500 K (Fig.6). This is assuming replication of the production described in section 2.2. The remaining heat is removed via dry (air) cooling (5.71 x 10\(^5\) m\(^2\) heat exchange area, 40 K effective \( \Delta T_{\text{air}} \), 2 kPa pressure losses [42], efficiency of fans and compressors assumed with 0.86 [43]). Although shown to be by a factor of 4-6 more energy-
intensive and by a factor of 4-12 more capital-intensive than wet cooling, dry cooling was employed due to the crucial role of water availability [44].

The N₂ is required only at industrial-grade purity (as sweep gas and feedstock) and is produced via membrane permeation (based on literature values for gas permeability and diffusivity for polysulfone fibers [45]) yielding at 0.2 MPa trans-membrane pressure an effective membrane area of ca. 8.48 x 10⁵ m². By-product O₂ (Table 1) is compressed isentropically and stored. NH₃ is separated from its synthesis vapor and liquefied via cooling and compression to 306 K and 2 MPa (see 2.2) and stored in steel tanks. H₂ and N₂ gas mixtures recovered from the NH₃ synthesis cycle are enriched with H₂ generated in the H₂ synthesis cycle and desiccated with a silica gel bed (0.4 g H₂O/g SiO₂ adsorption capacity [46], 200 K maximum ΔTgas between H₂O ad- and desorption, 10 kPa pressure drop assumed for all solid beds). Traces of Ar, CO₂, and Zn or MoO₃ vapor have been neglected for simplicity.

Given the net energy content of NH₃ and neglecting the energy stored in the separation of O₂ from air, conversion of solar energy to NH₃ was estimated with an efficiency of about 23-30% at maximum (LHV of NH₃ relative to the total energy requirement, including net electricity as solar heat-equivalent, for conservative or ideal operation, respectively) (Table 1). This energy is released in form of heat when NH₃ is combusted (e.g., as transportation fuel) [11] or in form of H₂ when NH₃ is used as a single-use hydrogen carrier [6, 7, 9]. The maximum efficiency as estimated is below 46.9% estimated for the NH₃ synthesis with natural gas at this scale (section 2.2) but within the range estimated for the industrial NH₃ synthesis via steam reforming with natural gas or coal respectively, i.e., 11-66% (Table 1).

The total electricity required for the proposed process (assuming as a worst-case scenario all grid-electricity is generated from coal-fired power plants emitting 0.91 t CO₂/MWh [47]) results in CO₂ emissions in the range of 0.62-1.08 t CO₂ / t NH₃. That is a net reduction of fossil CO₂ emissions by 50-71% relative to the current NH₃ synthesis with natural gas (see 2.2) or up to 96% when avoiding the use of coal feedstock in favor of biomass (see 1.).

3. Economic feasibility

The production of NH₃ presented in section 2 comprises two phases: (i) H₂ generation via a two-step solar thermochemical H₂O splitting cycle, and (ii) solar thermochemical NH₃ synthesis from desalinated water, air and H₂. It is envisioned that the required solar energy is
harvested with an array of heliostats concentrating solar radiation that is received by a reactor placed atop a central collector tower, i.e., “H₂ towers” if generating H₂, or “NH₃ towers” if generating NH₃ (Fig.2 and 5).

If the production described in section 2.2 is to be replicated by this process at “ideal operation”, it will require 33.2 t Mo, 17.0 t Zn, 0.48 km² lens area for concentration of sunlight, \( A_{lens} \), utilized by the NH₃ synthesis cycle, and 2.95 km² \( A_{lens} \) utilized by the H₂ synthesis cycle (Fig.5). On the other hand, “conservative operation” would lead to 239.1 t Mo, 17.0 t Zn, 1.11 km² \( A_{lens} \) to synthesize NH₃, and 3.10 km² \( A_{lens} \) to produce H₂. The increased amount of Mo does not significantly affect capital costs (see 3.1). However, the increased land requirements (increased totally by 22.7% for provision of sensible and latent heat and \( \Delta_{rxnH} \) of Eq.1) under the “conservative operation” lead to a significantly increased amount of capital that needs to be raised for reactors and solar concentrators. A process summary is given in Table 1 and Figure 6.

3.1 Data sources

The mass and energy balances were utilized to generate a generic list of components and equipment required for realizing the proposed process (see 2.3). Data estimates as realistic as possible were obtained from equipment manufacturers, service providers and operating facilities. A summary for fixed costs dependent on or independent of the NH₃ capacity or operational costs is given in Table 2, 3 or 4 respectively.

The operating costs including heliostat maintenance and service are assumed at 3% of total heliostat costs. Labor is assumed based upon plant size and skill level. Air compressors and water pump costs are based upon gas/liquid mass flow and stream conditions. Variable costs are assumed to grow at an annual inflation rate of 3%. Corporate income tax is set and maintained at its current level of 35%. NH₃ price is forecast using 10³ Monte Carlo simulations based on a 20-year historical mean and standard deviation of U.S. prices (Fig.7). The projected NH₃ price series drawn from the Monte Carlo model has a mean, median, or standard deviation of 522, 531 or 27 dollars/t NH₃ respectively. It is assumed that variable costs per unit will be linear in production, with scale effects influencing the linearity of per unit cost over different production levels.
3.2 Economics of Solar Thermochemical NH$_3$ Production

The economics do not assume any kind of regulatory or public support for the technology proposed here. No credit is taken or subsidies are assumed for reduction of CO$_2$ emissions, use of renewable vs. non-renewable resources in making the product, etc.

The NH$_3$ plant is assumed to be a profit-maximizing/cost minimization business. Thus, it will select its fixed and variable costs to produce at an output level that maximizes the return to its investment. This assumption implies that a principal constraint in building such a plant is available capital, treating costs of capital as part of operating costs. Economic theory suggests that the plant would be built if it is economically feasible. Economic feasibility is defined here to mean the ability of the plant to return a positive net present value at a specified discount rate over a reasonable lifespan of the project. “Reasonable lifespan” is defined to imply a period prior to the need for major capital retrofitting of the plant, assumed in this case to be 20 years.

An optimization approach is adopted to evaluate the economic feasibility of building a solar thermochemical NH$_3$ production plant using the foregoing technology. This approach is particularly helpful because of the physical plant constraints in the production process and the direct effects of these constraints on the plant’s profitability. It is assumed that the output of NH$_3$ in each scenario, $Z_{lm}$, is defined as:

$$Z_{lm} = \delta_{lm} X_N, \quad l=1,2; m=1,2$$

where $\delta_{lm}$ is a yield constant in t/h per tower based on the technical efficiency of the towers, and $X_N$ is the quantity of NH$_3$ towers (see 3.). The objective function of the optimization model is to maximize operational profits, $\pi$, by selecting the optimum number of $X_N$ given the expected market price, $p$, of NH$_3$ and in cognizance of plant operational costs. The model is evaluated under two operational, $l$, and cost, $m$, conditions, yielding four scenarios:

- Scenario 1: ideal operation, conservative costs ($l, m = 1$)
- Scenario 2: ideal operation, optimistic costs ($l = 1, m = 2$)
- Scenario 3: conservative operation, conservative costs ($l = 2, m = 1$)
- Scenario 4: conservative operation, optimistic costs ($l, m = 2$)

Since the only choice variable is $X_N$, the optimization model may be presented as:
\[
\max_{X_N} \pi_{lm} = pZ_{lm}(X_N) - C(Z_{lm}) - F_{Z_{lm}}
\]

\[
s.t. \ \rho_l X_N \leq X_H
\]

\[
K_{lm, j_m} - F_{Z_{lm}} \geq b_{N_{lm}} X_N + b_{H_{lm}} X_H
\]

\[
X_N, X_H \in [0, \infty)
\]

\[
X_N, X_H \equiv I
\]

where \(C(Z_{lm})\) is the variable costs for producing \(Z\) output and \(F_Z\) is the fixed costs under each scenario. The number of \(H_2\) towers is a fixed proportion, \(\rho\), of the number of \(NH_3\) towers. This constraint is set as an inequality because it is technically feasible to produce more \(H_2\) than required. The second constraint in Eq.6 stipulates that available capital, \(K_{lm}\), less the associated fixed costs, \(F_Z\), must be higher than the acquisition costs of the two types of central tower systems. Available capital is evaluated \(j\) times with particular characteristics to assess the effect of capital on the optimum output and profits. Since the optimization model is essentially choosing output using the number of \(NH_3\) and \(H_2\) towers, it is plausible to present all plant fixed costs, i.e., the coefficients \(b_{N_{lm}}\) and \(b_{H_{lm}}\), in terms of towers. The final two constraints indicate that the number of towers is always a positive non-zero integer to ensure that production does occur. Table 5 provides a summary of the assumptions underlying the computed scenarios.

With regard to assumed costs, cost and quantity estimates for Scenario 1 - the baseline scenario - are provided in Table 2. Note that the product and sum of each tower specific column and the cost column gives the value of \(b_{N11}\) and \(b_{H11}\). In Scenario 2 the heliostat price is decreased from 164 to 90 dollars/m² [33] and the price of heat exchangers for temperatures up to 2000 K is reduced by two thirds from 467 to 155 dollars/m². This results in an overall cost per tower of 13.9 million dollars per \(NH_3\) tower or 53.3 million dollars per \(H_2\) tower. That is cost reductions relative to Scenario 1 of 3.76 or 28.1 million dollars per \(NH_3\) or \(H_2\) tower respectively. It should be noted that the cost of the \(NH_3\) tower accounts for major components of the overall process (Table 2 and 4). This shows that the impact of the heliostat investment costs for the \(H_2\) production cycle is quite significant. Correlating linearly with the area estimate for harvesting solar radiation (see 2.3 and Fig.6), about 74-86% (conservative to ideal operation) of the total heliostat investment costs is absorbed in the construction of the \(H_2\) production cycle.

With regard to assumed process operation, Scenario 3 uses the cost estimates as shown in Table 2. However, the decrease in \(Y_3\) and the increase in \(Y_4\), results in an increased amount of heat and \(H_2\) required for recovering an increased amount of Mo from its oxide. This together
with the decrease in $r_{gas}$ results in an increased amount of gas that has to be processed increasing the number of required NH$_3$ or H$_2$ towers respectively. This, altogether, reduces the overall costs per NH$_3$ tower from 17.6 (Scenario 1) to 16.0 million dollars (Scenario 3) and increases the overall costs per H$_2$ tower from 81.3 to 92.9 million dollars. Scenario 4 uses the same quantities per tower as that of Scenario 3 with the cost estimates of Scenario 2 resulting in investment costs of 13.4 million dollars per NH$_3$ tower and 59.4 million dollars per H$_2$ tower.

3.3 Model Evaluation and Profitability

The economic feasibility of the plant is determined by its ability to generate a positive net present value (NPV). Given that the only choice variable in the optimization model is NH$_3$ output, the feasibility analysis presented here is conducted under optimal conditions by multiplying the prevailing price of NH$_3$ (see 3.1, Fig.7) by the optimum quantity to get total revenue and adjusting that by operating costs in each period.

The plant is assumed to operate 365 days per annum (in agreement with the annualized solar insulation data employed, see 2.3), running a single labor shift. Assuming an inflation rate of 3%, the cash flows from operations under the four scenarios are estimated. The NPV is then calculated for each scenario using a 10% discount rate. Total operating costs for the year includes labor, service contracts and utilities costs, presented on a per tower basis, as per discussion in section 3.2. The initial price of the O$_2$ byproduct is assumed at its current market price of about 21 dollars/t, and allowed to grow naively at the assumed inflation rate of 3% per annum over the simulated 20 year period. Thus, the total plant revenue is enhanced by the revenues emanating from the sale of O$_2$ in each period. Net revenues are adjusted for corporate taxes to provide the annual cash flow in each period. The economic feasibility based on NPV is presented as:

\[
NPV_{lm} = -Z_{lw} + \sum_{T=1}^{20} (\pi_{lmT}(1 + D)^{-T})
\]

where $D$ is the discount rate and $T$ is the relevant year.

Figure 8 displays the NPV and initial plant costs from Scenario 1 or 4 respectively. There exists a pattern of oscillating profitability in both scenarios shown. Scenario 1 shows a negative NPV at the initial output iteration increasing overall as output increases to an NPV of approximately 116.9 million dollars. The 4$^{th}$ iteration’s NPV is lower than the 3$^{rd}$ iteration’s
since, at this point, an increase in output requires an additional H₂ tower and thus allows for only a slight increase in overall output from the 3rd to the 4th iteration with a 116.7 million dollar increase in overall plant costs. This result also appears, albeit at different iterations, for Scenario 4 stemming from the same additional H₂ tower construction requirements. However, Scenario 4 approaches, yet does not reach, a positive NPV within the domain of the model. Since the model is maximizing profit, points at a lower NPV than the previous iteration would not be chosen and output would revert to the previous point. Thus, under the assumptions of Scenario 4 the model suggests the plant does not exist. Figure 8 shows furthermore the total plant costs vs. output for Scenario 1 or 4 respectively. This illustrates at points where there is excessive unused H₂ output can be increased for a relatively low cost.

Scenario 2 (not shown) yields similar results with Scenario 1, however, the NPV remains at or above 0 within the domain of the model. In contrast, the NPV of Scenario 3 (not shown) is negative at the minimum output simulated and tends further below zero as output increases. This shows the profitability of the plant regardless of the component costs or operational assumptions tends to increase non-linearly in Scenarios 1, 2 and 4.

The high construction cost of the H₂ towers prevents smooth output scaling. It is for this reason that an attempt was made to determine an optimum plant size based upon NPV return per initial capital dollar invested. For this the plant size is limited to model iterations producing approximately 1,324 t NH₃ per day (see 2.2) or less using the ratio of NPV over total initial plant costs. This results in an optimum plant size occurring in all four scenarios at the point where H₂ use is maximized or, from constraint 1 in Eq.6, \( X_H - \rho X_N \) is minimized. In Scenario 1 this point is found to yield an NPV return per initial investment dollar of 15.2%. In Scenario 4 the optimum configuration yields a return of -2.1%. Any deviation from this point within the upper and lower boundaries of the model will reduce the NPV to the initial capital ratio. These results are summarized for all four scenarios in Table 6.

The implications of an optimum plant size as well as the oscillating profit function are potential barriers to entry and deterrents in plant scale changes. The lowest initial required capital to reach an optimum plant size occurs in Scenario 1 at an initial cost of 769.9 million dollars and an output of 902.2 t NH₃ per day. Given this large initial capital requirement for efficient operation operating under the competitive market assumption will in fact generate a fairly large barrier to entry into the market. The oscillation of the profit function implies that
careful consideration must be taken before increasing the NH$_3$ capacity of any plant operating at optimum or at least at a relative maximum in the profit function. Increasing the NH$_3$ output from either of these points by a fraction will most likely increase the per unit production costs and may even reduce overall profits.

3.4 Sensitivity to the NH$_3$ sales price

The sensitivity of the presented model to major operational and economic variables is demonstrated analyzing Scenario 1 to 4 (compare section 2.3 and 3.2) in section 3.3. To assess the effect of the NH$_3$ sales price a sensitivity analysis was conducted determining the required NH$_3$ price to yield a NPV of 0 for all four scenarios within the domain of the model (Fig.9).

At an optimum output of 902.2 t NH$_3$ per day, a zero NPV is achieved in Scenario 1 or 2 by decreasing the Monte Carlo simulated 20 year prices (Fig.7) by 13.5% or 38.8% respectively (Fig.9A). That is, the possibility of yielding a negative NPV with a fall in the NH$_3$ sales price is much greater in Scenario 1 than in Scenario 2. On the other hand, at an optimum output of 911.2 t NH$_3$ per day, Scenario 3 or 4 would require NH$_3$ sales prices increased by 39.5% in Scenario 3 or 2.19% in Scenario 4 respectively (Fig.9B). This indicates that in Scenario 4 slight variations in NH$_3$ prices will result in economic feasibility of the proposed concept. Given in particular the high cost assumptions for acquiring heliostats, Scenario 3 requires a relatively large increase in NH$_3$ price to “break even”.

4. Conclusions and Outlook

The solar thermochemical synthesis of ammonia using a molybdenum-based reactant was presented and analyzed from a technical and an economic perspective. Major conclusions are:

- It appears technically feasible to form NH$_3$ with a reaction cycle conducted at near 0.1 MPa and at $\leq 1500$ K and without natural gas or solid reducing agents. This may allow synthesis of artificial nitrogen fertilizer without sophisticated machinery and less depended on the volatility of the natural gas price. As outlined for the U.S., geographical regions with high annual insolation and a relative close supply of coastal or fresh water appear suited for this technology.

- Maximum energy efficiencies of converting solar radiation to the lower heating value of NH$_3$ were estimated (23-30%) between the efficiency of the industrial
NH\textsubscript{3} synthesis employing coal (about 11%) or natural gas (up to 66%). As an aside, this approaches the DOE performance target for solar thermochemical H\textsubscript{2} (i.e., 30% by 2017 or > 35% by 2020 respectively) [33] and includes convenient storage of H\textsubscript{2} in form of NH\textsubscript{3}. In the future, research addressing yield and kinetics of the NH\textsubscript{3} formation via materials design [21], heat integration [48], and solar-to-hydrogen technology [33] will be critical for approaching efficiencies realized with the Haber-Bosch process.

- Indirect fossil CO\textsubscript{2} emissions (from coal-derived grid-electricity) are in the range of 4-50% of the CO\textsubscript{2} emitted by the current industrial NH\textsubscript{3} synthesis employing a coal or natural gas feedstock. Yet, no special monetary benefits for technologies utilizing renewable resources via regulations for example for CO\textsubscript{2} emissions are disregarded in the analysis presented.

- The cost of heliostats is a major factor determining the economic feasibility of the proposed technology. About 74-86% of the heliostat capital investment is absorbed for H\textsubscript{2} production. Thus, low cost heliostats (i.e., 90 dollars/m\textsuperscript{2}) or reduced H\textsubscript{2} reactor costs (< 16 million dollars, as estimated here), or replacing H\textsubscript{2} with another gaseous reducing agent, would result in WHAT?? (Scenario 4) or augment (Scenario 1 or 2 at > 450 t NH\textsubscript{3} per day) the return of investment.

- The sales price increase of NH\textsubscript{3} required for Scenario 4 to break even (2.19%) is below the standard deviation of the Monte Carlo Price simulation (5.19%). Which means what? The present simulation suggests economic feasibility of the proposed technology even under conservative assumptions at 534 ± 28 dollars per ton NH\textsubscript{3}. If natural gas prices rise break even will be possible at production levels below 900 t NH\textsubscript{3} per day.

- Production at small scale (down to 144-178 t NH\textsubscript{3} per day when employing only a single H\textsubscript{2} tower) would reduce initial capital requirements (e.g., 770 million dollars at 902 t NH\textsubscript{3} per day for Scenario 1) and facilitate market entry. Fertilizer production in regions with relatively undeveloped infrastructure for example in developing countries with significant population growth might then be conceivable.
Acknowledgments

This material is based upon work supported by National Science Foundation Grant # 0903701: “Integrating the Socioeconomic, Technical, and Agricultural Aspects of Renewable and Sustainable Biorefining Program, awarded to Kansas State University.” Funding by the Center for Sustainable Energy, Kansas State University is kindly acknowledged.
References


### Tables

#### Overall process mass balance

<table>
<thead>
<tr>
<th>Raw materials</th>
<th>m (t/d)</th>
<th>T (K)</th>
<th>p (MPa)</th>
</tr>
</thead>
<tbody>
<tr>
<td>air</td>
<td>1,970</td>
<td>300</td>
<td>0.1</td>
</tr>
<tr>
<td>water</td>
<td>2,418</td>
<td>293</td>
<td>0.1</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Products</th>
<th>m (t/d)</th>
<th>T (K)</th>
<th>p (MPa)</th>
</tr>
</thead>
<tbody>
<tr>
<td>NH₃ (99 wt% NH₃ in H₂O)</td>
<td>1,344</td>
<td>306</td>
<td>2</td>
</tr>
<tr>
<td>O₂ (82 wt% O₂ in N₂)</td>
<td>555</td>
<td>300</td>
<td>15</td>
</tr>
<tr>
<td>air (75 wt% O₂, 12 wt% H₂O)</td>
<td>2,489</td>
<td>413</td>
<td>0.1</td>
</tr>
</tbody>
</table>

#### Comparison of total energy requirements (GJ/t NH₃)

<table>
<thead>
<tr>
<th>Process</th>
<th>Lower heating value (LHV)</th>
<th>Gibbs free energy of mixing for O₂ separation</th>
</tr>
</thead>
<tbody>
<tr>
<td>Solar thermochemical NH₃</td>
<td>56.4 - 70.6</td>
<td></td>
</tr>
<tr>
<td>Natural gas/steam reforming &amp; Haber-Bosch</td>
<td>28 - 40.1</td>
<td></td>
</tr>
<tr>
<td>Coal gasification &amp; Haber-Bosch</td>
<td>47.6 - 165.9</td>
<td></td>
</tr>
<tr>
<td>Lower heating value (LHV) of NH₃</td>
<td>18.6</td>
<td></td>
</tr>
<tr>
<td>Gibbs free energy of mixing for O₂ separation</td>
<td>0.2</td>
<td></td>
</tr>
</tbody>
</table>

Table 1: Total heat and electricity input; a, ideal operation (Y₃ = 100 mol%, Y₄ = 0 mol%, r_gas = 90 mol%, see 2.1) to conservative operation (Y₃ = 13.9 mol%, Y₄ = 9.1 mol%, r_gas = 66.7 mol%, see 2.1); b, taken from [14] and see 2.2; c, [6, 14]; d, [49].
Table 2: Cost estimates and necessary quantities for each NH$_3$ and H$_2$ tower (scenario 1 conditions, M marks million, websites information retrieved in April 2011, HX abbreviates heat exchanger). Total costs can be found by the product of each quantity column element and its related cost per unit.

<table>
<thead>
<tr>
<th>Item category</th>
<th>Unit</th>
<th>Cost per unit (U.S. dollars)</th>
<th>Quantity per NH$_3$ tower</th>
<th>Quantity per H$_2$ tower</th>
<th>Primary source</th>
</tr>
</thead>
<tbody>
<tr>
<td>Mo reactant</td>
<td>kg</td>
<td>36,000</td>
<td>1,174</td>
<td>2,288</td>
<td>Mark Jensen, CF Industries, Plant Manager (April 8, 2011)</td>
</tr>
<tr>
<td>Zn reactant</td>
<td>kg</td>
<td>234</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Silica gel</td>
<td>kg</td>
<td>211</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>NH$_3$ reactor</td>
<td>each</td>
<td>8.0 M</td>
<td>1</td>
<td>1</td>
<td>Wayne French, VP Tank Builders Inc. (April 8, 2011)</td>
</tr>
<tr>
<td>H$_2$ reactor</td>
<td>each</td>
<td>16.0 M</td>
<td>1</td>
<td>1</td>
<td></td>
</tr>
<tr>
<td>Water tank</td>
<td>m</td>
<td>0.6 M</td>
<td>0.16</td>
<td></td>
<td><a href="http://www.lme.com/minormetals/">http://www.lme.com/minormetals/</a></td>
</tr>
<tr>
<td>Water pumps</td>
<td>m</td>
<td>69</td>
<td>199</td>
<td>1.4</td>
<td><a href="http://www.lme.com/zinc.asp">http://www.lme.com/zinc.asp</a></td>
</tr>
<tr>
<td>Heliostats</td>
<td>m$^2$</td>
<td>164</td>
<td>17,006</td>
<td>397,039</td>
<td>Kolb, G. et al. 2006. “Heliostat Cost Reduction”, project working paper</td>
</tr>
<tr>
<td>Land</td>
<td>m$^2$</td>
<td>4.9 M</td>
<td>21,257</td>
<td>496,299</td>
<td>Southern California location assumed, <a href="http://www.-bajarealestatgroup.net/baja_real_estate/viewcategory/6/">http://www.-bajarealestatgroup.net/baja_real_estate/viewcategory/6/</a></td>
</tr>
<tr>
<td>Support tower</td>
<td>each</td>
<td>1.3 M</td>
<td>1</td>
<td>1</td>
<td>Mid-Atlantic Regional Space Port, Wallops Island, VA, Construction Manager (April 5, 2011)</td>
</tr>
<tr>
<td>HX &lt; 2000 K</td>
<td>m$^2$</td>
<td>467</td>
<td>1,944</td>
<td></td>
<td>Dennis L. Youchison, Sandia National Laboratories, Fusion Technology Dept., Albuquerque, NM</td>
</tr>
<tr>
<td>HX &lt; 1500 K</td>
<td>m$^2$</td>
<td>467</td>
<td>2,992</td>
<td></td>
<td></td>
</tr>
<tr>
<td>HX &lt; 500 K</td>
<td>m$^2$</td>
<td>50</td>
<td>20,160</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Air separation</td>
<td>each</td>
<td>0.8 M</td>
<td>0.14</td>
<td></td>
<td>John Font, Innovative Gas Systems (April 6, 2011)</td>
</tr>
<tr>
<td>Trucks or trailers</td>
<td>each</td>
<td>0.3 M</td>
<td>0.59</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Blowers</td>
<td>each</td>
<td>30,000</td>
<td>0.78</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Storage tanks</td>
<td>m$^2$</td>
<td>60</td>
<td>139.4</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Engineering costs</td>
<td></td>
<td>10%</td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

http://www.saginawpipe.com/steel_pipe_chart-3.htm

assumed

Kolb, G. et al. 2006. “Heliostat Cost Reduction”, project working paper
Southern California location assumed, http://www.-bajarealestatgroup.net/baja_real_estate/viewcategory/6/
Mid-Atlantic Regional Space Port, Wallops Island, VA, Construction Manager (April 5, 2011)
Dennis L. Youchison, Sandia National Laboratories, Fusion Technology Dept., Albuquerque, NM
John Font, Innovative Gas Systems (April 6, 2011) assumed
http://metalbuildingdepot.com/specials/default.aspx

assumed at 10% total tower construction cost
<table>
<thead>
<tr>
<th>Item category</th>
<th>Cost per item (million U.S. dollars)</th>
<th>Primary source</th>
</tr>
</thead>
<tbody>
<tr>
<td>Compressor (NH₃)</td>
<td>0.15</td>
<td>Mark Jensen, CF Industries, Plant Manager (April 8, 2011)</td>
</tr>
<tr>
<td>Compressor (O₂)</td>
<td>0.15</td>
<td>assumed</td>
</tr>
<tr>
<td>Compressor (air)</td>
<td>0.10</td>
<td>&quot;</td>
</tr>
<tr>
<td>Blower (silica bed)</td>
<td>0.03</td>
<td>&quot;</td>
</tr>
<tr>
<td>Storage container (NH₃)</td>
<td>10.00</td>
<td>Mark Jensen, CF Industries, Plant Manager (April 8, 2011)</td>
</tr>
<tr>
<td>Storage container (O₂)</td>
<td>14.00</td>
<td>&quot;</td>
</tr>
</tbody>
</table>

Table 3: Costs of each fixed plant component independent of scale.
<table>
<thead>
<tr>
<th>Input</th>
<th>Cost (U.S. dollars/unit)</th>
<th>Quantity per NH₃ tower</th>
<th>Quantity per H₂ tower</th>
</tr>
</thead>
<tbody>
<tr>
<td>Electricity</td>
<td>41.79 MWh⁻¹</td>
<td>1</td>
<td>-</td>
</tr>
<tr>
<td>Water</td>
<td>0.8 t⁻¹</td>
<td>-</td>
<td>2.29 t</td>
</tr>
<tr>
<td>Labor (reactors)</td>
<td>75,000 year⁻¹</td>
<td>1</td>
<td>-</td>
</tr>
<tr>
<td>Labor (compressors)</td>
<td>100,000 year⁻¹</td>
<td>0.1</td>
<td>-</td>
</tr>
<tr>
<td>Labor (piping)</td>
<td>60,000 year⁻¹</td>
<td>0.25</td>
<td>-</td>
</tr>
<tr>
<td>Service (heliostats)</td>
<td>3% of total heliostat cost</td>
<td>varies</td>
<td>varies</td>
</tr>
<tr>
<td>Total variable costs (NH₃ tower)</td>
<td>133,747 year⁻¹</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Total variable costs (H₂ tower)</td>
<td>1,653,155 year⁻¹</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

Table 4: Operating costs per tower. The Master Control system labor was assumed at 200,000 dollars. Water cost per ton from [50], electricity cost per MWh from [51], other costs assumed.
Table 5: Combination of scenario conditions and parameters (M marks million).

<table>
<thead>
<tr>
<th>Scenario matrix</th>
<th>$l = 1$</th>
<th>$l = 2$</th>
</tr>
</thead>
<tbody>
<tr>
<td>$m = 1$</td>
<td>$\delta = 6.35$ $\rho = 0.262$</td>
<td>$\delta = 2.97$ $\rho = 0.146$</td>
</tr>
<tr>
<td></td>
<td>$j = $150M + $50M$ to max($1,100M$)</td>
<td>$j = $100M + $100M$ to max($1,600M$)</td>
</tr>
<tr>
<td>$m = 2$</td>
<td>$\delta = 6.35$ $\rho = 0.262$</td>
<td>$\delta = 2.97$ $\rho = 0.146$</td>
</tr>
<tr>
<td></td>
<td>$j = $150M + $50M$ to max($1,100M$)</td>
<td>$j = $200M + $100M$ to max($1,600M$)</td>
</tr>
<tr>
<td>Category</td>
<td>Scenario 1</td>
<td>Scenario 2</td>
</tr>
<tr>
<td>---------------------------</td>
<td>------------</td>
<td>------------</td>
</tr>
<tr>
<td>Number of NH₃ towers</td>
<td>19</td>
<td>19</td>
</tr>
<tr>
<td>Number of H₂ towers</td>
<td>5</td>
<td>5</td>
</tr>
<tr>
<td>Output (t NH₃/d)</td>
<td>902.2</td>
<td>902.2</td>
</tr>
<tr>
<td>Total plant cost (M dollars)</td>
<td>769.9</td>
<td>558.3</td>
</tr>
<tr>
<td>NPV (M dollars)</td>
<td>116.9</td>
<td>337.1</td>
</tr>
<tr>
<td>NPV to initial capital ratio (%)</td>
<td>15.2</td>
<td>60.4</td>
</tr>
</tbody>
</table>

Table 6: To outline the optimum plant size: Overall summary of tower quantity, NH₃ output, build costs, net present value (NPV), and the NPV to investment ratio at the scenario-specific configuration optimum (M marks million).
Figure 1: Concept for solar thermochemical NH₃ synthesis near atmospheric pressure using a molybdenum reactant.
Figure 2: Conceptual implementation of solar thermochemical NH$_3$ synthesis coupled with solar thermochemical H$_2$ synthesis.
Figure 3: To assess the temperatures that are required for thermodynamic feasibility of the reaction cycle proposed: equilibrium composition of (A) MoO$_2$ reduction (Eq.1) and (B) N$_2$ fixation via Mo nitridation (Eq.2) (see 2.1) as a function of temperature at 0.1 MPa.
Figure 4: To assess the temperatures that are required for thermodynamic feasibility of the reaction cycle proposed: equilibrium composition of Mo$_2$N hydrolysis forming MoO$_2$, NH$_3$ and H$_2$ (Eq.3), Mo$_2$N oxidation forming MoO$_2$, N$_2$, and H$_2$ (Eq.4) and thermal dissociation of Mo$_2$N forming Mo and N$_2$ (see 2.1) as a function of temperature at 0.1 MPa.
Figure 5: Process schematic of solar thermochemical NH₃ synthesis and on-site H₂ production (dry cooling system not shown).
### Solar energy (MW)

<table>
<thead>
<tr>
<th>Solar energy (MW)</th>
<th>Ideal operation</th>
<th>Conservative operation</th>
</tr>
</thead>
<tbody>
<tr>
<td>a) Radiation rejected at all NH₃ towers at 1500 K</td>
<td>(107)</td>
<td>(248)</td>
</tr>
<tr>
<td>b) $\Delta_{rxnh}$ of MoO₂ reduction at 1500 K</td>
<td>(374)</td>
<td>(864)</td>
</tr>
<tr>
<td>c) Radiation rejected at all H₂ towers at 400-2000 K</td>
<td>(669)</td>
<td>(685)</td>
</tr>
<tr>
<td>d) $\Delta_{rxnh}$ of ZnO reduction at 2000 K</td>
<td>(1,960)</td>
<td>(1,960)</td>
</tr>
</tbody>
</table>

### Electricity & process heat (MW)

<table>
<thead>
<tr>
<th>Electricity &amp; process heat (MW)</th>
<th>Ideal operation</th>
<th>Conservative operation</th>
</tr>
</thead>
<tbody>
<tr>
<td>e) Electricity for all operations except cooling</td>
<td>(63) (59)</td>
<td>(110) (101)</td>
</tr>
<tr>
<td>f) Electricity for cooling fan</td>
<td>(1,481)</td>
<td>(2,221)</td>
</tr>
<tr>
<td>g) Heat removed via dry cooling at 300-340 K</td>
<td>(317)</td>
<td>(451)</td>
</tr>
<tr>
<td>h) $\Delta_{sensh}$ for heating ZnO at 400-2000 K, solar energy</td>
<td>(2,597)</td>
<td>(3,895)</td>
</tr>
<tr>
<td>i) heat integrated at 293-2000 K</td>
<td>(864)</td>
<td>(685)</td>
</tr>
</tbody>
</table>

Figure 6: Total power requirements at industrial production scale (producing 1,324 t NH₃ per day, within 7.48 h/d operation on average, see 2.1, 2.2 and 2.3). All power values given in brackets are in MW/plant.
Figure 7: Actual average annual NH\textsubscript{3} sales price reported by the USDA between 1999 and 2010 (filled circles) and the Monte Carlo 20-year price simulation with a standard deviation of 27.13 and a mean of 522.24 dollars/t NH\textsubscript{3} (empty circles).
Figure 8: NPV and total initial plant costs as a function of NH$_3$ output (for descriptions of Scenarios 1 or 4 see 2.3 and 3.2).
Figure 9: Sensitivity of (A) Scenario 1 or (B) Scenario 4 to variations in the NH₃ sales price Monte Carlo-simulated over a 20 year plant lifespan (diamonds mark simulations with baseline-NH₃ sales prices, see 3.1 and 3.2, circles mark simulations with in- or decreased NH₃ sales prices, as indicated, to break even at “optimum plant size” shown with filled symbols, see 3.4).